BRIGHTON ED TESTING WITH ASAHI MONOVALENT SELECTIVE MEMBRANES

by

Dave Williams

D & N ENGINEERING
2108 S. Coors Cir.
Lakewood, Colorado 80228

Contract No. 1425-5-PG-81-08300

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Water Treatment Technology Report No. 14

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This report summarizes work done by Reclamation in pilot testing an electrodialysis (ED) water treatment system with specia membranes tailored for nitrate removal from water. The pilot unit was tested at well no. 7 at Brighton, Colorado, a municipa water well with nitrate concentration ranging from 10 - 20 mg/L nitrate as nitrogen. The testing toolc place between late 199: and early 1994. Product water from the ED unit had a nitrate concentration of 7.5 mg/L as nitrogen, well under the MCL o. 10 mg/L. Water recovery was 91 percent. Feed water was filtered with a 10 micron cartridge filter prior to the ED system Small quantities of sulfuric acid were added to control pH of the cathode wash streams to prevent scaling. No other scale inhibitors or chemicals were added. This report details how initial flowrates and process settings were established, hou automatic control functioned, operational issues, and chemistry results from sampling done on 1/24/94.

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Bureau of Reclamation
Mission Statement

The mission of the Bureau of Reclamation is to manage, develop, and protect water and related resources in an environmentally and economically sound manner in the interest of the American public.

U.S. Department of the Interior
Mission Statement

As the Nation's principal conservation agency, the Department of the Interior has responsibility for most of our nationally-owned public lands and natural resources. This includes fostering sound use of our land and water resources; protecting our fish, wildlife, and biological diversity; preserving the environmental and cultural values of our national parks and historical places; and providing for the enjoyment of life through outdoor recreation. The Department assesses our energy and mineral resources and works to ensure that their development is in the best interests of all people by encouraging stewardship and citizen participation in their care. The Department also has a major responsibility for American Indian reservation communities and for people who live in island territories under U.S. Administration.

Disclaimer

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Ed Burke	Director of Utility Operations, City of Brighton

Special recognition belongs to the City of Brighton for their donation of pilot plant space and invaluable assistance on the project.

Table of Contents

Glossary

All concentrations (mg/L of nitrate as nitrogen)

C_F	Feed steam concentration mg/L of nitrate as nitrogen
C_P	Product stream concentration
C_{waste}	Waste stream concentration (same as anode wash outlet stream)
C_{Ci}	Concentrate inlet stream concentration
C_{Co}	Concentrate outlet stream concentration
C_{Di}	Diluate inlet stream concentration
C_{Do}	Diluate outlet stream concentration

Flowrates

Q_{FD}	**Flowrate** of feed stream to diluate tank
Q_{FC}	**Flowrate** of feed stream to concentrate tank
Q_F	**Flowrate** of total feed stream
Q_P	**Flowrate** of product stream
Q_{Waste}	**Flowrate** of waste stream
Q_{Di}	**Flowrate** of diluate inlet stream
Q_{Do}	**Flowrate** of diluate outlet stream
Q_{Ci}	**Flowrate** of concentrate inlet stream
Q_{Co}	**Flowrate** of concentrate outlet stream

Nomenclature

FB	Feed and Bleed process for electrodialysis
PFD	Process Flow Diagram
ED	Electrodialysis
P&ID	Piping and Instrumentation Diagram
NIST	National Institute of Standards and Technology

1 Executive Summary

This report summarizes work done by the Bureau of Reclamation (Reclamation) in pilot testing an electrodialysis (ED) water treatment system with special membranes tailored for nitrate removal from water. The pilot unit was tested at well no. 7 at Brighton, Colorado, a municipal water well with nitrate concentration ranging from 10 - 20 mg/L nitrate as nitrogen. The testing took place between late 1993 and early 1994.

Product water from the ED unit had a nitrate concentration of 7.5 mg/L as nitrogen, well under the MCL of 10 mg/L. Water recovery was 91 percent. Feed water was filtered with a 10 micron cartridge filter prior to the ED system. Small quantities of sulfuric acid were added to control pH of the cathode wash streams to prevent scaling. No other scale inhibitors or chemicals were added. The unit was run on an intermittent basis during the month of December 1993, but was run continuously for approximately one week (between 1/18/94 and 1/24/94) before samples were taken.

The City of Brighton donated space in their Reverse Osmosis (RO) Pilot Test Building, adjacent to well no. 7, for testing of the ED process. The city currently treats water using a RO system that achieves 80% water recovery using a scale inhibitor. Since Brighton water was also hard, there were additional water quality issues the city considered in choosing the RO process. The ED process used did not achieve the same level of mineral removal as RO, but the ED process is optimized for nitrate removal. The ED stack and membranes were built by the Asahi Glass Co. of Japan. The stack was run on a modified feed and bleed process. It has one electrical stage that does not reverse polarity.

This report details how initial flowrates and process settings were established, how the system was controlled automatically, operational issues, and chemistry results from the sampling done on 1/24/94. The engineering work was done by Dave Williams while employed as a chemical engineer at the Bureau of Reclamation. It is the author's opinion that this process should be considered as a competitive technology with other treatments when the principal contaminate is nitrate.

2 Objective

Reclamation's initial objective for this project was to establish field experience with this type of water treatment, and to compare the process with the other nitrate removal technologies such as RO and ion exchange. If this technology proves promising, Reclamation would consider Asahi ED when searching for treatment alternatives for other small communities. In 1993, as part of the Small Communities Desalting program, Reclamation funded a small project to purchase and test an ED unit from Asahi. One goal was to acquire enough information and experience to allow Reclamation to estimate treatment costs for small community ED systems. Therefore, much of the work done at Brighton was geared toward process modeling, predicting operation, and verifying these predictions. Both the author and supervisor Larry Haugseth agreed that it was important to develop design understanding and techniques, not just to demonstrate operation of the process.

Initially the ED pilot system was assembled in Denver, and water was trucked in from Brighton for testing. But it was found that nitrate concentrations declined relatively quickly with time, and more long term operation with consistent feed concentrations were required to get accurate results. The City of Brighton, Colorado, located north of the Denver metro area, graciously offered space in their RO pilot test facility at the corner of 4th and Jesup Streets in Brighton.

The long-term objective was to obtain data necessary to compare cost of the existing RO system at Brighton with the Asahi ED process. Although this is a worthy objective, it is beyond of the scope of this report. The objective of this report is to document the work done on the project and offer design guidelines to use when pilot testing an ED system.

3 Design of Pilot Plant

3.1 Principles of Feed and Bleed and Asahi ED Information

3.1.1 Asahi ED Stack Specifications

The DB-0 electrodialysis stack was purchased by Reclamation for $10,500 US on **9/24/93,** and has the following specifications:

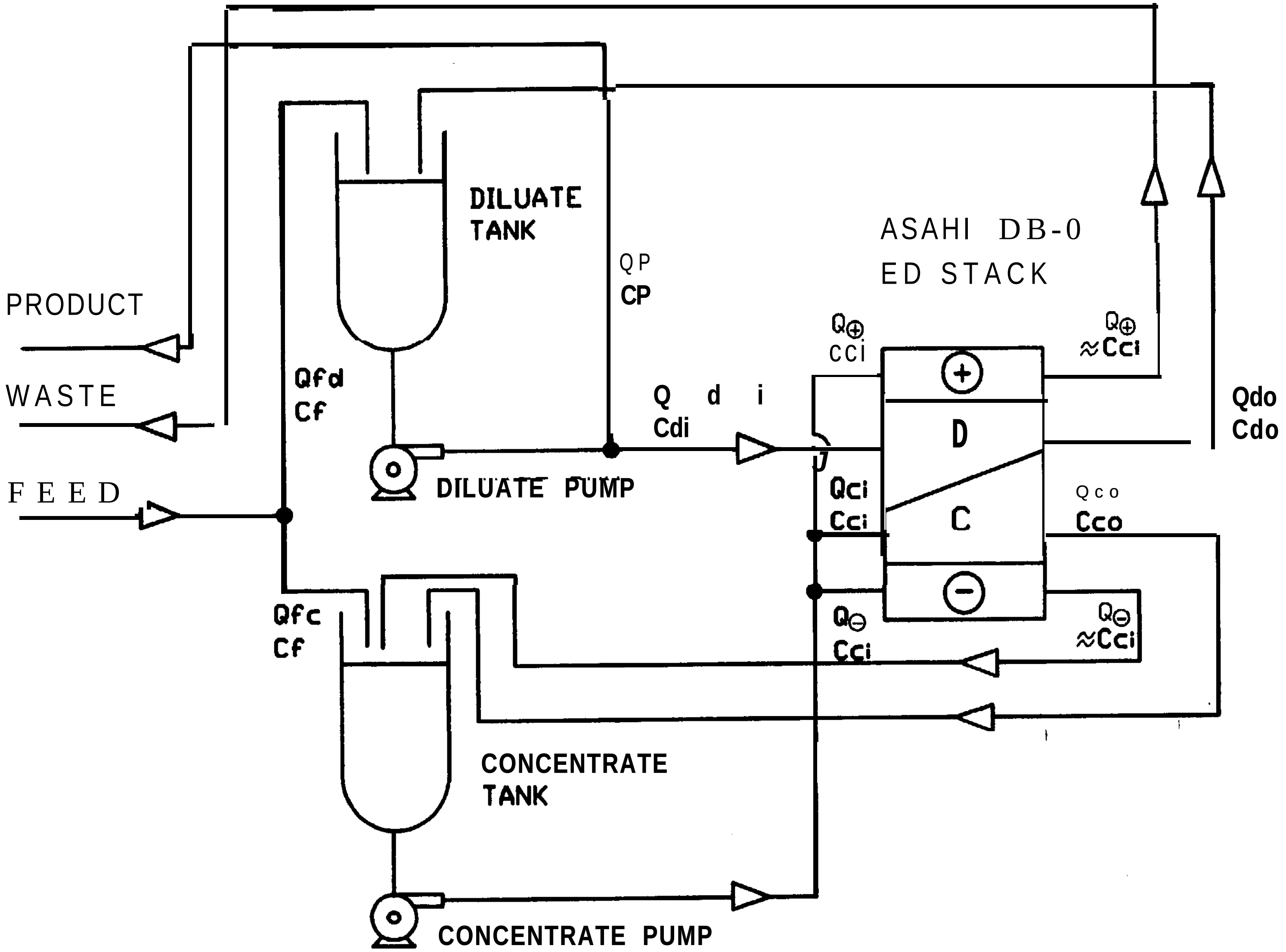
Membrane types:	ASV(monovalent selective anion exchange) CMV(cation exchange)
Membrane size:	7.1 inches x 21.7 inches **(18x55cm)**
Effective Membrane **Area:**	79 sq. inches (0.051 m^2) per cell pair
Number of Cell Pairs:	100
Product Flow:	2 to 6 gal/min (7.6 - 22.7 L/min)
Electrical Stages:	1

Although **Asahi** can provide a pilot unit complete with pumps and controls for pilot testing, due to limited funds, only the stack was purchased.

3.1.2 Feed and Bleed Process

There are numerous process flow configurations that can be used for ED systems [1], [2]. It is not the purpose of this report to explain the basics of ED, but the reader is referred to the many excellent papers on the subject. Reference [2] and the **Ionics** ED brochure [4] are

FIGURE 1 - PROCESS FLOW DIAGRAM



good places to start reading about ED.

Figure 1 illustrates the feed and bleed configuration used by Asahi, and adopted by the author for pilot studies at Brighton. Feed and bleed (FB) typically refers to using a single stage stack with continuous recycle of product water to achieve a desired salinity. Asahi utilizes a modified FB process where both concentrate and diluate are recycled. Some authors in [1] and [3] recycle only the product. Recycling both streams allows more control of the process. This can become especially important at high water recoveries (> 80%) to control scaling and polarization [2]. Other advantages of FB are that the stack operates at a low current density and the membranes operate under steady-state conditions [1]. Disadvantages of FB are higher electrical requirements than a multistage stack, more complicated piping and control for the various recycle streams [1], and high recirculation rates can cause water heating. It is important to note that when comparing power requirements between FB and multistage configurations, the multistage process should have lower power consumption than a FB process. Mintz [3] does a sample calculation in his paper comparing the power requirements for continuous process, batch recirculation, feed-and-bleed and internally staged stack. He concludes that the feed-and-bleed process consumes significantly more power than any of the other processes (up to 5 times as much power consumption). FB is used for pilot testing because it simulates staging or multiple stacks. It is especially important when scaling up a pilot study to a large commercial system not to confuse pilot FB power consumption to that of a large multistage stack design. However, data such as polarization and ultimate water recovery are quantities that can be used in designing large scale systems.

The process flow used for pilot studies was developed from drawing M2001-1 provided by Asahi. This drawing, included in the appendix of this report, specified major equipment and equipment capacities. The equipment data provided by Asahi was for maximum capacities only, so it was necessary to perform preliminary calculations before pilot testing to establish operating parameters for the specific system flow rates. The process flow diagram (PFD) shows the flows that were used while testing in the Reclamation Water Treatment Engineering and Research Pilot Plant Lab. The process stream nomenclature used here is maintained throughout the project.

3.1.3 Understanding the Process Flow Diagram and the ED Process

Various assumptions were made in determining the process flow settings.

- No significant amount of water transfers between compartments. Therefore, the flowrate of Q_{Di} is equivalent to Q_{Do}. Likewise, the flow rate Q_{Ci} is equivalent to Q_{Co}. Only minerals transfer from diluate compartment to the concentrate compartment.

- No significant ion transport occurs in the electrode wash streams. Although the pH can change, the majority of demineralization occurs through the diluate/concentrate compartments.

Both of these assumptions were verified in the field. To verify the first assumption, water was pumped through the concentrate compartment while the diluate lines were monitored for leakage. Then the same was done with the diluate compartment. A very small amount of water was observed to have leaked from one compartment to the other. Chemistry data was used to verify that insignificant amounts of demineralization occurs in the electrode cavity.

By inspection of the PFD it can be seen that two flowrates are equal,

$$Q_{FC} = Q_{+} = Q_{Waste}$$

and,

$$Q_{FD} = Q_{P}$$

The feed stream is split into two streams, and the feed stream to the concentrate tank is set to have a flowrate equal to the waste stream. The feed stream flowing to the diluate tank is set to have a flowrate equal to the product stream flowrate. If these conditions are not met, either tank will overflow.

Water from the concentrate tank is used to flush both electrodes (cathode - and anode +). Water from both diluate and concentrate compartments are recycled back to their respective tanks. To set flowrates from either the diluate pump or the concentrate pump, streams from the discharge side of each pump were returned back to the tank. This simple technique avoids costly variable speed motors and controllers. However, this method does skew the parasitic power measurements. These streams are not shown on the PFD but are shown in the process and instrumentation diagram (P&ID).

3.2 Calculations and Assumptions

The following is the procedure used by the author to set initial conditions for the DB-0 ED system. The material balances and concentrations refer to mg/L of nitrate as nitrogen. These equations and design decisions were made in Denver before testing began.

3.2.1 Establishing Known and Unknown Quantities

Usually the nitrate concentration of the feed stream is known. The designer would choose the quantity of product and waste, and choose the nitrate concentration of the product as a performance specification. The remaining parameters would be calculated, and are summarized below.

Given	Performance Spec	Quantity to be calculated
C_F (feed conc.)	Q_P (product flowrate)	I (stack current)
	Q_W (waste flowrate)	Q_D (diluate flowrate)
	C_P (product conc.)	Q_C (concentrate flowrate)
		C_{Ci} (concentrate inlet conc.)
		C_{Co} (concentrate outlet conc.)
		C_{Di} (diluate inlet conc.)
		C_{Do} (diluate outlet conc.)

Well no. 7 at Brighton was the feed water for the pilot studies. Nitrate concentration varies throughout the year at this site from 10 to 20 mg/L NO, as N, so the baseline feed value for design purposes was chosen to be 15 mg/L. Since the maximum contaminate level (MCL) for drinking water purposes is 10 mg/L, a product concentration of 5 mg/L appeared to be a reasonable value. Asahi specifications called for 2-6 gal/min (7.6-22.7 L/min) product flow, so a value of 4 gal/min (15.1 L/min) was chosen for product flow and to demonstrate a high water recovery, a 0.5 gal/min (1.9 L/min) was chosen for waste stream flow. This would give an 89% waste recovery for the system.

3.2.2 Calculating Diluate Flowrate Q_{Di}

The diluate flowrate Q_{Di}, can be found by writing a nitrate material balance on the diluate tank.

$$Q_{FD}C_F + Q_{Do}C_{Do} = Q_{Di}C_{Di} + Q_PC_P$$

$Q_{Di} = Q_{Do}$ from the first assumption, $Q_{FD} = Q_P$ and $C_{Di} = C_P$, from observation of the PFD,

$$Q_PC_F + Q_{Di}C_{Do} = Q_{Di}C_P + Q_PC_P$$

rearranging terms,

$$Q_PC_F - Q_PC_P = Q_{Di}C_P - Q_{Di}C_{Do}$$

gives a final result,

$$Q_{Di} = Q_P (C_F - C_P)/(C_P - C_{Do}) \tag{1}$$

All quantities are known or specified except C_{Do}, the concentration of the product water as it exits the stack. This concentration is a function of the electrical settings of the stack. C_{Do} was chosen to be 3 mg/L nitrate as nitrogen which, at first glance, may appear to be a random selection, but the following explanations show how this value was chosen.

Since ions move from the diluate compartment to the concentrate compartment, C_{Do} could not be larger than C_{Di} (or the product concentration). So, C_{Do} must be less than 5 mg/L. If this concentration was chosen to be a very small quantity, larger and larger electrical currents would be required. Because the electrical power is what moves the ions, to get more change in concentration across the diluate compartment would require more electrical power. This could mean a very large DC power supply. Also, concentration polarization becomes a problem with very dilute streams. On the other hand, if C_{Do} were chosen to be close to 5, this would equate to a very large diluate pump. Since it represents a very small change in concentration magnitude, large flowrates would be required to move a specific amount of ions. This can also be inferred by inspection of the design equation (1). After considering these two extremes, a midpoint value of 3 mg/L was chosen for C_{Do}. It may not be possible to have complete control of C_{Do} ; this concentration may not be attainable due to polarization of the water, but the value was used for design and initial testing.

Given that Q_P was specified to be 4 gal/min (15.1 L/min), and using design equation (1),

$$Q_{Di} = 4(15-5)/(5-3)$$

$$Q_{Di} = 20 \text{ gal/min } (75.7 \text{ L/mill})$$

3.2.3 Current Calculations

In the ED process the relationship between electrical current and the amount of minerals removed from the water is described by Faraday's Law. The author used an equation from **[2]**.

$$E = F \times F_t \, \Delta N/kI$$

where:
E = current efficiency
F^* = Faraday constant (96500 amp-sec/equiv.)
F_t = Total product flow (L/sec)
DA = change in feed concentration (equiv./L)
k = number product stream compartments
I = current (amperes)

Current was calculated to establish a target value. One assumption made was that the overall demineralization would follow the nitrate demineralization. It should be understood that this is an over simplification, since some ions are transferred more easily than others. But a target value is required before pilot studies and to size the DC power supplies.

Using Faraday's Law, the current was calculated for water sample H-3356 (appendix)

where: $\Delta N = C_{Di} - C_{Do} = C_P/3 - C_P/5$

assuming a 85% current efficiency and using 100 cell pairs.

$$Z = \frac{96500\ amp*sec}{eq} * \frac{10\ gal}{min} * \frac{3.785\ L}{gal} * \frac{1\ min}{60\ sec} * \frac{13.68x10^{-3}\ eq}{L} * [\frac{1}{3} - \frac{1}{5}] * \frac{}{.85 * 100}$$

I = 2.6 amps

Since the Asahi ASV membranes are monovalent selective, this current would be a worst case value. If no SO_4^{-2} anions were transferred, the current consumption would decrease. BY assuming that sulfate was not in the feed water, the electrical current was found to be:

I = 1.7 amps

In the field, the voltage was adjusted to attempt to reach 1.7 amps. When running an ED system one should be aware that this voltage may not be attainable due to polarization. Polarization could be occurring if the rate of increase in current with increasing voltage drops off. In addition, there is an absolute limit to the voltage that can be applied. Excessively high voltage can damage membranes [4]. For this project, no detailed polarization studies were performed except for field monitoring. To perform a polarization study for this water, a larger DC power supply would be required. In designing ED equipment for field studies, it is important that relatively large voltages (for this system-up to 400 volt) be available to measure polarization. The procedure in the field was to increase the voltage slowly while monitoring resistance (voltage/current) to maintain a linear relationship. Mason and Kirkham [2] indicate that due to the many factors affecting polarization, experimental determination is the preferred method for design.

3.2.4 Calculating Concentrate Flowrates and Concentrations Q_C, C_{Ci}, and C_{Co}

calculating c,:

A nitrate material balance done on the entire system gives:

$$Q_FC_F = Q_PC_P + Q_{waste}\ C_{waste} = Q_PC_{Di} + Q_{Waste}C_{Ci}$$

solving for C_{Ci},

$$(2) \qquad C_{Ci} = (Q_FC_F - Q_PC_P)/Q_{waste}$$

for Q_F = 4.5, Q_{Waste} = 0.5, Q_P = 4, and $C_P = C_F/3$,

$$C_{Ci} = [(4.5xC_F)-(4xC_F/3)]0.5$$

$$C_{Ci} = 6.33\ C_F$$

For a value of 15 mg/L feed, C_{Ci} would be 95 mg/L NO_3 as N.

Using the same reasoning as for the diluate outlet concentration, the concentrate outlet concentration should be larger than 95 mg/L. If C_{Co} is chosen to be close to 95 then a very large concentrate flowrate would be required to remove the ions. Also if C_{Co} was to be large, then one of the constituents could reach a limit of solubility (or scale). Using this reasoning C_{Co} was chosen to be 110 mg/L.

Calculating Q_{Ci}

Q_C is found from a material balance on the concentrate tank.

$$Q_{FC}C_F + Q_{Co}C_{Co} + Q_-C_{Ci} = (Q_+ + Q_- + Q\&a$$

rearranging and realizing that $Q_{Co} = Q_{Ci}$,

$$Q_{FC}C_F + Q_{Ci}C_{Co} = (Q_+ + Q_{Ci})C_{Ci}$$

gives a final result,

(3) $$Q_{Ci} = (Q_{FC}C_F - Q_+C_{Ci})/(C_{Ci} - C_{Co})$$

inserting values,

$$Q_{Ci} = [(0.5\ \text{gal/min}*15\ \text{mg/L})-(0.5\ \text{gal/min}*95\ \text{mg/L})]/(95\ \text{mg/L}-110\ \text{mg/L})$$

Q_{Ci} = 2.7 galhnin (10.2 Lhnin)

Both pump flowrates were within the maximum capacity specified by Asahi. These settings provide a framework for initial testing of the system.

3.2.5 Calculation Prediction Summary

Flowrates	Electrical Settings
Q_{Ci} = 2.7 galhnin (10.2 Lhnin)	I = 1.7-2.6 amps
Q_{Di} = 20 gal/min (75.7 L/min)	
Q_P = 4.0 gal/min (15.1 Lhnin)	
Q w = 0.5 galhnin (1.9 Wmin)	
Q_F = 4.5 gal/min (17.0 L/min)	
Q_- = 0.5 galhnin chosen (1.9 L/min)	

The current is achieved by adjusting the voltage upwards to reach the target amperage.

4 Pilot Plant

4.1 Description of Equipment

The P&ID drawing shows major equipment. No detailed bill of materials was generated by the author. Pumps were chosen by consideration of maximum capacities provided by Asahi. Most equipment was surplus and was salvaged from existing stock at Reclamation.

Diluate Pump	PVC case 1-1/12" x 1" discharge 1HP 120VAC single phase
Concentrate Pump	PVC case 1/2" suction x 1/2" discharge 3/4HP 120VAC single phase
Tanks	30 Gallon capacity, Plastic,
DC Power Supply	Sorenson model DCS-150-7 (150v/7A) HP Linear supply (unknown model)
Diluate Flowmeter	Pilot tube type F-300 Blue White Ind.
Concentrate Flowmeter and Electrode Wash Flowmeters	Rotameter type

The power supplies were wired in series to provide a higher voltage. Many of the process lines were simply looped into the tanks to make sampling easier. Feed water pretreatment was a 10 micron filter. Occasionally, due to the nutrient rich water, it became necessary to shock chlorinate with approximately 5 mg/L chlorine to kill bacteria growth. Bacteria growth was most prevalent when the stack was shut down for long periods of time (e.g. 1 week) with concentrated water still in the membranes. The pressure drop across the stack did not increase significantly over the period of operation. However, longer run times would be required to determine if bio-fouling of the membranes was a problem.

Since the DB-0 is not a reversing stack, acid was injected at the cathode (-) to control the basic reaction.

4.2 Control and Operation Issues

As previously mentioned, to maintain the levels in the diluate and concentrate tanks, the feed flowrate (Q_{FD} and Q_{FC}) must match the flowrate of the product and waste streams, respectively.

Because of variation in well pressure, it was difficult to maintain tank levels over long periods of time. To alleviate the constant adjustments, a control scheme was designed by the author to monitor tank overflow with a flowswitch. The switch was chosen such that small overflows would not actuate, but large flows (greater than about 4 gal/min) would shut down the system. System shutdown consisted of closing the 3/4 inch, normally closed (NC), inlet solenoid valve and turning off the pumps and rectifier. To monitor leaks and to protect the stack from a no flow condition, two pressure switches were installed on each pump. Also one pressure switch was utilized on the feed line to detect when the city well was shutdown. Loss of pressure from any of these three switches would shut down the system. The electrical logic schematic shows details of the shutdown and control logic. The shutdown logic preformed well without any problems.

Pressure gauges were installed on the inlet side of the stack to monitor pressure drop across the membranes. If inlet pressures were increasing, shock chlorination or acid descaling was used to clean the membranes. Asahi claims that these membrane can survive acidity in the range of 2-10 pH. Acid descaling was used when the unit was shut down for long periods of time and concentrated water had been allowed to stand in the membranes. During normal shutdowns, if the stack was flushed clean, no noticeable pressure drop occurred across the membranes and acid treatment was not required.

4.3 Acid Injection at the Cathode

As mentioned previously, acid was injected at the cathode to control pH at the concentrate tank to approximately 4-6. The positive displacement diaphragm pump (LMI) with frequency and stroke adjustment, was used to maintain a consistent injection quantity. This proved to require periodic attention and a closed loop method was considered by the author but not implemented. 93% sulfuric acid was diluted to about 30%, to make a more easily injected solution. It was difficult to control the pH with the 93% acid; also, the more dilute acid is safer to handle and minimizes the heat of dilution at the cathode. Acid injection was found to be 52 ml/hour when sampling was taken on 1/24/94.

4.4 Power Measurements

Electrical power to the stack was measured by a Fluke model 8060A digital multimeter (DMM) which was calibrated via a NIST traceable standard on 8/9/93 in the Reclamation electrical power lab. All voltage and currents to the stack were measured with this meter. Parasitic electrical power from pumps was not measured. Since flowrates and pressures are known, pump power could be estimated, but due to the large recycle rates involved in the FB process, it was decided by the author that it would not be a valid comparison to other processes. Also the small pumps employed are not as efficient as larger pumps that would be used in commercial systems.

FIGURE 2 - P&ID

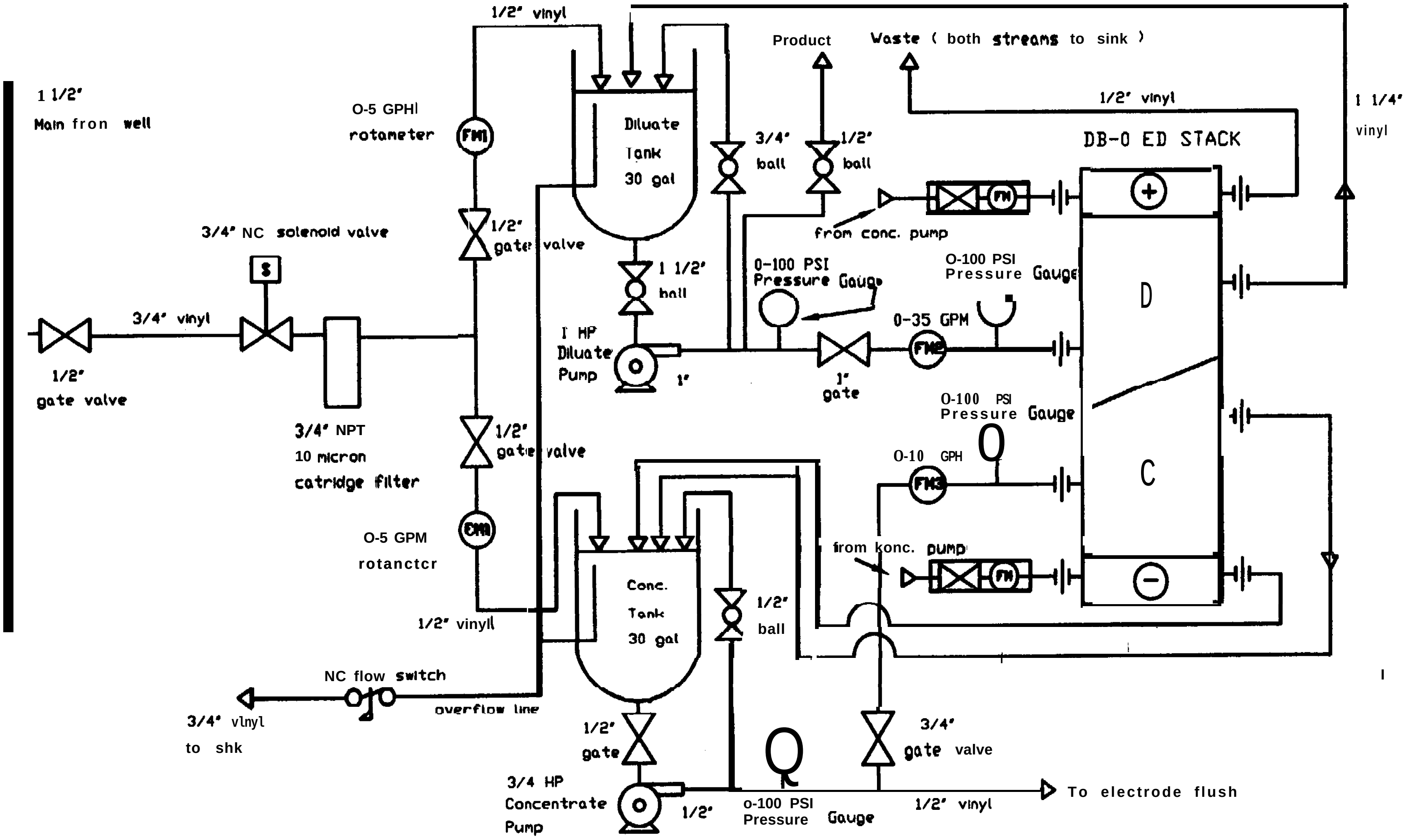
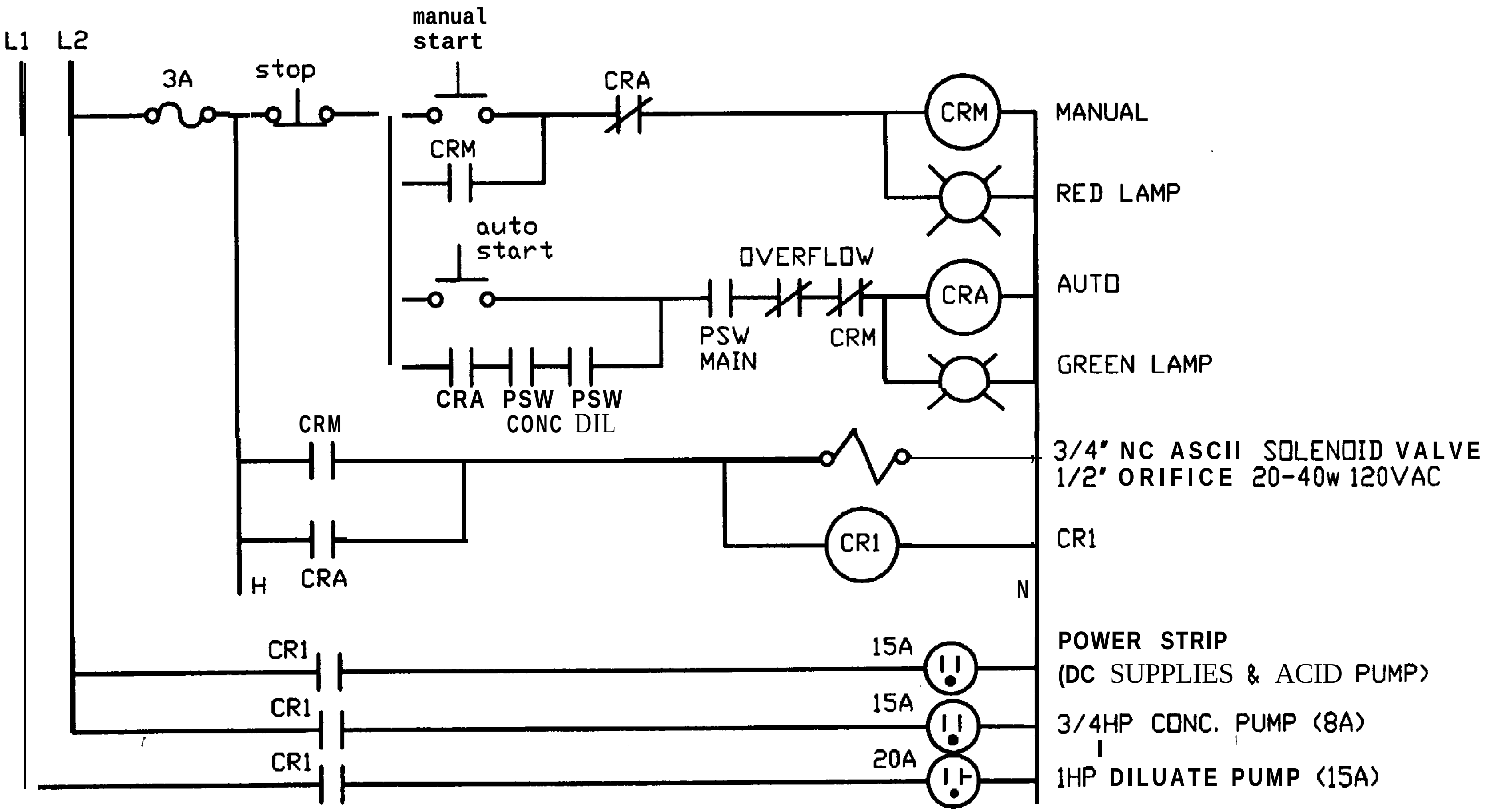

1/2" vinyl
1 1/2"
Main fron well
O-5 GPHI
rotameter
FMI
Diluate
Tank
30 gal
3/4"
ball
1/2"
ball
Product
Waste (both streams to sink)
1/2" vinyl
1 1/4"
vinyl
DB-0 ED STACK
+
from conc. pump
FM
3/4" NC solenoid valve
S
1/2"
gate valve
1 1/2"
ball
0-100 PSI
Pressure Gauge
O-100 PSI
Pressure Gauge
D
3/4" vinyl
1 HP
Diluate
Pump
1"
0-35 GPM
FM2
1"
gate
1/2"
gate valve
O-100 PSI
Pressure Gauge
3/4" NPT
10 micron
catridge filter
1/2"
gate valve
O-10 GPH
FM3
C
O-5 GPM
rotanctcr
FMI
from konc. pump
FM
Conc.
Tank
30 gal
1/2"
ball
−
1/2" vinyll
NC flow switch
overflow line
3/4" vinyl
to shk
1/2"
gate
3/4"
gate valve
3/4 HP
Concentrate
Pump
1/2"
o-100 PSI
Pressure Gauge
1/2" vinyl
To electrode flush

FIGURE 3 – ELECTRICAL LOGIC



NOTES: 1) PSW CONC. IS A PRESSURE SWITCH MOUNTED ON THE DISCHARGE OF THE CONCENTRATE PUMP

2) PSW DIL IS A PRESSURE SWITCH MOUNTED ON THE DISCHARGE OF THE DILUATE PUMP

3) OVERFLOW IS A NC FLOWSWITCH MOUNTED ON THE OVERLFLOW LINE FROM BOTH TANKS

4) 3/4' ASCO SOLENOID VALVE IS MOUNTED ON THE FEED LINE

5 Data

5.1 Compilation of Chemistrv Data

Major Cations/Anions (mg/L)

BOR Sample #	H-4698	H-4699	H-4701	H-4695	H-4696	H-4697	H-4700
Process stream	C_F	C_P	C_{Waste}	C_{Ci}	C_{Co}	C_{Do}	C-
Flowrates (gal/min)	4.68	4.25	0.433	2.6	2.6	20	0.67
pH	6.66	6.6	3.1	3.82	5.55	6.57	2.17
Conductivity (us/cm @25C)	1260	988	2950	2950	3190	923	7080
Suspended solids	0.357	0	0.877	0.531	0.862	**0.696**	**1.05**
Dissolved solids (180C)	884	653	1970	2220	2550	620	2410
Sum of ions	890	717	1720	1790	2070	660	2640
Ca^{+2}	112	73	307	341	394	58.7	357
Mg^{+2}	26.2	19	66.1	72.4	82.7	16.4	75.2
Na^{+}	127	116	173	188	202	103	194
K^{+}	4.23	3.55	10.8	10.6	11.9	4.42	10.9
CO_3^{-2}	0	0	0	0	0	0	0
HCO_3^{-}	310	243	0	0	94.3	226	0
so;*	210	213	781	787	810	215	1610
Cl-	100	49.1	381	392	472	36.2	395

Nitrate Results (mg/L)

BOR Sample #	H-4705	H-4706	H-4708	H-4707	H-4703	H-4704	H-4707
NO_3^- as N	5.9	7.5	65	73.8	87.3	5.55	76.4

Selected Metals (mg/L)

BOR Sample #	H4714	H-4715	H-4717	H4711	H4712	H-4713	H-4716
Ba^{+2} (ug/l)	253.2	62.59	241.6	253.2	294.1	54.23	262.1
Fe	49.67	9.112	55.62	49.67	51.97	11.49	127.4
Mn	<4	<4	<4	<4	<4	<4	<4
Sr	4478	954.7	4338	5254	5254	849.4	5007

5.2 Chemistrv Data Quality

Two standard solution samples were submitted along with other samples for quality assurance.

BOR Sample #	H-4709	H-4710
BOR lab results NO; as N(mg/L)	9.70	96.9
standard solutions (HACH)(mg/L)	10.0	100

6 Comparison of Data to Predictions

Table 1. Comparison of data with predicted values.

Description	Quantity	Prediction	Actual	Comments
Feed **Flowrate**	**Q_F (gal/min)**	4.5	4.68	calculated value
Feed Concentration	**C_F (mg/L)**	15	15.9	given by well
Dil. **Flowrate** Inlet	**Q_{Di}(gal/min)**	20	20	difficult to calibrate
Diluate (or Product). Concentration Inlet	**$C_{Di}=C_P$ (mg/L)**	5	7.5	
Diluate Concentration Outlet	**C_{Do} (mg/L)**	3	5.55	
Conc. Flowrate Inlet	**Q_{Ci} (gal/min)**	2.7	2.6	calibrated OK
Conc. Concentration Inlet	**C_{Ci} (mg/L)**	95	73.8	
Conc. Concentration Outlet	**C_{Co} (mg/L)**	110	87.3	
waste **Flowrate**	**C_+ (gal/min)**	0.5	0.433	calibrated OK
Cathode Wash **Flowrate**	**C_- (gal/min)**	0.5	0.67	calibrated OK
Product **Flowrate**	QP **(gal/min)**	4.0	4.25	calibrated OK
Stack Voltage	v (Volts)	?	300	maximum voltage
Stack Current	I (Amps)	1.7-2.6	1.14	more voltage required?
Stack Power	P (watts)	?	342	
Acid Injection			**52ml/hr**	
Diluate Pressure	**psig**	?	11	
Concentrate Pressure	**psig**	?	10	

6.1 Discussion

The electrical current calculation did not agree with the experimental value for two reasons. The calculation was performed making the assumption that all ions are removed with equal efficacy. Also, 300 volts was the maximum voltage attainable with the power supplies available. Higher voltages and higher corresponding currents may have improved product concentrations with the possibility of reaching the target of 5 mg/L nitrate as nitrogen. Of course, another way to improve the product concentration would be to change flowrates.

The nitrate concentrations on the concentrate streams, both inlet and outlet, appear to have lower values than predicted. However, all concentration calculations were based on a predicted product concentration of 5 mg/L. Since the resulting product concentration was found to be 7.5 mg/L, the other concentrations are expected to vary. The appendix shows the design equations calculated with actual values (see appendix).

When reviewing the design equations and material balances, it is important to keep in mind that the ideal conditions, $Q_{Fd}=Q_P$ and $Q_{Fc}=Q_{Waste}$ are not always met. In practice there is always some slight overflow of the tanks. It was found this was the preferred method of operation, to avoid any tank running empty. If the pressure of the well varied, the flowrate would change to each tank and eventually a tank would empty, shutting down the system on low pressure. This slight overflow condition introduces error in the material balances though. To maintain tighter control, an investment in better equipment, such as flow control valves and pressure regulators, would be required. Q_{Fd} and Q_{Fc} were measured during this project but not recorded; in the future it would be a good idea to record these values to help in resolving material balance errors. Other errors in the material balances could be attributed to variations in chemistry results and inaccuracies in flow measurements.

6.2 Current Efficiency Spreadsheet for Concentrate Compartment (Lotus 123)

Calculation of current for Asahi ED Stack
CONCENTRATE COMPARTMENT

Ions	**Equiv.Wt (grams/eq.)**	**C_{Ci} (mg/L)** H-4695	**C_{Ci} (meq/L)**	**C_{Co} (mg/L)** H-4696	**C_{Co} (meq/l)**
Ca^{+2}	20.00	341.00	17.05	394.00	19.70
Mg^{+2}	12.20	72.40	5.93	82.70	6.78
Na^{+}	23.00	188.00	8.17	202.00	8.78
K^{+}	[illegible]	[illegible]	0.27	[illegible]	0.30
sum of cations			31.43		35.57
CO_3^{-2}	30.00	0.00	0.00	0.00	0.00
HCO_3	61.00	0.00	0.00	94.30	1.55
SO_4^{-2}	48.00	787.00	16.40	810.00	16.88
Cl^{-}	35.50	392.00	11.04	472.00	13.30
NO_3^{-} as N			73.80		87.30
NO,‘ as **ION**	**62.00**	326.83	5.27	386.61	6.24
sum of anions			32.71		37.95
Normality(N)			32.07		36.76
TOTAL		2117.83		2453.51	

ΔN (meq/L) = -4.69 Flow paths= 100
Flow(gal/min) = 2.60 Cell pairs= 100

change in equivalents = **Flowrate(L/min)*ΔN (eq/L)**
=3.785*2.60*4.69/1000
=0.0462 eq./min

6.3 Current **Efficiency Spreadsheet** for Diluate **Compartment** (Lotus 123)

Calculation of current for Asahi ED Stack
DILUATE COMPARTMENT

Ions	**Equiv.Wt (grams/eq.)**	Product **(mg/L)** H-4699	Product **(meq/L)**	C_{Do} **(mg/L)** H-4697	C_{Do} **(meq/L)**
Ca^{+2}	20.00	73.00	3.65	58.70	2.94
Mg^{+2}	12.20	19.00	1.56	16.40	1.34
Na^{+}	23.00	116.00	5.04	103.00	4.48
K^{+}	39.10	3.55	0.09	4.42	0.11
sum of cations			10.34		8.87
CO_3^{-2}	30.00	0.00	0.00	0.00	0.00
HCO_3	61.00	243.00	3.98	226.00	3.70
so,”	48.00	213.00	4.44	215.00	4.48
Cl^{-}	35.50	49.00	1.38	6.20	1.02
NO_3^{-} **as N**		7.50		5.55	
NO_3^{-} as ION	62.00	33.21	0.54	24.58	0.40
sum of anions				10.34	9.60
Normality(N) (average)			10.34		9.24
TOTAL		749.76		684.30	

AN **(meq/l)** = 1.10
Flow(gal/min) = 20.00

Flow paths= 100
Cell pairs= 100

Change in **equivalents/min** = **Flowrate(L/min)*ΔN (eq/L)**
= **3.785*20.0*1.10/1000**
= 0.0833 **eq/min**

Averaging the concentrate and diluate compartments:

Average equivalents **transferred/min** = **(0.0462+0.0833)/2** = **0.0648eq./min**

E = **Faraday(amp-sec/eq)*Flowrate(L/sec)*ΔN(eq/L)/current (amps)/flowpaths**
= **96500*0.0648/60/1.13/100**
E = 0.922

7 Conclusions and Recommendations

Other Data and Samples

Water samples were taken after initial startup in Denver, but the author has omitted discussing results. While being tested in Denver, the stack removed nitrate as expected, but the author does not believe the unit was run long enough with consistent feed concentrations to provide useful information. Reclamation continued to run the pilot system in Brighton after the author was no longer employed at Reclamation, and that data has not been considered in this report.

Current Efficiency

The current efficiency, calculated in section 6.2, was found to be 92%. This value was derived by averaging the current efficiencies between the concentrate and diluate compartments. This method was chosen since the efficiency is sensitive to slight variations in chemical results and errors in flowrates. Typical commercial ED systems have current efficiencies between 80-95%.

Power Consumption and PostTreatment

DC power consumption was found to be 1.5 KWH/1000 Gallons of Product. Although no anti-scalants were required for the process, due to the low pH of the waste water, post treatment may be required before disposal.

Recommendations for Further Study

One of the disadvantages of ED is that numerous process streams samples are required to verify predications. Of course this increases the cost of pilot studying ED systems. A hand-held nitrate probe was used at Brighton, but it was difficult to obtain consistent readings. Perhaps with more field experience, consistent readings are possible or, better field type measurement equipment could be purchased.

Since only one sample point was taken, longer testing duration and more frequent samples were required. The City of Brighton at one time offered space in their large RO treatment plant, located east of town, for continued testing of the process. The City eventually sold the RO Pilot Testing facility on Jesup Street and Reclamation had to cease testing and remove equipment.

It is premature to recommend this process for commercial water treatment processes without further study. But some conclusions can be made. High recoveries (> 80%) are possible with this technology, and with ion selectivity, overall power consumption can be lower than conventional ED.

References

1. **Shaffer** L.H. and **Mintz** M.S., *Principles of Desalination*, Chap. 6, pp 257-357, Academy Press 1980, ISBN o-12-656701-8

2. Mason E.A. and **Kirkham** T.A., "*Design of Electrodialysis Equipment*", Chem. Eng. Progress Symposium Series,vol. 55, no.24 pp 173-189, 1959

3. **Mintz** M.S., *Electrodialysis Principles of Process Design*, (complete text included in appendix)

4. Meller Floyd H., *Electrodialysis (ED)& Electrodialysis Reversal (EDR) Technology*. **Ionics** Inc. 1984.

Appendix A

Field Data Taken on **1/24/95** during sampling

Qdi = 20 galhnin
Q_{ci} = 2.6 gal/min
QP = bucket checked to 4.25 gal/min
Q_+ = bucket checked to 0.433 gal/min
$Q-$ = 0.433 gal/min

V = 300.0 volts
I = 1.13 - 1.14 amps

Concentrate tank pH = 4 .99 with hand held probe

(1) $$Q_{di} = QP\,(C_F - C_P)/(C_P - C_{Do})$$

using actual values,

$$Q_{di} = 4.25(15.9-7.5)/(7.5-5.55)$$

$$= 18.3 \text{ gal/min}$$

(2) $$C_{ci} = (Q_F\, C_F - Q_P C_P)/Q_{waste}$$

$$= (4.68)(15.9) - (4.25)(7.5)/(0.433)$$

$$= 98.2 \text{ mg/L}$$

(3) $$Qci = Q_{FC}\, C_F - Q_+ C_{Ci} /(\, C_{Ci} - C_{Co})$$

$$= (0.433)(15.9) - (0.433)(73.80/(73.8-87.3)$$

$= 1.86$ gal/min assuming $Q_{FC} = Q +$

in actual operation $Q_{FC} > Q_+$ to maintain slight overflow of tank.

APPENDIX B
USBR CHEMISTRY LABORATORY
REPORT OF MAJOR CATION AND ANION ANALYSES
07/26/1993

PROJECT.. .... **NITRATE ED**
SAMPLING DATE........................ 07/12/1993

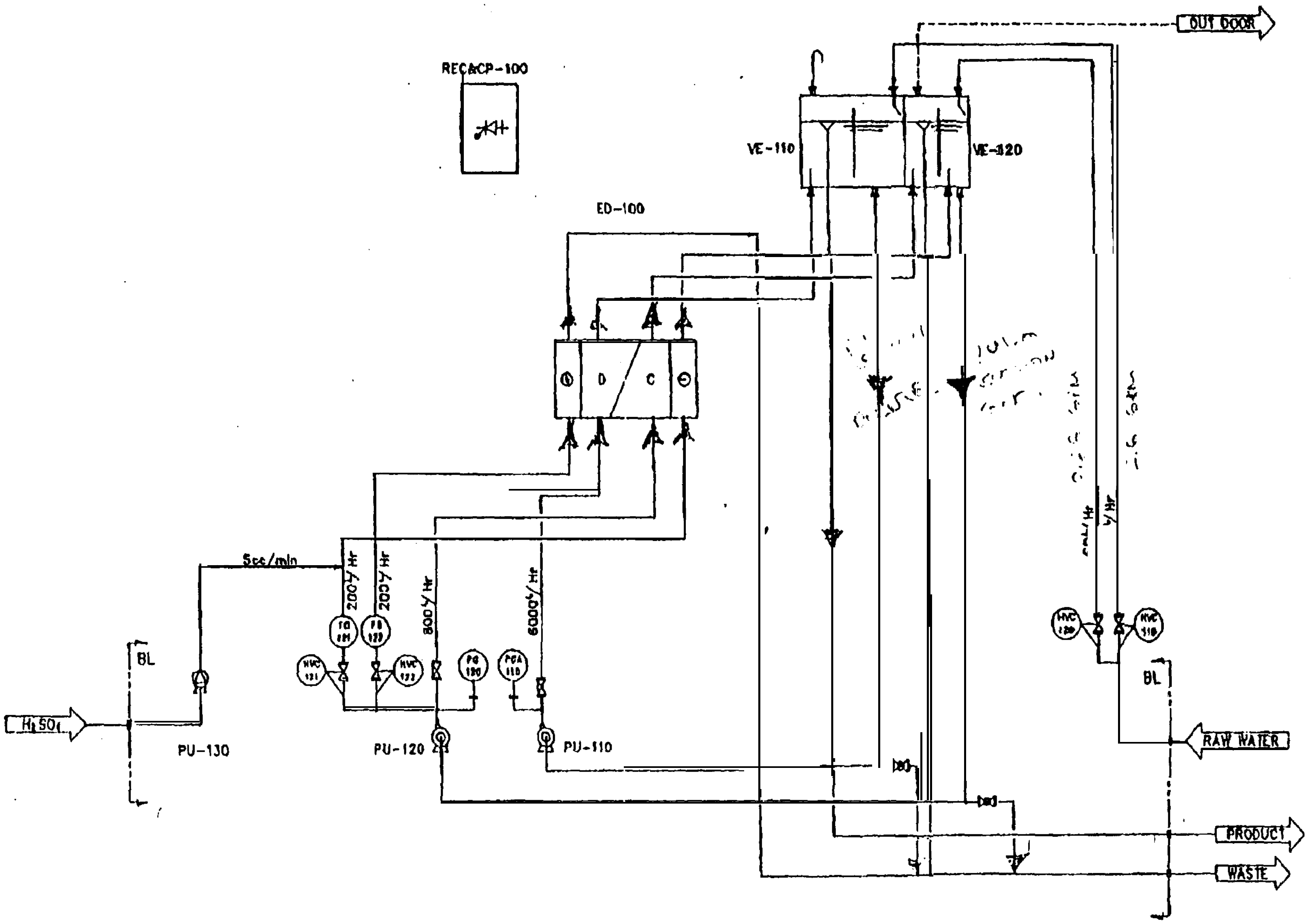
H- 3356 BRIGHT0 WELL NO. 7

pH			6.17E+000
Conductivity			1.16E+003 uS/cm @ 25C
Suspended solids			5.22E-001 mg/L
Dissolved solids-180C			8.42E+002 mg/L
Sum of cations+anions			9.18E+002 mg/L
Calcium	5.79E+000 meq/L	....	1.16E+002 mg/L
Magnesium	2.36E+000 meq/L	....	2.87E+001 mg/L
Sodium	5.48E+000 meq/L	....	1.26E+002 mg/L
Potassium	1.10E-001 meq/L	....	4.30E+000 mg/L
Carbonate	0.00E+000 meq/L	....	0.00E+000 mg/L
Bicarbonate	5.29E+000 meq/L	....	3.23E+002 mg/L
Sulfate	4.73E+000 meq/L	....	2.27E+002 mg/L
Chloride	2.62E+000 meq/L	....	9.30E+001 mg/L

NO3 as N = 13.4 mg/L

NO_2 as N = less than 0.1 mg/L

NH4 as N = less than 0.1 mg/L



ITEM No,	ED-100	REC&CP-100	M-110	M-120	PU-110	PU-120	PU-130
EQU'T NAME	ELECTRODIALYZER	RECTIFIER & CONTROL PANEL	DILUATE TANK	CONCENTRATE TANK	DILUATE PUMP	CONCENTRATE PUMP	H_2SO_4 PUMP
CAPACITY	DB-0 (100)	DC100Vx3A	50ℓ	5ℓ	6m³/Hrx15mH	1m³/Hrx5mH	5cc/min
MATERIAL							
QUANTITY	1	1	1	1	1	1	1

NOTE WETTED PART PLASTICS

ITEM	NAME	MATERIAL	NO SET	REMARK

DRAWN T.Ueno
DESIGNED
CHECKED S.Otsuka
APPROVED T.Hayama

MESSRS. B & R USA
DB-0 TYPE ELECTRODIALYZER
COMPLETED UNIT FLOW SHEET

AGC ASAHI GLASS CO., LTD.

DATE JAN.22,1993
SCALE
DOC. NO. M2001-1 ◇

APPENDIX D

ELECTRODIALYSIS

PRINCIPLES OF PROCESS DESIGN

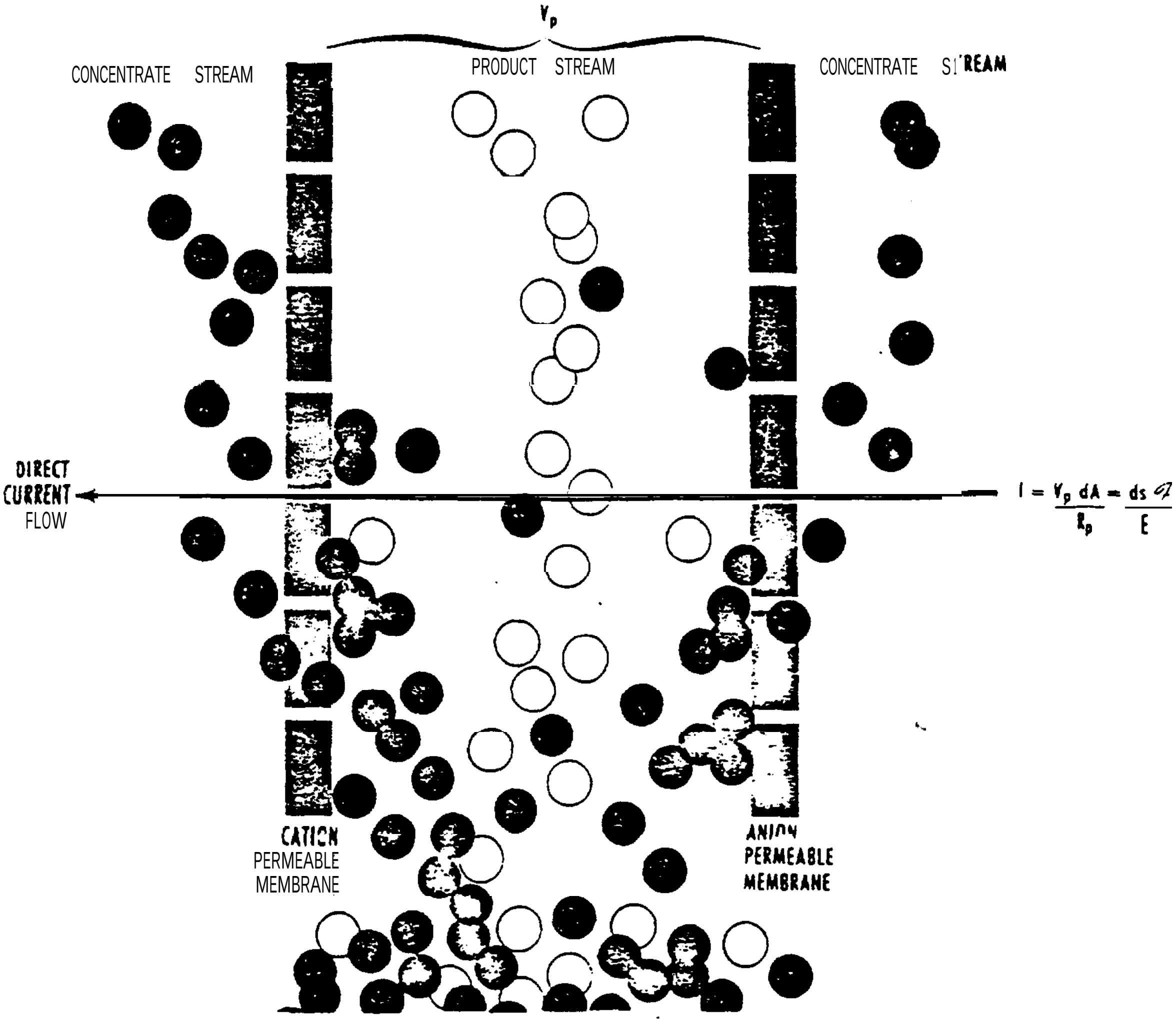


Figure 7. Demineralization by electrodialysis in a single membrane pair. Basic performance equations of such a pair can be extended to complex engineering processes

M. S. **MINTZ**

Criteria for economic optimization are presented in the form ‘of comparative performance equations for various methods of operation

From the chemical engineering viewpoint, electrodialysis is, strictly speaking, a unit operation in which ion transport through a membrane is achieved with an electrical driving force. As. a mass transfer operation it is thus distinguished from unit processes involving ‘distinct chemical transformations of the materials treated.

Demineralization by electrodialysis using alternately arrayed anion and cation selective membranes has been the most emphasized application of this unit operation, howcva, and particularly in the saline water conversion field, has become known simply as “the electrodialysis process.” Within this definition of the electrodialysis process, however, lie several alternate operational methods which have also earned the title of “processes.” They arc represented, in essence, by different flow sheets and associated energy and material balances.

The basic paformancc equations for multicompartment electrodialysis with ion selective membranes have also been developed and presented in several forms by researchers in this field, but their utility in practical clcctrodialysis process design and optimization is usually limited to a single method of operation.

One of the objects of this papa is to extend and modify these equations for practical engineering process designs such as continuous, batch recirculation, feed-and-bleed, and, internally staged systems, and present them in a form that will permit comparative economic evaluation of the processes they represent. This is achieved in most cases by considering plants of equivalent membrane area and product throughput, operating at constant voltage. Further optimization of plant size and energy requirement is then possible through the use of the power equation presented for each process. Other process refinements such as variable voltage operation and heat exchange of the feed and product streams arc best considered on their individual merits for each application.

In the course of the development of these equations, it is also shown that only two sets of empirical data are required to substantially characterize all of the methods of operation for any particular application. First, a relation must be established between solution concentration and membrane stack resistance. The data should, of course, be obtained ova the concentration range in question and at the soiution temperatures likely to be encountered. Second, a relation must be established between maximum practical current density and solution velocity for a given cell design. This type of data may readily be obtained during the resistance correlation runs by varying the flow rate through the cells at each concentration level. The electrochemical limitation on current density will depend on the nature of the solution being treated, and, for design purposes, will usually be chosen with a liberal safety factor.

Comparative Process Evaluation

In addition to the economic optimization of physical plant size and energy requirements, then arc several technical advantages and disadvantages inherent in each process which must be considered in determining an over-all plant design. Although a detailed discussion of each item is beyond the scope of thii paper, the principal factors appear with the flow diagrams of the processes on the next page (Figures 2–6).

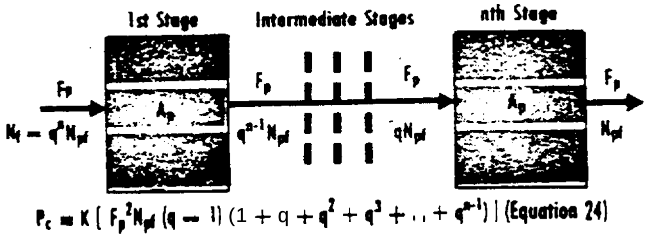
Basic Performance Equations

The performance of a multiple chamber electrodialysis stack is best understood by first considering a differential area of a single membrane pair across which a saline sotution is flowing and through which a diict current is being passed (see Figure 1).

The current is related to the transfer of salt by:

$$I = \frac{V_s\, dA}{R_p} = \frac{ds\, \mathfrak{F}}{E} \qquad (1)$$

The current efficiency term, E, is related in part to the pcrmsclcctivity of the membranes used, but it is also a function of the solution concentration range in which they arc applied. For practical design purposes, the



CONTINUOUS PROCESS *Figure i*

For large scale applications of electrodialysis in which the desired capacity **or** *demineralization* **range** *is well beyond* **the capabilities** *of a single or modular size membrane stack, suitable combinations of stacks in parallel and series can readily meet the desired specifications. Arranging membrane stacks hydraulically* in *parallel will proportionally increase plant capacity, and* arranging *them* **in** *series will permit progressively higher over-all demineralization ranges. Once this process is in its equilibrium operating condition, its power demand is constant. During start-up, however, low* **solution** *resistance will* **cause a** *peak power demand unless a preventive procedure or control instrumentation is used. Interstack pumps may also be required.*

ADVANTAGES: *Minimum pipe and valve* **size,** *minimum control instrumentation,* **no** *recirculation reservoir,* **minimum** *power requirements, stage voltages readily adjusted to conform with actual stack resistances, all membranes operate at a single equilibrium concentration.*

DISADVANTAGES: *Production rate must be balanced with flow velocity, feed water salinity w temperature changes require plant adjustment, process is sensitive to changes in flow rate, increasing membrane resistance may put plant out of operation.*

current efficiency is determined empirically by any method of stack operation and the use of the following relations :

$$E = \frac{\text{No. of chemical equivalents of salt effectively transferred}}{\text{No. of electrical equivalents passed through membrane pairs}} \qquad (2)$$

$$E = \frac{\mathfrak{F} F_t \Delta N}{k I} \qquad (3)$$

where :

F_t = Total product stream flow rate in l./sec.

ΔN = Difference between feed and product normalities, equiv./l.

It should be noted that current efficiency, as defined above, is independent of stack resistance. voltage or power requirements, membrane area, and method of stack operation. The effects of water transfer by osmosis and electroendosmosis, and salt transfer by back diffusion arc also included in the current efficiency term when it is evaluated empirically in the above described manner.

The relation between the rate of salt transfer, the product stream flow rate, and the change in concentration of the product stream is given by:

$$dS = -F\,dN_d \qquad (4)$$

Substituting Equation 4 in Equation 1 gives:

$$\frac{E\,V_p\,dA}{F\mathfrak{F}} = -R_p dN_d \qquad (5)$$

As a first approximation, and to demonstrate the utility of the performance equations, it is now assumed that the resistance of each square centimeter of a cell pair, R_p, is completely controlled by the product stream concentration and that it is inversely proportional to this concentration. Expressed mathematically:

$$R_p = \frac{a}{N_d} \qquad (6)$$

Substituting Equation 6 in Equation 5 gives:

$$\frac{E\,V_p\,dA}{F\mathfrak{F}} = -a\frac{dN_d}{N_d} \qquad (7)$$

Integrating Equation 7 over the effective area of a single membrane pair, A_p, and over the desired product stream desalination range gives:

$$\frac{E\,V_p}{F\mathfrak{F}} \int_0^{A_p} dA = -a \int_{N_f}^{N_p} \frac{dN_d}{N_d} \qquad (8)$$

or:

$$\frac{E\,V_p\,A_p}{F\mathfrak{F}} = R_p N_d \ln\frac{N_f}{N_p} \qquad (9)$$

Thus, if the desalination ratio, N_f/N_p, and product flow rate per cell pair, F, are given, the required effective cell area, A_p, may be calculated. Conversely, if the effective cell area is known, the desalination ratio for any given flow rate may be calculated. In either case, however, it is necessary to know or estimate an average value for $R_p N_d$ and E, and to choose a value of V_p.

Since V_p cannot be arbitrarily chosen because of polarization limitations, it is more convenient to rearrange Equation 9 and substitute the product of cell width, W, and cell length, L, for A_p, giving:

$$\ln\frac{N_f}{N_p} = \frac{E}{\mathfrak{F}}\left(\frac{V_p}{R_p N_d}\right)\left(\frac{W}{F}\right)L \qquad (10)$$

For a given cell thickness, t, the ratio F/W is proportional to linear velocity, v. Expressed mathematically:

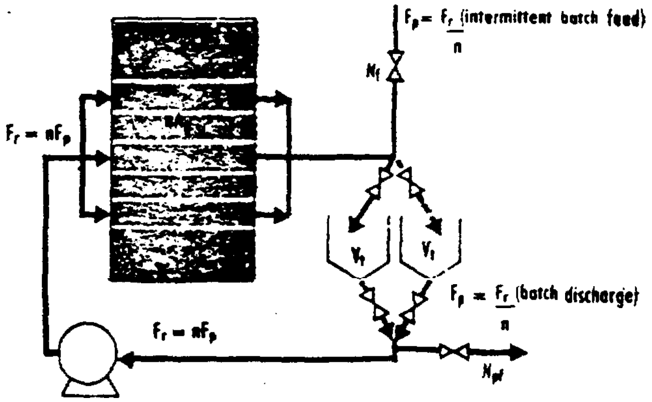
$$\frac{F}{W} = \frac{vt}{1000} \qquad (11)$$

The ratio $V_p/R_p N_d$ may also be converted to a parameter which is indicative of the limiting polarization conditions :

$$\frac{V_p}{R_p N_d} = \frac{CD}{N_d} \qquad (12)$$

where values for CD and N_d exist at the same point in the cell. They are not average conditions Under the assumed conditions of constant V_p and constant $R_p N_d$, however, the ratio CD/N_d is also constant.

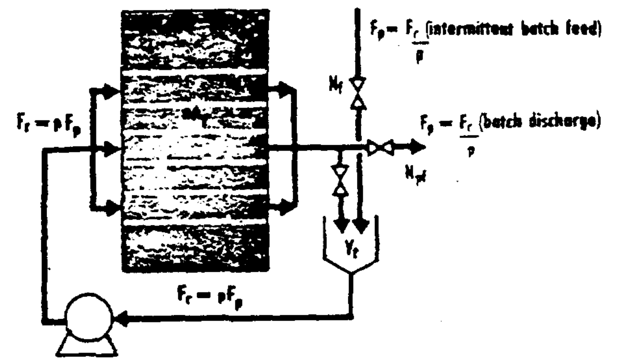
The term $1/R_p N_d$ may also be recognized as being proportional to equivalent conductance which is in fact independent of concentration for dilute electrolytes.



$$P_{bs}\ K\left[F_p^2 N_{pf}(q-1)(1+q+q^2+q^3+\ldots+q^{n-1})\right]$$ (Equation 30)

BATCH RECIRCULATION PROCESSES *Figures 3 and 4*

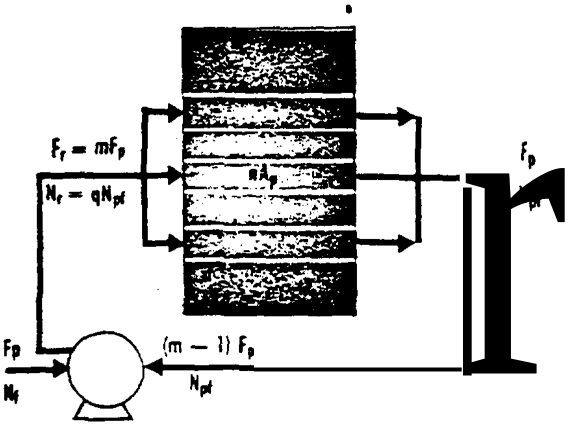
In a batch recircuiation process, a fixed volume of jccd *water* is *pumped from a* holding *lank through a mnbranr stack and back to the holding* tank until *tk desired degree* Of *demineralization is achieved. The power requirement for such a process is Largely dependent on wktkr or not tk intermediate product leaving the stack at any* given *time* is mixed with *the* water in *the* holding tank. Since various degrees *of* mixing are possible, *it* is most *convenient to consider the extreme* cases *of no mixing and* complete *mixing. The case* of *no mixing can be* achieved in *practice* by using *two* holding tanks *and having them serve alternately* as *feed and product* reservoirs. *In any* constant voltage *batch recirculation process, however, there will be a peak power demand at tk beginning of the cycle because of low solution resistance.*



$$P_{bm} = \frac{(p+1)K}{n}\ F_p^2 N_{pf}(q-1)(1+q+q^2+\ldots+q^{n-1})$$ (Equations 24 and 54)

ADVANTAGES: *Optimum velocity is independent of production rate, changes in feed water salinity* or *temperature only modify the* production rate, *variations* in *recirculation* rate *do not affect* performance, changes in *membrane properties only affect* production rate.

DISADVANTAGES : *Larger pipe, valve, and pump sizes; higher* power *requirement* and sometimes *higher average power* requirement; *greater degree* of *instrumentation; recirculation tank required; membranes are never at equilibrium; variable current density through* membranes; *performance difficult* to cvaluatc.



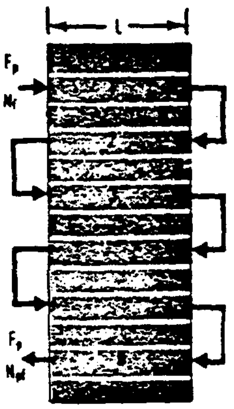
$$P_{fb} = \frac{K}{n} m^2 F_p^2 N_{pf}(q-1)$$ (Equation 64)

FEED-AND-BLEED PROCESS *Figure 5*

For small demineralization ranges or where large changes in feed solution concentration *are encountered* and a continuous product *flow* is &sired, *a feed-and-bleed* process is sometimes used. *Any* conventional *multicell stack may be used in this process. A recirculation pump is* required. *however, to* blend *the raw* fttd *solution with a portion* of *the* product solution. *The blended solution then becomes the* actual *feed solution to the membrane stack and the stack duty is equivalent to that of the last stage of a multistack continuous* process or *the fast* pass *of* a *batch recirculation process. The production rate, however, is only that of the portion oj the product stream which is "bled" out of the* recirculation Loop. *The power demand for this* process is. also constant *under equilibrium* operating *conditions. Since several feed-and-bleed loops may be connected in* series *to give tht effect oj multiple staging with lower recycle rates, and the product flow is continuous, some authors* have *called* this *process "the* continuous process."

ADVANTAGES: *Provides continuous product, process may be adapted to feed water of any salinity, membranes are at a single equilibrium condition, variation in performance* is *readily observed,* minimum *absolute current density.*

DISADVANTAGES: *Highest power consumption, recirculation system* requires *sensitive instrumentation*, largest *recirculation* rate.



$$L_t = gL = \frac{l(q^n-1)}{e^X-1}$$

$$\text{Where } X = \frac{E}{\mathcal{F}}\frac{CD}{K}\cdot\frac{W}{F}\,l$$ (Equation 71)

INTERNALLY STAGED STACK DESIGN *Figure 6*

At low production rates, it often appears uneconomical to *provide either the recirculating pumps necessary for the* batch and *feed-and-bleed processes,* or *the separate electrodes* and interconnecting piping *necessary for a multistack continuous* process. *Under* these conditions. *a* stack *design combrising several infernal stages may be used. In a stack design of this type, the product stream makes several passes* in *series between a single set oj electrodes and is thereby &mineralized to a greater extent than would be possible* in *a* single *pass through a stack of tkc same cross sectional area. Since this process is, by* its *very* nature. *a continuous one. its power demand will be constant under* equilibrium *operating* conditions. *During startup. however, low solution r&lance will cause a peak power demand. For a small plant size il is often* mow *economical* to *design for this* peak power than to install rontrois *to avoid it.*

ADVANTAGES: *No repressurizing* pumps. *high desalination range, single set of electrodes. single operating voltage, product flow is continuous.*

DISADVANTAGES: *Large membrane area* per *unit of product, high operating pressure, performance sensitive to variations* in *flow rote and membrane resistance.*

Substituting Equation 12 in Equation 10 gives:

$$\ln \frac{N_f}{N_p} = \frac{E}{\mathfrak{F}}\left(\frac{CD}{N_d}\right)\left(\frac{W}{F}\right) L \qquad (13)$$

From an empirically determined relationship between CD/N_d and linear velocity, values of CD/N_d and W/F may now be chosen without recourse to resistance data or voltage requirement. When these parameters are chosen, the desalination ratio is simply a function of cell length. A plot of Equation 13 for a velocity of 10 cm./sec. in a 0.1 cm. cell and at an assumed current efficiency of 90% is given in Figure 7. As an example, assume that the limiting CD/N_d is 1 .O at a linear solution velocity of 10 cm./sec. In a cell 0.1 cm. thick this corresponds to an F/W of 0.001 l./sec.-cm. Assuming an average current efficiency of 0.90, the cell length required for a four to one desalination ratio would he calculated as follows:

$$L = \frac{96.5}{0.9}(1.385) = 148.5 \text{ cm.}$$

Equation 13 is thus seen to be of general utility in determining the effect of current efficiency, CD/N, linear velocity, and cell length on the desalination ratio. It must be emphasized, however, that this equation was developed for a single cell pair across which a constant voltage was applied. It may be extended to describing the conditions in each cell pair of a parallel flow membrane stack only if uniform flow distribution and a uniform voltage per cell pair can safely be assumed.

Further refinement of the performance equations developed in this section for use in process design would require a more exact relation between cell resistance and product stream concentration than that assumed in Equation 6. This relation may take the form:

$$R_p = a\,(N_d)^b \qquad (14)$$

or

$$R_p N_d = a + bN_d + cN_d^2 + dN_d^3 \ \ldots \text{ and so forth} \qquad (15)$$

Substitution of either Equation 14 or Equation 15 in Equation 5 presents no problem in integration and the resulting performance equation is readily converted to the form of Equation 13. In order to simplify the analyses presented in later sections, however, the approximation for R_p defined by Equation 6 will be retained.

Basic Power Equations

The direct current electrical power requirement for a single cell pair is given by the relation:

$$P = V_p I \qquad (16)$$

Solving Equation 9 for V_p gives:

$$V_p = \mathfrak{F}\frac{F}{A_p}\left(\ln \frac{N_f}{N_p}\right)\left(\frac{R_p N_d}{E}\right) \qquad (17)$$

The current passing through the cell pair may also be expressed in terms of the product flow rate and it. change in concentration as follows:

$$I = \frac{\mathfrak{F}}{E} F\,(N_f - N_p) \qquad (18)$$

Substituting Equations 17 and 18 in Equation 16 gives:

$$P = \mathfrak{F}^2 \frac{F^2}{A_p}(N_f - N_p)\left(\ln \frac{N_f}{N_p}\right)\left(\frac{R_p N_d}{E^2}\right) \qquad (19)$$

Since Equation 19 expresses the power requirement on an absolute basis, it is usually rearranged to express the power requirement on the basis of a unit product flow rate as follows:

$$U = \frac{P}{F} = \mathfrak{F}^2 \frac{F}{A_p}(N_f - N_p)\left(\ln \frac{N_f}{N_p}\right)\left(\frac{R_p N_d}{E^2}\right) \qquad (20)$$

With the units defined in the nomenclature section, U is in w.-hr./l. sec. The equation may be presented in American engineering units and the constants expressed numerically by the following conversions:

F (in l./sec.) = 1.05×10^{-3} F' (in gal./hr.)

A_p (in sq. cm.) = 930 A'_p (in sq. ft.)

ln (base e) = 2.3 log (base 10)

$\mathfrak{F}$ = 96.500 amp. sec./equivalent

Then :

$$U' = 25.4 \frac{F'}{A'_p}(N_f - N_p)\left(\log \frac{N_f}{N_p}\right)\left(\frac{R_p N_d}{E^2}\right) \qquad (21)$$

where U' now is the power per unit product flow rate in kw. hr. per 1000 gal.

Equation 21 is useful for estimating the power requirements for a given desalination range when $R_p N_d/E^2$ for this range is reasonably constant.

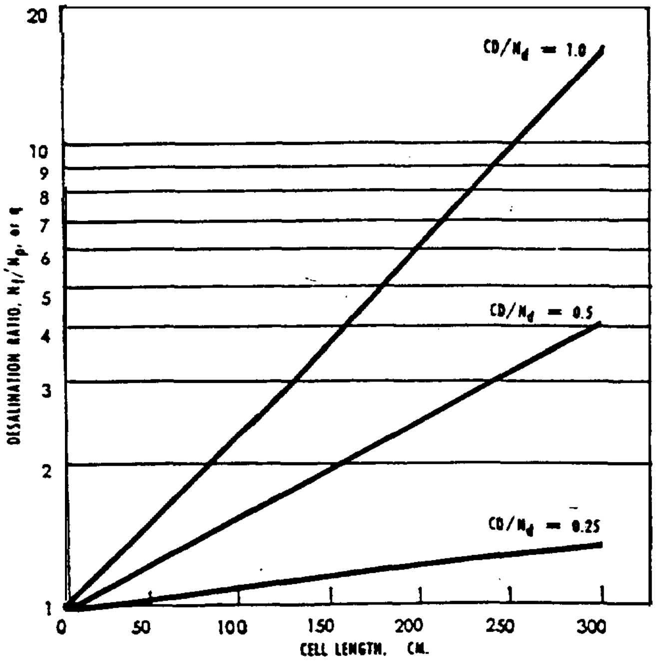


Figure 7. Electrodialysis performance in a single pass. Equation 73 is plotted for the example given above

to develop design methods for engineering systems

It should be noted, however, that the power equations in this section were developed for a single cell pair in which the desired change in product concentration was achieved in a single pass. The total power requirement for a multiple cell-pair stack of parallel construction would simply be that multiple of the single cell pair power requirement, whereas the power requirement per unit product flow rate would remain unchanged.

If two or more membrane stacks are placed in series to achieve a greater degree of demineralization than can be achieved in a single pass through one stack, the power requirement for each stack may be calculated by Equations 19 to 21 and the total power will, of course, be their sum.

Under these conditions, the product from the first stack becomes the feed to the second stack. It is interesting to note in this connection that if the same voltage is applied to each stack and $R_p N_d$ is substantially constant, the demineralization ratio, N_f/N_p, will be the same in each stack.

Batch Recirculation without Mixing (Slug Flow)

For batch recirculation without mixing (Figure 3) it is immediately apparent that multiple passes through a single stack are equivalent to passing the water to be demineralized through several stacks in series. If the single stack is of *the* same size as the stacks arranged in series, however, the production rate for the single stack will be reduced by a factor equal *to* the number of passes required. In order to properly compare the batch and continuous processes, therefore, they should be based on the same production rate and the same total membrane stack area. For this purpose, let the desired production rate be F_p liters per second and let each cell pair have A_p square centimeters effective area (Figure 2). Assuming that n passes are required to achieve the desired demineralization in a multistage continuous plant and that a constant V_p will be used to give a constant desalination ratio, $q = N_f/N_p$, the power requirement for the continuous plant will, from Equation 19, be given by:

$$P_c = \mathfrak{F}^2 \frac{F_p^2}{A_p} N_{pf}(q-1)(\ln q)\left(\frac{R_p N_d}{E^2}\right)\text{(nth stage)}$$

$$+ \mathfrak{F}^2 \frac{F_p^2}{A_p} qN_{pf}(q-1)(\ln q)\left(\frac{R_p N_d}{E^2}\right)\text{(n-1 stage)}$$

$$+ \mathfrak{F}^2 \frac{F_p^2}{A_p} q^2 N_{pf}(q-1)(\ln q)\left(\frac{R_p N_d}{E^2}\right)\text{(n-2 stage)}$$

$$+ \ldots \qquad \text{(other intermediate stages)}$$

$$+ \mathfrak{F}^2 \frac{F_p^2}{A_p} q^{n-1} N_{pf}(q-1)(\ln q)\left(\frac{R_p N_d}{E^2}\right)\text{(1st stage)} \tag{22}$$

Letting:

$$\frac{\mathfrak{F}^2}{A_p}(\ln q)\left(\frac{R_p N_d}{E^2}\right) = K \tag{23}$$

$$P_c = K\,[F_p^2 N_{pf}(q-1)(1+q+q^2+q^3+\ldots q^{n-1})] \tag{24}$$

For the batch recirculation plant with the same equivalent total area nA_p, and slug flow (Figure 3). the power requirement would be the average of the power requirement for each pass, that is:

$$P_{bs} = \left(\frac{1}{n}\right)(P_n + P_{n-1} + P_{n-2} + \ldots P_1) \tag{25}$$

and

$$P_n = \mathfrak{F}^2 \frac{(nF_p)^2}{nA_p} N_{pf}(q-1)(\ln q)\left(\frac{R_p N_d}{E^2}\right) \tag{26}$$

$$P_{n-1} = \mathfrak{F}^2 \frac{(nF_p)^2}{nA_p} qN_{pf}(q-1)(\ln q)\left(\frac{R_p N_d}{E^2}\right) \tag{27}$$

$$P_{n-2} = \mathfrak{F}^2 \frac{(nF_p)^2}{nA_p} q^2 N_{pf}(q-1)(\ln q)\left(\frac{R_p N_d}{E^2}\right) \tag{28}$$

Similar equations can be written for other intermediate passes.

$$P_1 = \mathfrak{F}^2 \frac{(nF_p)^2}{nA_p} q^{n-1} N_{pf}(q-1)(\ln q)\left(\frac{R_p N_d}{E^2}\right) \tag{29}$$

Then nF_p represents recirculation flow rate of the product stream for a production rate, F_p, l./sec.; and nA_p represents total effective membrane area of the cells used which are equivalent to that required for the continuous process at production rate, F_p.

Therefore:

$$P_{bs} = K\,[F_p^2 N_{pf}(q-1)(1+q+q^2+q^3+\ldots q^{n-1})] \tag{30}$$

The equivalence of Equations 30 and 24 prove that the average power demand for a batch recirculation process with slug flow is equal to that for a continuous process. A peak power demand occurs during batch recirculation, however, and is given by Equation 29 as the power demand for the first pass. The ratio of this peak power demand to the average power demand or that for continuous operation is given by the ratio of Equation 29 to Equation 30 or 24:

$$\frac{P_1}{P_c} = \frac{P_1}{P_{bs}} = \frac{nq^{n-1}}{(1+q+q^2+q^3+\ldots q^{n-1})} \tag{31}$$

A plot of Equation 31 is given in Figure 8.

As an example, assume that two stacks are required for the continuous process (n = 2), and that the desalination ratio, N_f/N_p, for each stack is two (q = 2). The ratio of peak power to continuous power would then be:

$$\frac{P_1}{P_c} = \frac{P_1}{P_{bs}} = \frac{2(2)^{2-1}}{(1+2)} = 1.33$$

Batch Recirculation with Mixing

For batch recirculation with mixing, the usual simplifying assumptions of constant voltage per cell pair, V_p, and constant R, N_d may be made. But since the concentration of the brine in the holding tank changes continuously rather than in discrete steps, the change in concentration with treated volume must be analyzed on a differential basis (see Figure 4).

The constant desalination ratio, q = N_f/N_p, may also be expressed as a constant fractional demineralization f, as:

$$f = \frac{N_f - N_p}{N_f} = \frac{q-1}{q} \qquad (32)$$

Equating the salt transferred out of the holding tank to that transferred *out* of the product stream during the electrodialysis of a differential volume gives:

$$-V_t\,dN_t = f\,N_t\,dV_t \qquad (33)$$

integrating Equation 33 over the desired desalination range, N_f to $q\,N_{pf}$ (as will be explained later), gives:

$$\ln \frac{N_f}{qN_{pf}} = f\,\frac{V_r}{V_t} \qquad (34)$$

The ratio V_r/V_t represents the number of times the tank volume must be recirculated to achieve the desired degree of demineralization. In this sense it is equivalent to the number of passes or the multiple of the actual production flow rate, F_p. Expressed mathematically:

$$\frac{V_r}{V_t} = p \qquad (35)$$

Substituting Equation 35 in Equation 34 gives:

$$p = \left(\frac{1}{f}\right) \ln \frac{N_f}{qN_{pf}} \qquad (36)$$

Since the power requirement for this process also changes continuously rather than in discrete steps during each pass, the average power must be obtained by a relation of the type :

$$P_m = \frac{1}{\theta}\int_0^\theta P\,d\theta \qquad (37)$$

Dividing both sides of Equation 33 by $d\theta$ gives:

$$-V_t\,\frac{dN_t}{d\theta} = f\,N_t\,\frac{d\,V_t}{d\theta} \qquad (38)$$

but:

$$\frac{dV_t}{d\theta} = F_r = p\,F_p \qquad (39)$$

Substituting Equation 39 in Equation 38 and integrating over the period of time, θ, required to desalinate from feed concentration, N_f, to some intermediate concentration, N_t, gives:

$$-\int_{N_f}^{N_t} \frac{dN_t}{N_t} = \frac{f\,p\,F_p}{V_t}\int_0^\theta d\theta \qquad (40)$$

or:

$$\ln \frac{N_f}{N_t} = \frac{f\,p\,F_p\,\theta}{V_t} \qquad (41)$$

or:

$$N_t = N_f e^{-fpF_p\theta/V_t} \qquad (42)$$

Equation 42 simply shows the concentration in the holding tank, N_t, as a function of feed concentration and time.

The current flowing through the stack at any time may be related to the concentration in the holding tank and the fractional demineralization as follows:

$$I = \frac{\mathfrak{F}\,\Delta N\,F_p}{E} = \frac{\mathfrak{F}\,f\,N_t\,p\,F_p}{E} \qquad (43)$$

where AX equals the change in normality of the product stream during one pass through the stack.

Substituting Equations 16, 17, 42, and 43 in Equation 37 gives:

$$P_m = \frac{1}{\theta}\int_0^\theta V_p\,I\,d\theta = \frac{V_p}{\theta}\int_0^\theta \frac{\mathfrak{F}\,F_p\,f\,N_f e^{-fpF_p\theta/V_t}}{E}\,d\theta$$
$$= \frac{V_t\,\mathfrak{F}^2\,p\,F_p}{\theta\,n\,A_p}(\ln q)\,\frac{R_p\,N_d}{E^2}\,N_f\left(1 - e^{-fpF_p\theta/V_t}\right) \qquad (44)$$

Choosing time, θ, such that the concentration of the solution in the holding tank reaches qN_{pf} (a solution requiring one more pass to achieve the desired product N_{pf}):

$$N_f e^{-fpF_p\theta/V_t} = q\,N_{pf} \qquad (45)$$

$$\theta = V_t\,\frac{\ln \dfrac{N_f}{qN_{pf}}}{f\,p\,F_p} \qquad (46)$$

Substituting Equations 45 and 46 in Equation 44 gives:

$$P_m = \frac{\mathfrak{F}^2 p^2 F_p^2 f}{n\,A_p}(\ln q)\,\frac{R_p N_d}{E^2}\left(\frac{N_f - q\,N_{pf}}{\ln \dfrac{N_f}{q\,N_{pf}}}\right) \qquad (47)$$

Substituting Equation 32 in Equation 47 gives:

$$P_m = \frac{\mathfrak{F}^2 p^2 F_p^2}{n\,A_p}(\ln q)\,\frac{R_p N_d}{E^2}(q-1)\,N_{pf}\left(\frac{q^{n-1}-1}{\ln q^{n-1}}\right) \qquad (48)$$

Inspection of Equation 48 and comparison with Equations 26 to 29 show that the average power demand for the batch recirculation process with mixing is the log mean average of its peak and minimum power demands. The last term in Equation 48 depicts this comparison as the log mean average of q^{n-1} and 1 .0.

It will be noted that the average power demand, P_m, was derived for a demineralization range of N_f to qN_{pf} rather than from N_f to N_{pf}. This was done because it would be impractical to continue recirculation beyond the point when one more pass through the stack would produce the desired product, N_{pf}. In a practical process, the stack effluent would be diverted to storage at this point and the power demand during the last pass would be the same as that for the last pass with slug type flow; that is. Equation 26 would be applicable.

The average power for both steps in the batch recirculation process with mixing may now be expressed as follows:

$$P_{bm} = \frac{1}{p+1}\left[\frac{\mathfrak{F}^2 (p+1)^2 F_p^2}{n\,A_p}(\ln q)\,\frac{R_p\,N_d}{E^2}N_{pf}(q-1)\right] \times \left[1 + \frac{p\,(q^{n-1}-1)}{\ln q^{n-1}}\right] \qquad (49)$$

$$P_{bm} = \frac{1}{n(p+1)}\left[K\,(p+1)^2\,F_p^2\,N_{pf}\,(q-1)\right] \times \left[1 + \frac{p\,(q^{n-1}-1)}{\ln q^{n-1}}\right] \qquad (50)$$

The ratio of average power demand, P_{bm}, to that for the continuous process is given by the ratio of Equation 50 to Equation 24.

$$\frac{P_{bm}}{P_c} = \frac{(p+1)\left[1 + \frac{p(q^{n-1}-1)}{\ln q^{n-1}}\right]}{n(1+q+q^2+q^3+\ldots q^{n-1})} \qquad (51)$$

Using Equation 36 to substitute for p in the bracketed term of the numerator and observing the following identities:

$$p = \frac{\ln \frac{N_f}{q N_{pf}}}{f} = \frac{\ln\left(\frac{q^n N_{pf}}{q N_{pf}}\right)}{\left(\frac{q-1}{4}\right)} = \frac{q \ln (q^{n-1})}{q-1} \qquad (52)$$

and

$$\frac{q^n - q}{4-1} = q + q^2 + q^3 + \ldots q^{n-1} \qquad (53)$$

giving the rather surprising simplification:

$$\frac{P_{bm}}{P_c} = \frac{p+1}{n} \qquad (54)$$

A plot of Equation 54 is given in Figure 9.

Choosing the same example as for the previous section, assume that two stacks are required for continuous operation ($n = 2$) and that the desalination ratio, N_f/N_p, for each stack is two ($q = 2$). The number of passes, p, required in the batch recirculation process with mixing to bring the concentration from N_f to qN_{pf} will be given by Equation 36:

$$P = \frac{\ln \frac{q^2 N_{pf}}{q N_{pf}}}{\frac{q-1}{q}} = \frac{q \ln q}{q-1} = \frac{2 \ln q}{1} = 1.386$$

$$\frac{P_{bm}}{P_c} = \frac{2.386\left[1 + \frac{1.386(2-1)}{\ln 2}\right]}{2(1+2)} = \frac{2.386\left(1 + \frac{1.386}{0.693}\right)}{2(3)} = 1\ 193$$

The peak power demand for the batch recirculation process with mixing will occur at the very beginning of the batch cycle when the water concentration in the holding tank is N_f or $q^n N_{pf}$, and that leaving the stack is $q^{n-1} N_{pf}$.

$$P_{bmp} = \frac{\mathfrak{F}(p+1)^2 F_p^2}{n A_p} q^{n-1} \bar{N}_{pf} (q-1)(\ln q) \frac{R_p N_d}{E^2} \qquad (55)$$

$$P_{bmp} = \frac{K(p+1)^2 F_p^2}{n} q^{n-1} N_{pf} (q-1) \qquad (56)$$

The ratio of this peak power demand to the power required for continuous operation is given by the ratio of Equation 56 to Equation 24.

$$\frac{P_{bmp}}{P_c} = \frac{(p+1)^2 q^{n-1}}{n(1+q+q^2+q^3+\ldots q^{n-1})} \qquad (57)$$

A plot of Equation 57 is given in Figure 10.

For the example cited above, in which $n = 2$ and $q = 2$. the ratio of peak power for this process to that required for the continuous process would be:

$$\frac{P_{bmp}}{P_c} = \frac{(1.386+1)^2 (2)^{2-1}}{2(1+2)} = 19$$

Feed-and-Bleed

In a feed-and-bleed process, the rate of flow of the feed-and-bleed streams are equal and equivalent to the production flow rate, F_p. The recirculation flow rate

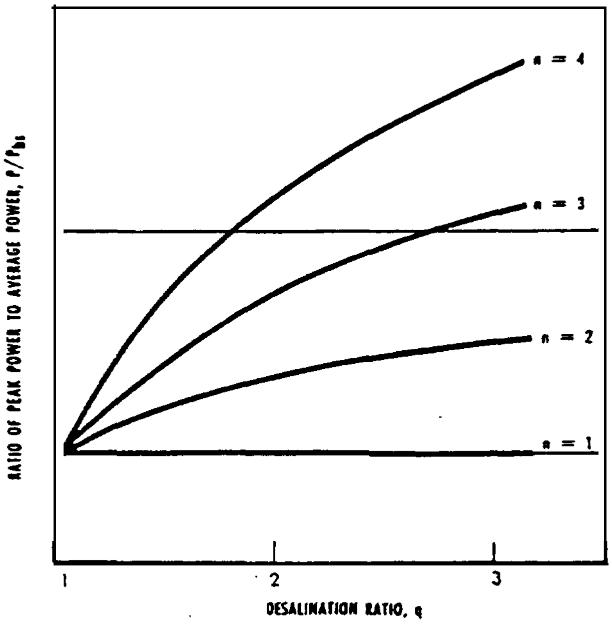


Figure 8. Batch recirculation process without mixing. Equation 31 is plotted

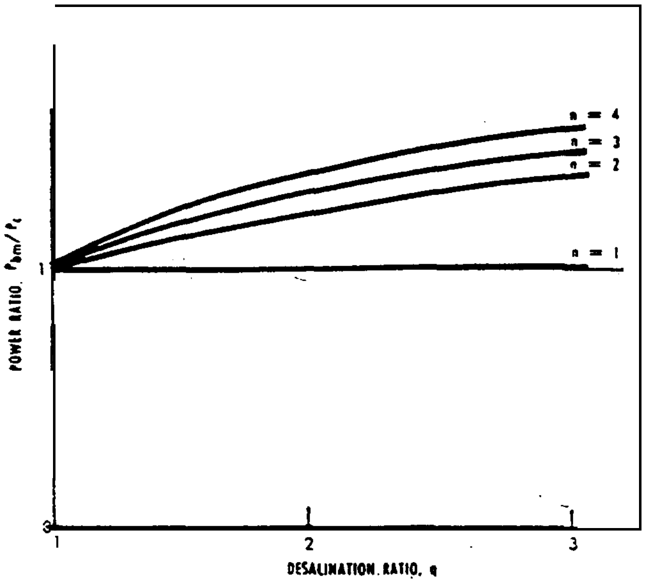


Figure 9. Batch recirculation process with complete mixing. Average power requirement is compared with that of the continuous process (Equation 51)

(Continued on next page)

will, in the general case, be higher and may be related to the production flow rate by:

$$F_r = m F_p \tag{58}$$

The value of m and consequently of F_r may be determined for any given desalination range by a material balance around the recirculation pump (see Figure 5).

$$F_p N_f + (m - 1) F_p N_{pf} = m F_p N_r$$

$$m = \frac{N_f - N_{pf}}{N_r - N_{pf}} \tag{59}$$

Since the water leaving the stack in this process must be of the final product composition, N_{pf}, the composition of the recirculated solution being fed to the stack, N_r, must be that which can be demineralized in one pass through the stack. Expressed mathematically:

$$N_r = q N_{pf} \tag{60}$$

The feed water concentration may also be related to the final product concentration by the number of passes which would be required in a continuous process:

$$N_f = q^n N_{pf} \tag{61}$$

Substituting Equations 60 and 61 in Equation 59 gives:

$$m = \frac{N_{pf} (q^n - 1)}{N_{pf} (q - 1)} = \frac{4'' - 1}{q - 1} \tag{62}$$

The power requirement for this process is constant and is given:

$$P_{fb} = \frac{\mathfrak{F}^2 (mF_p)^2}{n A_p} N_{pf} (q - 1)(\ln q) \frac{R_p N_d}{E^2} \tag{63}$$

or:

$$P_{fb} = \frac{K}{n} m^2 F_p^2 N_{pf} (q - 1) \tag{64}$$

The **ratio of** P_{fb} to the power demand for a continuous process is given by the ratio of Equation *64* to Equation *24:*

$$\frac{P_{fb}}{P_c} = \frac{m^2}{n (1 + q + q^2 + q^3 + \ldots q^{n-1})} \tag{65}$$

A plot of Equation 65 is given in Figure 11.

Using the same example as in previous sections, assume that two stacks are required for continuous operation ($n = 2$) and that the desalination ratio, N_f/N_p, for each stack is **two** ($q = 2$). The multiple of the production flow rate is calculated from Equation 62 as:

$$m = \frac{2^2 - 1}{2 - 1} = 3$$

and

$$\frac{P_{fb}}{P_c} = \frac{3^2}{2(1 + 2)} = 1.5$$

Internally Staged Stack Design

In **an internally** staged stack. the extent of demineralization is **not** equivalent to that of multiple passes at **constant** CD/N_d **because** the **current** density is constant **throughout** the stack while the product concentration is lower in each successive pass. Furthermore, since the **maximum** CD/N_d **will occur in the last pass, it usually** serves as the basis for' internally staged stack design. See Figure 6.

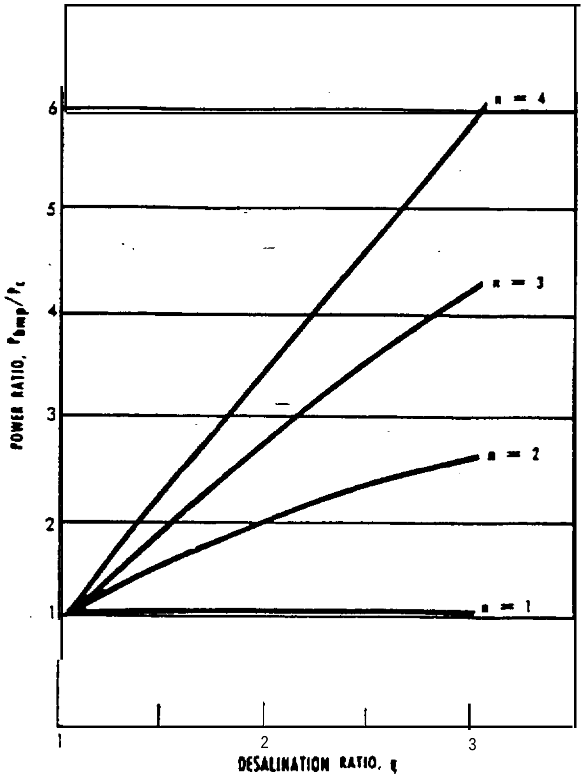


Figure 10. Batch recirculation process with mixing. Peak power requirement is compared with power for continuous process (Equation 57)

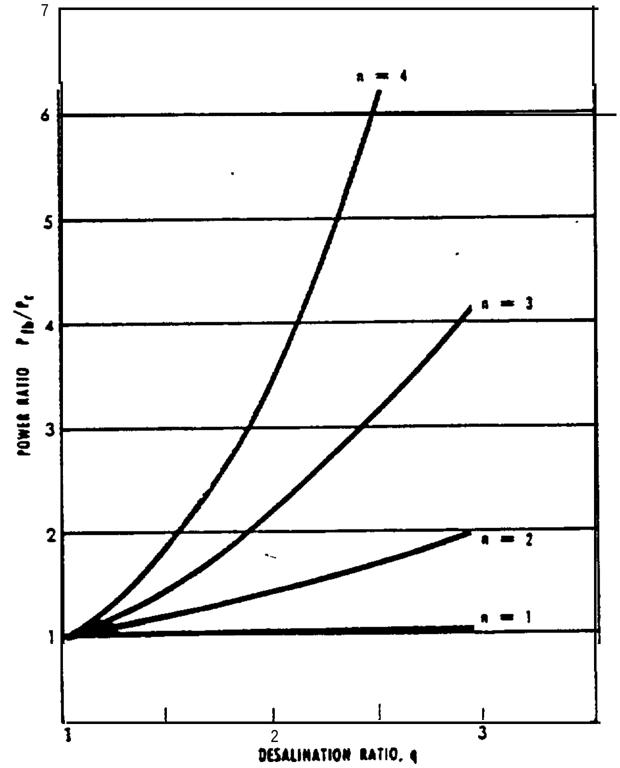


Figure 11. Feed-and-bleed process. Power demand is compared with that of the continuous process (Equation 65)

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EXAMPLE

Assume that a cell 40 cm. wide, 100 cm. long, and 0.1 cm. thick is available for which polarization studies have shown that a 50% demineralization (N_p/N_f = 0.5) may be achieved ar a solution velocity of 10 cm./sec. Equation 13 may be used to back calculate the average CD/N_d LO which these data correspond :

$$\ln \frac{N_f}{N_p} = \ln q = \ln 2 = \frac{0.90}{96.500}\left(\frac{CD}{N_{dd}}\right)(1000)\ 100$$

$$\frac{CD}{N_d} = 0.743$$

Assume now that it is desired to demineralize a 0.04 N solution to 0.005 N. For natural brines, this is roughly equivalent to demineralizing a 2400 p.p.m. water to 300 p.p.m. or an d-fold reduction in concentration. The only additional empirical data now required is cell resistance as a function of solution concentration. Taking this as $R_p N_d = 6$ for the given example, and noting that at 10 cm./sec. the flow rate through the given cell will be 40 ml./sec. or 0.040 l./sec. (approximately 40 gal./hr.) and the power rquircmcnts for the various processes may be calculated as follows:

(1) For the continuous process, using Equation 24 and

$q = 2$ $\quad E = 0.9$ $\quad N_f$ = 0.04 equiv./l.

$n = 3$ $\quad F_p$ = 0.040 l./sec. $\quad N_{pf}$ = 0.005 equiv./l.

$R_p N_d = 6$ ohm-q. cm.^{2} equiv./l. $\quad A_p$ = 4000 sq. cm.

$$P_c = \frac{(96{,}500)^2}{4.000}\ (0.693\left(\frac{6}{0.0.8}\right) \times$$
$$[(0.04^2)\ (0.005)(1)\ (1 + 2 + 4)]$$

$P_c = 667$ w.

Since the production rate is 40 ml./sec. or 38.1 gal./hr., the direct current energy requirement for this example may be expressed as:

$$U'_c = \frac{667 \text{ w.}}{38.1 \text{ gal./hr.}} = 17.5 \text{ kw.-hr./1000 gal.}$$

Energy conversion losses for the production of suitable d.c. power, pumping energy, and other plant power requirements must, of course, be added to the d.c. energy requirements when a complete economic analyses is desired.

(2) For the batch recirculation process without mixing, Equation 30 gives the same average power as that for the continuous process:

$$P_{bt} = 667 \text{ w.}$$

(3) For the batch recirculation process with mixing, the number of recirculation passes required may first be calculated bv Equation 52:

$$p = \frac{2\ (\ln 4)}{1} = 2.77$$

The average power requirement may then be found from Equation 54 :

$$P_{bm} = \frac{(2.77 + 1)\ (667)}{3} = 839 \text{ w}$$

It may also be noted that since a total of 3.77 passes are rquircd for this process instead of 3 for the foregoing processes, the pumping energy requirement will also be at least commensurately higher.

(4) For the feed-and-bleed process, the multiple of the production flow rate required may first be calculated by Equation 62:

$$m = \frac{8 - 1}{2 - 1} = 7$$

The power requirement may then be found from Equation 65:

$$P_{fb} = \frac{(49)\ (667)}{3\ (1 + 2 + 4)} = 1555 \text{ w}$$

(5) If an internally staged stack design were used for this example the number of stages required (using one cell pair per stage) quid be found by Equation 70:

$$g = \frac{8 - 1}{2 - 1} = 7$$

The power requirement may then be found by the application of Equation 19 to each stage noting that the change of concentration for each stage is 0.005N:

$$P = \frac{(96{,}500)^2\ (0.04)'}{4000}\ (0.005)\left(\ln\frac{0.010}{0.005} + \ln\frac{0.015}{0.010} + \ln\frac{0.020}{0.015} + \ln\frac{0.025}{0.020} + \ln\frac{0.030}{0.025} + \ln\frac{0.035}{0.030} + \ln\frac{0.040}{0.035}\right)\frac{6}{0.81}$$

$$\cdot P = \frac{3\ (667)}{7} = 286 \text{ w.}$$

'This iow power requirement is not directly comparable to that for the other processes, however. because 7 cells are used instead of 3. Equation 19 shows this inverse relationship between power requirement and ceil area.

Applying Equation 13 to the last cell pair of an internally staged stack:

$$\ln\frac{qN_{pf}}{N_{pf}} = \ln q = \frac{E}{\mathfrak{F}}\left(\frac{CD}{N}\right)\left(\frac{W}{F}\right)\ L \qquad (66)$$

where qN_{pf} equals normality of product stream entering the last cell pair.

The current required to effect the change in concentration $q\,N_{pf} - N_{pf}$, in the last cell is given by:

$$I = \frac{\mathfrak{F}}{E}\,F_p\,(q\,N_{pf} - N_{pf}) \qquad (67)$$

Substituting Equation 66 in Equation 67 gives:

$$I = \frac{\mathfrak{F}}{E}\,F_p\,N_{pf}\left(e^{\frac{E}{\mathfrak{F}}\left(\frac{CD}{N}\right)\left(\frac{W}{F}\right)L} - 1\right) \qquad (68)$$

Since this same current flows in series through all of the cells. the number of cells required to achieve the over-all cicsalination range is given :

$$g\,I = \frac{\mathfrak{F}}{E}\,F_p\,(q^n\,N_{pf} - N_{pf}) \qquad (69)$$

Here 9 " N_{pf} represents normality of feed water in terms of the final product normality and the number of passes that would be required in a multistack continuous process.

Dividing Equation 63 by Equation 68 gives:

$$g = \frac{q^n - 1}{\left(e^{\frac{E}{\mathfrak{F}}\left(\frac{CD}{N}\right)\left(\frac{W}{F}\right)L} - 1\right)} \qquad (70)$$

(Continued on next page)

. Since CD/N and W/F are usually fixed by empirically determined polarization data, Equation 70 may be used to calculate the number of cell pairs required fur a given desalination range as a function of cell length, L.

inspection of Equation 70 shows that as the cell length decreases toward zero, the number of cells required approaches infinity. If the number of cells is set equal to one, Equation 70 simply degenerates into Equation 13 and defines the cell length required to achieve the desired desalination range in one pass. Values of L which indicate that only a fraction of a cell pair is required may be interpreted as meaning either that the desired desalination will bc achieved in that fraction of the chosen path length or that a greater degree of desalination will be achieved than that which is desired.

The total length of the desalination flow path required in an internally staged stack is simply the product of the cell length and the number of cells:

$$L_t = gL = \frac{L(q^n - 1)}{\left(e^{\left(\frac{E}{\mathfrak{F}}\right)\left(\frac{CD}{N}\right)\left(\frac{W}{F}\right)L} - 1\right)} \qquad (71)$$

Fur a given cell width and gasket design, L_t becomes a measure of the total quantity of the membrane and gasket materials required in the stack. It is interesting to note, therefore, that although the number of cells continues to increase with decreasing cell length, the value of L_t in Equation 71 approaches a finite limit.

Expressed mathematically :

$$\lim_{L \to 0} L_t = \frac{L(q^n - 1)}{\left[1 + \frac{E}{\mathfrak{F}}\frac{CD}{N}\frac{W}{F}L\right] - 1} = \frac{q^n - 1}{\frac{E}{\mathfrak{F}}\left(\frac{CD}{N}\right)\left(\frac{W}{F}\right)} \qquad (72)$$

The power requirement for an internally staged stack cannot be compared directly to that of the continuous multistack process &cause the effective area per unit production rate will always be higher in the former case. This is due to the fact that the CD/N in all but the last cell is below the maximum design value. The total power requirement may be calculated, however, by making stage to stage material balances and applying Equation 19 to each cell pair.

NOMENCLATURE

a = empirical constant dependent on cell thickness
A = effective membrane area, sq. cm.
A_p = effective area of one cell pair. sq. cm.
A'_p = effective area of one cell pair, sq. ft.
b = empirical constant dependent on cell thickness and membrane properties
c = empirical constant dependent on cell thickness and membrane properties
CD = current density, amp./sq. cm.
d = empirical constant dependent on cell thickness anb membrane propertics
E current efficiency expressed as a fraction. dimension&
f I fractional dcmincralization per pass
F = product stream flow rate for one cell pair, l./sec.
$\mathfrak{F}$ = Faraday's constant, 96,500 amp.-sec./equiv.
F_p = production rate, l./sec.
F_r = recirculation Row rate of product stream for a production rate F_p, l./sec.
F_t = total product stream Bow rate, l./sec.
g = number of cells required in an internally staged stack to achieve the desired dcminrralization range
I 3 current, amp.
k = number of product stream compartments
$K = \frac{\mathfrak{F}^2}{A_p} \ln q \frac{R_p N_d}{E^2}$
L = cell length, cm.
L_t = total length of flow path. cm.
m = multiple of the production flow rate rquircd for the feed-and-bleed process
n = number of stages required for desired range of demineralization in a continuous process
N = solution normality, equiv./l.
ΔN = change in normality of product stream during one pass through the stack
N_d = product (diluting) stream normality. equiv./l.
N_f = normality of feed to a stack
N_p = normality of product from a stack
N_r = normality of recirculated solution fed to the stack in the feed-and-bleed process
N_t = normality of solution in product recirculation tank
N_{pf} = nomialiry of final product
p = number of passes rrquirrd for batch rrcircularion with mixing
P = direct current power for a single cell pair. w.
P_c = direct current power for a continuous multistage process at constant voltage with one cell pair per stage, w.
P_m = average direct current power demand for batch recirculation with mixing, w.
P_n = power demand for nth pass for batch rccircuiarion at constant voltage without mixing, w.
P_{bm} = average power demand for a batch recirculation process at constant voltage with mixing, followed by a single pass discharge at product conccnrration. w.
P_{bmp} = peak power demand for batch recirculation with mixing and product discharge during final pass, w.
P_{bs} = average direct current power for a batch recirculation process with sfug flow, w.
P_{fb} = direct current power for the feed-and-bleed process, w.
4 = daalination ratio, N_f/N_p
R_p = resistance of one square centimeter of one cell pair, ohm-sq. cm.
s = rate of salt transfer, equiv./sec.
t = cell thickness, cm.
U = power per unit product flow rate. w.-hr./l. sec.
U' = power per unit product flow rate. kw.-hr.: 1000 gal.
v = superficial linear velocity, cm. /sec.
V_p = voltage across one cell pair, V.
V_r = total volume recirculated in batch process durine desalination fmm N_f Co qN_{pf}
V_t = volume of holding tank
W = cell width, cm.
θ = time

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