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DEVELOPMENT OF AN ELECTROLYTIC CELL FOR REMOVAL OF NICKEL FROM HRT FUEL SOLUTIONS

E. L. Youngblood

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Chemical Technology Division

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ABSTRACT

An HRT prototype electrolytic cell with a mercury cathode and a platinum anode was tested for nickel and manganese removal from simulated HRT fuel solution. Approximately 90% nickel removal and 80% manganese removal was accomplished in 290 amp-hr per liter. The manganese removal required replacement of the liquid mercury cathode after one half of the nickel was removed. In the course of this electrolysis, essentially all of the copper was removed and 60% of the uranium was reduced to the four valence state. Further work will be required to improve the radiation stability of the insulators and to reduce the corrosion of the cell.

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1.0 INTRODUCTION

This report contains the results of an engineering test of a full-scale, 30-gal-capacity, electrolytic cell for batch removal of nickel from a concentrated[1] HRT-type fuel solution. The corrosion of stainless steel components in the HRT core and blanket systems introduces nickel into the fuel solution. The nickel concentration of the fuel is limited because of the possibility of fuel solution instability[2] and the unproductive capture of neutrons by the nickel.

Except for the initial equipment testing, all runs were made with either a light- or heavy-water simulated fuel solution of the approximate composition 0.045 M $NiSO_4$-0.27 M $CuSO_4$-0.03 M $MnSO_4$-0.34 M UO_2SO_4-0.24 M H_2SO_4. In the HRT system a solution of this approximate composition would be produced by evaporation of the fuel. The nickel concentration in the test system represents about 250 ppm nickel in the original core solution. The test equipment was sized to process 30-gal batches of concentrated fuel solution, which would be the approximate volume of the HRT core and blanket fuel after concentration, assuming that the blanket concentration is about one-third that of the core. A method of regenerating the mercury cathode to remove the deposited metals without removing the amalgam from the cell was tested, but conditions for replacing the copper and reoxidizing the uranium for re-use of the fuel were not investigated in this series of tests.

The nickel removal process described was developed by the Chemical Technology Division Chemical Development Section A. R. A. McNees of that section, W. D. Burch of the Pilot Plant Section, H. O. Weeren of the Process Design Section, and P. A. Haas of the Unit Operations Section contributed to the design of the electrolytic cell used in the test described in this report.

2.0 EQUIPMENT AND PROCEDURE

2.1 Equipment

The equipment used in the test electrolytic nickel-removal system consisted of the fuel, off-gas, and regeneration systems (Fig. 1). The fuel solution system contained a 30-gal storage tank from which the fuel solution was circulated through the electrolytic cell by a Carpenter 20 stainless steel Chempump equipped with aluminum oxide bearings. A condenser and a 7-gal condensate tank were connected to the tank to produce condensate for washing the uranium from the cell prior to the mercury regeneration step. The system was constructed primarily of 300 series of stainless steels.

Gases produced in the cell during nickel removal were discharged to the off-gas system through a line attached between the fuel solution tank and the condenser. An air purge was introduced into the fuel solution tank to dilute the hydrogen or deuterium, produced in the cell by the decomposition of water, to below the flammability limits (about 4% H_2). A Lectrodryer and Drierite column were placed on the air inlet and exit lines to prevent contamination of the D_2O fuel solution by H_2O.

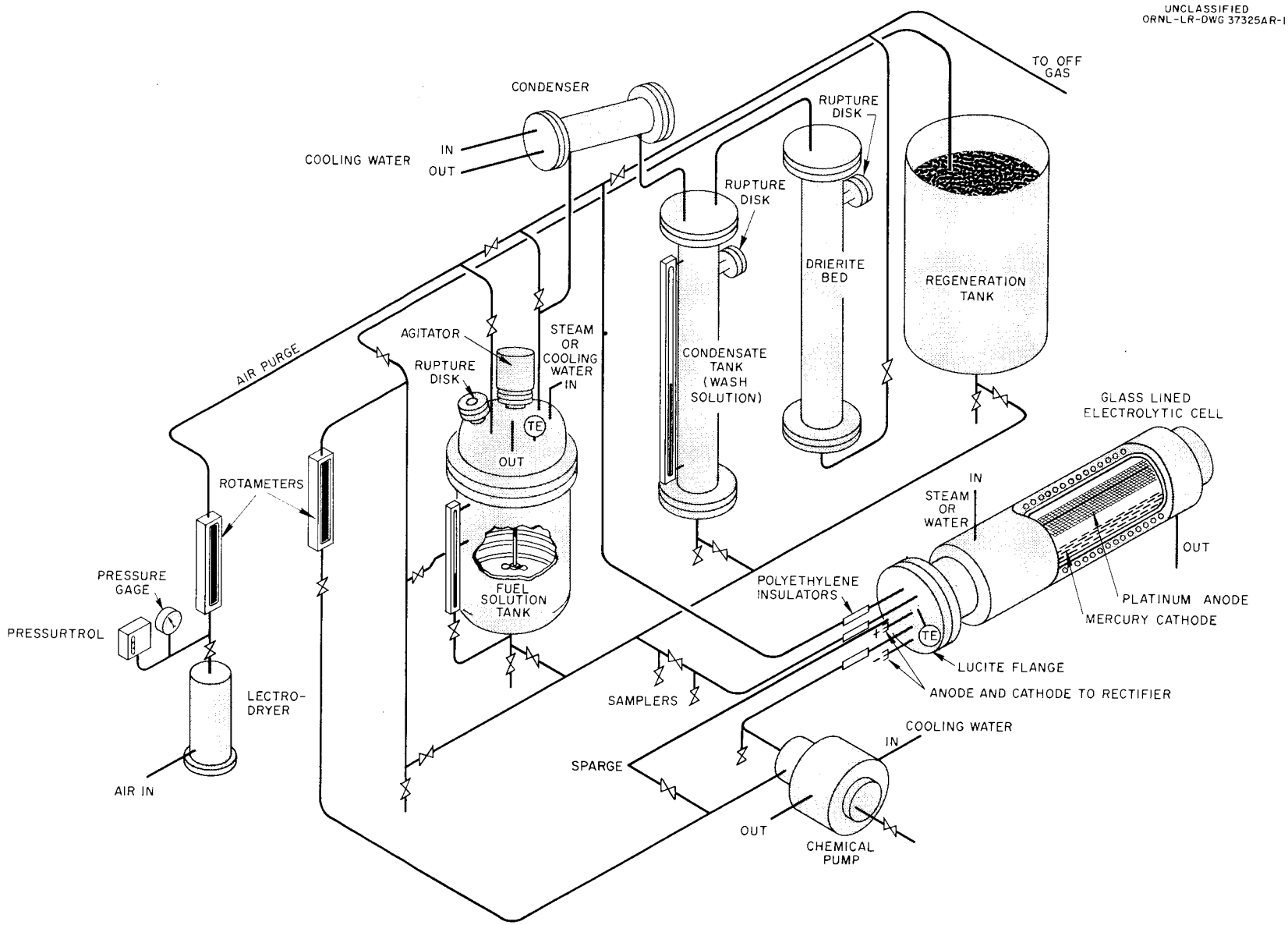

Fig. 1. Electrolytic nickel removal system.

A 55-gal drum was used as a storage tank for the mercury-regenerating solution. The Chempump used to circulate the fuel solution during nickel removal was also used to circulate the regenerating solution through the cell. During regeneration gases from the cell were diluted with air and vented directly to the off-gas system.

The electrolytic cell was constructed from a horizontal glass-lined tank 6 in. dia and 48 in. long. Over the end of the tank was placed a Lucite flange containing six 3/4-in. Swagelok fittings through which all electrical and piping connections to the cell were made (Fig. 2). The cathode was a pool of approximately 0.6 gal of mercury in the bottom of the cell, with an area of 1370 cm^2. The anode, located approximately 3 in. above the cathode, was constructed of 45-mesh platinum gauze supported by Lucite supports attached to the fuel inlet tube (Fig. 2 or ORNL Dwg. D-21343). The anode area was 1187 cm^2 (4 by 46 in.) in runs 1-5; for runs 6-9 the area was decreased to 297 cm^2 by folding the platinum screen. The electrical terminals for both the cathode and anode were made of 3/4-in. 347 stainless steel rods. Six-inch sections of 3/4-in. polyethylene tubing were used as electrical insulators in the solution lines to the cell. The nominal cell current and voltage were 200-300 amp and 6-20 volts. Direct current was provided by a silicon rectifier unit capable of an output of 400 amp at 25 volts.

Solution entered the cell at the rear and was discharged at the front, at the Lucite flange, just above the mercury level. In most runs part of the fuel solution (about 0.4 gpm) from the discharge of the circulating pump was introduced under the mercury cathode to provide agitation. The cell was jacketed to permit heating or cooling.

2.2 Procedure

The operating conditions used were those which had been found feasible in laboratory studies[4] and are tabulated, together with solution compositions, in the appendix. The fuel solution was circulated from its tank through the cell at about 1.5 gpm while a current of 200-300 amp was passed through the solution in the cell. The cell was maintained at essentially room temperature. The rate of nickel removal was determined by periodically sampling the contents of the fuel solution tank. Initial runs (1-3) to test equipment performance were made with a light-water solution containing nickel, copper, and manganese sulfates and sulfuric acid but no uranium. Runs 4-7 were made with a light-water fuel solution containing uranium in addition to the above components. Runs 8 and 9 were made with a heavy-water fuel solution of similar composition.

Although the system was designed to process 30-gal batches of concentrated solution, smaller batches were processed in most runs to decrease the operating time required. Runs 1, 3, 4, 6, and 7 were made on an intermittent basis, with the equipment being operated only during the day shift. Runs 2, 5, 8, and 9 were made on a continuous basis.

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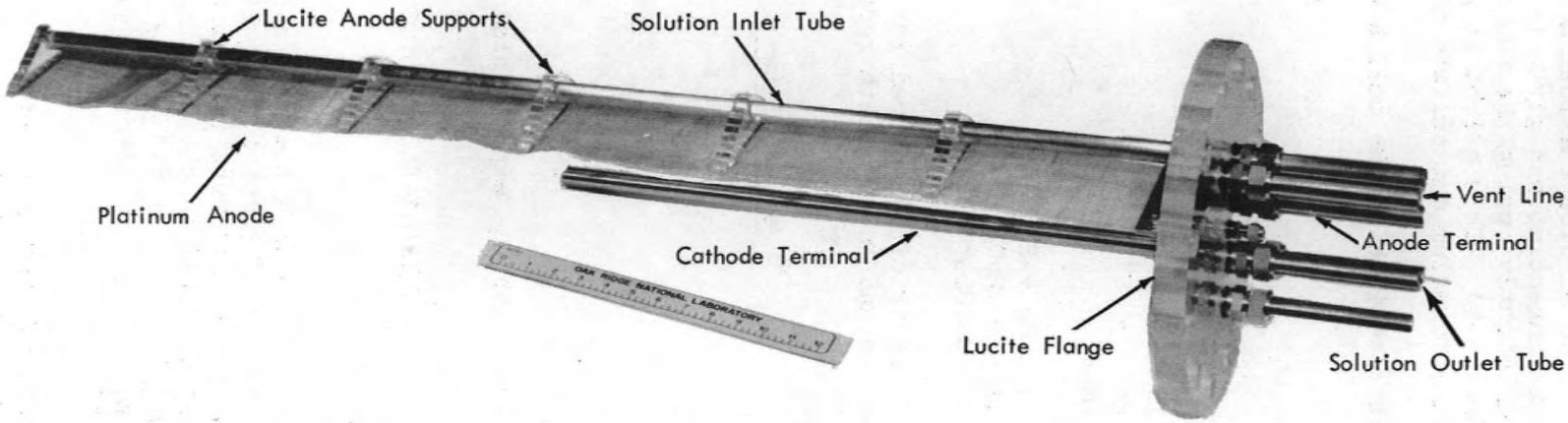

Fig. 2. Electrolytic cell flange and anode.

The amount of nickel removed from the fuel solution was slightly greater than that indicated by the concentrations given in the appendix since the solution was concentrated slightly during the run by evaporation and decomposition of the water (Sec. 3.7). This effect is also evident from the slight increase in concentration of uranium and sulfate ions in some runs.

Some of the probable reactions in the cell are:

	Reaction	Reaction Potential,[3] volts
Cathode:	$Ni^{++} + 2e \longrightarrow Ni$	-0.23
	$Cu^{++} + 2e \longrightarrow Cu$	+0.346
	$2H^{+} + 2e \longrightarrow H_2$	0.000
	$2D^{+} + 2e \longrightarrow D_2$	-0.044
	$UO_2^{++} + 4H^{+} + 2e \longrightarrow U^{4+} + 2H_2O$	+0.041
	$Mn^{++} + 2e \longrightarrow Mn$	-1.029
Anode:	$2OH^{-} - 2e \longrightarrow 1/2\ O_2 + H_2O$	
	$Mn^{++} + 2H_2O - 2e \longrightarrow MnO_2 + 4H^{+}$	
	$U^{4+} + 2H_2O - 2e \longrightarrow UO_2^{++} + 4H^{+}$	

The theoretical amount of electricity required for complete removal of nickel, copper, and manganese and for complete reduction of U(VI) to U(IV) is:

	amp-hr/liter	amp-hr/30 gal
0.045 M $NiSO_4$	2.4	273
0.27 M $CuSO_4$	14.5	1640
0.03 M $MnSO_4$	1.6	182
0.34 M UO_2SO_4	18.2	2062
	36.7	4157

These values may be compared with the experimental values reported below.

3.0 EXPERIMENTAL RESULTS

3.1 Rate of Nickel Removal

The amount of nickel removed from the fuel solution was approximately a linear function of the amount of electricity passed through the solution (Fig. 3). The removal of nickel accounts for approximately 1% of the current passed through the solution. Based on the data from runs 8 and 9, approximately 213 amp-hr/liter is required to remove 87% of the nickel from a heavy-water fuel solution. This is equivalent to a decrease from 250 ppm to 33 ppm in the original HRT-type fuel.

Comparison of the rate of nickel removal in runs 1 and 2 (Fig. 3a) in which there was no uranium in the solution with runs on uranium fuel solutions indicates that the uranium is responsible for a considerable decrease in the

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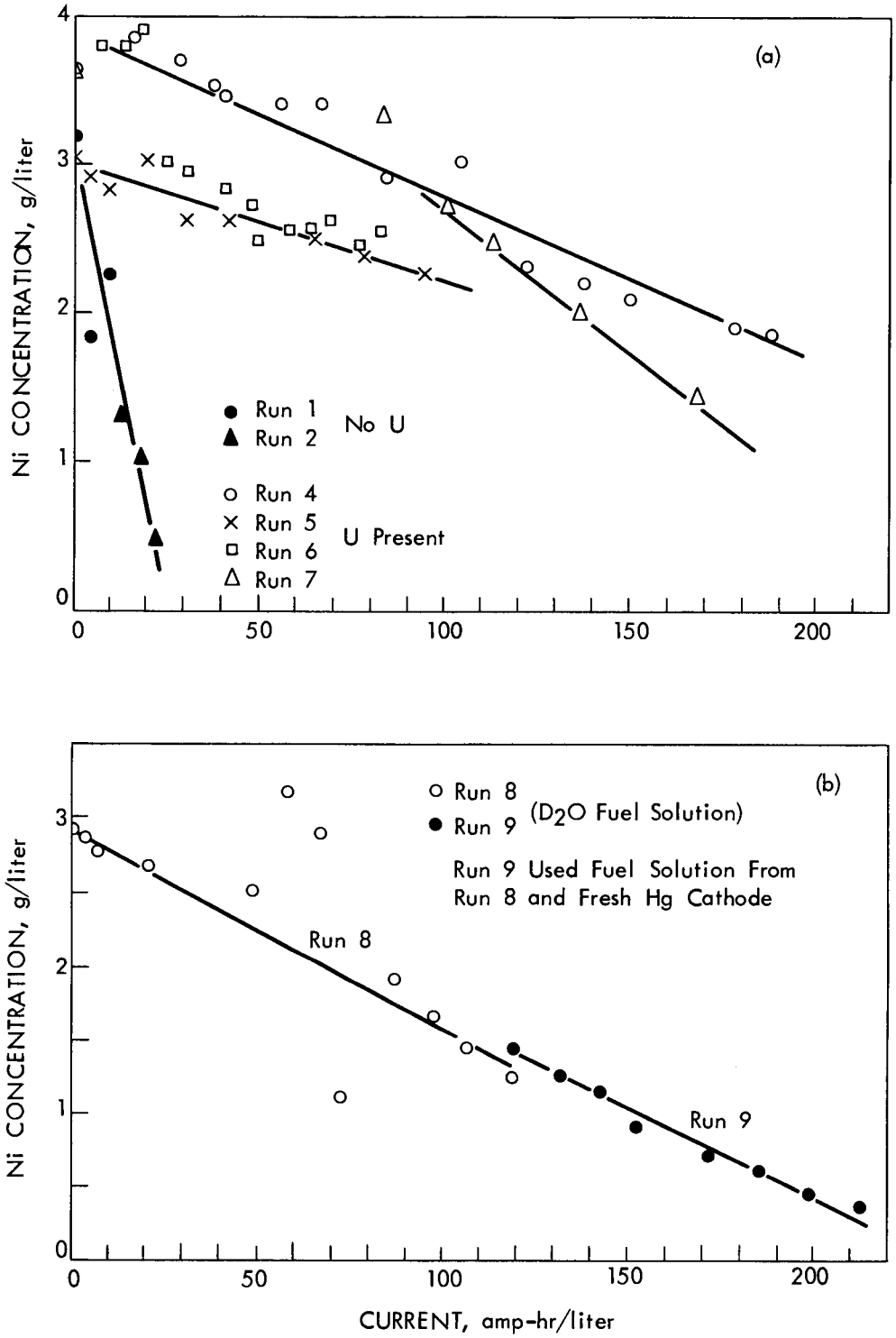

Fig. 3. Concentration of nickel in fuel solution as a function of the amount of electricity passed through solution. (a) H_2O solutions; (b) D_2O solutions.

rate of nickel removal. The considerably lower rate in run 3, in which no uranium was used, is thought to be due to an electrical short circuit in the cell.

At a cell current of 300 amp (the upper value used in the test cell), approximately 80 hr of operation would be required to remove 87% of the nickel from a 30-gal batch of heavy-water fuel solution, but this time might be considerably decreased with high cell currents. Mercury loadings of up to 5% deposited metals were obtained during the tests; however, the amalgam began to solidify at concentrations above about 1%. In order to determine whether the amalgam loading affected the removal of nickel in the cell, runs 7 and 9 were made with fresh mercury cathodes and fuel solution that had been processed in previous runs to remove most of the copper. The change from a cathode containing approximately 3.3% deposited metals at the end of run 8 to a fresh mercury cathode at the beginning of run 9 did not significantly change the rate of nickel removal (Fig. 3b). However, the nickel removal rate after the change to a fresh mercury cathode in run 7 was somewhat greater than that in run 6, in which the amalgam loading was about 5% deposited metal (Fig. 3b). This indicates that excessively high amalgam loadings may decrease the nickel removal rate.

3.2 Rate of Copper Removal

The rate of copper removal from the fuel solution was dependent on the concentration of copper in the solution (Fig. 4). The rate in the test cell for both light and heavy water in the range from 17 to approximately 0.1 g/liter can be expressed by

$$C = C_o e^{-0.058A}$$

where C and C_o are the final and initial copper concentrations, respectively, in grams per liter and A is the amount of electricity passed through the solution in ampere-hours per liter.

Since copper removal is dependent on the concentration of the solution, the efficiency of copper removal varies throughout a run. In run 8 the efficiency for copper removal varied from 57% during the first hour of operation to essentially zero as the copper content of the solution decreased. About 74 amp-hr/liter was required to decrease the copper concentration from 14.8 to 0.11 g/liter in run 8, in which a heavy-water fuel solution was used. Because of the higher removal rate, essentially all copper was removed from the fuel solution during the nickel removal process.

3.3 Rate of Manganese Removal

Manganese was removed only after a considerable amount of the copper and nickel had been removed. The only runs in which appreciable amounts of manganese were removed from the fuel solution were run 2, on a fuel solution containing no uranium, and run 9, in which the heavy-water fuel solution from run 8 was further processed with a fresh mercury cathode. The manganese concentration during runs 8 and 9 as a function of the amount of electricity passed through the solution is shown in Fig. 5. Seventy-six per cent of the

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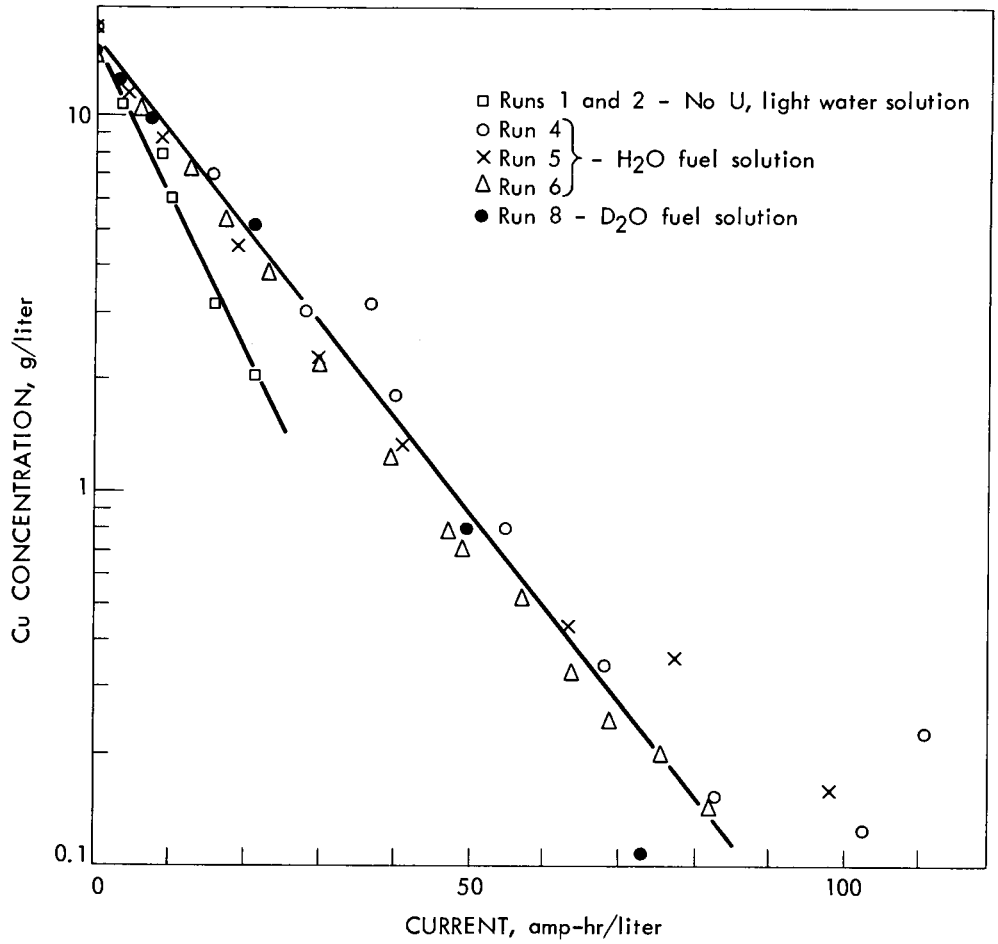

Fig. 4. Concentration of copper in fuel solution as a function of amount of electricity passed through solution.

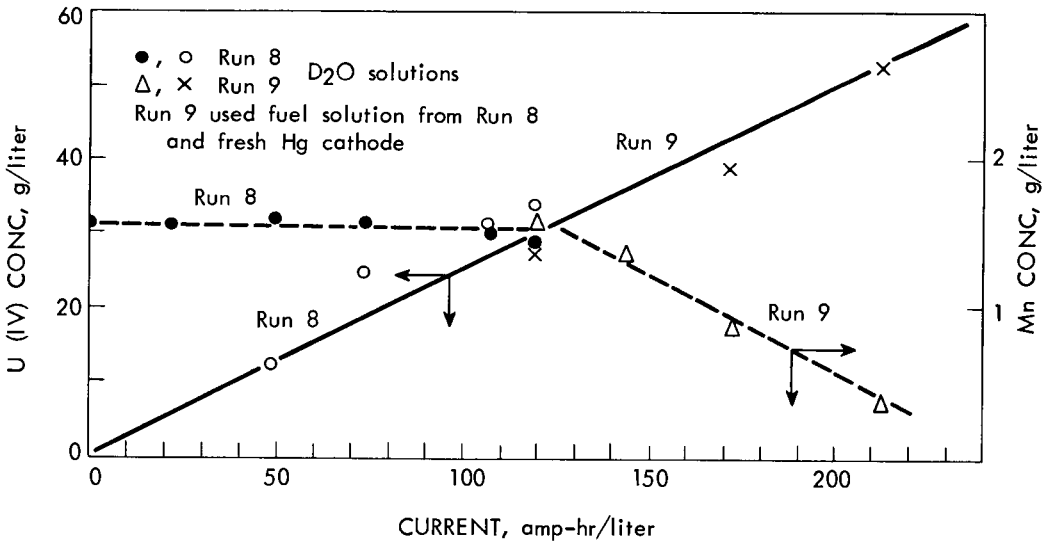

Fig. 5. Concentration of (---) manganese and (——) U^{4+} ions in fuel solution as a function of amount of electricity passed through solution.

manganese was removed during 213 amp-hr/liter of processing time in runs 8 and 9 as compared to 87% removal of nickel during the same period. However, since manganese removal was apparently initiated by the use of a fresh mercury cathode in run 9 (Fig. 5), the amount of manganese removed in a continuous run of the same duration as runs 8 and 9 might vary considerably from the value reported above.

3.4 Uranium Reduction

Uranium is reduced from the VI to the IV valence state at the cathode of the electrolytic cell during nickel removal. Since the UO_2^{++} ions are in competition with nickel ions for electrons at the cathode, it appears likely that nickel removal would be aided by the decrease in UO_2^{++} ions which occurs during operation as uranium is reduced. However, since the U^{4+} produced remains in solution, it may be reoxidized by contact with the anode. Thus the rate of U^{4+} buildup in the system is dependent on anode and cathode reactions, which may be influenced by their relative areas. In runs 4 and 5, in which an anode area of 1187 cm^2 was used, the U^{4+} concentration did not exceed 3-5% of the total uranium concentration. In run 6 the anode area was decreased to 297 cm^2 by folding the platinum screen while the cathode area remained constant. This resulted in U^{4+} concentrations of 19% of the total uranium. In runs 8 and 9, in which a heavy-water fuel solution was used along with the smaller anode, 61% of the uranium was reduced in the two runs (Fig. 6). A considerably higher percentage, > 90%, of the uranium was reduced in laboratory studies.[4] Because of the effect of other variables, the effect of anode area has not been evaluated directly. Increased current efficiencies should result from less reoxidation of U(IV) to U(VI) for smaller anode areas.

3.5 Mercury Entrained in Fuel Solution

Although the fuel solution discharge line from the cell was only about 0.25 in. above the mercury cathode, very little difficulty was encountered with mercury entrainment. Only infrequently were drop amounts of mercury observed in the fuel solution system outside the cell although no mercury de-entrainer was used in the system.

Analysis of fuel solutions for dissolved mercury showed:

Run No.	Final Hg Conc of Fuel, g/liter	Run No.	Final Hg Conc of Fuel, g/liter
4	0.026	7	<0.005
5	0.07	8	<0.004
6	<0.004	9	<0.004

A possible source of dissolved mercury in the fuel solution is mercury droplets deposited on the anode either during filling and assembly of the cell or during circulation of the fuel solution. However, mercury dissolved anodically should be redeposited at the cathode. As shown above, a substantial decrease in the mercury content of the solution between runs 5 and 6 corresponded to the decrease in anode area which was made between these runs.

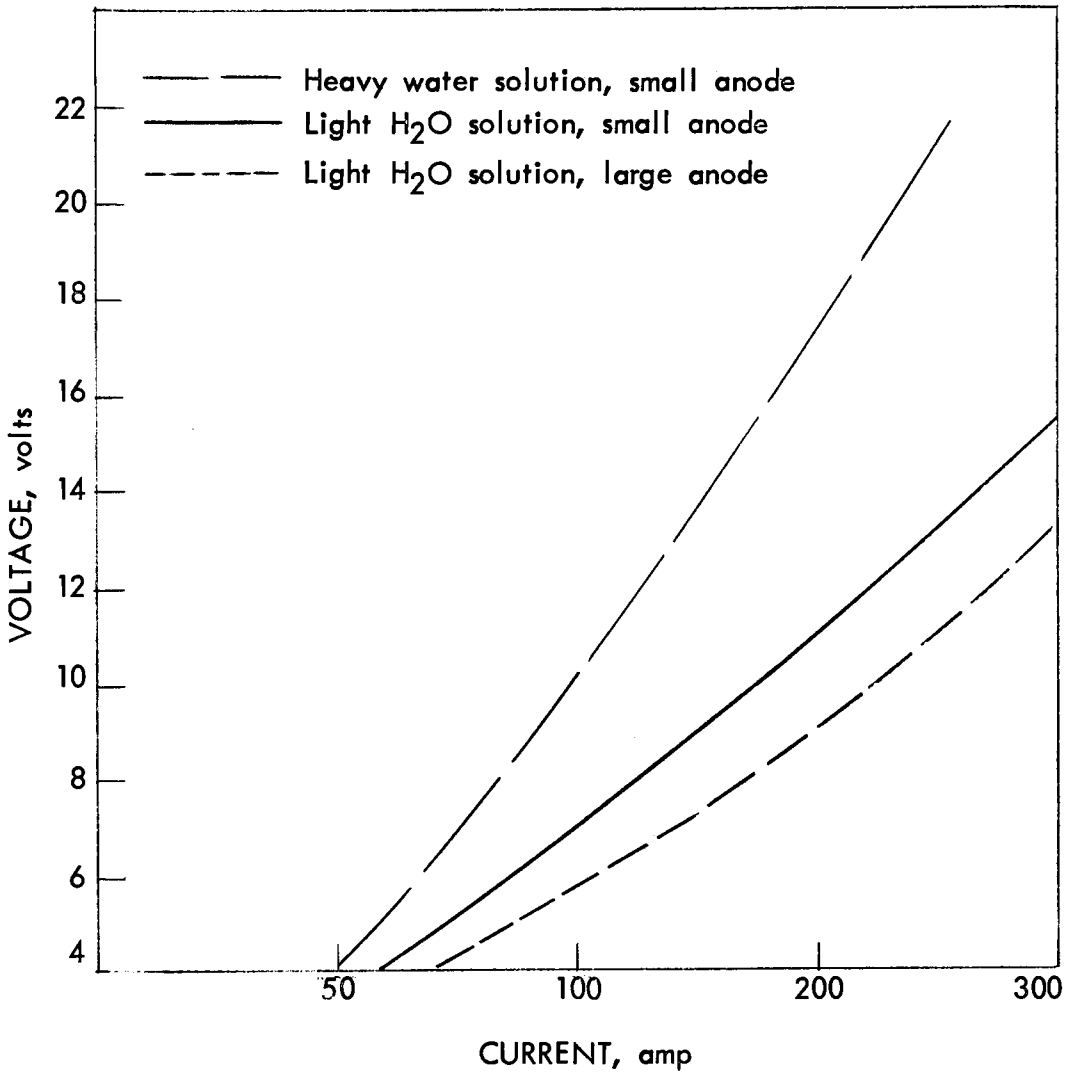

Fig. 6. Typical electrical conductance in electrolytic nickel removal cell.

3.6 Uranium Losses

Uranium can be lost from the system either by being deposited in the mercury cathode or as a result of incomplete washing of the fuel solution from the cell prior to the mercury-regeneration step. The amount of uranium lost to the regeneration solution was studied in run 6. The cell had previously been washed three times with about 5 gal of condensate produced by boiling the fuel solution. Of the 2 g of uranium found in the regenerating solution from this run, approximately 1 g can be accounted for by uranium deposited in the amalgam, which was decreased from 0.003% to 0.0008% during regeneration. The remaining gram of uranium apparently resulted from incomplete removal of fuel solution from the cell.

Typical amounts of uranium deposited in the mercury cathode for other runs were:

Run No.	Final U Conc of Amalgam, wt %	U Loss, g
7	0.0120	4.0
9	0.00016	0.05

3.7 Heavy Water Losses and Dilution

Heavy water may be lost from the system by decomposition during electrolysis and by evaporation by the air purge used to dilute the off-gas. The heavy water loss by decomposition may be as high as 0.37 g/amp-hr if all the current is used in decomposing water. However, the fraction of the current used in water decomposition varies with the solution composition during electrolysis, as evidenced by the change in gas composition during run 4 from an initial composition of 92.4% O_2 - 7.6% H_2 to 41.5% O_2—58.5% H_2 after most of the copper had been removed from solution. The gas evolution rate for a heavy-water fuel solution, measured at the end of run 9 with a wet-test meter, was 0.5 liter/min at a current of 250 amp (39.6% O_2, 60% D_2). This rate would result in a loss of about 15 ml/hr of D_2O, or about 400 ml during the run.

A total of about 3 liters of D_2O was lost during run 9, indicating that a substantial fraction was lost by evaporation to the air purge. During run 8 a total of 6 liters of D_2O was lost during the nickel removal step. The D_2O lost by evaporation might be considerably decreased by modification of the air purge system to eliminate contact of the purge stream and the fuel solution or by adding a moisture collection system to the exit gases.

3.8 Electrical Conductance of the Solution

The only outward difference in performance of the cell with light- and heavy-water fuel solution was the lower electrical conductivity of the D_2O solution (Fig. 6). In general, conductance of the solution increased with increasing temperature and increased as the copper and nickel ions in the solution were replaced by hydrogen ions during the course of the run.

3.9 Mercury Regeneration

In the operation of the cell it is desirable to remove the deposited metals from the mercury cathode after each run without removing the mercury from the cell. In preliminary tests the mercury was regenerated by circulating 1 M HNO_3—1.5% H_2O_2 through the cell. Application of a small reverse current during the regeneration increased the rate of dissolution of the nickel and copper.

The increase in concentration of nickel and copper (the chief components to be removed from the mercury) in the regenerating solution is shown in Figs. 7 and 8, respectively. The steep portions of the curve indicate the time during which the solution was circulated. During the remaining time the solution remained stagnant above the amalgam in the cell. Typical copper removal rates when the fuel solution was circulated were 11 g/hr. At this

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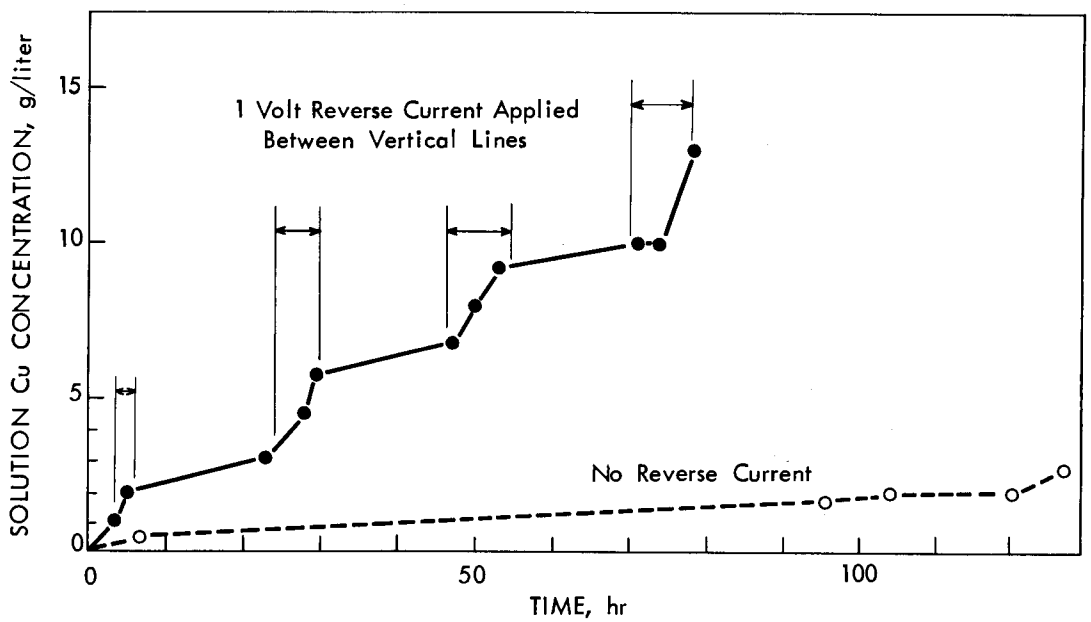

Fig. 7. Copper concentration of regenerating solution (20 gal) vs time.

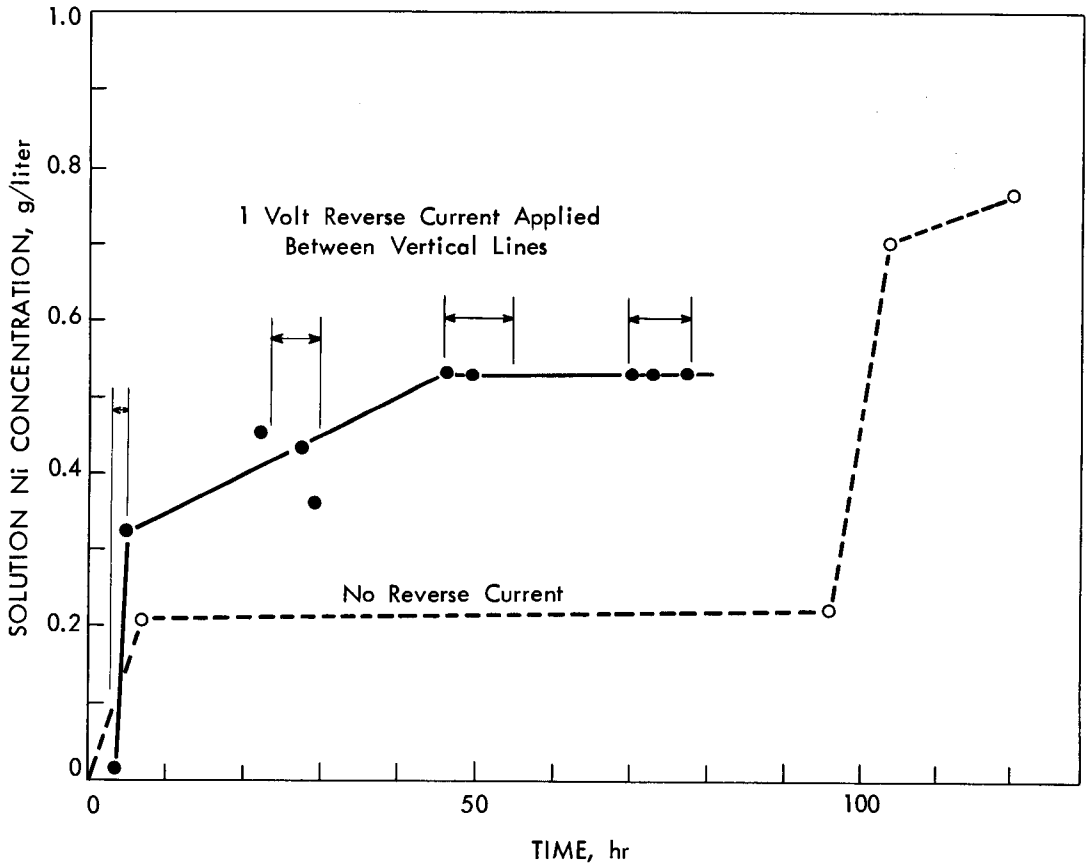

Fig. 8. Nickel concentration of regenerating solution (20 gal) vs. time.

rate about 180 hr would be required to remove the copper deposited from 30 gal of fuel solution. The nickel removal rate was approximately 3 g/hr, requiring about 100 hr for removal of the nickel deposited from 30 gal of fuel.

In the run in which the reverse current was applied the copper was removed from the cathode at a rate of about 29 g/hr, requiring approximately 70 hr to remove the copper deposited in the mercury from 30 gal of fuel. The concentration of copper in the mercury was decreased from about 4% to 0.44% and the nickel from 0.2% to 0.002%. Because of the higher concentration of copper in the amalgam, it is the limiting component in the regeneration step.

In both cases there was considerable dissolution of mercury by the regenerating solution, with up to 10% of the mercury being dissolved. Further testing of the regeneration system is required to determine its effectiveness under repeated loading and regeneration cycles.

3.10 Equipment Performance

The performance of the equipment in the test system was generally satisfactory. The chief difficulties were caused by solid amalgam formation at the cathode, which sometimes touched the anode, causing electrical short circuits and anodic corrosion of the stainless steel components in the vicinity of the anode. A minor limitation of the cell was the necessity of operating below 300 amp to prevent the electrical terminal from becoming hot enough (above 120°C) to damage the Lucite flange over the face of the cell.

Short-circuiting of the electrical system in the cell by amalgam growth from the cathode to the anode occurred at a mercury loading of about 2% deposited metals in runs 1-3. By introducing a stream of fuel solution (about 0.4 gpm) from the circulating pump discharge under the mercury cathode to provide agitation, higher amalgam loadings were achieved and short circuits were largely eliminated. In run 6, thirty gallons of fuel solution was processed to give an amalgam loading of about 5% deposited metals without short-circuiting occurring. By agitation of the mercury through a pipe containing 1/16-in.-dia holes and extending under the mercury the length of the cell, the solid amalgam formations were spread more uniformly over the surface of the cathode, thus preventing large formations from touching the anode. Stagnant areas and areas in which pipes or other objects in the cathode extended near the surface of the mercury were particularly susceptible to amalgam growth. However, by careful cell design and by limiting the amalgam loading, short-circuiting should not present an operating difficulty in future cell designs.

Anodic corrosion was a serious problem where stainless steel components were attached directly to the anode, on the fuel solution inlet tube located about 1 in. above the anode, and on a thermocouple well placed between the cathode and anode. These components were made anodic by conduction through the fuel solution. Corrosion of the stainless steel components attached to the cathode permitted processing only about one batch of fuel solution before replacement was required. A platinum plate gave protection to the

stainless steel in one run but peeled off during the regeneration step, making the electrode subject to attack in subsequent runs. Coating the stainless steel with polyethylene largely eliminated the anodic corrosion, but because of the instability of plastics in a high radiation field a different material would be required in a cell for reactor operation.

The plastic electrical insulators used in the solution lines are also undesirable for use in a reactor system in which a high radiation field is present. In run 5 an attempt was made to decrease the number of lines entering the cell and eliminate the plastic insulators by introducing the fuel inlet stream under the mercury cathode and allowing the solution and gas to leave the gas through a single line at the top of the cell. However, it was necessary to replace the insulator in the solution exit line after about 2 hr of operation because of a large copper deposit on the line in the cell which threatened to short-circuit the cell electrical system. The design of a cell for reactor use would require special emphasis to prevent stray electrical current, which would cause either anodic corrosion or deposition of copper on lines where short-circuiting might occur.

No corrosive attack was noted on the stainless steel components outside the cell or the pipes that were submerged in the mercury cathode. The Carpenter 20 Chempump, equipped with aluminum oxide bearings, gave good performance throughout the test.

4.0 REFERENCES

1. Homogeneous Reactor Project Quarterly Progress Report for Period Ending April 30, 1959, ORNL-2743, p. 202.

2. Ibid. for Periods Ending April 30 and July 31, 1958, ORNL-2561, pp. 147-150.

3. "Handbook of Chemistry and Physics," 36th ed., pp. 1630-2, Chemical Rubber Publishing Co., Cleveland, Ohio.

4. R. A. McNees, ORNL, personal communication, April, May, June, 1959.

5.0 APPENDIX

Conditions, analyses, and results from nine experimental runs are summarized here in tabular form. These data were presented in graphical form and discussed in Sect. 3.0.

Table 1. Run Conditions

Run No.	Current Density, amp/cm^2 Cathode	Current Density, amp/cm^2 Anode	Avg. Cell Voltage, volts	Avg. Cell Current, amp	Avg. Cell Temp, °C	Vol, gal Fuel Solution	Vol, gal Hg Cathode
1	0.15	0.17	10	200	20	25	0.63
2[a]	0.15	0.17	10	200	20	25	0.63
3	0.15	0.16	6	212	20	25	0.63
4	0.16	0.19	10	223	30	15	0.63
5	0.16	0.19	10	226	25	22	0.66
6[b]	0.16	0.73	10	218	30	30	0.64
7[a]	0.18	0.84	13	250	35	10.6	0.65
8	0.15	0.71	19	212	30	20	0.71
9[a]	0.18	0.81	17	241	30	19	0.66

[a] Used fuel solution from preceding run with fresh mercury cathode.

[b] Feed prepared by adding $Cu(OH)_2$ and $NiSO_4$ to solution from runs 4 and 5.

Table 2. Fuel Solution Analyses

Run No.	Current, amp-hr	Solution Concentration, g/liter Cu	Ni	Mn	Total U	U(IV)	Hg	SO_4, M	D_2O, %
1	0	14.6	3.16	1.55	0	0	0		0
	310	10.9	1.82	1.43					
	771	7.88	2.25	1.43					
2	0	7.88	2.25	1.43	0	0	0		0
	203	6.05	1.33	1.37					
	748	3.30	1.00	1.03					
	1236	2.03	0.486	0.89					
3	0	13.5	1.09	1.54		0	0	0.57	0
	730	12.0	1.07	1.60					
	1511	10.1	0.86	1.60					
	2347	9.0	0.72	1.60					
	3150	7.2	0.74	1.63					
4	0	17.3	3.64	1.85	97	0	0	1.00	0
	873	7.00	3.88	1.88					
	1567	3.00	3.69	1.89					
	2050	3.20	3.5	1.94					
	2224	1.8	3.45					1.00	
	2728	0.8	3.40						
	3685	0.338	3.40						
	4635	0.150	2.92					1.00	
	5792	0.125	3.01	2.28				1.10	
	a	0.175	2.34	1.71				0.95	
	6779	0.225	2.31						
	7659	0.213	2.18					0.95	
	8313	0.150	2.10				0.065		
		0.150	2.14						
	9938	0.10	1.90	1.85					
	10046	0.125	1.86	1.88		5.0	0.026	1.07	
5	0	17.13	3.04	1.82	82.9	0	0	1.02	0
	350	11.63	2.92						
	740	8.70	2.82						
	1570	4.50	3.01	1.60					
	2471	2.25	2.62					0.96	
	3388	1.31	2.62	1.71			0.09		
	5289	0.46	2.50	1.77					
	6429	0.33	2.41	1.77					
	7791	0.20	2.29	1.82	85.0	2.2	0.07	1.01	

[a] Water added to make up for loss by evaporation and decomposition.

Table 2. Continued

Run No.	Current, amp-hr	Solution Concentration, g/liter Cu	Ni	Mn	Total U	U(IV)	Hg	SO_4, M	D_2O, %
6	0	15.3	3.64	1.71	83.2	2.5	0.028	1.07	0
	675	10.4	3.78						
	1406	7.3	3.78	1.71					
	1919	5.35	3.88						
	2572	3.8	2.98						
	3377	2.2	2.95					1.08	
	4467	1.22	2.83						
	5295	0.78	2.70	1.73	81.8	14.0	<0.004	1.11	
	5549	0.70	2.48	1.77	83.5	20.9		1.12	
	6427	0.52	2.55	1.80					
	7324	0.33	2.55	1.82				1.15	
	7849	0.245	2.62	1.87					
	8575	0.205	2.43	1.87					
	9255	0.145	2.55	1.88	76.3	14.5	<0.004	1.16	
7	0	0.285	3.33	2.11	96.8	17.5		1.30	0
	713		2.72						
	1193		2.48						
	2223		2.04	2.28		25.8			
	3343	0.045	1.46	2.35	104	25.9	<0.005	1.38	
8	0	14.8	2.92	1.54	79			0.89	90
	282	12.3	2.87						
	558	9.8	2.79						
	1557	5.1	2.68	1.54					
	3790	0.8	2.53	1.60	81	12.4	<0.005	0.89	
	4460		3.16						
	5145		2.92						
	5605	0.108	1.12	1.55		24.8			
	6753		1.94						
	7453		1.65						
	8158		1.46	1.48	84.5	31.0	<0.002	0.95	
	9123	<0.01	1.26	1.44	84.2	36.9	<0.004	0.95	
9	0	0.160	1.49	1.55	83.6	27.1	<0.004	0.95	86
	893		1.26						
	1843	0.038	1.13	1.36					
	2828		0.92						
	3821	0.010	0.73	0.855	87.4	38.8		0.98	
	4814		0.63						
	5788		0.46						
	6829	<0.01	0.38	0.371	88.0	53.9	<0.004	1.00	

Table 3. Amalgam Analyses

Run No.		Amount, wt %				
		Cu	Ni	Mn	Total U	Hg
1	Final conc.	0.92	0.18	0.007		97.6
2	Final conc.	1.69	0.28	0.16		97
4	Initial conc.	0.19	<0.005			99
	After 5792 amp-hr	1.89	0.007	<0.01	0.0002	97.6
	Final conc.	2.53	0.12	<0.005	0.0006	96.95
6	Initial conc.	<0.01	0.0003			99.1
	After 5295 amp-hr	4.59	0.11	<0.005	0.003	95.1
	Final conc.	4.97	0.36			
7	Initial conc.	0.38	0.22		0.00012	97.9
	Final conc.	0.32	0.40	0.13	0.012	97
8	Initial conc.	0.01	0.0007			99.8
	Final conc.	2.98	0.33			
9	Initial conc.	0.38	0.025		0.009	99.2
	Final conc.	0.52	0.27	0.25	0.00016	98.1

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