WASTE TREATMENT AT THE RADIOCHEMICAL ENGINEERING DEVELOPMENT CENTER

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ABSTRACT

At the Radiochemical Engineering Development Center (REDC), irradiated targets are processed for the recovery of valuable radioisotopes, principally transuranium nuclides. A system was recently installed for treating the various liquid alkaline waste streams for removal of excess radioactive contaminants at REDC. Radionuclides that are removed will be stored as solids, and thus the future discharge of radionuclides to liquid-low level waste tank storage will be greatly reduced. The treatment system is of modular design and is installed in a hot cell at REDC, where preliminary testing is in progress. The module incorporates the following: (1) a resorcinol-formaldehyde resin column for cesium removal, (2) a cross-flow filtration unit for removal of rare earths and actinides as hydroxide, and (3) a waste solidification unit. Process flowsheets for operation of the module, key features of the module design, and its computer-assisted control system are presented. Good operability of the cross-flow filter system is mandatory to the successful treatment of REDC wastes. Results of tests to date on the operation of the filter in its slurry collection mode and its slurry washing mode are presented. These tests include the effects of entrained organic solvent in the waste stream to the filter.

INTRODUCTION

The Radiochemical Engineering Development Center (REDC) at Oak Ridge National Laboratory (ORNL) processes highly irradiated targets for the separation and

recovery of valuable radioisotopes, principally transuranium (TRU) nuclides. Presently a waste treatment system module has been fabricated and installed to pretreat essentially all liquid low-level waste (LLLW) generated at REDC. The module incorporates (1) a resorcinol-formaldehyde (RF) resin ion-exchange column, (2) a cross-flow filtration unit, and (3) a waste solidification unit.

The waste treatment strategy is to remove the precipitated solids from the neutralized waste streams by processing the waste streams through a cross-flow filter, with the filtrate going to the LLLW and the solids going to solid waste storage. Neutralized waste is first treated using cross-flow filtration to remove the insoluble hydroxide of the transuranium and rare earth elements as a concentrated slurry with concentrations as high as 10–20% by weight. The filtrate can be subsequently treated using an RF resin column to remove cesium. The concentrated slurry and the neutralized cesium eluted from the RF resin column are then transferred to the in-cell solidification unit. Waste canisters resulting from this treatment system will be stored on-site while awaiting final disposal.

To carry out the waste treatment strategy, three REDC waste pit tanks (F-111, F-115, and F-126) are used in conjunction with the waste treatment module to characterize, neutralize, and store the liquid waste. These tanks are shown in Figure 1 along with the flowsheet for the in-cell waste treatment module. The 550-gallon capacity F-115 tank is used to collect various wastes from the many REDC processes for acidification and characterization sampling. From F-115 the acidic waste is transferred in batches into a caustic heel in the 125-gallon F-111 tank for neutralization and sampling. The neutralized waste is then pulled up from F-111 to the waste treatment module feed tank, T-780, in 13- to 15-L batches. The neutralized solution is filtered through the cross-flow filter with the filtrate routed to the third tank, F-126. F-126 is a 1200-gallon tank that collects all REDC LLLW for sampling and transfer to the ORNL LLLW collection and treatment system. If the filtrate contains cesium, the module offers an option of treating the filtrate through an ion-exchange column; however, the majority of the cesium-contaminated waste is generated during target dejacketing operations and will be transferred directly through the ion-exchange column after filtration by conventional means. The concentrated slurry from the filtration system is periodically transferred from

FIGURE 1.   Waste pretreatment schematic for REDC.

the T-780 feed tank to a solidification unit, where the slurry is dried using a clam shell heater.

EXPERIMENTAL METHODS

Waste Treatment Module

The approach taken to treat waste at REDC was to build a waste treatment module using existing equipment and framework designs that could be installed in a hot cell position in Building 7920. The installation of this system required the removal of the old equipment rack in the right position in Cubicle 7 and replacement with the new waste treatment rack module. The module contains all in-cell equipment necessary for the operation of the waste treatment system described previously.

        The module is operated using a computer system to automate routine operations to
the extent allowed by available cell penetrations for instrumentation. This automation
reduces the need for constant attention by operating personnel so that normal processing
operations can continue with minimal impact from the waste treatment system. Automated
tasks include waste transfers to the module, pump flow rate control, and back flushing of
the filter on a set time schedule. The computer system is programmed to activate control
valves, solenoid valves, and pressure regulators to allow a task to occur. The computer is
also programmed to monitor the waste operation and warn operators of unusual events or
operations that may require special attention. A more detailed description of the
equipment and operation is provided as follows.

Removal of Cesium by Ion Exchange

        A 4-in.-diameter, 36-in. length of glass pipe was employed to fabricate the 7.2-L
capacity cation-exchange column for the waste treatment module. The column was
designed to operate using gravity flow feed with a resin bed volume ranging from 3 to 4 L.
An RF resin bed will be employed to remove cesium from REDC waste.

        The RF resin was developed at the Westinghouse Savannah River Laboratory and
is highly specific for the removal of cesium from an alkaline waste of high sodium
content.[1] It is particularly advantageous for REDC because it permits treatment of the
highly alkaline (2 M NaOH) aluminum decladding wastes. This resin has received
considerable study.[2,3,4] The resin is a condensation polymer of resorcinol and
formaldehyde, and like other phenolic resins (such as Duolite CS-100), it can be expected
to suffer loss of capacity by oxidation of its phenolic (OH⁻) groups. The oxidation
instability has been the largest objection to both widespread use and large-scale application
of phenolic resins. Additional degradation by slow hydrolysis may also occur. In typical
usage of Duolite CS-100 at the ORNL wastewater treatment plant, there was extreme
capacity loss after 39 cycles. This is not necessarily an objection at REDC, where waste
to be treated is relatively small in volume. Limited operational life (~5 cycles) of RF
resins could be accommodated since only a few liters (<10) per annum would be
consumed.

Removing cesium from an LLLW stream using the RF resin has been demonstrated at REDC.[5] The two in-cell demonstrations resulted in standard breakthrough curves for loading performance and elution profiles for the cesium stripping performance. The feed solutions were alkaline dejacketing solutions created during the dissolution of Mark 42 target segments. The feed solutions typically had an [OH⁻] range from 1 to 2 M with gamma activity levels of 2.6×10^8 Bq/mL for the first run and 3.4×10^7 Bq/mL for the second run. The distribution coefficient (K_d) of the resin was calculated for each run. The distribution coefficient is defined using the following equations.

$$K_d = C_R/C_A,$$

where

C_R = Cs activity/gram of resin,

C_A = Cs activity/mL of solution in equilibrium with the resin.

Stripping the cesium from the resin was done with 2 M HCl, which converts the resin to an H⁺ form. The resin was converted back to the Na⁺ form between runs by washing the resin with 12 resin column volumes (CV) of 2 M NaOH using an upflow path to fluidize the resin for maximum liquid-to-resin contact. The resin remained in the caustic wash for ~18 hours before loading to ensure that the resin was fully converted back to the Na⁺ form used for cesium loading.

Cross-Flow Filtration

A cross-flow filtration unit has been employed for removal of solids. The inertial cross-flow filter is a nonclogging filter designed for particulate or sludge concentration and liquid recovery that contrasts with dead-end filtration, where the process stream passes perpendicular to the plane of the filter medium and the particles accumulate on the filter medium surface.[6] The filter consists of three Mott Metallurgical Corporation inertial cross-flow filter elements (3/8-in. OD, 1/4-in. ID, 24-in. length, 0.5-μm-pore size, 316 SS porous metal filters) housed in 1½-in. OD 316 SS tubing (0.065-in. wall) with ½-in. OD 316 SS tubing connections for feed flow (0.065-in. wall) and a 3/8-in. OD 316 SS tubing connection for filtrate flow.

The cross-flow filter is operated using a slurry velocity of ~20 ft/s and applying back pressure on the filter, which is referred to as the filter pressure. With a differential pressure across the porous wall, a portion of the liquid phase of the main stream will flow through the porous wall into the annulus as clear filtrate and exit the outlet. The main stream is returned to the feed tank, resulting in an increased concentration of solids.

Figure 2 demonstrates the typical filtration pattern of a cross-flow filter. The saw-tooth effect is created by back flushing the filter. Back flushing is necessary to dislodge the surface membrane that forms on the filter tubes. Typically, a new filter will have high filtrate rates until a surface membrane is formed on the filter tubes. The filtrate rate will decline and eventually go to a steady state. Back flushing the filter clears the surface membrane and increases the filtrate rate. Therefore, a periodic back flush is used to increase the filtrate flow. Back flushing the filter is carried out using air pressure over a filtrate reservoir to force filtrate back through the filter pores. The frequency of the back flush can range from 1- to 30-min intervals. A 15-min back flush interval was used for the filter test and module check-out tests.

The cross-flow filter was evaluated using a filter test loop. A 22-L tank was used as the feed tank. The slurry was pumped though the loop using an air-driven double-diaphragm pump. The pump flow rate was controlled using an air regulator to adjust the 100-psig supply pressure. A restricting valve downstream of the cross-flow filter was partially closed to develop back pressure in the pump loop which is the differential pressure across the filter or the filtration pressure. The filter was back flushed using an air reservoir set at 70–80 psig to drive the filtrate back through the filter elements.

A synthetic feed solution was used in testing the cross-flow filter. The feed consisted of ~1% wt. each of $Al(OH)_3$, $ZrO(OH)_3$, and $Fe(OH)_3$ suspended in a 0.1 M NaOH and 0.5 M $NaNO_3$ solution. The metal hydroxide feed was developed by dissolving the nitrate form of the materials in water. The salts were then converted to hydroxides by precipitating the solution with 2 M NaOH solution to yield the metal hydroxide forms. The liquid was then decanted, and the precipitate was added to 0.1 M in NaOH and 0.5 M
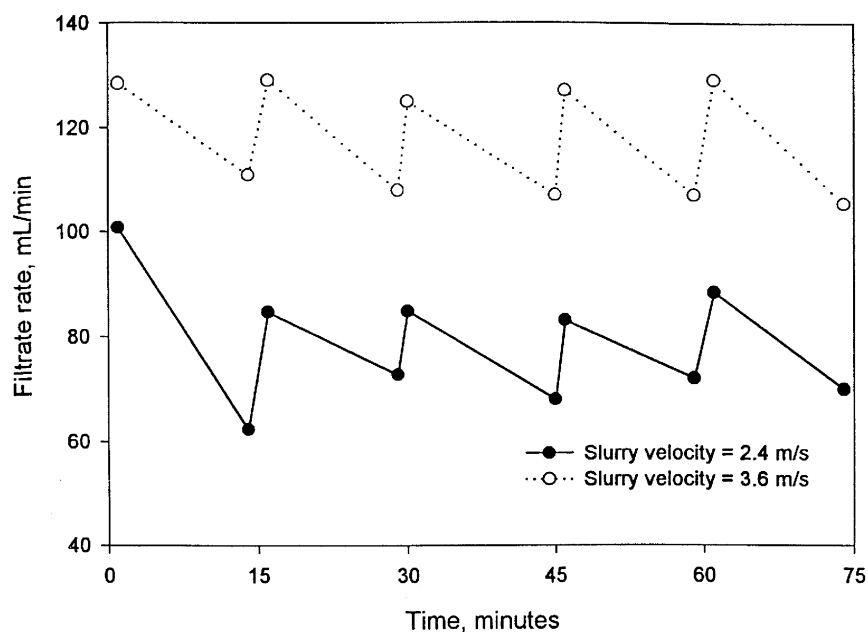
FIGURE 2.   Typical filtrate patterns of a cross-flow filter.

in NaNO₃ to give the 3% wt. metal hydroxide. This synthetic feed solution was used on all cross flow filter tests and module tests.

Typical flux behavior is shown in Figure 3. Normally, the characteristics of a cross-flow filtration system will indicate that as the slurry concentration increases, the filtrate flux will decrease due to the buildup of solids in the main feed stream. The slurry flow rate was set by the amount of pressure selected to operate the diaphragm pump. The process parameters investigated were maximum pump operating pressure and the filtration pressure. The maximum operating pressure is the pressure developed when the pump is deadheaded. The filtration pressure is the back pressure applied to the filter by a restricting valve downstream of the filter. The filtrate rate is recorded as the filtrate flux, $L/(min \cdot m^2)$, which is the filtrate rate relative to the surface area of the filter element.
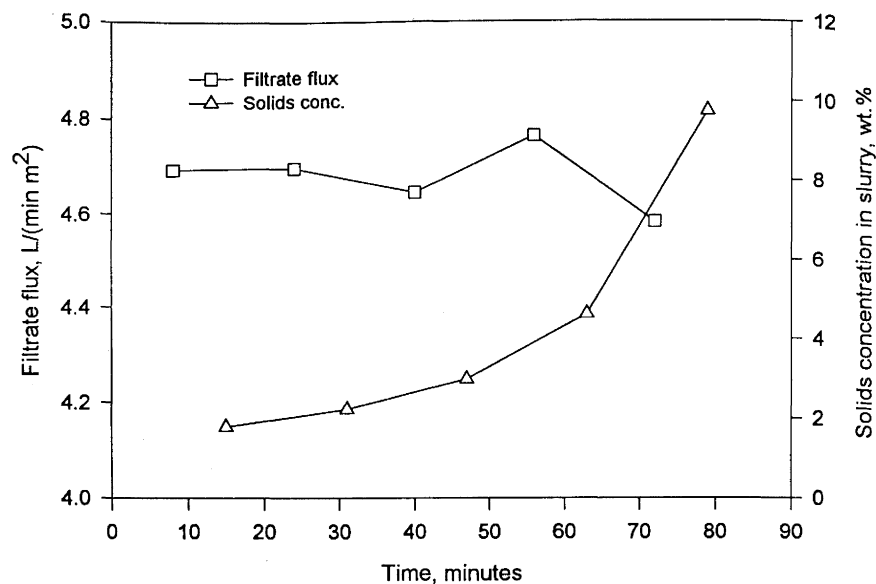
FIGURE 3. Typical filtrate flux and solids concentration profiles.

The normal process schemes at REDC involve organic extractions; therefore, it is reasonable to expect that organic reagents will be entrained into the aqueous waste streams. The most common organic extraction process used at REDC is the Cleanex process, which is a versatile solvent extraction process for recovery and purification of lanthanides, americium, and curium. The Cleanex extraction process uses di (2-ethylhexyl) phosphoric acid (HDEHP) as the organic extractant. The effects of organic contamination on the 3 wt.% synthetic slurry were investigated using the Cleanex solvent extractant HDEHP in a normal paraffin hydrocarbon (NPH). To evaluate the effects of organic contamination, the cross-flow filtration system was operated using an upstream pressure of 49 psig with a back pressure of 40 psi, and the filtrate side of the filter routed back to the vented feed tank to recycle the filtrate. The filter was back flushed at 15-min intervals. The slurry flow rates were set at 10.2 ± 1.1 L/min. Filtrate measurements were taken at fixed increments after each back flush. Total volume of

filtrate removed was determined by direct volumetric measurement and by intermittently measuring the decrease in feed volume.

Four sets of tests were used to examine the effect of organic contamination on the filtrate flux. First, two baseline filtrate flux measurements were performed using water and a 0.1 M NaOH solution. A second set of tests was performed to determine the flux potential for similar alkaline solutions contaminated with Cleanex extractant. The organic solution was added incrementally to 10 L of 0.1 M NaOH in the filtration test loop between runs to yield suspensions of 0.1, 0.2, and 0.4 vol.% Cleanex extractant.

A third set of tests was conducted to gauge the effects of small amounts of dissolved and suspended Cleanex extractant using the synthetic metal hydroxide feed. Cleanex extractant was added incrementally to the filtration test loop to yield 0.1, 0.2, 0.4, and 1.0 vol.% Cleanex extractant in the 3 wt.% metal hydroxide slurry feed.

The fourth set of tests was conducted to concentrate the slurry by not recycling the filtrate back to the feed tank, thus allowing the solids concentration to increase in the feed tank. The 3 wt.% metal hydroxide feed was concentrated without Cleanex extractant and with a 1.0 vol.% Cleanex extractant. The filter and filtration loop were cleaned between each run using a 5 wt.% solution of HNO_3.

Waste Solidification

Waste solidification is accomplished by diverting the concentrated slurry from the filtration loop to a canister. The canister volume is 4 L. The canister (solids can) is positioned in a clam shell heater which evaporates the slurry to dryness. A conductance probe is used to indicate when the slurry level in the can reaches 80% of the canister volume so slurry transfer operations can be terminated before the solids can is overfilled.

The clam shell heater is operated from 110 to 140°C to deliver a solids-can temperature ranging from 95 to 105°C. A programmable controller is used to slowly ramp up to the desired temperature range. Slurry evaporation rates averaged 3 L of water per 24 hours. The solids can will be heated 24 hours per day to allow the solids to bake during any delayed transfer intervals. Once the solids can is filled to an ~2-L volume with dried solids, the solids will be baked for ~5 days before the solids can is removed from the

heater. When the solids can is removed from the module, it is capped and vented using a certified HEPA filter to prevent material releases.

Sodium Nitrate Washing

To avoid the buildup of soluble salts, in particular sodium nitrate, the solids will be washed before the concentrated slurry is transferred to the solids can. The concentrated slurry at its minimum volume is diluted with water and then reconcentrated using the cross flow filtration system. The dilution and reconcentrating may be repeated as many times as necessary to remove the soluble salts to the desired level. This technique may be described in terms of a dilution factor (DL), which is the total volume of the wash (water plus slurry) divided by the initial slurry volume. The dilution factor is

$$DL = V_t/V_s,$$

where

V_t = total volume of the water wash and concentrated slurry,

V_s = initial volume of the slurry.

Then the concentration of the soluble salts after one wash is

$$C_1 = C_0/DL_1,$$

where

C_0 = original concentration,

DL_1 = Dilution factor for first wash.

If the same dilution factor is used for each wash, it follows that

$$C_n = C_0/(DL)^n,$$

where

n = number of washes,

C_n = concentration after n washes.

Waste Treatment Control System

The routine operation of the waste treatment equipment has been automated by the use of an industrial computer system. All the electrical components of the waste treatment

system, with the exception of the waste solidification evaporation system, are connected to the electronic inputs and outputs of the industrial computer. The electrical components of the waste treatment system are electronic differential-pressure transmitters, an electronic pressure transmitter, pressure switches, electronic conductance probes, electric solenoid valves, an electronic controlled pressure regulator, and a custom photometal instrument control panel comprising electronic indicators, lights, and switches. The evaporation system of the waste solidification unit is monitored and controlled by both a temperature controller and a high-temperature alarm indicator/switch.

A conductance probe is used in the solids transfer operation. It was not feasible to have liquid-level probes in the can since the probes would plug with dried solids. A conductance probe was installed in the can at a level of ~3.2 L (~80% can volume) to prevent overfilling the can. When the transferred slurry reaches the 3.2-L level, an alarm is activated indicating that the transfer must be terminated. Transferring the concentrated slurry is a manual operation that is critical. If the solids can is overfilled, the excess slurry enters the vent line and could possibly plug the module's off-gas line. The conductance probes are critical to the solids transfer operations. A check of the conductance probe's circuitry is also performed to make certain that the probes are functioning.

Back flushing the filter is an automated operation that is programmed to occur at 15-min intervals whenever the filtration pump loop is operating above 5 psi. Back flushing is accomplished using air over the filtrate to push the filtrate back through the pores of the filter to remove the surface membrane on the filter. An air volume under high pressure is captured between two solenoid valves, while a third solenoid valve, downstream of the filtrate, is closed. The trapped air is then released into the filtrate line and forces the filtrate back through the filter. A panel switch can also manually back-flush the filter when pump loop is operating above 5 psi.

RESULTS AND DISCUSSION

Cesium Removal by Ion Exchange

Two in-cell tests were performed to evaluate the RF resin using a 1.5-L resin bed to remove the cesium from REDC Mark 42 target dejacket waste. The cesium loading

characteristics of the in-cell tests are shown in Figure 4, where the ratio of the ^{137}Cs effluent concentration (C) to ^{137}Cs feed concentration (C_0) is profiled with the dejacketing solution throughput. The initial breakthrough point for the runs was in the range of 85–92 CV. Run 1 reached only a 17.7% breakthrough point due to an excess of the resin. Run 2 used a larger volume of feed material and exhibited a fully developed breakthrough curve similar to the bench-scale run performed in the laboratory using a 4.45-mL fixed-bed column of the RF resin. The bench-scale test and Run 2 reached the 50% breakthrough point at ~110 CV. The distribution coefficient (K_d) values for the bench-scale test and Run 2 were based on the 50% breakthrough point and found to be 237 and 283 mL/g, respectively.

Figure 5 profiles the stripping data from the in-cell runs. The resin was stripped with a 2 M HCl solution. Samples were pulled as the strip solution exited the column. Due to erratic flow, caused by siphoning at the start of Run 1, the elution peak went undetected; however, a complete elution profile was obtained during Run 2. The highest cesium concentrations were found as 1–2.5 CV of the strip solution passed through the resin bed, indicating that the vast majority of the material is actually removed from the resin in less than 3 CV.

The stripping data revealed that a 1.5-L resin bed was loaded with 1130 Ci of ^{137}Cs and 146.9 Ci of ^{134}Cs during Run 1 and 1160 Ci of ^{137}Cs and 141 Ci of ^{134}Cs during Run 2. The resin loaded 27.6 g of cesium during Run 1 and 28.4 g of cesium during Run 2. Samples of the eluted resin bed revealed that the percentage of cesium remaining on the resin was 0.006% of the cesium loaded. Essentially all of the cesium was removed during the stripping operation.

In summary, the RF resin demonstration indicated that a 1.5-L resin bed averaged loading ~1300 Ci of cesium or ~30 g of total cesium. The initial cesium breakthrough occurred as 85–92 CV of alkaline waste passed through the resin bed. The majority of the cesium was stripped from the resin bed with only 2.5 CV of 2 M HCl. Based on the in-cell demonstration, a 3- to 3.5-L resin bed of RF would process the cesium produced during the dissolution of two to three Mark 42 target segments.
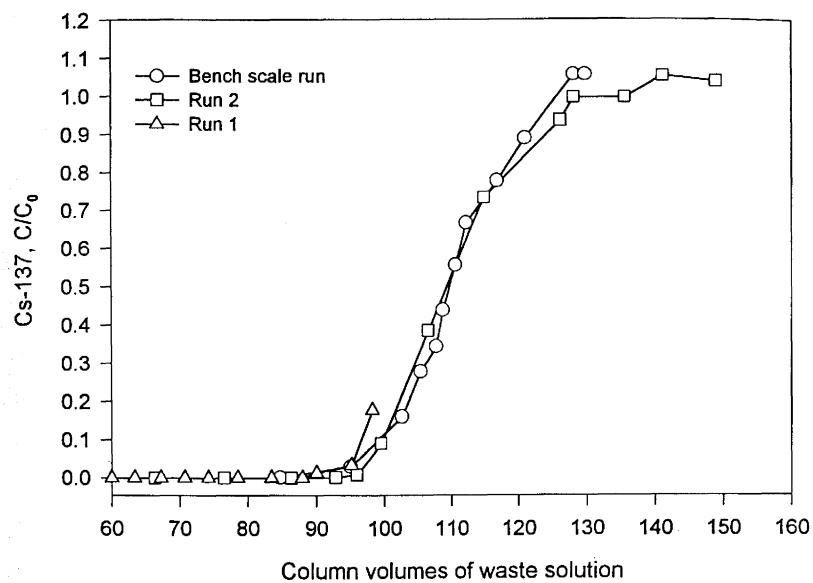
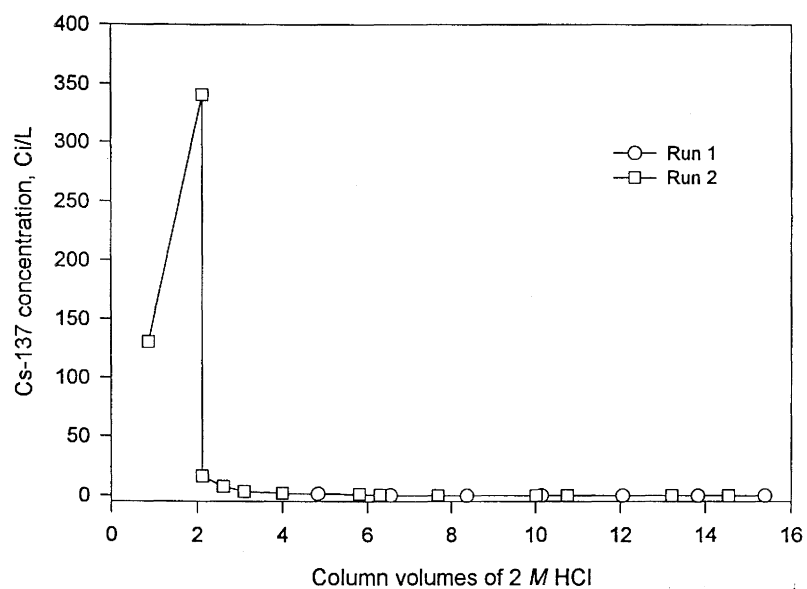FIGURE 4.   Cesium loading profiles of two in-cell runs and a bench-scale run.

FIGURE 5.   Cesium concentration profiles during resin elution with 2 M HCl.

Cross-Flow Filtration

Laboratory tests indicated that the maximum filtrate flux is achieved using a differential pressure range of 10–20 psi between the upstream pressure and the back pressure. Figure 6 shows the typical effects that the differential pressure between the maximum pump operating pressure and the filtration pressure have on the filtrate flux. Based on these results, the cross-flow filter used on the module will be operated using a pump pressure of ~60 psig with a back pressure range of 40–45 psi.

The baseline test revealed that the addition of NaOH to water would affect the filtrate rate. A comparison of the baseline results for water and 0.1 M NaOH indicated that the filtrate flux was ~90% lower for the NaOH solution. This reduction in flux was attributed to excessive foaming during the NaOH solution tests.

The second set of tests where Cleanex organic was incrementally added to a 0.1 M NaOH solution did not show a clear effect on the filtrate rate. The filtrate samples taken at 8 min and 12 min after a back flush indicated that there is no effect on the filtrate flux when the organic concentration ranges from 0 to 0.4 vol.% Cleanex extractant. The results are shown in Table 1.

It was observed that upon mixing in the filtration test loop, the Cleanex solution organic appeared to form a fairly stable opaque emulsion in the 0.1 M NaOH. As with the 0.1 M NaOH solutions, considerable foaming was observed with the Cleanex-contaminated NaOH solutions while circulating in the filtration test loop, and the presence of the organic appeared to compound this effect. As expected, the filtrate samples from these suspensions had the same appearance as the opaque, emulsified feed, indicating that the cross flow filtration unit was not separating the organic from the feed. Phosphate and total organic carbon analysis of samples of feed and permeate from these runs were inconclusive in determining the distribution of Cleanex extractant in the feed and the filtrate. The filtrate flux did not appear to be appreciably affected by the addition of Cleanex solution to the NaOH solution.

The third set of tests used the 3 wt.% metal hydroxide slurry contaminated with the Cleanex organic. The filtrate flux decreased slightly as the concentration of Cleanex extractant increased. The average filtrate flux for each simulated waste slurry test is
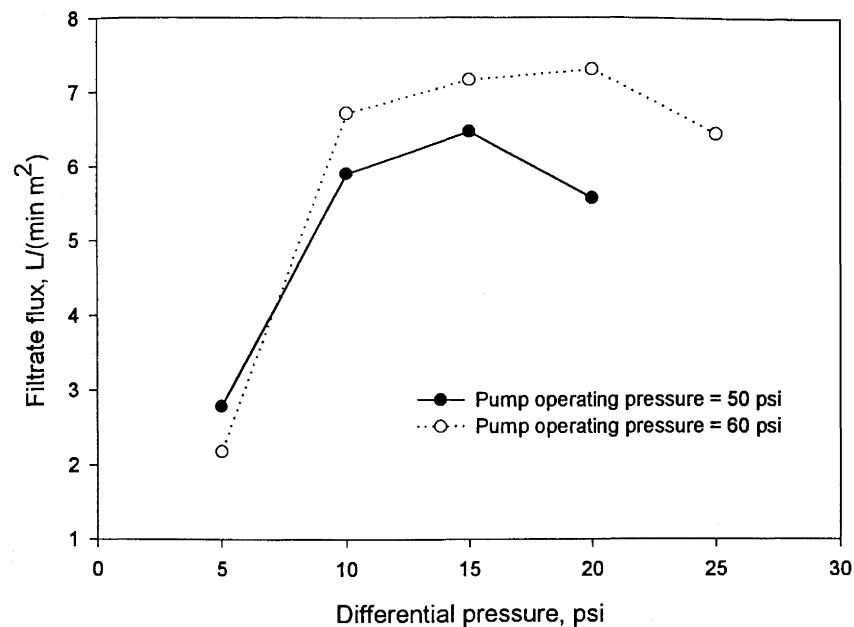
FIGURE 6. The effects of the differential pressure between the pump pressure and the filtration pressure on the filtrate flux.

TABLE 1. AVERAGE FILTRATE FLUX FOR 0.1 M NAOH SOLUTIONS OF VARYING CONCENTRATIONS OF CLEANEX SOLUTION.

Cleanex solution conc., % vol. of 1 M HDEHP in NPH in 0.1 M NaOH	Average Filtrate Flux [a]		
	1 min a.b.f. L / (min·m²)	8 min a.b.f. L / (min·m²)	12 min a.b.f. L / (min·m²)
0.0% [b]	22.5	10.0	8.7
0.1%	16.3	8.0	6.3
0.2%	21.4	11.9	9.5
0.4%	17.5	10.8	8.8

[a] Values are for the average flux of runs measured 1, 8, and 12 min after a back flush (a.b.f.).

[b] 0.1 M NaOH solution with no organic.

plotted in Figure 7. Unlike tests with 0.1 M NaOH contaminated with Cleanex solution, the organic-contaminated metal hydroxide slurries did not exhibit any foaming.

The fourth set of tests was conducted using the 3 wt.% metal hydroxide slurry contaminated with 1 vol.% Cleanex organic and removing the filtrate in order to concentrate the solids in the slurry feed tank. The filtrate flux for each run is plotted in Figure 8, and the slurry concentration profiles are shown in Figure 9. The average filtrate flux for both the organic-contaminated and nonorganic-contaminated runs was found to increase slightly between the runs and was attributed to the cleaning process. The filter element characteristics may have been changed after each HNO_3 cleaning cycle. Similar flux patterns were observed for both slurries from run to run.

Results from these experiments indicate that the organic solutions will inhibit the filtrate flux. These results were expected. Organic solutions commonly emulsify and clog filtration systems. The filtrate flux measured when the solids were concentrated in the feed tank ran about 1.5–2.0 L/(min·m^2) less initially than the runs not contaminated with Cleanex extractant. After 30–35 min into the runs, the filtrate flux for the nonorganic runs was reduced below the filtrate flux for the runs which contained organic. This observation is not surprising and can be attributed to a faster build up of solids in the feed tank when the filtrate rate is higher. The build up of solids will degrade the filtrate rate. The solids concentration reached 10 wt.% in ~34 min for the nonorganic runs compared with ~47 min for the runs containing organic.

Sodium Nitrate Washing and Solidification

Three water washes of the synthetic feed reduced the feed nitrate content from 25,000 μg/mL to 780 μg/mL, resulting in the removal of 97% of the $NaNO_3$. The sodium nitrate washing profiles are shown in Figure 10. It was determined that one water wash with a dilution factor of 5.3–5.7 will reduce the nitrate content of the feed by ~65–75%.

Testing of the solidification unit revealed that a solids-can temperature of 95–100°C is needed to slowly evaporate the water from the solids can. The slurry evaporation rates averaged 3 L of water per 24 hours when the solids can temperature was maintained at 100°C. The solids can will be heated 24 hours per day to allow the solids to bake during any delayed transfer intervals.
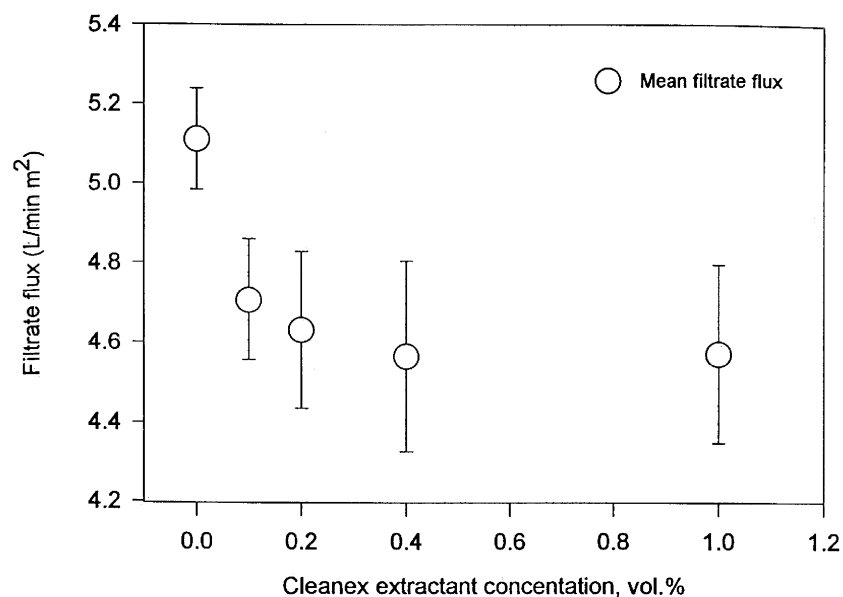
FIGURE 7. Filtrate flux profile using a synthetic feed contaminated with Cleanex extractant.

MODULE STATUS AND FUTURE WORK

The waste treatment module has been fabricated and installed. Equipment checkout is scheduled for October and November of 1997. The module is being operated using a personal computer with Visual Basic programs. The final control system will be a panel board on the master control room panels of Building 7920. The installation of the control panel and the programming will continue through the equipment checkout and testing. The waste system is scheduled to begin hot testing operations in the spring of 1998.

Based on testing with synthetic feeds and organic contaminants, the cross flow filter will be operated to deliver a minimum filtrate rate of 200 mL/min. The slurry pump
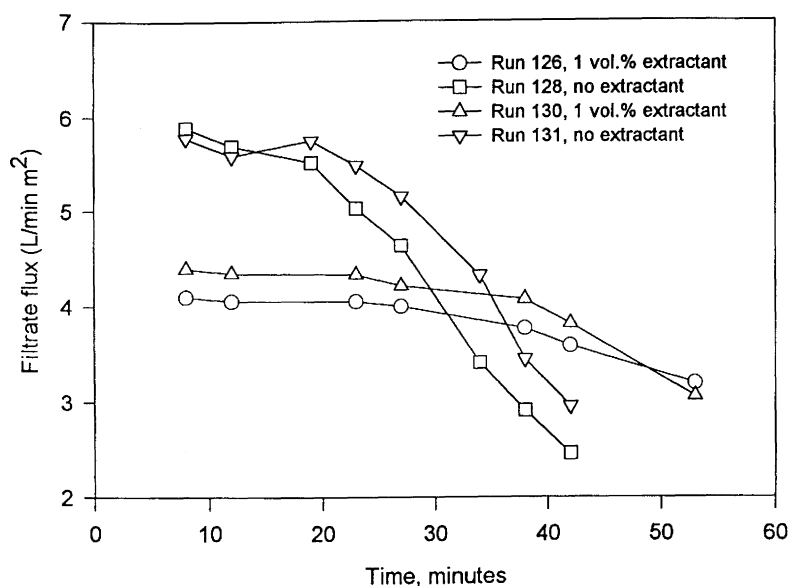
FIGURE 8.   The effects of contaminating the synthetic feed with 1 vol.% Cleanex extractant.

will be operated using 60- to 65-psi air pressure to pump the slurry through the filtration loop. The back pressure applied to the filter will range from 40 to 45 psi. Assuming that a waste tank is filled to 80% capacity, the F-115 acid waste tank would be processed through the module in ~140 hours. The concentrated slurry will be washed in the T-780 feed tank with three 13- to 14-L water washes to reduce the sodium content by ~97% before transferring the slurry to the solids can. To help prevent the transfer of organic solution to the waste treatment module, a 50-L heel will be kept in the F-115 acidification waste tank.

Check out of the ion-exchange column is scheduled to begin during the summer of 1998. The ion-exchange column will contain a 3- to 4-L volume of an RF resin capable of removing ~3000 Ci of cesium or processing ~3 Mark 42 target segments. The resin will be stripped with 10–15 L of 2 M HCl.
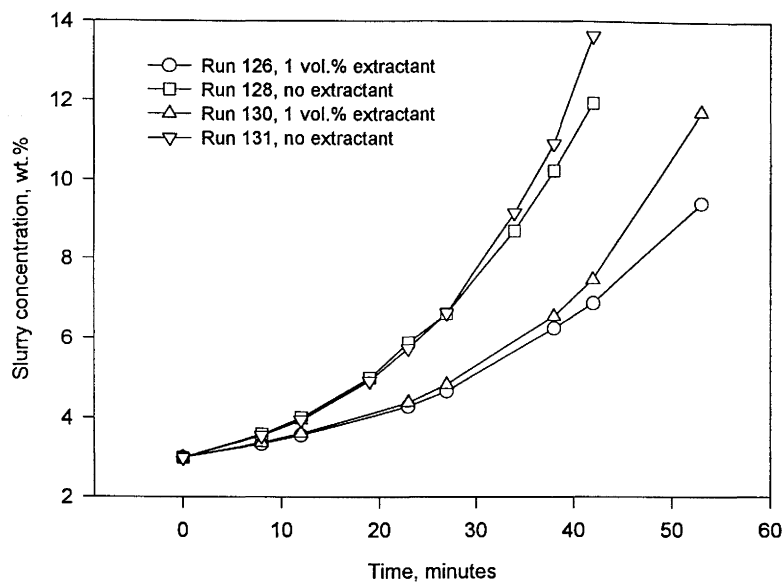
FIGURE 9.   The effects of 1 vol.% Cleanex extractant on the solids buildup rate.
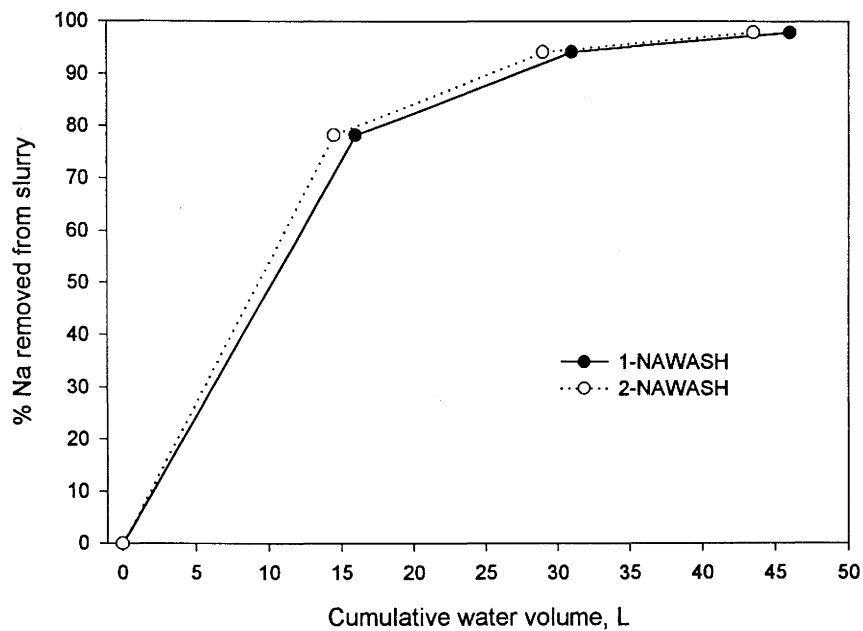
FIGURE 10.   Sodium nitrate removal using three water washes.

Future work on the waste treatment system will include developing processing alternatives, programming the computer control system, developing decontamination methods for the removal of the solids can, and developing methods to remove the solids can from the building. Alternative processing methods may allow waste to be transferred directly to the waste module, thus bypassing the F-115 acidification tank. Programming the operations will continue as the module is checked out for operations. Decontamination of the solids can will involve developing an electropolishing system that will allow the can to be removed from the cubicle free of surface contamination. The solids can will be removed from Cubicle 7 via a slug chute that is normally used to drop materials into the cubicle. The slug chute must be kept free from contamination in order for operators to use the chute. The methods for removing the solids can have not been finalized. An independent carrier that will mate to the existing slug chute will have to be fabricated. The carrier will then be used to transport the canisters to a storage cask or transportation carrier. These remaining tasks will be focused on during 1998.

ACKNOWLEDGMENTS

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