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LaVerkin Springs, Utah On-Site Pretreatment and Desalting Processes Evaluations

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Bureau of Reclamation

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LAVERKIN SPRINGS, UTAH ON-SITE PRETREATMENT AND DESALTING PROCESS EVALUATION

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16. ABSTRACT

During December 1979 through August 1980, pilot plants and laboratory facilities were operated at the LaVerkin Springs, Utah site to characterize the water and to evaluate various pretreatment processes and desalting processes. The pretreatment processes studied included aeration-precipitation, lime treatment, lime-soda ash treatment, and lime-ion exchange treatment. Desalting at 80 and 90 percent water recovery levels was studied with both electrodialysis and high pressure reverse osmosis process equipment. Also, a "spiractor" was evaluated at the site as a lime-soda ash reactor and as a calcium sulfate desupersaturator. Data and results obtained from the test program will be used for feasibility study, cost estimates, and development and design of any future water collection system and pretreatment and desalting plant installations at the LaVerkin Springs site.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS-- /water pretreatment/ desalination/ *aeration softening/ *lime softening/ *soda ash treatment/ *ion exchange/ *Spiractor/ water hardness/ dissolved carbon dioxide/ *electrodialysis/ *reverse osmosis/ total dissolved solids/ calcium hardness/ calcium seals/ water quality

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LAVERKIN SPRINGS, UTAH ON-SITE PRETREATMENT AND DESALTING PROCESSES EVALUATIONS

by
R. J. Eisenhauer

April 1985

Applied Sciences Branch
Division of Research and Laboratory Services
Engineering and Research Center
Denver, Colorado

UNITED STATES DEPARTMENT OF THE INTERIOR ★ BUREAU OF RECLAMATION

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The following personnel from the Lower Colorado Region assisted in coordinating and implementing setup and startup at the test site:

- Ken Trompeter, Head of Desalting Engineering Branch
- Sus Suemoto, Yuma Project Office engineer
- Pete Esselman, Yuma Desalting Test Facility site manager

The following Engineering and Research Center, Division of Research and Laboratory Services personnel contributed to the success of the site pretreatment and desalting process studies:

- Roy J. Eisenhauer, chemical engineer, principal investigator, and technical manager for the LaVerkin Springs test program from May 1976 to September 1980
- Bob Schiller, lead technician, directed construction of the ion exchange pilot plant, a package chiller unit, and various other equipment units used at the test site
- Ed Backstrom, section head, lead task force that planned and coordinated the test program
- Dr. John Kaakinen, chemical engineer, optimized the design and operation of the ion exchange and electrodialysis pilot plants, principal investigator for the factorial design ion exchange experiments
- Larry Haugseth, chemical engineering unit supervisor, provided technical editing of the entire report
- Bill Boegli, environmental engineer, developed conceptual design for the ion exchange brine regeneration process
- Charles Goodner, technical specialist and resident engineer, coordinated activities between the Bureau and contractor personnel
- Dr. Charles Moody, chemical engineer, consulted on computer calculations of reverse osmosis membrane performance parameters

The following Engineering and Research Center, Division of Design personnel contributed to design on pretreatment, reverse osmosis, and ion exchange pilot plants and site startup assistance:

- Harry Jong, Sanitary and Utilities Section, unit supervisor
- Marilyn Hudson, design engineer
- Clyde Herrington, design engineer
- Wayne Lorenz, design engineer
- Erick Rogers, design engineer
- Don Vernon, design engineer
- Larry Wyeno, design engineer
- Frank Lietz, Technical Engineering Analysis Section, unit supervisor
- Ernst Ewoldsen, reverse osmosis process engineer
- Wolfgang Sattler, Steel Pipe and Special Equipment Section, technical specialist, reverse osmosis equipment design engineer

ABBREVIATIONS USED IN THIS REPORT

A	aeration
A-L	aeration — partial lime treatment
A-L-SA	aeration-lime-soda-ash treatment
A-L-SA-IX	aeration-lime-soda ash-ion exchange treatment
A-L-IX	aeration-lime-ion exchange treatment
CW	Characterization of water
DSI	Desalination Systems, Incorporated
E&RC	Engineering and Research Center, Bureau of Reclamation
ED	electrodialysis
ED-B	electrodialysis brine
ED-P	electrodialysis product
EDPP	electrodialysis pilot plant
EF	effluent filtrate
EP&ID	equipment, piping, and instrument diagram
F	filtration
FC	flocculation and clarification
FSD	Fluid Systems Division
IX	ion exchange
IXPP	ion exchange pilot plant
JTU	Jackson turbidity units
L	partial lime treatment
L-SA-PP	lime-soda ash pilot plant
LT	lime treatment
LTPP	lime treatment pilot plant
LT-SA	lime treatment-soda ash
LVS	LaVerkin Springs
O&M	operation and maintenance
OCC	operations control center
PRC	Planning Research Corporation
R	raw water
RO	reverse osmosis
ROPP	reverse osmosis pilot plant
RR	recycle regenerant
S&A	structural and architectural
SA	soda ash treatment
SCPR	specific cell pair resistance
SCR	solids contact reactor
SHMP	sodium hexametaphosphate
SP	spiractor
SS	selective scaling
TDS	total dissolved solids
USBR	Bureau of Reclamation
USGS	U.S. Geological Survey
YDTF	Yuma Desalting Test Facility
YHR	Yuma High Recovery test program

CONVERSION FACTORS AND SI[1] UNIT PREFIXES

To convert from	To	Multiply by (except where noted)*
cubic meter (m^3)	liter (L)	1.00×10^3
cubic meter (m^3)	gallon (gal)	2.642×10^2
cubic meter per second (m^3/s)	liter per minute (L/min)	6.000×10^4
(m^3/s)	acre-foot per day (acre-ft/d)	7.000×10^1
(m^3/s)	gallon per minute (gal/min)	1.585×10^4
(m^3/s)	gallon per day (gal/day)	2.282×10^7
cubic meter per second per pascal [$m^3/(s \cdot Pa)$]	gallon per day per pound per square inch (gal/day)/(lb/in^2)	1.574×10^1
degree centigrade (°C)	degree fahrenheit (°F)	1.8°C + 32*
gram	pound (lb)	2.205×10^{-3}
meter (m)	inch (in)	3.937×10^1
meter of water (head) at 16°C	foot of water (head)	3.281
meter of water (head) at 16°C	inch of mercury	9.797×10^3
meter per second (m/s)	gallon per minute per square foot (gal/min)/ft^2	1.473×10^3
(m/s)	Gallon per day per square foot (gal/day)/ft^2	2.120×10^6
pascal (Pa)	pound per square inch (lb/in^2)	1.450×10^{-4}
radian per second (rad/s)	revolution per minute (rpm or r/min)	9.551
siemen per meter (S/m)	microsiemen per centimeter (μS/cm)	1.000×10^4
(S/m)	micromho per centimeter (μmho/cm)	1.00×10^4
square meter (m^2)	square foot (ft^2)	1.076×10^1

SI UNIT PREFIXES

Prefix	Symbol	Multiplication factor
mega	M	1 000 000 = 10^6
kilo	k	1 000 = 10^3
centi	c	0.01 = 10^{-2}
milli	m	0.001 = 10^{-3}
micro	μ	0.000 001 = 10^{-6}

[1] International System of Units

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INTRODUCTION

A Colorado River Water Quality Improvement Program

The Bureau of Reclamation initiated a Water Quality Improvement Program in early 1971. The objective of this program was, and still is, to maintain salinity concentrations at or below the levels presently found in the lower main stem of the Colorado River. Control of salts from point sources has been proposed. LaVerkin Springs was one of the point sources under study. The LVS (LaVerkin Springs) flow into the Virgin River which flows into the Colorado River. The purpose of this study was to evaluate the technical and economic feasibility of desalting and to determine the appropriate pretreatment systems to use for the disposal of water by deep well injection or energy development.

1. *Project area.* The project area is near St. George, Utah shown on the vicinity map (fig. 1). The LVS are located on the Washington County map between the towns of Hurricane and LaVerkin. The springs discharge into the Virgin River a short distance upstream of the highway bridge for Utah State Highway No. 17.

2. *Project plan.* A report on controlling salts contributed by LVS was issued in December 1981 [1], which proposed alternatives for controlling the salt content of the Virgin River. The proposed project of building a pretreatment and desalting plant at the site would prevent most of the salts contributed by LVS from entering the Virgin River and being carried into the Colorado River.

Several alternative plans for preventing water from entering the Virgin River were also studied [2]*. These alternate plans include disposal of water by deep well injection, use of water as powdered coal transport media, use of water as secondary coolants in binary cooling towers, and use of water in solar salt gradient ponds. Water used for energy development would require a pretreatment plant at the LVS site, but would eliminate the requirement for a desalting plant.

Under the proposed project desalting plan, the springflows would be collected and conveyed to a desalting facility that would have a pretreatment unit and an ED (electrodialysis) or RO (reverse osmosis) desalting plant. The product water from the desalting plant would be returned to the Virgin River below the project, while the brine containing most of the salt would be conveyed to a brine pond, where the water would be evaporated. The locations of the project features, which include dams, a bypass facility, a pumping unit, a desalting facility, and a brine pond, are shown on the detailed insert of figure 1.

The river water would be separated from the springflows by constructing a diversion dam in the canyon upstream from the springs. This dam would divert the flows of the river into a conduit which would convey them around the springs area. The conduit would be sized to carry all of the normal riverflows and most of the floodflows of the river. A control dam would be constructed immediately below the springs area to collect the spring discharges in a small pool, thus providing a forebay from which the saline water would be pumped to a desalting facility. Water would not be pumped from the forebay during occasional periods of extreme floodflows which exceed the capacity of the bypass conduit.

The product water would be returned to the river system west of the desalting plant. The brine from the desalting unit and the sludge from the lime-soda pretreatment unit would be conveyed to a lined evaporation pond, where the brine discharge would be stored while the liquid evaporated and the salt precipitated out. There would be adequate capacity in the pond to store the estimated salt accumulation for more than 200 years.

If LVS water is disposed of by deep well injection or used for energy development such as for cooling towers or salt gradient ponds, tha spring water would require only pretreatment. Consequently, there would be no need for disposing of desalted water and brine water at the site. However, calcium carbonate sludge disposal would still be required at or near the site. The quantities of sludge requiring disposal would depend on the degree of pretreatment required for the chosen alternate method of disposal. The pretreated water would be pumped through a pipeline to the area of deep well injection or to the area where it would be used for energy development.

B Spring Water Supply

1. *LaVerkin Springs setting.* — The springs issue at intervals from many points in and along the river for a distance of about 2000 ft. The located 20 separate springs are shown in figure 2. During periods of low riverflow (June through October), most of the hot springs water can be observed emerging from fissures and joints on both banks of the stream, from the riverbed, and from a tunnel in the limestone. The springs are

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* Numbers in brackets refer to entries in bibliography

hot, varying from about 100 to 109 °F, and are heavily mineralized. A strong odor of hydrogen sulfide gas is present in the vicinity of the springs. Tests for this gas indicate a concentration of about 4 mg/L as the springs emerge. The only direct usage of the spring water is by the Pah Tempe Hot Springs Resort, where the water is used for baths and therapeutic purposes.

2. *Water Quantity.* — Because of the great number of spring outlets, many of which are in the riverbed, direct measurement of spring discharges is not practical. Determinations of the discharges have been made by measuring the flow of the river both above and below the springs area. Measurements that reflect spring discharges with reasonable accuracy were made when the flow in the river above the springs was low. Table 1 lists under the heading "Estimated Discharge" the measurements of the stream made over the period of 1943-1972. After studying these measurements, it was decided to use springflows of a maximum flow of 0.34 m³/s (12 ft³/s) for all designs.

C. Test Site Description

1. *Site preparation.* — The test site was located adjacent to the Virgin River in Timpoweap Canyon approximately one-half mile downstream from the LaVerkin Springs and the well, identified as LV102, in figure 2. The site, prior to modification, consisted of a low-lying alluvial bar at a bend in the river directly below the Utah State Highway No. 15 bridge that connects the city of Hurricane with the town of LaVerkin. During the last week of May 1979, site preparation was initiated. A local contractor filled, leveled, and compacted the site to an elevation above the probable flood level of the Virgin River. A rough entrance dirt road was scraped level and compacted. A brine pond and an intermediate and a long-term storage sludge pond were excavated. Later, the site was graveled and compacted.

2. *Test equipment.* — The test equipment at the site consisted of trailer-mounted treatment equipment and ground-mounted process water tanks. The Spiractor installed for Phase II was mounted on a small concrete pad adjacent to the electrodialysis trailer, and the final-configuration aeration tank was also mounted on a concrete pad.

The following trailers and quipment were installed at the test site for the phase I testing which began in December 1979 (more detailed descriptions are given in appendix A):

Laboratory trailer;
OCC (Operations Control Center) trailer;
Pretreatment trailer and associated equipment;
Filter trailer;
Shop trailer;
Ion exchange trailer and associated equipment;
ED trailer and associated equipment;
Storage trailer; and
Office trailer.

All process pipe interconnecting test trailers, associated quipment, and ponds were laid underground in 2-foot-deep trenches and covered over with dirt.

After the program was expanded to include the phase II testing, the following were received and installed in April 1980:

RO trailer and associated equipment; and
Permutit Spiractor and associated equipment.

3. *Final equipment and piping layout.* — The final site equipment layout is diagrammed in figure 3. A partial piping layout is also included as figure 4. This diagram shows all major site piping.

4. *Photographs of test site.* — Figures 5a and 5b are panoramic views of the operating test site. The photographs were taken in February 1980 prior to the arrival of the additional phase II equipment. The brine pond and the sludge ponds can be seen in the foreground in figure 5a and to left in figure 5b. The layout of equipment can be identified by reference to figure 3. The flow of the Virgin River is apparent along the left side in figure 5a and behind the test equipment in figure 5b. The buildings on the other side of the road in the upper third of the photograph (fig 5a) belong to the Pah Tempe Hot Springs Resort. All the springs are located adjacent to the spa and toward the rear of the photographic view (fig 5a) for 2000 ft. A pump house can be seen at the end of the dirt road that runs up the left side of the photograph (fig 5a). Raw spring water was pumped through 3-in PVC (polyvinyl chloride) pipe from a well at the pump house to the test area.

Table 1. — *Estimated quantity of springflow, LaVerkin Springs Unit, Utah [1]*

Date	Estimated discharge			Data source
	Below springs (ft³/s)	Above springs (ft³/s)	Spring discharge (ft³/s)	
July 12, 1943	12.54	1.00	11.54	USBR[1]
July 23, 1943	12.50	1.50	11.0	USBR
July 16, 1947	11.44	—	11.44	USBR
Sept. 10, 1956	—	—	10.0	USGS[2]
Aug. 10, 1960	—	—	10.6	USGS
Aug. 1, 1963	—	—	10.2	USGS
Aug. 21, 1963	—	—	11.6	USGS
(No date)	—	—	10.0	USGS
June 21, 1972	14.18	3.25	10.93	USBR
July 12, 1972 a.m.	13.29	1.44	11.85	USBR
July 12, 1972 p.m.	11.88	0.94	10.94	USBR
July 13, 1972	12.18	0.71	11.47	USBR

[1] Bureau of Reclamation
[2] U.S. Geological Survey

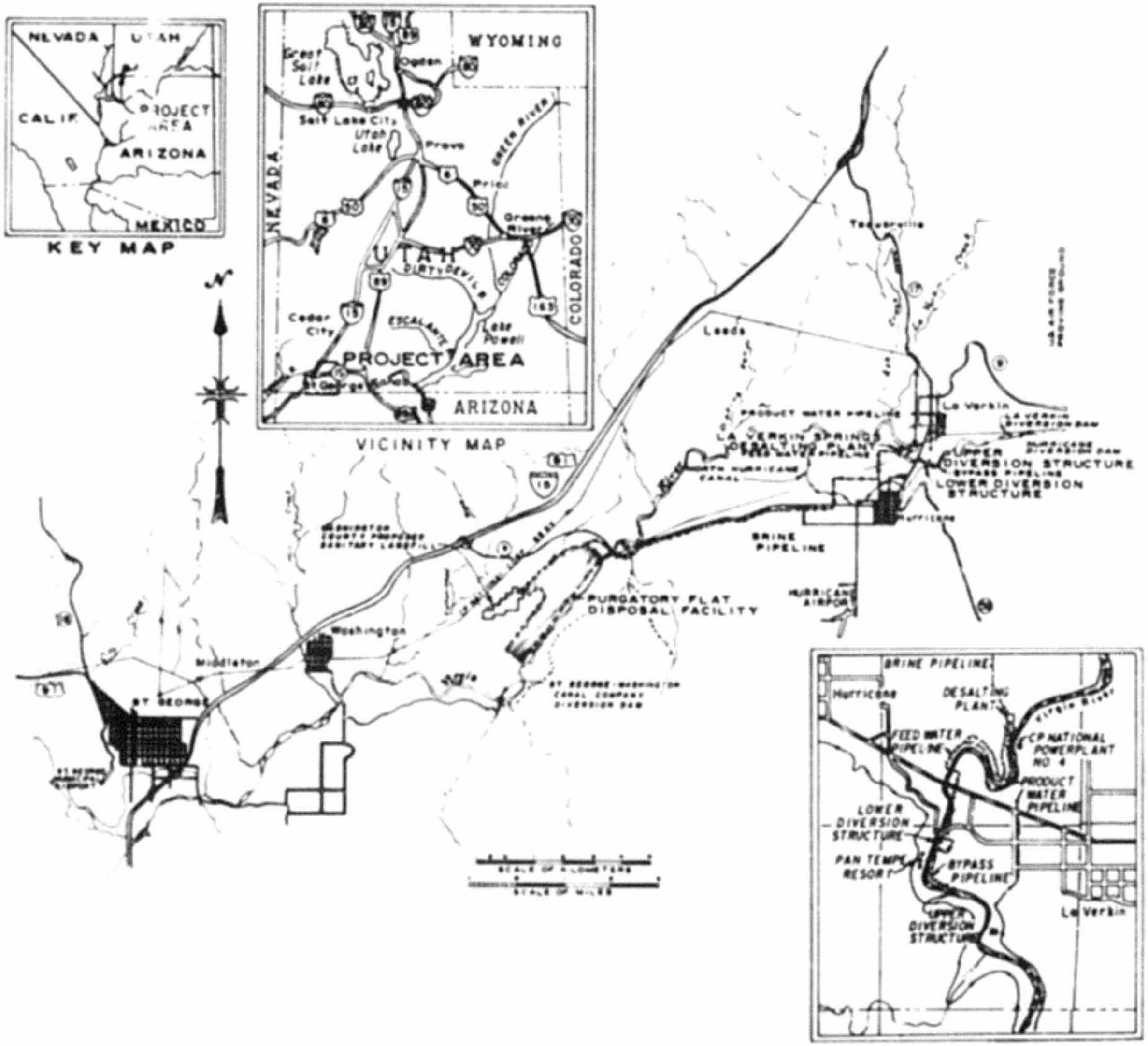

Figure 1.—Area map.

NOTES

The staff gages and flumes were temporary and used only during the duration of the study. Source 1-West cave is the main source of water for the resort.

Figure 2.–Spring locations.

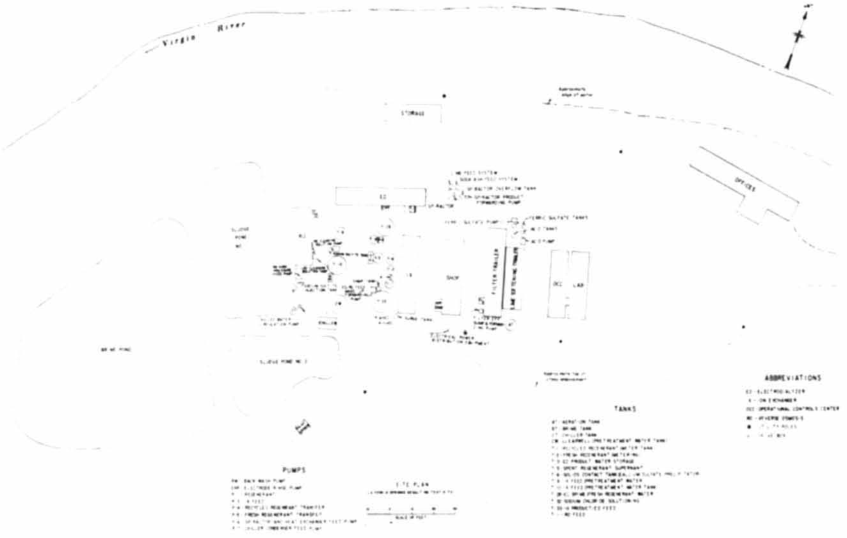

Figure 3.–Site plan.

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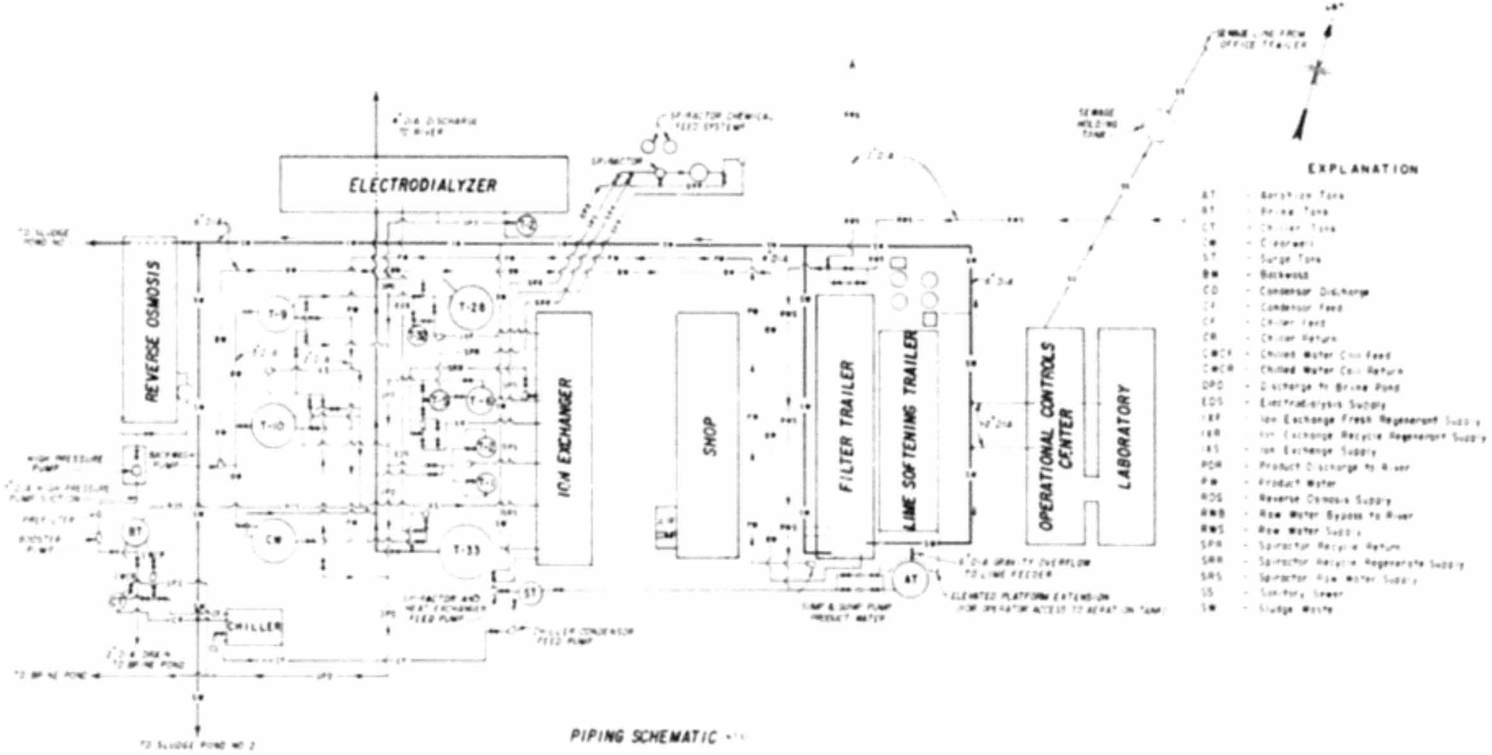


Figure 4.—Piping schematic.

Figure 5a.–The LVS test site upriver view, February 1980. Photo P801-D-80756.

Figure 5b.–The LVS test site river bank view, February 1980. Photo P801-D-80757.

TEST SCHEDULES AND TASKS

The operating contractor, PRC (Planning Research Corporation), completed site preparations and started installation of test equipment on October 1, 1979. The first phase of testing began October 29, 1979, and continued to April 21, 1980. The second phase of testing began on April 27, 1980, and terminated on August 31, 1980. During September 1980, PRC dismantled equipment, and a summary data report [3] was issued in August 1981.

A. Phase I Test

A schedule of the test program for phase I of the task performed is shown in figure 6. A summary of the task descriptions for phase I is given in table 2.

1. *Task CW (characterization of water).* — The purpose of task CW was to thoroughly chemically characterize raw LVS water as it pumped from a well and after it passed through each stage of pretreatment and desalting. Water characterization tasks CW-IA, IB, and IC were accomplished by logging and averaging eight hourly raw water influent and process stream control and analyses and control instrument readings. Task CW-IIA was accomplished by submitting weekly samples of river water, raw water, and water after each stage of processing to an offsite analytical laboratory for cation and anion analysis. Task CW-IIB covered special sampling and analysis required for the IX (ion exchanged process study. Task CW-IIC covered trace metal, organic, and special offsite analysis every 4 weeks for the same streams covered by CW-IIA.

2. *Task LT (lime treatment process study.* — The LT operation had the following objectives:

   Demonstrate the feasibility of aeration and LT of LVS water;

   Operate for several weeks at the different practical variable levels to determine the optimum conditions for achieving minimum lime dosage, pH, residual calcium, residual silica, and turbidity;

   Operate for 3 months at the established best conditions to obtain data for plant design and for calculations of O&M (operation and maintenance) and capital cost and to provide softened feed water for the IX operation; and

   Demonstrate aeration removal of dissolved carbon dioxide from raw water and the precipitation of calcium carbonate originally held in solution by carbon dioxide.

   Descriptions of task LT-I, -II, -III, and -IV are given in table 2.

3. *Task ED (electrodialysis) process study.* — The ED operation had the following objectives:

   Determine the performance characteristics of the ED system at various levels of water recovery;

   Demonstrate the feasibility of high recovery desalting of LVS pretreated water;

   Operate for 3 months to obtain data for plant design and for calculations of O&M and capital cost;

   Produce brine of the quantity and concentrations required for regenerant needed for the (IX) parametric study; and

   Collect performance data while operating the ED system at 80 and 90 percent water recovery.

   Descriptions of tasks ED-I, II, III and IV are given in table 2.

4. *Task IX (ion exchange) process study.* — The IX operation had the following objectives:

   Demonstrate that IX with desalting process brine regeneration will remove residual calcium in lime-treated water meeting feed water specifications for membrane desalting;

   Optimize the IX process by performing a statistical factorial experiment;

   Demonstrate the viability of regeneration of IX resin with brine from 90 percent water recovery membrane desalting; and

   Operate for 3 months at different cycle conditions to obtain data for plant design and for calculations of O&M and capital cost.

   Descriptions of tasks IX-IA, B, and C are given in table 2. Task IX-II was a 2^4 factorially designed experiment. The IX system was operated through 18 different run conditions. The results of this factorial-designed experiment have been reported by Kaakinen [4].

B. Phase II Test

A schedule and a task description table of the test program for phase II of the task performed are shown in figure 7.

1. *Task LT-SA (lime treatment-soda ash) process study.* — The purpose of lime treatment task V was to establish the lime treatment operation at a design flowrate of 38 L/min before proceeding

with soda ash addition study of phase II. (Throughout phase I, it was necessary to operate the LT equipment at a flow rate of 57 L/min or 150 percent of design basis to provide enough lime-softened water to perform all the task under phase I.) Task LT-V also provided equilibrium data for later O&M cost calculations. Data were obtained by continuing operation of LT without soda ash addition for 48 hours at the following equilibrium conditions:

- Raw water: Influent flow = 170 L/min (45 gal/min)
  Return to river = 132 L/min (35 gal/min)
- Train effluent: Flow = 38 L/min (10 gal/min)
- Aerated unit effluent: $CO_2 < 20$ mg/L
- Rapid mix tank: Flow = 38 L/min
  Lime feed rate control 9.8 pH
- Flocculator tank: Speed = 140 r/min
- Clarifier tank: Speed = 0.6 r/min
  Sludge recycle = 19 L/min (5 gal/min)
- Ferric Sulfate concentration: 7.5 mg/L

Soda ash addition was phased into the lime treatment as task LT-SA-I. The chemical treatment was continued as Task LT-SA-II to optimize the operation and to make necessary equipment modification for the lime-soda ash treatment process. Task LT-SA-III was a study of parameters for obtaining optimum operations data for O&M cost calculations.

2. *Task ED process study.* — Summary descriptions of task ED-I, II, III, and IV are given in table 2. After completing task ED-IV from phase I, the ED system was shut down to perform phase II task ED-V, an internal inspection of the cell stack. Since the IX resin was to be regenerated with RO (reverse osmosis) brine instead of with ED brine, the ED system was left shut down until August 10, 1980. The ED system was operated the last 3 weeks of the test program at the same conditions as phase I task ED III to provide reject brine for regeneration of IX resin.

3. *Task IX process study.* — Summary descriptions of phase I tasks IX-I and II are given in table 2. The system was operated as phase II task IX-III according to the table on figure 7, with lime-soda ash treated water to minimize the residual calcium concentration in the chemically treated water and to determine the regeneration efficiency of RO brine. The IX system was operated as task IX-IV the last 3 weeks of August 1980 to test the effectiveness of the Spiractor in removing calcium sulfate from recycle regenerant brine.

4. *Task RO (reverse osmosis) evaluation of membranes.* — Description of tasks RO-I, II, and III are given in the table of figure 7. The RO desalting operation had the following objectives:

Demonstrate the feasibility of high recovery desalting of LVS pretreated water;

Produce brine of the quantity and concentration required to regenerate IX resin;

Operate at steady-state conditions to obtain data for plant design and for calculations of O&M and capital cost; and

Evaluate new, high-temperature RO membranes.

5. *Task SP (spiractor) process study.* — Descriptions of phase II tasks SP-I, II, III, and IV are given in table of figure 7. The SP operation had the following objectives:

Demonstrate lime-soda ash treatment of LVS water with nucleation and densification of precipitated calcium carbonate on sand particles; and

Demonstrate calcium sulfate desupersaturation of ED-produced reject brine for IX regenerant by nucleation on calcium sulfate pellets.

6. *Task SS (selective scaling) problem.* — At the outset of the operation at the LVS site, considerable problems were experienced with scale formation in pipes, valves, instrument sensors, and aeration equipment located between the raw water aeration tank and the rapid mix tank of the LT train. Task SS-I modified the aeration equipment and changed the aeration operation to eliminate shutdowns caused by excessive calcium carbonate scale buildup on the internal surfaces of equipment in front of the LT train.

7. *Task CW.* — The CW task described in phase I was continued through phase II.

Table 2. — *Description of tasks for the phase I test program*

Task	No	Source	Description
CW. characterization of spring water	IA	Raw water	Site instrument readings and control analyses, once per 8-hour shift
	IB	Aerated water	Site instrument readings and control analyses, once per shift
	IC	Lime-softened water	Site instrument readings and control analyses, once per shift
	IIA	River water, raw water, aerated clarifier effluent, stabilized effluent, filter effluent, ED (electrodialysis) feed, ED brine, and ED product	Samples for complete analysis by an offsite laboratory, weekly
	IIB	IX (ion-exchange) data run samples	Special samples for partial and complete analysis by an offsite laboratory, monthly
	IIC	Same streams as IIA	Special samples for heavy metals, organics, and radionuclides, monthly
LT. Lime pretreatment process demonstration and lime softening of water for Task LVS-IX	I	LTPP (lime treatment pilot plant) 38-L/min design basis flow rate	Hydraulically test and shakedown LTPP, including aeration system
	II	LTPP	Start lime addition and stabilization operation at baseline conditions
	III	LTPP	Establish dual media sand filter operation
	IV	LTPP	Produce lime-treated water for IX study at 57L/min and 9.8 pH
ED. electrodialysis cell performance study and brine production for IX regenerant	I	EDPP (electrodialysis pilot plant) Ionics Aquamite V, two electrical stages, six hydraulic stages, 24-L/min nominal flow rate	Hydraulically test and shakedown EDPP system
	II	EDPP	Performance test of EDPP at 73, 80, 85, 90, and 93 percent water recoveries
	III	EDPP	Operate the EDPP to produce brine as required for regeneration of the IX resin
	IV	EDPP	Analytical and operational data acquisition for performance study of EDPP at 80 and 90 percent water recovery levels
IX. ion exchange process optimization study	IA	IXPP (ion exchange pilot plant), two columns, 36-cm diameter by 2-m high, packed with 100 mL of IX resin	Renovation and installation of IXPP at LVS site to meet specific requirements of test conditions
	IB	IXPP	Hydraulically test IXPP by going through dummy cycles with water
	IC	IXPP	Shakedown and optimize IXPP operation by softening lime treated water and regeneration with 10 percent sodium chloride
	II	IXPP	Determine feasibility of using IX to remove permanent hardness calcium from lime treated water by performing a statistically designed experiment

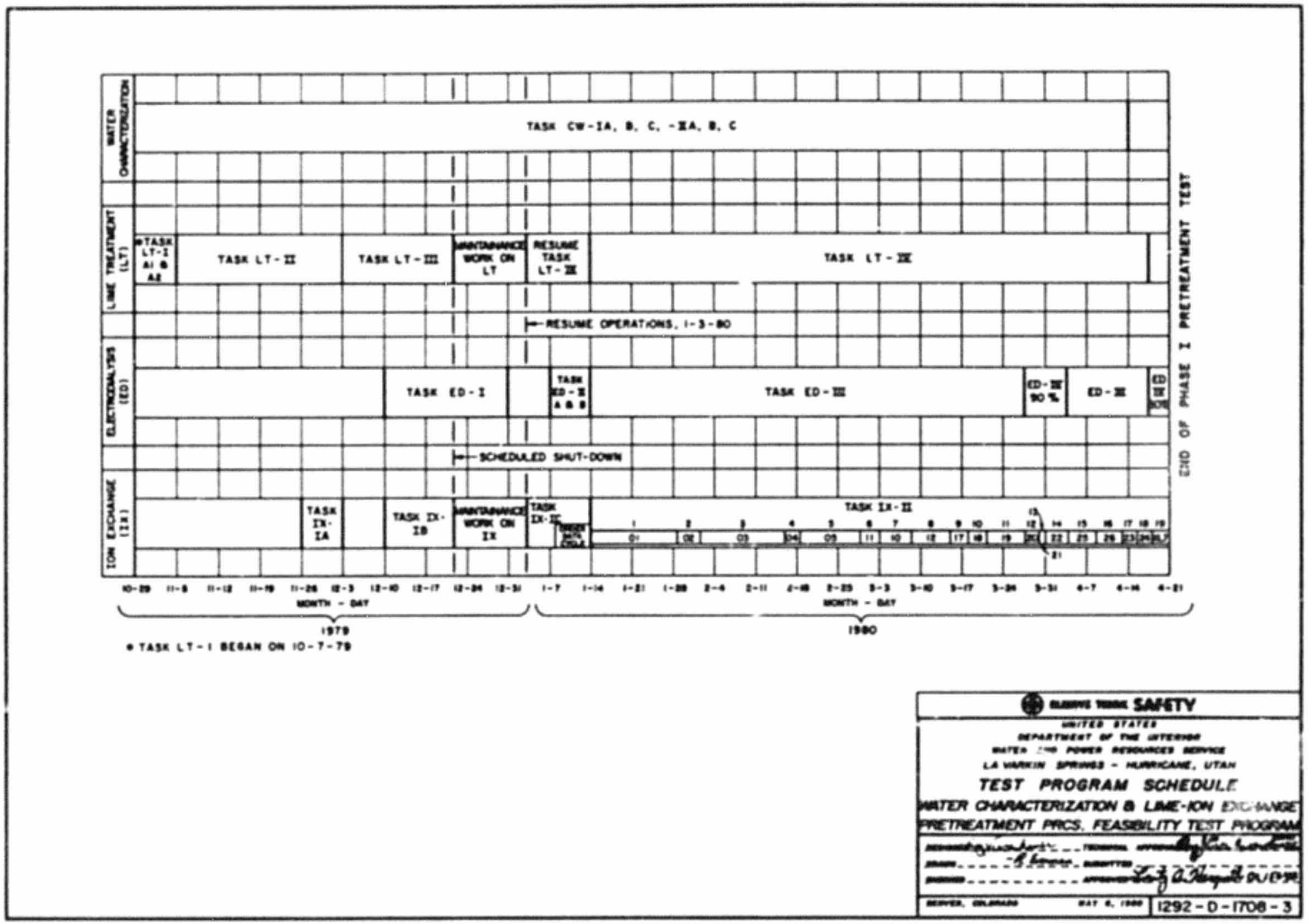
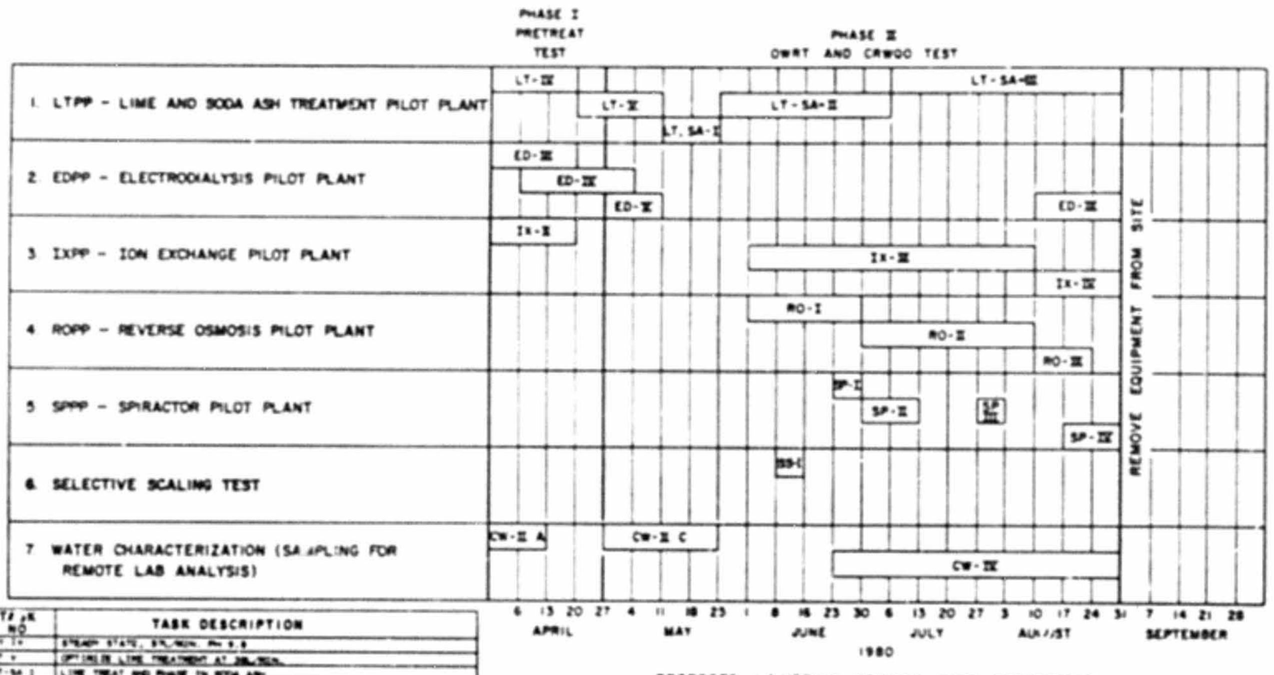


Figure 6.—Phase I test schedule.



PROPOSED LA VERKIN SPRINGS SITE OPERATIONS

ITEM NO	TASK NO	TASK DESCRIPTION
1	LT-IV	STEADY STATE, 57L/MIN, PH 9.8
2	LT-V	OPTIMIZE LIME TREATMENT AT 38L/MIN.
3	LT-SA I	LIME TREAT AND PHASE IN SODA ASH
4	LT-SA II	LIME-SODA ASH OPTIMIZATION AT 38L/MIN.
5	LT-SA III	LIME-SODA ASH PRETREATMENT FOR RO AND ED
6	ED-III	ELECTRODIALYSIS OPERATION FOR IX STUDY
7	ED-IV	ED PERFORMANCE TEST FOR 80% AND 85% RECOVERY
8	ED-V	DISASSEMBLY OF STACK FOR INTERNAL INSPECTION
9	IX-II	PARAMETRIC STUDY OF REJECT BRINE REGENERATION
10	IX-III	CALCIUM POLISHING WITH SODIUM CHLORIDE REGENERATION
11	IX-IV	CALCIUM SULFATE SCALING FOR SP-IV
12	RO-I	HYDRAULICALLY TEST AND SHAKEDOWN RO SYSTEM
13	RO-II	MEMBRANE PERFORMANCE TEST AT 5520 kPa
14	RO-III	MEMBRANE PERFORMANCE TEST AT 4210 kPa
15	SP-I	HYDRAULICALLY TEST AND SHAKEDOWN SPIRACTOR
16	SP-II	SPIRACTOR LIME-SODA TREATMENT AND SLUDGE REDUCTION
17	SP-III	SHAKEDOWN OF SPIRACTOR SYSTEM
18	SP-IV	CALCIUM SULFATE DESATURATION OF 40,000 MG/L BRINE
19	SS-I	SELECTIVE SCALING DURING RAW WATER AERATION
20	CW-IIA	COMPLETE ANALYSIS OF RAW, AERATED, LT, IX, AND RO WATERS
21	CW-IIC	ANALYSIS OF WATER FOR VARIOUS CONTAMINANTS
22	CW-IV	COMPLETE ANALYSIS OF RAW, AERATED, LT-SA, IX, AND RO WATERS

8-19-80 B-	DESCRIPTION UPDATED

ALWAYS THINK SAFETY

UNITED STATES
DEPARTMENT OF THE INTERIOR
WATER AND POWER RESOURCES SERVICE
LA VERKIN SPRINGS - HURRICANE, UTAH

TEST PROGRAM SCHEDULE
WATER CHARACTERIZATION, PRETREATMENT,
DESALINATION AND DESATURATION TESTS

DESIGNED .... TECHNICAL APPROVAL ....
DRAWN ..... R. Grenone SUBMITTED ....
CHECKED .... APPROVED ....

DENVER, COLORADO APRIL 8, 1980 1292-D-1708-5

Figure 7.—Phase II test schedule.

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CHARACTERISTICS OF SPRING AND PROCESSED WATERS

A major objective of the LVS site test program was to characterize the spring water as it was pumped from a well and as it emerged from various pretreatment and desalting processes.

A. **Summary**

The following characteristics were obtained for the LVS waters:

- The raw water contained 750 mg/L of dissolved carbon dioxide. Typically, LVS water contained 400 mg/L of calcium temporary hardness, and 570 mg/L of calcium and magnesium permanent hardness.
- About 50 percent (200 mg/L) of the calcium temporary hardness can be precipitated by air sparging the raw water or by holding the water in an open reservoir for 48 hours.
- The remaining 200 mg/L of calcium temporary hardness can be removed by partial lime treatment.
- About 80 percent of calcium-magnesium permanent hardness can be removed by following partial lime treatment with either ion exchange or soda ash treatment.
- A 90-percent water recovery by ED (electrodialysis) of pretreated LVS water containing 30 mg/L of calcium, 58 mg/L of magnesium, and 8710 mg/L of TDS (total dissolved solids) was demonstrated. The ED product water contained 1370 mg/L of TDS, and less than 30 mg/L of calcium and magnesium. The 10 percent of the water recovered as brine, contained 94 900 mg/L of TDS, about 900 mg/L of calcium and magnesium, and was very prone to precipitation of calcium sulfate. A total of 7480 mg/L TDS was removed from feed water.
- Further characteristics of LVS water as it emerged from a well pump (raw water) and of the water through various stages of processing are summarized in table 3.

B. **Methods**

Analytical methods and instrumentation used to characterize LVS raw water and water from each step of processing are summarized in appendix B.

The CW (water characterization) phase was scheduled between November 29, 1979, and August 28, 1980. Water was characterized by process control analyses and control instrument readings once every 8 hours on influent raw water, aerated feed water, and lime treated water and by once-a-week total analysis of all process influent and effluent streams. Since a focal point of the pretreatment test program was to study the effects of ten operating conditions on process parameter responses, extensive analyses were needed for each change in a process condition. Special analyses were made to determine the presence of trace metals and contaminants.

1. *Analyses of shift samples.* — Task CW-I provided characteristic data on the water by once-a-shift control readings and analyses. Process control water samples were taken in 250-mL capped plastic bottles and taken to the laboratory trailer for the following analyses:

• Raw water:	Temperature pH Carbon dioxide Sulfide Calcium Alkalinity Magnesium
• Aerated water:	pH Carbon dioxide Sulfide Calcium Alkalinity Magnesium Turbidity
• Lime-softened water:	pH Calcium Alkalinity Magnesium Turbidity

2. *Analyses of weekly samples.* — Task CW-II provided characteristic data on the water by sending samples to a fully equipped offsite laboratory for analyses. The following procedures were used for weekly analyses:

a. Duplicate samples were collected in 250-mL capped plastic bottles and shipped each Monday morning. One of the samplings was adjusted to pH 2 with nitric acid before sealing. For each sampling, initial pH and temperature of the water were determined and so marked on the bottle.

b. The samples which were collected weekly and sent to an offsite laboratory for analysis included:

River water
Raw (well) water
Aerated water
Clarifier effluent
Stabilized clarifier effluent
Filter 3 or 4 effluent
Ionics ED feed (IX product)
Ionics ED brine
Ionics ED product

c. The following analyses were made by the offsite laboratory on the unacidified but filtered weekly samples:

pH
Conductivity
Sodium
Potassium
Calcium
Magnesium
Iron
Manganese
Strontium
Carbonate
Bicarbonate
Chloride
Sulfate
Silicates
P-alkalinity
T-alkalinity

3. *Analyses of acidified samples.* — The following analyses were made by the offsite laboratory on the acidified weekly samples:

Calcium
Magnesium
Iron
Manganese
Strontium

4. *Analyses of IX samples.* — Samples were analyzed for each IX process condition level [4] data run. During each service, backwash, and regeneration mode of the data run, 12 to 24 separate 250-mL influent and effluent samples were taken. Portions of each of the individual samples of influent or effluent were appropriately blended to produce 5 to 7 duplicate 250-mL "time weighed average" composite samples of each mode.

Duplicate composite samples of a service mode, a backwash mode and a regeneration mode for each data run were analyzed for:

Calcium
Magnesium
Sodium

5. *Analyses for trace metals and contaminants.* — Special analyses were made for possible trace metals and contaminants in the raw water, filter effluent, and ED product.

The offsite laboratory required three 1-liter samples of each stream twice, 4 weeks apart, for the following analyses:

Arsenic
Barium
Cadmium
Chromium
Copper
Lead
Mercury
Selenium
Silver
Zinc
Total phosphates
Total nitrogen
Insecticides
Radionuclides

C. **Results and Discussion**

The 9-month test program was completed at the LaVerkin Springs site on August 30, 1980. Details of the test program are available [3]. Summary data of water characteristics from the site test program relevant to this report are included in table 3. A separate report [5] evaluated raw water data in table 3 with data from February 11 through November 6, 1972, for 14 springs at the LVS site.

1. *Raw water.* — When LVS water is pumped from a well, the raw water typically contains 750 mg/L of dissolved CO_2 and has a pH near 6.0. Also, the raw water typically contains the following concentrations (in mg/L): calcium, 820; magnesium, 150; bicarbonate, 1266; sulfate, 1860; sodium, 2220; and chloride, 3345. Essentially, 400 mg/L of calcium are associated with the bicarbonate as temporary hardness, and 420 mg/L of calcium are associated with chloride and sulfate as permanent hardness. The raw water contains significant amounts of silica, boron, strontium, and sulfide in concentrations of 40, 5.0, 10.2, and 4.9 mg/L, respectively. All trace elements in the raw water are in concentrations of less than 0.1 mg/L. A measurable amount of radioactivity of 33 pCi/L of radium-228 is present in the water. At the bottom of table 3, the solubility of calcium sulfate in the raw water is expressed as percent water recovery; i.e., for line item 45, solubility at 20 °C is 23.0 percent. This means, when raw water is fed directly to an RO or ED process, calcium sulfate could precipitate in the brine if the process was operated at a water recovery rate greater than 23 percent. Anotner interpretation is that raw water should not be desalted by either ED

or RO without sufficient pretreatment. Also, raw water has a high calcium carbonate scaling tendency due to dissolved CO_2.

2. *Aerated water.* — Before coming to the surface as a spring, LVS's raw water is under subterranean pressure. After the water comes to atmospheric pressure at the surface, CO_2 will escape from the water. Standing water tests were performed at the LVS site to determine the amount of CO_2 that could be removed from the raw water with ponding [6]. The tests consisted of filling four 1-liter beakers with raw water and allowing the beakers to stand quiescent. After standing 0, 24, 48, and 72-hours, analyses were made of filtered aliquot. The following changes occurred in the water:

	Number of hours quiescent			
	0	24	48	72
pH	6.2	7.3	7.5	7.6
Ca, mg/L	820	806	725	677
Free CO_2, mg/L	712	56	18	17

Test data showed that an aliquot of LVS raw water after standing for 24 hours would reduce in CO_2 content from 712 to 56 mg/L. After 48 hours, CO_2 reached a near-minimum level of 18 mg/L. Although at 24 hours, analysis showed 93 percent of the CO_2 escaped solution, 98 percent of the calcium was still in solution. At 48 and 72 hours, 88 and 82 percent of the calcium remained in solution, respectively. The chemical change was verified by physical observations. At 24 hours, calcium carbonate precipitate could not be observed. At 48 hours, a light coating of white precipitate was observed. At 72 hours, a markedly visible coating was seen.

During the site test program, a continuous process was optimized and demonstrated for air purging CO_2 from the spring water. The characteristics of water after thorough aeration are shown in column 5 of table 3. Aeration reduced CO_2 from 750 to 18 mg/L and raised pH from 6.0 to 7.4. At the same time, calcium was reduced from 820 to 600 mg/L (or to 73 percent) by calcium carbonate precipitation. The TDS was reduced by 500 mg/L, to 9054 mg/L. No appreciable changes occurred in concentrations of magnesium, sodium, potassium, sulfate, or chloride. Aeration reduced silica from 40 to 30 mg/L, boron from 5.0 to 4.3 mg/L, strontium from 10.2 to 6.4 mg/L, and sulfide from 4.9 to 0.7 mg/L. Aeration improved the calcium sulfate solubility slightly, i.e., for line item 45 of table 3, calcium sulfate solubility increased at 20 °C from 23.0 to 37.2 percent water recovery. This indicates that after aeration, calcium sulfate would precipitate if desalting water recovery exceeded 37.2 percent.

3. *Aerated-lime treated water.* — During the LVS site test program, a continuous process was optimized and demonstrated for partial lime treatment of aerated water. The characteristics of LVS water after A (aeration) and subsequent partial L (lime) treatment are summarized in column 6 of table 3. Partial L treatment to an adjusted pH of 9.5 removed the rest of the calcium temporary hardness from solution by preciptation as calcium carbonate. Permanent calcium hardness remained at a level of 420 mg/L.

Overall, A-L treatment removed 49 percent of the calcium from solution. The A-L treatment also changed water characteristics by reducing the concentrations of silica from 40 to 15 mg/L, magnesium from 150 to 128, bicarbonate from 1266 to 0, strontium from 10.2 to 4.7, and sulfide from 4.9 to ND (not detectable). The TDS was reduced from 9507 to 8530 mg/L. As indicated by line item 45 of table 3, precipitation of calcium sulfate is not likely since the percent water recovery possible before calcium sulfate precipitation occurs is 49.9 percent at 20 °C. However, A-L pretreatment does not produce sufficiently softened feed water for high recovery ED or RO desalting.

4. *Aerated-lime-ion exchange treated water.* — Following removals of temporary calcium hardness by aeration and lime treatment, a process was optimized and demonstrated for IX (ion exchange) treatment. The characteristics of water after A-L-IX treatment are summarized in column 7 of table 3. Before IX treatment, pH was adjusted from 9.5 to 7.5 with sulfuric acid. Sodium-calcium ion interchange is very effective in removing permanent hardness. Calcium was reduced from 420 to 30 mg/L and magnesium was reduced from 128 to 58 mg/L. Ion exchange did not significantly affect the concentration of silica or boron, but reduced strontium from 4.7 to 0.4 mg/L. The TDS increase from 8530 to 8710 mg/L was primarily due to divalent calcium exchange by monovalent sodium. The calcium sulfate precipitation tendency of A-L-IX treated water is essentially nil. At 20 °C, 92 percent of the water could be removed from the treated solution before any calcium sulfate would precipitate. The A-L-IX pretreated water is acceptable as feed for high recovery ED or RO desalting.

5. *Aerated-lime-soda ash treated water.* — One of the objectives of the LVS site test program was to compare IX treatment with SA (soda ash) treatment for removal of calcium permanent hardness. During the test program, a continuous process was optimized and demonstrated for SA treatment of A-L treated water. The characteristics of IX treated water are compared with SA treated water in columns 7 and 8 of table 3. The IX treatment reduced calcium to 30 mg/L, and SA treatment reduced calcium to 60 mg/L. Also, IX treatment reduced magnesium to 58 mg/L and SA treatment reduced magnesium to 93 mg/L. Neither treatment had a significant effect on silica or boron. The IX treatment reduced strontium to 0.4 mg/L compared to 1.0 mg/L for SA. The A-L-SA treatment reduced radium-226 from 33 to 1.5 pCi/L. The total hardness of either IX or SA treated water is sufficiently reduced for use as feed water in a desalting process.

6. *Aerated-lime-soda ash-ion exchange (polished) treated water.* — Characteristics of A-L-SA-IX treated water were also determined during the LVS site test program. Polishing with IX essentially removed all silica, calcium, and strontium as indicated by column 9 of table 3.

7. *Electrodialysis desalted water.* — An objective of the LVS site test program was to optimize and demonstrate an electrodialysis desalting process. The characteristics of ED-P (electrodialysis product) water at 90 percent water recovery are shown in column 10 of table 3. Characteristics of the ED-B (electrodialysis brine) are shown in column 11. The feed water used for ED desalting was A-L-IX pretreated. The TDS in the ED-P was 1370 mg/L and in the ED-B was 94 900 mg/L. About 88 weight percent of the calcium and about 76 weight percent of the magnesium ended up in the brine. Boron ended up in nearly equal concentrations in the product water and brine. All the remaining radium-226 ended up in the brine. Compared with raw water and the aerated water, ED-B has a higher tendency to precipitate calcium sulfate. Referring to line item 45, if the ED-B were to pass through another desalting step, calcium sulfate would precipitate when water recovery exceeded 15.4 percent. Another interpretation would indicate calcium sulfate precipitation will occur in a brine pond when more than 15.4 percent of the water evaporates.

Table 3. — *Characteristics of LVS water through stages of processing*

(1)	(2)	(3)		(4)	(5)	(6)	(7)	(8)	(9)	(10)	(11)	(12)
										Electrodialysis (90 percent water recovery)		
Item	Characteristic	Units		Raw water (R)[1]	Aerated water (A)[2]	Lime-treated water (A-L)[3]	Lime-ion exchange (A-L-IX)[4]	Lime-soda-ash treated (A-L-SA)[5]	Lime-soda-ash IX treat (A-L-SA-IX)[6]	Product (ED-P)[7]	Brine (ED-B)[8]	Item
1	pH	—		6.0	7.4	9.5	7.5	10.2	7.6	7.4	7.7	1
2	TDS[9] (calculated)	mg/L		9243	8754	8430	8930	8282	8627	1370	91 900	2
3	TDS (evaporated at 105°C)	mg/L		9507	9054	8530	8710	9062	9407	1370	94 900	3
4	Conductivity at 25 °C	mS/L		1480	1410	1280	1320	1380	1450	264	8 590	4
	Major constituents											
5	Silica	mg/L		40	30	15	15	2	2	6	8	5
6	Calcium	mg/L		820	600	420	30	60	6	6	366	6
7	Magnesium	mg/L		150	144	128	58	93	46	20	557	7
8	Sodium	mg/L		2220	2220	2300	2970	2690	2970	460	32 100	8
9	Potassium	mg/L		182	181	180	153	188	153	23	2 650	9
10	Bicarbonate	mg/L		1236	795	ND	46	ND	10	20	83	10
11	Carbonate	mg/L		ND	ND	26	0	43	ND	0	0	11
12	Sulfate	mg/L		1560	1850	1850	1970	1970	1970	181	22 200	12
13	Chloride	mg/L		3345	3330	3340	3340	3350	3390	653	36 400	13
14	T-alkalinity as $CaCO_3$[10]	mg/L		1038	652	50	38	70	8			14
15	P-alkalinity as $CaCO_3$[11]	mg/L		40	ND	30	0	34	ND			15
16	Free carbon dioxide	mg/L		750	18	ND	0	ND				16
17	E.F. [TDS (calculated)/cond.][12]			6.2	6.2	6.2	6.8	6.0		5.2	11.0	17
18	E.F. [TDS (evaporated)/cond.]			6.4	6.4	6.3	6.6	6.6		5.2	11.0	18
19	Anions	mg/L		153.9	145.4	133.3	142.2	134.5	130.2	[illegible]		19
20	Cations	mg/L		153.6	143.1	131.0	143.0	132.7	137.2	[illegible]		20
	Trace elements		RCRA[13]									
21	Arsenic	mg/L		1.0								21
22	Barium	mg/L	100.0	0.09								22
23	Boron	mg/L		5.0	4.3		4.1	4.2		3.5	2.9	23
24	Cadmium	mg/L	1.0	0.004								24

25	Chromium	mg/L	5.0	0.002								25
26	Iron	mg/L		<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	0.9	26
27	Lead	mg/L	5.0	0.012								27
28	Mercury	mg/L	0.2	0.0038							0.0055	28
29	Manganese	mg/L		0.013								29
30	Selenium	mg/L	1.0	<0.02								30
31	Silver	mg/L	5.0	<0.0005								31
32	Aluminum	mg/L		<0.1								32
33	Copper	mg/L		0.001								33
34	Nickel	mg/L		0.01								34
35	Zinc	mg/L		<0.1								35
36	Strontium	mg/L		10.2	6.4	4.7	0.4	1.0	<0.3	<0.1	4.0	36
37	Sulfide	mg/L		4.9	0.7	ND		ND		ND		37
	Trace anions											
38	Orthophosphate	mg/L		2.6								38
39	Nitrate	mg/L		1.8								39
40	Fluoride	mg/L		2.6			0.6					40
41	Total organic carbon	mg/L										41
	Radioactivity											
42	Radium 226	pCi/L	[14]5.0	33.3				1.5		0.0	2.5	42
43	Radium 228	pCi/L		0								43
	Desalting — % water recovery											
	Based on $CaSO_4$ solubility at											
44	0 °C	%		4.3	21.9	38.0	91.6	86.8	96.7	99.0	11.8	44
45	20 °C	%		23.0	37.2	49.9	92.0	88.2	96.6	99.1	15.4	45
46	50 °C	%		28.2	41.6	53.5	92.4	88.9	96.8	99.1	19.7	46
47	90 °C	%		16.1	31.7	45.6	92.2	87.9	97.0	99.1	15.2	47

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[1] (R) — Raw water
[2] (A) — Aerated water
[3] (A-L) — Aerated-lime treated water
[4] (A-L-IX) — Aerated-lime-ion exchange treated water
[5] (A-L-SA) — Aerated-lime-soda ash treated water
[6] (A-L-SA-IX) — Aerated-lime-soda ash-ion exchange treated water
[7] ED-P — Electrodialysis-product
[8] ED-B — Electrodialysis-brine
[9] TDS — Total dissolved solids
[10] T — Total alkalinity
[11] P — Phenolphthalein end point
[12] E.F. — Effluent filtrate
[13] RCRA — Resources Conservation Recovery Act
[14] 5 pCi/g (pico Curies per gram)

21A

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COMPARISON OF PRETREATMENT PROCESSES

The high salt content and moderate flow rates of the LaVerkin Springs significantly contribute to the salinity of the Colorado River. Consequently, methods of preventing the salts in the water from entering the Virgin River, which flows into the Colorado River, were evaluated [2]. Reverse osmosis and electrodialysis are feasible membrane methods of desalting for preventing saline spring water from contaminating the river. By membrane desalting, the TDS content could be reduced from 9500 to 1000 mg/L or less in 90 percent of the spring water. The 90 percent of low salt water then could be allowed to flow into the Virgin River. The 10 percent of high salt water (86 000 mg/L brine) could then be stored in lined ponds or disposed of by deep well injection. Other feasible methods of preventing salt in the water from reaching the Virgin River include use of the water as a powdered coal pipeline transport media, use of the water as a secondary coolant in binary cooling towers, or use of the water in solar salt gradient ponds. Depending on which method is chosen, raw water will require degrees of pretreatment at the LVS site to prevent scaling in intake and pipeline systems by postprecipitation of calcium carbonate and/or calcium sulfate.

The pretreatment required for each method of salt water disposal is as follows:

Membrane desalting	A (aeration, L (partial lime treatment), SA (soda ash treatment), FC (flocculation and clarification) and F (filtration) or A, LFC, and F followed by IX (ion exchange)
Deep well injection	A, FC, and F
Coal slurry transport media	A, FC, and F
Secondary coolant	A, L, FC, and F
Solar salt gradient pond	A, L, FC, and F

To meet the requirements for deep well injection or for coal slurry transport, aeration followed by flocculation, clarification, and dual media filtration is proposed, and lime and soda ash treatment would not be required. To meet requirements for secondary coolant as solar salt gradient pond water, the raw water is airblown to remove carbon dioxide, and then subsequently lime treated to a pH of 9.5, flocculated, clarified, and filtered (A, L, FC, and F). The A, L, SA, FC and F process has the same unit processes, except soda ash addition to precipitate the balance of the calcium follows the lime addition. Likewise, the A, L, FC, IX process goes through the same unit operations except after filtration the water passes through ion exchange to remove the balance of the calcium.

A. Effectiveness of Pretreatment Processes

An indepth discussion of the "characteristics of raw and processed waters" is given in an earlier section of this report. The effects of various pretreatment processes on the dissolved constituents in the water are summarized in table 4.

1. *Aeration process.* — Aeration of subterranean LVS water is very effective in reducing dissolved undesirable chemicals. Air sparging reduces CO_2 from 750 mg/L to 18 mg/L (table 4). As a result, 220 mg/L of calcium precipitates as calcium carbonate along with 10 mg/L of silica and 6 mg/L of magnesium. The reduction in scaling tendency is evident from item 10. Calcium sulfate solubilities were derived from a Marshall Program [7] desk computer tape. The calcium sulfate solubility is reported for raw water as 23.0 percent and for aerated water as 37.2 percent. This means that 23.0 percent of the raw water and 37.2 percent of the aerated water could be removed in a desalting process before precipitating calcium sulfate; i.e., the aeration process reduces the tendency of the water to form scale on equipment, but would result in fouling of desalting membranes at water recoveries in excess of 37.2 percent. The Marshall program determines these percents based on relative weight concentrations of each species of dissolved anions and cations. Consequently, the decrease in scaling tendency is reflected to a certain extent by the reduction of TDS from 9507 to 9054 mg/L.

Aeration is an economical process step. It precipitates 220 mg/L of calcium as calcium carbonate with low-cost air and saves the cost of the equivalent amount of lime not required for coprecipitation. Aeration is also effective in reducing the concentrations of the undesirable strontium and sulfide. Strontium, which is reduced from 10.2 to 6.4 mg/L, is undesirable because of its radioactivity. Raw LVS water has a very unpleasant "rotten egg" odor due to the presence of 4.9 mg/L of sulfide. Aeration eliminates the odor by reducing sulfide to a residual of 0.7 mg/L.

2. *Aeration - partial lime treatment process.* —After aeration, the remaining 180 mg/L of calcium associated with bicarbonate ion, 16 mg/L of magnesium, and 15 mg/L of silica can be coprecipitated by addition of lime to bring the solution pH to 9.5. The tendency to scale equipment surfaces would be further reduced as indicated by the 49.9 percent change in calcium sulfate saturation (item 10, table 4). Aeration-partial lime treatment would not be adequate to permit membrane desalting to

greater than 50 percent water recovery. Essentially, neutralization of calcium bicarbonate with calcium hydroxide (lime) has a stoichiometric effect on reducing TDS from 9054 to 8530 mg/L. Overall, the A-L process removes 977 mg/L of TDS from the raw water. Partial lime treatment further reduces concentrations of undesirable chemicals. Residual sulfide is eliminated and overall for the A-L process strontium in the raw water is reduced from 10.2 to 4.7 mg/L.

3. *"Permanent" calcium hardness removal processes.* — After removal of calcium bicarbonate from LVS water by A-L treatment, 420 mg/L of permanent hardness calcium remains in solution as chloride and sulfate salts. Before further processing by membrane desalting (reverse osmosis or electrodialysis), most of the salt-solubilized calcium must be removed. This can be accomplished by further processing A-L treated water through a sodium-calcium ion exchange resin or by calcium-magnesium precipitation with soda ash (sodium carbonate).

a. Aeration-lime-ion exchange process. — Processing A-L treated water through IX results in a residual calcium of 30 mg/L and magnesium of 58 mg/L (table 4). This degree of water softening would be required, if reverse osmosis or electrodialysis desalting is done at the LVS site. Desalting could be carried out to 92.0 percent (item 10) water recovery without calcium sulfate precipitation occurring.

b. Following A-L treatment with SA treatment is not quite as effective as IX softening. Soda ash resulted in a residual calcium of 60 mg/L and magnesium of 93 mg/L. Membrane desalting would be limited to an 88.2 percent water recovery. Soda ash was effective in removing all but 2 mg/L of silica, while IX did not effect the silica concentration.

c. Following the A-L-SA process with an IX polishing step reduced calcium to 6 mg/L and magnesium to 46 mg/L. Based on item 10 of table 4, IX polished water could be desalted at 96.6 percent water before any calcium sulfate fouling occurred.

B. Cost of Pretreatment Processes

The LaVerkin Springs site operating contractor included detailed pretreatment process cost estimates in their test site data acquisition reports [3]. The Feasibility Report [8] was used to determine the plant capacities. These pretreatment process cost estimates are enclosed as appendixes to this report.

Appendix C	Lime-soda ash pretreatment plant cost
Appendix D	Lime-ion exchange pretreatment plant cost
Appendix E	Partial lime pretreatment plant cost
Appendix F	Aeration-pretreatment plant cost

The cost factors developed in appendixes C, D, E, and F are compared for the A, A-L, A-L-IX, and A-L-SA processes in table 5. These factors were based on mid-1981 cost data for processing of LVS raw water at a rate of 0.34 m^3/s.

1. *Installed equipment cost and construction cost.* — Item 1 of table 5 compares the construction cost for the four processes. The derivation of plant construction costs are given in table 6. All four processes have the same cost for aeration towers, a ferric sulfate feed system, a chlorine feed system, and associated building. The aeration plant has the lowest cost because it does not require lime, soda ash, or sulfuric acid feed systems, a solids contact reactor, or gravity filter plant. Since the purpose of the aeration would be to produce nonscaling water to permit pipeline transport away from the LVS site, aeration tower effluent could be sufficiently clarified with a 140-foot-diameter clarifier. Solids contact reactors and gravity filters are not needed. Also, the aeration process will result in less sludge than the other three processes and filtration and sludge disposal cost would be only $119,000. Consequently, the total construction cost for the "A" plant is $1,468,000.

In addition to the aforementioned equivalent equipment cost, the A-L, A-L-IX, and A-L-SA process plants would have equivalent cost for a lime feed system, a sulfuric acid feed system, solid contact reactors, and a gravity filters package plant (table 6). The A-L-IX plant cost of $4,045,000 includes the A-L plant cost of $3,127,000 plus an additional cost of $856,000 for the ion exchange process. The A-L-SA plant cost of $3,488,000 includes the A-L plant cost of $3,127,000 plus an additional cost of $185,000 for soda ash addition, plus an additional cost of $176,000 ($356,000 less $180,000) for the increase in size of the filtration and sludge disposal system needed to handle the larger quantity of sludge generated by the soda ash.

2. *Capital cost.* — Total capital costs (item 2, table 5) for the A, A-L, A-L-IX, and A-L-SA process plants are $1,945,000, $4,143,000, $5,360,000, and $4,620,000, respectively. Total capital costs were derived from construction cost (item 1, table 5) by using a 32.5 percent contingency factor which included engineering, legal, fiscal, administrative, and other cost.

Annual capital costs (item 3, table 5) are, respectively, $214,000, $456,000, $590,000, and $509,000. Annual capital cost are based on 25-year life, 10 percent interest, and 0.1101 capital recovery factor.

3. *The O&M (operation and maintenance) cost.* — The O&M cost for A, A-L, A-L-IX, and A-L-SA plants are $185,000, $621,000, $675,000, and $2,111,000 (item 4, table 4), respectively. The breakdown of factors contributing to O&M costs are summarized in table 7.

The process of precipitation and isolation of calcium carbonate by air sparging, clarification, filtration, and sludge disposal would be totally automated and require a minimum of personnel to operate and maintain the plant. Personnel would include four shift process operators, a mechanic, an electrician, a clerk, a laborer, a process engineer, and a site superintendent. For the A plant maintenance, labor and energy annual costs are, respectively, $30,000, $101,000, and $18,000. The chemical cost for chlorine disinfectant and for ferric sulfate floculation aid is $35,000 per year.

Lime handling is labor and maintenance intensive. Consequently, A-L, A-L-IX, and A-L-SA processes would require four additional shift process operators and an additional laborer. Because chemistry of the process becomes more complicated, a chemist would be needed at the plantsite. Lime addition increases maintenance materials by $17,000 ($47,000 to $30,000), chemicals by $299,000 ($344,000 to $35,000), labor by $95,000 ($101,000 to $196,000) and energy by $36,000 ($54,000 to $18,000).

A completely automated self-contained add on ion exchange plant is proposed for the A-L-IX plant. Water would only be processed through IX as a pretreatment process for further desalting by either reverse osmosis or electrodialysis. Consequently, the ion exchange resin would be regenerated with waste salt brine from the desalting process and the chemical cost for the A-L and A-L-IX processes would be the same. Although completely automated, the IX plant would require the site staff to be increased by two process operators and one mechanic. The increase in O&M cost for the IX process over the A-L process would be $4,000 ($51,000 less $47,000) for maintenance, $25,000 ($221,000 less $196,000) for labor, and $14,000 ($68,000 less $54,000) for energy.

The handling and use of soda ash in the A-L-SA process would require two additional process operators and an additional laborer compared to the A-L process. Annual labor cost would be $44,000 ($240,000 less $196,000) higher. Energy would be $9,000 ($63,000 less $54,000) higher. The cost of soda ash would add $1,424,000 ($1,758,000 less $334,000) to the annual chemical cost.

4. *Annual cost.* — The total annual cost of capital and O&M for the A, A-L, A-L-IX, and A-L-SA processes are $399,000, $1,087,000, $1,265,000, and $2,619,000, respectively.

5. *Unit water costs.* — Cost per units of product water are compared as item 6a, b, and c of table 5 for the four pretreatment processes.

Table 4. — *Effect of processing[1] on dissolved constituents*

Constituent	Units	R	A	A-L	A-L-IX	A-L-SA	A-L-SA-IX
TDS (total dissolved solids)	mg/L	9507	9054	8530	8710	9062	9407
Bicarbonate	mg/L	1266	795	nil	46	nil	10
Calcium	mg/L	820	600	420	30	60	6
Carbon dioxide	mg/L	750	18	nil	nil	nil	nil
Magnesium	mg/L	150	144	128	58	93	46
Silica	mg/L	40	30	15	15	2	2
Strontium	mg/L	10.2	6.4	4.7	0.4	1.0	<0.3
Sulfide	mg/L	4.9	0.7	nil	nil	nil	nil
pH[2]	mg/L	6.0	7.4	9.5	7.5	10.2	7.6
Calcium sulfate saturation at 25° [7]	percent[3]	23.0	37.2	49.9	92.0	88.2	96.6

[1] Process — includes FC and F unit processes.

[2] pH — product water.

[3] Percent — quantity of water that can beremoved before reaching calcium sulfate saturation.

Table 5. — *Cost of pretreatment of LVS water by various methods[1]*

Item	Cost factor	Units (app. F)	A (app. E)	A-L (app. D)	A-L-IX (app. C)	A-L-SA
1	Construction cost	\$ x 10^6	1.468	3.127	4.045	3.49
2	Total capital cost[2]	\$ x 10^6	1.945	4.143	5.36	4.62
3	Annual capital cost[3]	\$ x 10^6	0.214	0.456	0.590	0.509
4	O&M cost	\$/yr x 10^6	0.185	0.621	0.675	2.111
5	Total annual cost[4]	\$/yr x 10^6	0.399	1.087	1.265	2.619
6	Unit costs:					
	a. \$/1,000 gal	\$	0.14	0.38	0.44	0.92
	b. \$/acre-ft	\$	42	125	146	301
	c. \$/m^3	\$	0.037	0.101	0.118	0.244

[1] Mid-1981 cost estimates for plants with flowrate of 0.34 m^3/s. All processes include clarification and dual media filtration.

[2] 32.5 percent contingency.

[3] 25-year life, 10% interest rate, 0.1101 capital recovery factor.

[4] Items 3 plus 4

Table 6. — *Installed equipment cost and total construction cost*

	Cost (\$ x 1000)			
	A (app. F)	A-L (app. E)	A-L-IX (app. D)	A-L-SA (app. C)
Aeration towers	675	675	675	675
Lime feed system	—	123	123	123
Soda ash feed system	—	—	—	185
Ferric sulfate feed system	28	28	28	28
Chlorine feed system	33	33	33	33
Sulfuric acid feed system	—	25	25	25
Upflow solids contact reactor	525	595	595	595
Gravity filters package plant	—	1380	1380	1380
Ion exchange processing plant	—	—	856	—
Filtration and sludge disposal	119	180	180	356
Administrative, laboratory, and maintenance facility	86	88	88	88
Total construction cost	1468	3127	4045	3488

Table 7. — *Operation and maintenance cost*

	Cost (\$ x 1000)			
	A (app. F)	A-L (app. E)	A-L-IX (app. D)	A-L-SA (app. C)
Maintenance materials	30	47	51	47
Chemicals	35	334	334	1758
Labor	101	196	221	240
Energy	18	54	68	63
Annual O&M cost	185	631	675	2111

EXPERIMENTAL AERATION PRECIPITATION PROCESS

The high CO_2 (carbon dioxide) content in LVS raw water could create a $CaCO_3$ (calcium carbonate) post-precipitation problem in a saline water collector system or pretreatment process intake system. The LVS aquifer water contains about 750 mg/L of dissolved CO_2. If the CO_2 is allowed to come out of solution, about 25 percent of the calcium present in the raw water will precipitate as $CaCO_3$. The mechanism of precipitation is for nucleation to occur on confining equipment wetted surfaces. A hard, tenacious scale builds up on all internal wetted surfaces. A process for removal of CO_2 by air sparging to prevent scale formation on the internal wetted surfaces of intake and process equipment was developed and designed at the test site.

A. Equipment Description

In the LVS site test, aeration was evaluated as a front-end unit process step in the A-L (aeration-lime) treatment process described in the next section. However, it will be treated in this discussion as the aeration floculation, clarification, filtration process shown in figure 8. Clarification and dual media filtration equipment are illustrated in the equipment description of the A-L process section of this report. The aeration-softening pretreatment equipment consists of:

- Aeration system
- Pretreatment system
  - Rapid mix tank
  - Flocculation tank
  - Clarifier
- Filters

Figure 9 shows relative elevation dimensions of the aeration, rapid mix, and flowlator tanks. Also, air distribution to the diffuser is shown. A photograph (fig. 10) of the installed aeration tank shows its physical relationship with the rapid mix tank. Figure 11 demonstrates design of the onsite fabricated air diffuser pipe network and figure 12 shows its construction.

The final design of the aeration unit developed during the LVS site test program is demonstrated in figures 9, 10, 11, and 12. An FRP (fiberglass reinforced polyester) tank 1.52 meters in diameter with a liquid volume of 4.8 m³ was used as the aeration vessel. Raw water is pumped directly from the well down the inner side wall of the aeration tank to 150 mm above the air diffuser. Aerated water flows by gravity from the surface level into the rapid mix tank, flocculator, clarifier, and dual-media filters, respectively. This equipment is described in the next section. Compressed air is sparged by two airblowers into the left and right sides (fig. 9) of the PVC pipe-fabricated air diffuser (fig. 11). The surface area of the diffuser is 1.47 m² or covers 90 percent of the 1.84 m² of aeration tank surface area.

B. Process Operations

According to the aeration softening system flow diagram (fig. 8), water is transferred from a 30-m depth by a submerged well pump into the aeration tank. Air mixes intimately with the water, driving CO_2 to dissolution and precipitating 550 mg/L of $CaCO_3$. Water overflows from the aeration tank into and through the rapid mix tank where flocculation of the calcium carbonate is promoted by the addition of 7.5 mg/L of ferric sulfate. The flow through the rapid mix tank, flocculator, clarifier, and dual-media filter is described in the next section.

The aeration tank is shown schematically in figure 13. At optimum conditions, water transfers from a spring well under pump pressure to the open aeration tank at a flow rate of 0.63×10^{-3} m³/s (10 gal/min). About 10 percent of the clarifier blowdown sludge recycles into the aeration tank to selectively seed the precipitation of calcium carbonate. Air flows from two blowers at 47.2×10^{-3} m³/s. Air release causes the turbulent, boiling action illustrated by the photograph (fig. 14) of the water surface in the aeration tank. Intimate mixing of air and water effectively sparges carbon dioxide and reduces the concentration in solution typically to 45 mg/L. Within the bulk of the water in the aeration tank, calcium carbonate precipitates by nucleation of recycled sludge particles from the clarifier. Nucleation on the sludge particles prevents buildup of a hard adhesive coating on the internal equipment wetted surfaces. Turbulence keeps calcium carbonate suspended and evenly dispersed in the bulk solution. The suspension overflows continually into the flocculator tanks where the calcium carbonate precipitate coagulates. A view of the clarifier, flocculator, and aeration tank is presented in figure 15.

C. Results and Discussion

A summary of the aeration systems operation at the LVS site for the period October 8, 1979, through August 29, 1980, is given in table 8.

1. *Development and design of the aeration process.* — Characterization studies of LVS were conducted between February 11 and November 6, 1972. The study included taking water samples about once each week from 14 identified surface springs and submitting them to an offsite laboratory for chemical analyses. The 1972 water samples were analyzed for pH, conductivity, TDS, major cations, and major anions. The analyses were statistically evaluated by Eisenhauer [5]. The earlier analyses did not test for dissolved CO_2. During site preparations for

the 1979-1980 site test program, the CO_2 content and scaling tendency of the springwater aquifer were discovered. Consequently, aeration process development and design were not preplanned prior to arrival at the test site on October 7, 1979. During startup operations, process development and design task for the aeration process were incorporated into the schedules according to figures 6 and 7 and table 2.

a. Original design operations. — The preliminary design of the mobile L-SA-PP (lime-soda ash treatment-pilot plant) is shown in appendix G. The SCR (solids contact reactor) and associated piping and controls shown were not part of the equipment operated at the LVS site. In early planning, an SCR was to operate in parallel to an in-line pretreatment system. Because of limited funds, the SCR was dropped from the program and all process studies were done with the in-line pretreatment system. The preliminary design did not initially include any means for preaeration of influent water to the L-SA-PP, but did include a raw water feed tank. Its purpose was to provide for 40 minutes of surge capacity between influent raw water flow and feed water flow to the lime or lime-soda ash reaction zone.

An air diffuser assembly was constructed from 1-inch, schedule No. 40, PVC, threaded pipe. Air holes were drilled 3/32-inch diameter, at 90° and 1-inch spacing in the pipe over the length of the cross shaped diffuser. Aeration was provided by an air blower with a capacity of 50 ft^3/m at 10 lb/in^2 and driven by a 1 hp, 3600 r/min motor.

Task LT-1, scheduled to occur between October 7 and November 5, 1979, covered the hydraulic testing and shakedown of the lime train and included the front-end aeration system. The purpose of this initial testing was to demonstrate the capability of the aeration system for reducing carbon dioxide levels to 20 g/m^3. The initial operation, conducted between October 8 and October 10, which used the 1.3-m^3-capacity aeration tank, was partially successful. The CO_2 concentration reduced from an average of approximately 605 g/m^3 in the raw water to 160 g/m^3 in the aeration tank effluent. Water flow to the aeration tank was regulated by float control meter item No. 23 at a rate of 0.00258 m^3/s or for a residence time of 0.14 hour. Earlier bench scale test had demonstrated with good air sparging dissolved CO_2 could be reduced to a level of 18 mg/L in 0.17 hour.

A 1-inch PVC pipe network diffuser similar in construction to that shown in figures 11 and 12, except with an overall diameter of 1.40 m, was installed in the raw water feed tank and the aeration process was operated at the conditions shown in table 8 between October 10 and 18. A 91.6 percent reduction in dissolved CO_2 was achieved for a residence time of 0.38 hour and a water feed rate of 0.95 m^3/s.

Operation with a new diffuser in the original design mode, although moderately successful in reducing CO_2 content to a residual level of 47 mg/L resulted in serious scale buildup on process equipment wetted surfaces. During the 10 days of operations, a hard tenacious coating of calcium carbonate about 3- to 6-mm thick formed on the sparger pipe network, the process wetted surfaces of the aeration tank, and associated pumps and pipewalls.

b. Intermediate design operation. — Operation of the aeration process in the original design mode did not consistently reduce the CO_2 content of the water to an acceptable concentration level and resulted in excessive calcium carbonate scale buildup on the process wetted surfaces of the sparger, aeration tank, transfer pump, and transfer pipe leading to the rapid mix tank. This led to the following changes which ultimately became the final operating design of the aeration system:

Standardize the feed water flow to the aeration tank at 0.63 m^3/s.

Increase the liquid column volume from 1.3 m^3 to 3.5 m^3, or residence to time from 0.38 to 1.54 hours.

Increase the air blowing capacity from 0.0236 m^3/s to 0.0472 m^3/s.

Increase the diameter of the air diffuser pipe network from 1.40 to 1.53 m.

Eliminate the hydraulic transfer system between the aeration tank and rapid mixing tank by relocating the aeration tank adjacent and at a higher overflow level than the rapid mix tank to permit gravity flow directly through an open channel, and

Provide for nucleation by recycling calcium carbonate particles to the aeration tank from the clarifer.

Converting the aeration process to final design mode required purchasing a 3.5-m^3 aeration tank, a second 23.6 m^3/s air blower, and additional pipes, valves, and controls. Also, fabrication of a diffuser pipe network and extensive reconstruction was required. To keep the development of the aeration process on schedule, a variation of the intermediate design was used until the needed equipment arrived at the site.

Since ion exchange system was not scheduled to start until November 26, tank T-6 from the IX process was temporarily substituted in the place of original aeration tank. The PVC pipe network diffuser was transferred from the smaller tank (1.3 m^3) to T-6 (3.5 m^3). The hydraulic transfer system 1-inch pipe was replaced by 2-inch pipe. It was hoped that the larger pipe would reduce the frequency of unscheduled shutdowns required for cleaning scale from pipe inner walls.

The conditions for operation of the aeration system between October 18 and 26 at intermediate design mode are summarized in table 8. Essentially, the larger tank permitted evaluation of the effectiveness of CO_2 purging for longer residence time. A high of 96.9 percent CO_2 reduction was demonstrated at a water feed rate of 0.00019 m^3/s and a residence time of 5.12 hours. Since the final design mode required operation of the aeration process at a water feed rate of 0.00063 m^3/s, the result shown for October 22 or 94.1 percent at 0.00069 m^3/s for 1.41 hours is more relevent. Twice during this mode, excessive scale buildup occurred internally in the hydraulic transfer system, and the system was shut down to disassemble and remove the scale from the inner surfaces of the pump and pipe with hydrochloric acid.

On October 25, tank T-6 was placed back in the IX system and aeration was continued with the original 1.3 m^3 tank in combination with other tanks as noted in table 8 for the period October 26, 1979, through March 3, 1980. Process development of the aeration process was held in abeyance for this transition period and resumed on March 4 after arrival and installation of final design aeration equipment.

c. Final design operation. — With one exception, the final equipment was installed for operation between March 4, 1980, and April 15, 1980. Because major onsite refabrication was required, no provision was made at that time to recycle calcium carbonate particles from the clarifier to the aeration tank. This aspect of the aeration process was later evaluated as separate task SS-I, selective scaling test. Aeration operation with final design equipment was mostly at a feed water rate of 0.00063 m^3/s to conform with water flow rates required for task LT and LT-SA. Between April 21 and June 11, a 98.6 percent CO_2 reduction (10 cm^3 CO_2 residual) was demonstrated for a residence time of 1.54 hours and an air flow rate of 0.0472 m^3/s.

The final configuration solved the scale coating problems between the aeration tank and rapid mix and flocculation tanks. This was accomplished by a direct gravity flow from water surface level into the rapid mix tank and into the flocculation tank. However, heavy calcium carbonate scale formed in the aeration tank and particularly on the surface of the air diffuser pipe network. The air sparger generally operated continuously for 7 days before excessive scaling reduced the air flow rate below desired 0.0472 m^3/s level. When scale formation became excessive, the diffuser was removed and replaced by a standby scale-free one. Changeout of the diffuser took about 1 hour. After removal, scale was dissolved from the diffuser by soaking in 5 percent sulfuric acid. The soaking tank was constructed of FRP and was 0.6-m deep and 1.8-m diameter.

2. *Task SS-I, selective scaling test.* — Task SS-I was performed the week of June 8, 1980 to determine the effect of recycling precipitated calcium carbonate from the clarifier into the front-end aeration tank (figs. 9 and 10). A scale-free air diffuser pipe network (figs. 11 and 12) was installed in the aeration tank on June 11, 1980. The operation was continued at the conditions shown in table 8 between June 11 and 17. During this operation, 2- to 4-percent by weight calcium carbonate sludge was recycled from the sludge discharge of the clarifier at a flow rate of 0.06×10^{-3} m^3/s into the bulk volume of the aeration tank. On June 17, 1980, the diffuser was removed and examined. The surface was free of calcium carbonate scale. Typically, the diffuser, after 6 days of operation, would have a hard scale coating about 3- to 4-mm thick over the surface. Instead of forming a hard tenacious coating on the diffuser surface, the calcium carbonate precipitated in the air and water mixing zone on the seed particles. Recycling sludge from the clarifier to the aeration tank needs to be incorporated into any future plant design.

D. Optimum Aeration Process Conditions

The aeration process optimum steady-state conditions for removal of carbon dioxide and precipitation of calcium carbonate were determined.

1. *Aeration tanks.* — Optimum conditions were as follows:

• Influent water	
Feed rate	0.63×10^{-3} m^3/s
pH	6.0
Carbon dioxide	750 mg/L
Calcium	820 mg/L
Bicarbonate	1266 mg/L
Magnesium	150 mg/L
TDS	9507 mg/L
• Blower air rate	0.0472 m^3/s
• Residence time	1.54 hours

- Sludge recycle	
Typical concentration	2.5 weight percent
Flow rate	0.06 x 10⁻⁶ m³/s

2. *Rapid mix tank.* — Optimum conditions were as follows:

- Ferric sulfate dosage	7.9 mg/L

3. *Flocculator.* — Optimum conditions were as follows:

- Residence time	1.00 hours
- Agitation intensity	
Volume/area	2.27 m³/1.77 m²
Agitator	0.25-m diameter, 1.5 pitch propeller
Energy	1/3 hp at 140 r/min
- Effluent pH	7.4

4. *Clarifier.* — Optimum conditions were as follows:

- Residence time	3.8 hours
- Agitation intensity	
Volume/area	8.72 m³/4.11 m²
Agitator	Rake, 2.21-m diameter
Energy	1/3 hp
- Sludge	
Control level	0.6 m
Recycle rate	0.00019 m³/s
Purge rate	1.4 x 10⁻⁵ m³/s
Concentration	2- to 4-weight percent
- Effluent	
Turbidity	60 JTU
pH	7.6
Carbon dioxide	16 mg/L
Calcium	600 mg/L
Bicarbonate	795 mg/L
Magnesium	144 mg/L
TDS	9054 mg/L

Table 8. — *Aeration system operation, summary of results LaVerkin Springs Test Site*

Dates of operation	Aeration tank cap (m³)	Raw water flow rate (m³/s x10⁻³)	Tank residence time (h)	Air flow rate (m³/s x10⁻³)	Air to liquid ratio	Raw water CO_2 Conc avg (g/m³)	Residual CO_2 Conc	Percent CO_2 reduction (%)	Operation mode
10/8-10/10	1.3	2.58	0.14	23.6	9.15	605	160	73.6	Original design
10/11	1.3	1.39	0.26	23.6	16.98	642	74	88.5	Original design
10/11	1.3	0.76	0.48	23.5	31.05	658	60	90.9	Original design
10/12-10/17	1.3	0.95	0.38	23.6	24.84	569	47	91.6	Original design
10/19	3.5	1.26	0.77	23.6	18.73	562	36	93.6	Intermediate design
10/22	3.5	0.69	1.41	23.6	34.20	580	34	94.1	Intermediate design
10/23	3.5	0.50	1.92	23.6	47.20	514	38	92.6	Intermediate design
10/23	3.5	0.38	2.55	23.6	62.11	488	18	96.3	Intermediate design
10/23	3.5	0.13	7.48	23.6	181.54	488	17	96.5	Intermediate design
10/24	3.5	0.19	5.12	23.6	124.21	423	13	96.9	Intermediate design
10/25	3.5	0.95	1.02	23.6	24.84	617	51	91.7	Intermediate design
10/26-11/9	1.3	0.63	0.57	23.6	37.46	642	84	86.9	Transition
11/9-1/17	4.8	0.63	2.12	23.6	37.46	708	24	96.6	Transition
1/17-3/4	1.3	0.63	0.57	23.6	37.46	764	74	90.3	Transition
3/4-4/15	3.5	0.63	1.54	23.6	37.46	810	67	91.7	Final design
4/15-4/21	3.5	0.63	1.54	47.2	74.92	777	40	94.9	Final design
4/21-6/11	3.5	0.63	1.54	47.2	74.92	712	10	98.6	Final design
6/11-6/17	3.5	0.63	1.54	47.2	74.92	722	24	96.8	Final design
6/17-7/30	3.5	0.63	1.54	47.2	74.92	731	37	94.9	Final design
7/31-8/2	3.5	1.26	0.77	47.2	37.46	716	66	90.8	Final design
8/2-8/6	3.5	0.63	1.54	47.2	74.92	735	69	90.6	Final design
8/6-8/22	3.5	0.95	1.02	47.2	49.68	702	46	93.4	Final design
8/27-8/29	3.5	0.63	1.54	47.2	74.92	691	37	94.6	Final design

The small aeration tank was utilized during this phase of testing, with one blower.

Tank T-6 (ion exchange spent regenerant tank) was used to provide a larger capacity (i.e., greater detention time).

The small tank was placed back in operation to allow tank T-6 to be installed in the ion exchange system.

Tank T-6 was again utilized in the aeration system, this time in series with the smaller tank. Both tanks were aerated using the one blower.

The system was out of service from December 21, 1979 through January 2, 1980, due to a scheduled Christmas shutdown.

Tank T-6 was returned to the ion exchange system since that testing had begun.

A new large capacity tank was installed. This tank utilized a gravity overflow of the aerated effluent directly into the lime feeder.

A second blower was placed in service at this time. The two blower configuration was utilized throughout the remainder of the test program.

Task SS-1

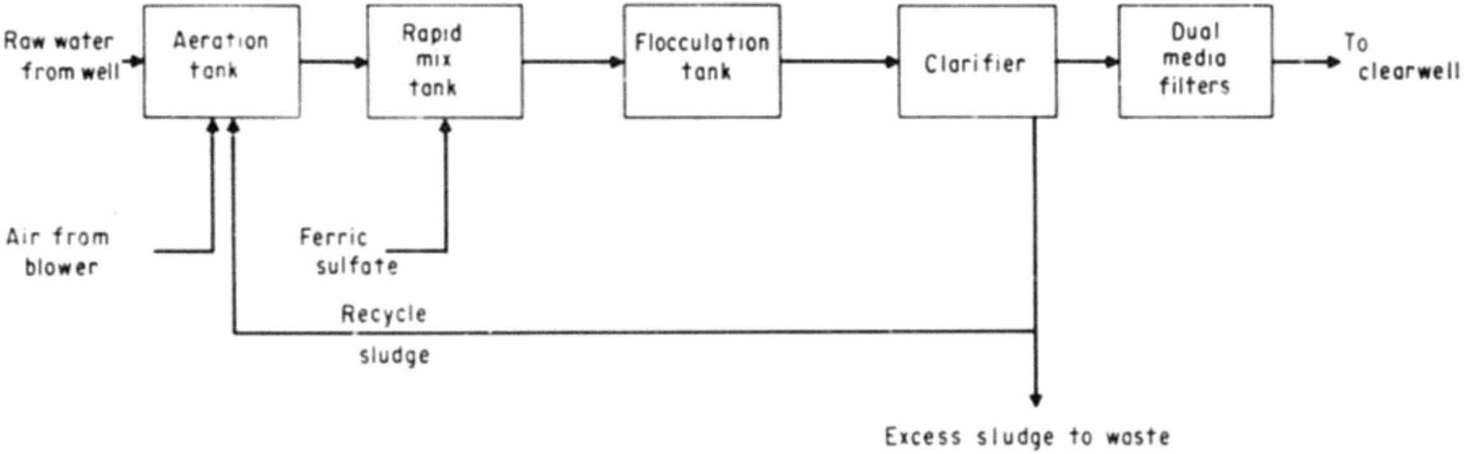


Figure 8.—Aeration softening system flow diagram.

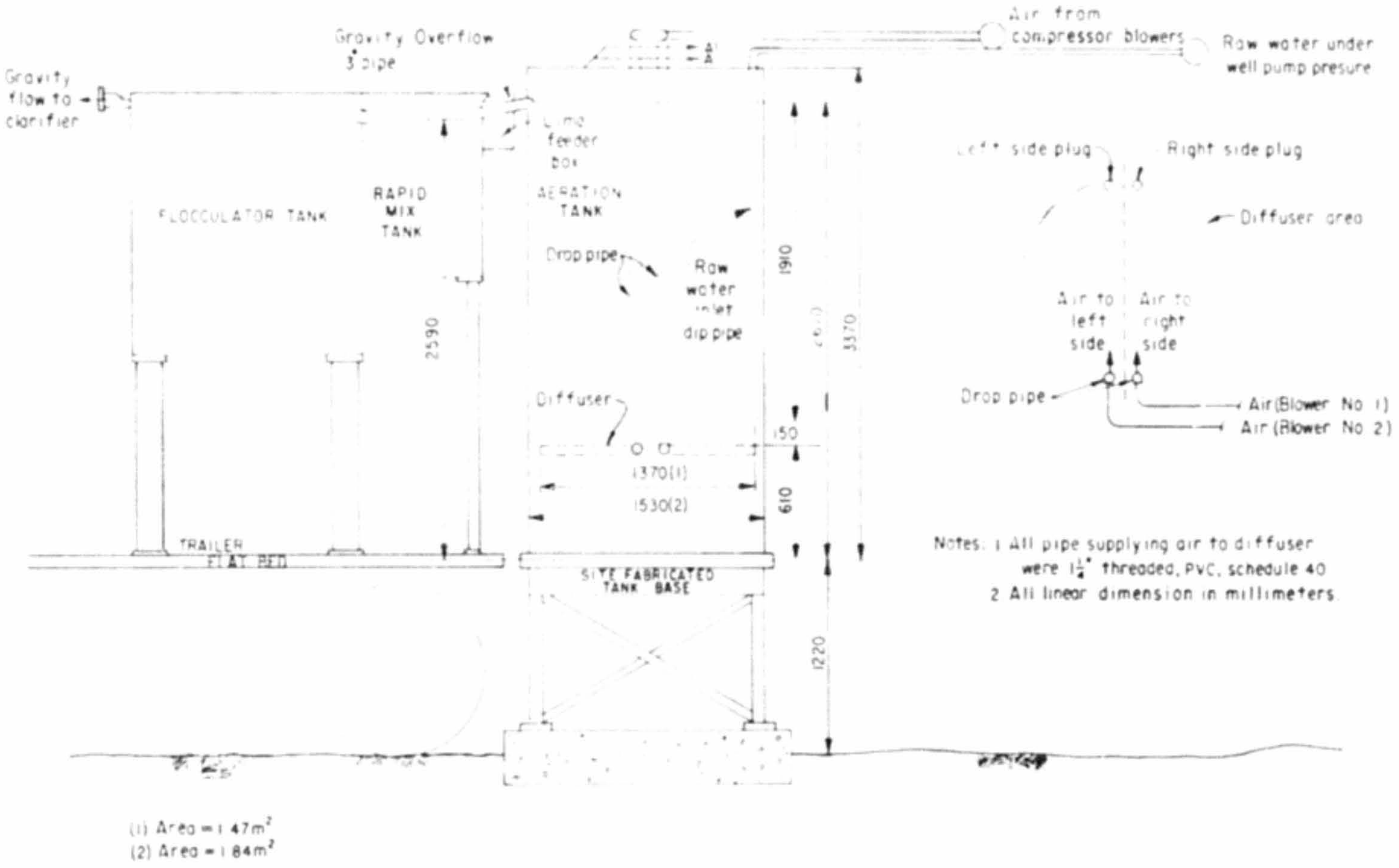
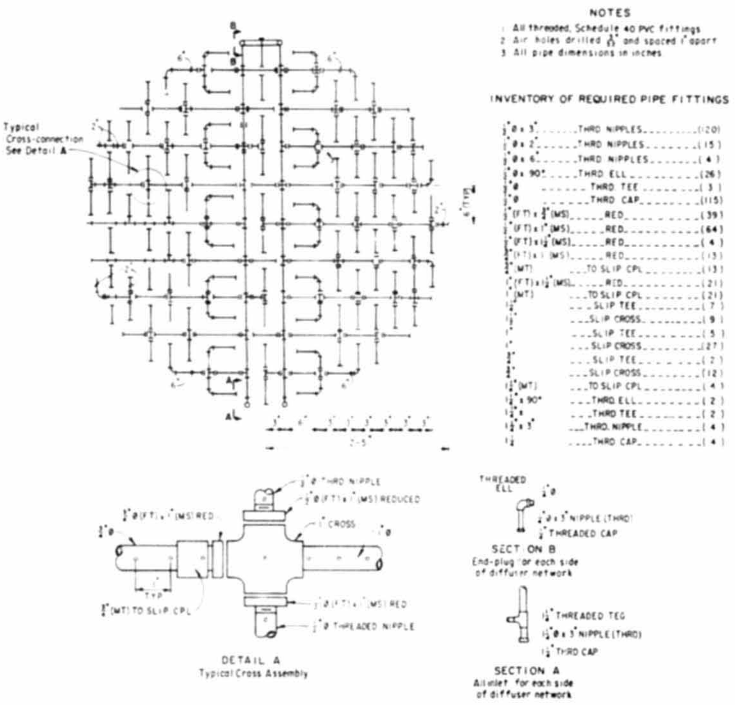


Figure 9.–Elevation of aeration equipment.

Figure 10.–Aeration tank. Photo P801-D-80758.



NOTES

1. All threaded, Schedule 40 PVC fittings
2. Air holes drilled 3/32" and spaced 1" apart
3. All pipe dimensions in inches

INVENTORY OF REQUIRED PIPE FITTINGS

½"Ø x 3"	THRD NIPPLES	(120)
½"Ø x 2"	THRD NIPPLES	(15)
½"Ø x 6"	THRD NIPPLES	(4)
½"Ø x 90°	THRD ELL	(26)
½"Ø	THRD TEE	(3)
½"Ø	THRD CAP	(115)
½"(FT) x ¾"(MS)	RED	(39)
½"(FT) x 1"(MS)	RED	(64)
½"(FT) x 1½"(MS)	RED	(4)
¾"(FT) x 1"(MS)	RED	(13)
¾"(MT)	TO SLIP CPL	(13)
1"(FT) x 1½"(MS)	RED	(21)
1"(MT)	TO SLIP CPL	(21)
1½"	SLIP TEE	(7)
1½"	SLIP CROSS	(9)
1"	SLIP TEE	(5)
1"	SLIP CROSS	(27)
¾"	SLIP TEE	(2)
¾"	SLIP CROSS	(12)
1½"(MT)	TO SLIP CPL	(4)
1½" x 90°	THRD ELL	(2)
1½" x	THRD TEE	(2)
1½" x 3"	THRD. NIPPLE	(4)
1½"	THRD CAP	(4)

DETAIL A
Typical Cross Assembly

SECTION B
End-plug for each side
of diffuser network

SECTION A
Air inlet for each side
of diffuser network

Figure 11.–Diffuser design.

Figure 12.–Diffuser pipe network. Photo P801-D-80759.

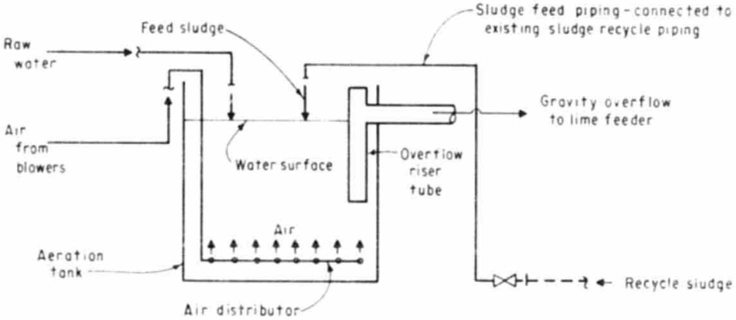


Figure 13.–Operation of the aeration tank to promote selective scaling.

Figure 14.–Aeration tank surface. Photo P801-D-80760.

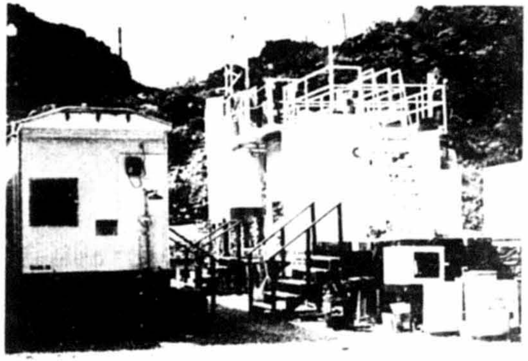


Figure 15.–View of clarifier, flocculator, and aeration tank. Photo P801-D-80761.

EXPERIMENTAL LIME SOFTENING PROCESS

Aeration of LVS raw water will reduce dissolved calcium from 820 mg/L to 600 mg/L and bicarbonate from 1266 mg/L to 795 mg/L. Subsequent additions of lime (calcium hydroxide) to increase water pH from 7.4 to 9.5 neutralizes the remaining bicarbonate and removes 180 mg/L of calcium. Consequently, 400 mg/L of calcium carbonate precipitates and 420 mg/L of calcium remains in solution. Equipment required for aeration processing was designed and constructed at the operating site. Preliminary design of the mobile LTPP (lime treatment pilot plant) was done by E&RC personnel. The operating contractor and Government engineers at the YDTF (Yuma Desalting Test Facility) finalized the design and constructed the LTPP. The process which removes bicarbonate by aeration and by subsequent lime treatments was developed and optimized at the LVS test site.

A. Equipment Description

Preliminary designs for the mobile LTPP in the lime treatment process studies are given in appendix G. Details of process, pipe, valves, instruments, and construction are include on the drawings. Two parallel lime pretreatment systems are shown with a common feed system and separate filter systems. The original scheme was to carry out process development optimization studies in a separate chemical treatment train with a separate rapid mix tank (18)*, flocculator tank (20), and clarifier (22), and at the same time, evaluate and demonstrate the design of a SCR (solids contact reactor) (12). The original scheme would have permitted concomitant or separate operations of the in-line train or the SCR. The money funded for the LVS site test program was not adequate for purchasing both the in-line and SCR equipment. Since the maximum process development design data would be obtained with the in-line chemical treatment train, a decision was made to fabricate the mobile chemical treatment pilot plant according to drawings G.21 through G.24 with the omission of the SCR. A free space was to be left for the SCR for future installation.

The equipment sequence for the lime softening pretreatment system is shown in figure 16 and figure G.6. Raw water enters the aeration tank. Processed water flows through the aeration tank, rapid-mix tank, flocculator, clarifier, and dual media filters. A description of the aeration equipment is omitted here since it was discussed earlier.

1. *Rapid-mix tank.* — The design basis for rapid-mix tank is given as equipment item (18) in preliminary design equipment location drawings (figures G.1 and G.2). Construction basis and details for the rapid-mix tank used at the LVS site are given in figure G.7. For rapid mixing of lime and soda ash with influent water, a box-type open tank was constructed of carbon steel with straight sides 686-mm long, 605-mm wide, and 1067-mm high. The rapid mix tank was welded on the upper corner of the flocculation tank as shown (fig. G.6). Internal surfaces were sandblasted and coated with a white enamel. A 1/3-hp, 1750 r/min propeller-type mixer provided agitation.

2. *Flocculator.* — The design for the flocculation tank is given as item (20) in figures G.1 and G.2. Construction details for the flocculation tank used at the LVS site are given in figure G.7. For flocculation of calcium carbonate precipitate, a box-type covered tank was constructed of carbon steel with straight sides 1371-mm long, 1473 mm wide, and 1524-mm high. Internal surfaces were sandblasted and coated with a white enamel. A 1/3-hp, 20-to 350-r/min, propeller-type mixer provided agitation.

3. *Clarifier.* — The design for the clarification tank is given as item (22) in figures G.1 and G.2. Construction details for the clarification tank used at the LVS site are given in figure G.8. The tank is constructed of carbon steel. The clarifier has a 2286-mm diameter by 2896-mm outer shell. A circular baffle 1067 mm in diameter and 1529-mm high is located in the center of the vessel. A rake-type paddle driven by a 1/3-hp motor and varidrive provides settling agitation.

4. *Dual-media filters.* — The design for dual-media filters is given as equipment items (13), (13A), (14), and (14A) in figures G.1 and G.2. Construction details for the filters are shown in figure G.9. The filters were fabricated by cutting and welding together ¼-inch carbon steel plate to give the construction details shown. All interior surfaces were sandblasted and coated with an epoxy-phenolic resin.

5. *Operations control center.* — The aeration lime treatment system was designed to be operated from a control center trailer. Figure G.5 shows plan, elevation, dimensions, and utilities for the control center trailer.

6. *Solids contact reactor.* — Although an SCR was not constructed for use at the LVS site, the vessel was designed. Figure G.3 is a shop drawing for fabrication of the SCR.

B. Process Operations

The operation of lime treatment can be followed by reference EP&ID (equipment, piping and instrumant

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*Equipment item number on figures G.1 and G.2.

diagram) process flow, figure G.4, EP&ID isometric as in figure G.6, process flow sheet figure 16, and equipment configuration figure 17. The lime treatment train initially consisted of a trailer-mounted "in-line" lime-softening pilot plant consisting of three major components: (1) a 0.37-m^3-capacity rapid-mix tank where aerated raw water, lime, ferric sulfate, and recycled sludge were mixed by a high-speed electric mixer, (2) a 2.3-m^3-capacity flocculation tank where coagulated solids were allowed to form with gentle mixing by a variable-speed electric motor-driven agitator, and (3) a 2.1-m-high by 2.3-m-diameter clarifier where solids settled by gravity and were raked to a sludge hopper located at the center of the clarifier. Settled sludge was continuously pumped from the center of the clarifier to the rapid-mix tank by a progressive-cavity sludge pump.

1. *Lime addition.* — Lime was added by a volumetric feed to the aerated raw water in the lime-solutioning tank. The lime-solutioning tank low flow rate used during the testing was 0.95 by 10^{-3} m^3/s maximum. The pH sensing automatically regulated lime feed rate. Flocculation tank pH was continuously monitored by a submersible pH probe equipped with an ultrasonic cleaner.

2. *Rapid-mix and flocculation.* — The rapid-mix tank opened to the flocculation tank through a 0.46-m-wide rectangular weir. A single-propeller-type, variable-speed flocculator was used to maintain particle suspension and to provide gentle mixing action to promote particle growth. The flocculation tank was connected to the feed well, located in the center of the clarifier, by a 0.1-m wide-influent trough.

3. *Sludge recycle and waste.* — Sludge was continuously removed from the clarifier by a progressive cavity sludge pump. The sludge was continuously recycled to the rapid-mix tank. Excess sludge was bled off through a sludge-wasting valve on the discharge side of the sludge pump. The amount of sludge wasted was determined by the solids level in the clarifier, which was observed by collecting samples from the clarifier sample taps located every 0.30 m from the bottom of the clarifier to the top. These sample taps were also used during the initial phases of operation for sludge wasting. This method later proved to be unsatisfactory and was discontinued in favor of wasting a portion of the recycle.

4. *Clarifier operation.* — Clear clarified water overflowed from the clarifier through 32 evenly spaced, submerged orifices along a peripheral effluent launder. Effluent was conveyed to the filters by gravity flow. Area of the clarification zone at a point 0.4 m below the effluent launder was approximately 3.2 m^2. Perimeter of the effluent launder was approximately 6.4 m. Total clarifier volume was approximately 8.8 m^3. Clarifier effluent was acidified prior to entering the filters. Sulfuric acid was injected immediately upstream of a 5.1-cm-diameter static mixer. Acid feed rate was automatically controlled to maintain a set point pH in the filter effluent.

Only one major change was made to the pretreatment equipment during the course of the operation. The clarifier feed well was originally designed and built to extend to within approximately 0.9 m of the bottom of the clarifier as shown in figure 18. The feed well was extended to within 0.38 m of the bottom of the clarifier on May 13, 1980. This modification was necessary to improve the filter effluent plugging factors by utilizing the filtering action created by forcing the clarifier flow to pass through the sludge blanket, as in an SCR.

5. *Filter operation.* — Figure 19 shows plan, elevation, dimension, volumes, and flow nozzle on the four filters located on the filter trailer. Filters 1 and 2 had a surface area of approximately 0.75 m^2 each. These filters were designed and constructed for use with an SCR which had originally been planned for use but was never purchased. The filters, although charged with media, were never used. Filters 3 and 4, which were constructed for use with the in-line system, each had a surface area of approximately 0.37 m^2. Filter media consisted of approximately 0.51 m of anthracite coal (effective size = 0.90-1.0 mm, uniformity coefficient = 1.5) and over 0.30 m of silica sand (effective size = 0.35-0.50 mm, uniformity coefficent = 1.6).

The filter underdrain was of a plenum chamber with plastic filter nozzles protruding slightly up into the silica sand media. A centrifugal pump was used for backwashing the filters. The backwash flow rate was varied according to the temperature of the backwash water. The rate varied from approximately 4.5 x 10^{-3} m^3/s at 10 °C to 5.9 x 10^{-3} m^3/s at 30 °C. The backwash cycle, including air-scour, backwash, and rinse, was manually controlled by a flow-control piping manifold (fig. 20).

C. Results and Discussion

Figure 17 shows the flow arrangement of the equipment used for lime-softening. As a first step in the process, raw water enters the aeration tank (1). Air from two blowers sparge into the water and removes carbon dioxide. Carbon dioxide solution concentration reduces from 750 mg/L to 18 mg/L and bicarbonate in solution drops from 820 mg/L to 600 mg/L. The bicarbonate reduction is accompanied by precipitation of calcium carbonate. To prevent scale crustation on wetted equipment surfaces, sludge is recycled from the clarifier as nucleation sites for calcium carbonate precipitate.

Aerated water flows through the lime feeder reservoir (2) and the rapid-mix tank (3) where lime is added and intimately mixed with the water. Lime addition is controlled by the pH electrode in the outflow end of the flocculator. Ferric sulfate addition occurs in the rapid-mix tank to promote flocculation. Sufficient resident time and optimum agitation completes precipitation in the flocculator (4).

A homogenous light dispersion of calcium carbonate and treated water continually overflows into the clarifier (5) center downflow baffle. As water rises slowly in the expanded annular area, calcium cabonate precipitate migrates to the bottom and is moved toward the center through by the sludge rake. A pump, which cycles off and on, removes the sludge from the clarifier. Effluent overflows the peripheral launder and flows by gravity to the dual media filters (6). Effluent is adjusted down in pH to prevent post precipitation of calcium carbonate on the filter media.

1. *Task LT-I — Startup of lime train.* — The first task was to start up the lime treatment system. For description of task, refer to the section titled "Test Schedule and Task Performed." Task LT-I began on October 7, 1979, and continued until November 15, 1981. The installed LTPP was hydraulically tested and the effectiveness of air sparging CO_2 (carbon dioxide) was evaluated. The system was started with approximately 40 L/min of raw water with direct water flow through the aeration tank, rapid mix tank, flocculator tank, and clarification tank.

Hydraulic testing operations continued with water as needed to correct leaks, condition equipment, and evaluate and adjust operations of pumps and flow control devices.

A preliminary aeration checkout was conducted to determine what modifications were needed to overcome the scaling problems and to achieve a carbon dioxide level of less than 20 mg/L. Changes in tanks, pipelines, and sparger design deemed necessary to achieve desired carbon dioxide levels and to assure continuous operations without scale blockage were made.

2. *Task LT-II — Demonstration of operation of lime train.* — Task LT-II began on November 7, 1979, and continued to December 3, 1979.

a. Lime startup. — Lime addition to the LTPP train and the effects of lime dosage on pH, calcium, magnesium, silica, flocculation, and effluent turbidity were determined. During this initial phase, the lime treatment was operated approximately at the following conditions:

• Raw water	Flow, L/min	151
	Return to river, L/min	125
• Train influent	Flow, L/min	20 to 30
• Aerated train influent	CO_2, mg/L	<20
• Rapid-mix tank	Flow, L/min	23 to 30
	Lime feed rate pH	9.5, 10.0, 10.4
• Flocculator tank	Speed, r/min	80 to 120
• Clarifier tank	Speed, r/min	0.6
	Sludge recycle, L/min	11

b. Records. — For task LT-II and all the future lime treatment operations, daily operation reports were maintained for raw water, aerated water, clearwell influent and effluent, and filters.

c. Filter startup. — The dual-media filters were put into operation. Feed water was started to the dual-media filters. After demonstrating acidified lime-treated clarifier, effluent was maintained with a target turbidity of less than 200 JTU's.

3. *Tasks LT-III and IV, steady-state water production.* — Combined tasks LT-III and IV began on December 3, 1979, and were shut down temporarily on December 22, 1979, for scheduled maintenance and Christmas leaves. Tasks LT-III and IV resumed on January 3, 1980, and continued until April 21, 1980. The objectives of tasks LT-III and IV included:

Demonstration of the operation of the A-L train at 150 percent of design basis flowrate for the flocculator, clarifier, and dual-media filters. The A-L train was operated at 150 percent design basis flow rate throughout tasks LT-III and IV to assure an adequate feed water supply for the concomitant testing of the IX process;

Data acquisition for operation of the LTPP;

Demonstration of steady-state operations at the optimum conditions established by task LT-II; and

Operate the LTPP, on lime-softened water mode to produce constant quality water for the IX process test program.

a. Train operation. — The lime train operating data for tasks LT-III and IV during the period from December 3, 1979, through April 21, 1980, are presented in table 9. During this period, the train operated at a flow rate of approximately 0.9×10^{-3} m^3/s and an average flocculation tank pH of 9.8. The average calcium reduction in the flocculation tank was 339 g/m^3 (848 g/m^3 as $CaCO_3$). The clarifier effluent pH averaged 9.7. Total alkalinity as $CaCO_3$ was reduced, on the average, from 956 g/m^3 in the aerated train influent to 58 g/m^3 in the clarifier effluent. The phenolphthalein alkalinity was zero in the aerated train influent and averaged 30 g/m^3 in the clarifier effluent. Therefore, the caustic alkalinity (excess lime) averaged 2.0 g/m^3 in the clarifier effluent.

b. Clarifier operation. — Figure 21 presents the plot of the clarifier effluent for its entire operation on lime softening from December 1, 1979, through May 22, 1980. Figure 21e covers the operation of the clarifier for the steady-state period March 29 to April 22, 1980. Chemical and physical data for the clarified water during the period are given in table 10. Table 11 is the data after acid stabilization of the clarifier effluent. The maximum desired turbidity was 200 JTU. Figure 21e indicates frequent excursion above the control point. The main reason was operating the lime train at 150 percent of design flow rate. The lime train was operated later at the design flow rate of 0.63×10^{-3} m^3/s (10 gal/min) during phase II, task LT-V.

c. Filter operation. — Operating data for filters 3 and 4 are summarized in table 12 for the period of December 5, 1979, to April 21, 1980. Physical and chemical data for the effluent of filters 3 and 4 are presented in tables 10, 13, and 14. The average filter effluent pH for both filters 3 and 4 was 7.0; the average calcium carbonate saturation pH was 7.2 for both filters 3 and 4. Since the average effluent pH was lower than both the calculated calcium carbonate saturation pH's, the Langelier Saturation Index was negative, indicating that the effluent was undersaturated with respect to calcium carbonate.

4. *Task LT-V, optimization of lime train operation at design basis.* — During phase 1 of the LVS test program, it was necessary to operate the lime train at 150 percent of design feed water flow rate or at 0.95×10^{-3} m^3/s (15 gal/min). At the beginning of phase II, during period April 21 through May 22, 1980, the lime train was operated at the design flow rate of 0.63×10^{-3} m^3/s (10 gal/min) as task LT-V.

The objective of task LT-V included:

Demonstration of steady-state operation of the A-L train at the design basis flow rate;

Data acquisition for operation of the LTPP;

Optimization of the process for lime treatment of LVS-aerated water; and

Establish lime treatment unit operation of the A-L-SA process.

On April 21, testing was initiated to better determine the optimum operating conditions for the lime-softening pretreatment system. The primary objective of this phase II testing was to reduce the clarifier effluent to below 200 JTU at a flow rate of 0.63×10^{-3} m^3/s, the design flow rate of train L-I. This testing continued until May 9 when L-SA optimization testing was initiated.

a. Test procedure demonstrated. — Task LT-V brought the operation of LTPP from the phase I flow rate of task LT-IV of 57 L/min (15 gal/min) to the design basis flow rate of 38 L/min (10 gal/min). The following optimum equilibrium conditions were demonstrated for a feed water rate of 38L/min and a lime pH adjustment of 9.8:

• Raw water	Flow, 170 L/min (45 gal/min) Return to river, 132 L/min (35 gal/min)
• Train effluent	Flow, 38 L/min (10 gal/min)
• Aerated train influent	CO_2, ≤20 mg/L
• Rapid mix tank	Flow, 38 L/min Lime feed rate, 9.8 pH
• Flocculator tank	Speed, 140 r/min
• Clarifier tank	Speed, 0.6 r/min Sludge recycle, 19 L/min (5 gal/min)
• Ferric sulfate	Conc. 7.5 mg/L

The LTPP train operation was optimized to minimize the following responses:

- Calcium, mg/L
- Turbidity of stabilized reactor effluent, JTU
- Turbidity of filter effluent, JTU
- Plugging factor of filter effluent, percent

b. Task test conditions. — A total of 12 test conditions were attempted in the phase II testing. Table 15 presents the various test conditions. All 12 tests were run at the design basis flow rate 0.6×10^{-3} m^3/s. All lime additions were controlled at pH 9.8. Operation of the lime train at pH 9.8 was determined in the earlier task to be the optimum lime feed rate that would assure the neutralization of all of the bicarbonate. Flocculator speed was tested at 75, 100, 120, and 140 r/min. Rake speed was tested at 0.5 and 0.6 r/min. Sludge recycle rate was tested at 0.2, 0.32, and 0.5×10^{-3} m^3/s. Clarifier sludge control level was tested 0.3, 0.6, and 0.9 m. Ferric sulfate additions were controlled at 0.65, 0.75, and 10.0 g/m^3. These variable levels were selected based on the optimum conditions already determined for operation of the LTPP at 150 percent of design flow rate and by the range limitations of the equipment.

c. Task results. — Table 16 presents operating data obtained for various conditions tested during this phase of the program. Clarifier effluent calcium, total hardness, total hardness, total alkalinity, and phenolphthalein alkalinity did not vary appreciably for any of the conditions tested. This was expected, since the primary objective of this phase of testing was to reduce the clarifier effluent turbidity.

Table 17 presents the physical and chemical properties of clarifier effluent before and after acid stabilization. Figure 22 presents the plot of the clarifier effluent turbidity for phase II. As can be seen from these data, the turbidities did not decline appreciably until May 13 after the feedwell was extended down to just above the rake. The average turbidity was reduced by approximately 40 JTU after this modification. The period between May 14 and May 19 represents the first time during the lime-softening operation that the average clarifier effluent turbidities were consistently below 200 JTU. These data indicate that the filtering effect of the sludge blanket is very useful in reducing turbidity.

Table 18 presents the summary operating data for filters 3 and 4. Physical and chemical data for the effluent of filters 3 and 4 are presented in table 19. Average filter effluent pH for filters 3 and 4 was 6.9; average calcium carbonate saturation pH for both filters 3 and 4 was 7.2. Since the average effluent pH was lower than calculated calcium carbonate saturation pH, the Langelier Saturation Index was negative, indicating effluent was undersaturated with respect to calcium carbonate.

D. Optimum Lime Treatment Conditions

Data in table 16 demonstrated test condition 4 in table 15 gives the highest amount of calcium removal from solution and the lowest clarifier effluent turbidity. Optimum results were as follows:

• Conditions:	Water flow rate	0.6×10^{-3} m^3/s
	Control pH	9.8
	Ferric sulfate dosage	7.5 g/m^3
	Flocculator speed	100 r/min
	Rake speed	0.6 r/min
	Sludge recycle rate	0.5×10^{-3} m^3/s
	Clarifier sludge level	0.6 m
• Responses:	Calcium removed by lime	330 g/m^3
	Clarifier effluent turbidity	188 JTU
	Excess lime	5 g/m^3

Table 9. — *Lime-softening steady-state operation, Task LT-III*

	Units	12/3/79 - 1/17-80 (tank T-6 and small tank used for aeration)			1/18/80 - 3/5/80 (small tank used for aeration)			3/8/80 - 4/21/80 (new tank used for aeration)			OVERALL SUMMARY		
		Max	Min	Avg	Max	Min	Avg	Max	Min	Avg	Max	Min	Avg
Aerated Train Influent													
Flow[a]	$m^3/s \times 10^{-3}$	1.0	0.7	0.9	0.9	0.8	0.9	0.9	0.9	0.9	1.0	0.7	0.9
pH (bench)		7.9	7.2	7.5	7.4	6.7	7.1	7.8	6.6	7.2	7.9	6.6	7.2
CO_2	g/m^3	45	9	23	110	42	73	220	17	58	220	9	51
Sulfides, total	g/m^3	3.0	0	1.4	2.4	0.5	1.1	3.6	0.2	1.0	3.6	0	11.2
Calcium	g/m^3	778	585	712	866	769	807	850	633	779	866	585	766
Hardness as $CaCO_3$	g/m^3	2640	1920	2400	2860	2380	2619	2780	2290	2552	2860	1920	2524
T-alk. as $CaCO_3$	g/m^3	996	652	845	1080	1012	1047	1068	720	976	1080	652	956
Temperature	°C	36.0	29.0	33.6	36.0	29.4	34.1	37.5	31.0	33.8	37.5	29.0	33.8
Flocculation Tank													
pH (in-line)		10.3	9.1	9.8	10.5	9.1	9.7	10.1	9.5	9.8	10.5	9.1	9.8
Solids conc.	vol %[b]	19	1	7.9	12	2	4.9	10	2	4	19	1	5.6
Clarifier													
0.3 m level	vol %[b]	5.5	0.1	1.7	2.0	0.2	0.8	1.5	Tr	0.7	5.5	0.1	1.1
0.6 m level	vol %[b]	5.0	0.1	1.1	3.0	0.2	0.5	1.5	0.2	0.5	5.0	0.1	0.7
Sludge Recycle													
Flow[a]	m^3/s	0.3	0.2	—	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.2	—
Sludge recycle solids	vol %[b]	87	6	37	59	7	22	35	9	16	87	5	25
Clarifier Effluent													
pH (bench)		10.1	9.3	9.7	10.1	9.4	9.7	10.1	9.4	9.8	10.1	9.3	9.7
Turbidity (in-line)	JTU	>500	60	—	>500	110	—	410	150	252	>500	60	—
Calcium	g/m^3	489	361	441	513	321	426	441	385	414	513	321	427
Hardness as $CaCO_3$	g/m^3	1800	1420	1642	1880	1500	1584	1620	1500	1546	1880	1420	1590
T-alk. as $CaCO_3$	g/m^3	96	44	59	76	44	57	72	48	57	96	44	58
P-alk. as $CaCO_3$	g/m^3	40	8	28	40	16	31	40	20	31	40	8	30
Summary													
Calcium removal	g/m^3			271			381			365			339
Calcium removal as $CaCO_3$	g/m^3			678			952			912			848
Hardness removal as $CaCO_3$	g/m^3			758			1035			1007			933
T-alk. removal as $CaCO_3$	g/m^3			7[illegible]6			990			919			898
Excess lime present in clarifier effluent	g/m^3			None			5			5			2

[a] Flow rates when in service.
[b] Determined after 15-minute settling time in a graduated cylinder.

Table 10. — *Lime-softening steady-state operation raw clarifier effluent composition*

Date Time Collected Date Analyzed at Denver	Units	3/25/80 0750 5/6/80	4/1/80 0740 6/3/80	4/7/80 0850 6/5/80	Summary Max	 Min	 Avg
pH		10.5 (9.6)*	7.4 (9.6)*	8.2 (9.6)*	10.5 (10.1)*	6.0 (9.5)*	7.2 (9.7)*
TDS (calculated)	g/m^3	8280	8580	8580	8690	8280	8543
TDS (evaporated)	g/m^3	8510 (8540)	9620 (8669)	8580 (8423)	9620 (8860)	8330 (8249)	8582 (8590)
Conductivity @ 25 °C	mS/m	1240 (1350)	1280 (1380)	1150 (1380)	1280 (1500)	1150 (1100)	1242 (1308)
Silica	g/m^3	11.0	9.3	7.0	17	0	10
Calcium	g/m^3	416	480	416	480	385	418
Magnesium	g/m^3	117	351	142	351	113	143
Sodium	g/m^3	2180	2020	2350	2450	2020	2315
Potassium	g/m^3	213	206	206	225	196	210
Iron, total	g/m^3	0.40	0.38	0.13	0.40	0.06	0.11
Manganese, total	g/m^3	0.02	0.02	0.01	<0.05	0	-
Strontium	g/m^3	6.0	5.2	6.0	7.0	4.6	5.9
Bicarbonate	g/m^3	95.2 (12.0)	40.9 (10)	48.2 (0)	95.2 (12)	26.8 (0)	46.3 (3.5)
Carbonate	g/m^3	21.0 (26.0)	0 (34)	0 (26.0)	37.2 (36)	0 (24)	12 (29)
Hydroxide	g/m^3	- (0)	0	4.0	4.6 (4.0)	0.9 (0)	2.9 (-)
Sulfate	g/m^3	1840	1850	1790	2020	1790	1925
Chloride	g/m^3	3380	3620	3620	3620	3380	3489
T-alkalinity as $CaCO_3$	g/m^3	- (54)	- (64)	- (56)	71 (64)	30 (50)	38 (55)
P-alkalinity as $CaCO_3$	g/m^3	- (22)	- (28)	- (34)	62 (34)	0 (20)	24 (28)
T-phosphorus as PO_4	g/m^3	-	-	-	1.5	0	0.4
Sulfides, total	g/m^3	-	-	-	-	-	-
E.F. [TDS(calculated)/cond.]		6.7	6.7	7.5			
E.F. [TDS(evaporated)/cond.]		6.9 (6.3)	7.5 (6.3)	7.5 (6.1)			
Σ Anions	eq/m^3	135.86	141.17	140.09			
Σ Cations	eq/m^3	130.86	146.08	139.66			
Control value	eq/m^3	+2.26	-2.14	+0.19			

*Analyses conducted at LVS test site

Table 11. — *Lime-softening steady-state operation stabilized clarifier effluent composition*

Date Time Collected Date Analyzed at Denver	Units	3/25/80 0750 5/6/80	4/1/80 0745 6/3/80	4/7/80 0850 6/5/80	Summary Max	 Min	 Avg
pH		9.0 (6.7)*	4.7 (6.4)*	8.0 (6.3)*	9.0 (7.4)*	4.7 (6.0)*	5.9 (6.6)*
TDS (calculated)	g/m³	8470	8520	8400	8920	8400	8642
TDS (evaporated) @ 105°C	g/m³	8660 (8688)	9370	8660	9370	8450	8684
Conductivity @ 25 °C	mS/m	1240 (1370)	1380 (1400)	1150 (1400)	1380 (1500)	1150 (1100)	1246 (1321)
Silica	g/m³	15.5	9.1	7.0	18	5.0	11
Calcium	g/m³	432	608	448	608	420	446
Magnesium	g/m³	107	205	122	205	107	133
Sodium	g/m³	2270	2020	2380	2440	2020	2315
Potassium	g/m³	213	203	201	225	201	211
Iron, total	g/m³	0.39	0.49	0.20	0.49	0.06	0.15
Manganese, total	g/m³	0.02	0.02	0.01	<0.05	0	-
Strontium	g/m³	5.9	5.3	5.5	7.8	4.7	5.8
Bicarbonate	g/m³	123 (39.0)	1.83(37)	51.2 (32)	55.5	32	45
Carbonate	g/m³	7.8 (0)	0	0	0	0	0
Sulfate	g/m³	1920	1920	1640	2220	1640	2005
Chloride	g/m³	3380	3550	3550	3610	3340	3483
T-alkalinity as $CaCO_3$	g/m³	-(32)	(30)	(26)	45	26	37
P-alkalinity as $CaCO_3$	g/m³	-(0)	(0)	(0)	0	0	0
T-phosphorus as PO_4	g/m³	-	-	-	1.4	0	0.5
Sulfides, total	g/m³	-	-	-	-	-	-
Hydroxide	g/m³	-	- (0)	- (0)	-	-	-
E.F. [TDS(calculated)/cond.]		6.8	6.2	7.3			
E.F. [TDS(evaporated)/cond.]		7.0 (6.3)	6.8	7.5			
Σ Anions	eq/m³	137.48	140.03	135.04			
Σ Cations	eq/m³	134.46	140.38	140.54			
Control value	eq/m³	+1.35	-0.15	-2.50			

*Analyses conducted at LVS test site

Table 12. — *Lime-softening steady-state operation summary operating data, filters 3 and 4*

Filter no.		3			4		
From		12-5-79			12-8-79		
To		4-21-80			4-21-80		
Duration (hr), typical		72			72		
Number of filter runs		17			17		
Final wash-water turbidity, JTU, low/high		1.0/29.0			3.7/22.0		
		Maximum	Minimum	Average	Maximum	Minimum	Average
Filter effluent (pipeline header)							
Flow x 10^{-3}	m³/s	0.9	0.5	0.9	0.9	0.5	0.9
pH (approx.)		7.0	4.9	5.9	7.3	4.2	5.8
Turbidity	JTU	>500	45	100	190	19	95
Calcium	mg/L	521	353	436	473	405	433
Hardness as $CaCO_3$	mg/L	1900	1500	1614	1780	1500	1603
T-alk. as $CaCO_3$	mg/L	48	12	27	68	8	27
P-alk. as $CaCO_3$	mg/L	ND	ND	ND	ND	ND	ND
Filter effluent							
pH		8.0	6.0	7.0	7.2	6.8	7.0
Turbidity	JTU	4.2	0.00	0.26	3.50	0.00	0.25
Plugging factor	percent	-	-	-	-	-	-
Calcium	mg/L	505	361	440	480	413	438
Hardness as $CaCO_3$	mg/L	1800	1540	1624	1820	1540	1617
T-alk. as $CaCO_3$	mg/L	76	24	48	78	24	48
P-alk. as $CaCO_3$	mg/L	ND	ND	ND	ND	ND	ND
Iron	mg/L	-	-	-	-	-	-
Manganese	mg/L	-	-	-	-	-	-
Free Cl_2 residual	mg/L	1.6	0.0	0.5	1.1	0.2	0.6
Total Cl_2 residual	mg/L	3.9	0.2	1.4	4.0	0.2	1.5
Temperature	°C	35.5	25.0	29.7	32.5	25.0	29.1
Total head loss	m	3.0	0.14	0.35	0.55	0.03	0.36

Flow rates when in service.

Table 13. — *Lime-softening steady-state operation effluent composition, filter 3*

Date		1/7/80	1/15/80	1/21/80	1/28/80	2/19/80	2/25/80	3/3/80	3/10/80	3/17/80	Summary		
Time collected		0530	0800	0815	0830	0945	0900	0920	0845	0845			
Date analyzed at Boulder City		1/15/80	1/23/80	2/4/80	3/11/80	3/17/80	3/18/80	3/21/80	3/21/80	5/1/80[c]	Max	Min	Avg
ph		7.0	7.4	7.9	7.6	7.7	7.9	7.8	7.7	7.5	7.9	7.0	7.5
TDS (calculated)[a]	g/m³	8750	8610	8850	8690	8550	8520	8670	8600	8100	8850	8100	8593
TDS (evaporated @ 180 °C)	g/m³	8640	8590	8630	8660	8450	8570	8620	8610	8630[d]	8660	8450	8600
Conductivity @ 25 °C	mS/m	1230	1220	1220	1250	1230	1240	1260	1260	1190	1260	1190	1233
Silica	g/m³	17	16	12	5	11	11	12	15	8.0	17	5	12
Calcium	g/m³	425	420	435	465	385	445	435	425	432	465	385	430
Magnesium	g/m³	137	146	128	113	119	128	128	128	127	146	113	128
Sodium	g/m³	2340	2320	2440	2310	2320	2320	2380	2350	2180	2440	2180	2329
Potassium	g/m³	215	225	225	205	205	205	205	205	210	225	205	211
Iron, total	g/m³	0.08	0.06	0.07	0.05	0.07	0.06	0.05	0.07	0.03	0.08	0.03	0.06
Manganese, total	g/m³	0.04	0.04	<0.06	<0.05	<0.06	<0.05	<0.05	<0.05	<0.04	<0.05	<0.05	-
Strontium	g/m³	6.2	5.0	6.5	6.3	7.3	6.8	6.0	6.0	5.14	7.3	5.0	6.0
Bicarbonate	g/m³	43.9	43.3	54.3	43.9	43.3	50.6	44.5	46.4	48.8	54.3	43.3	46.6
Carbonate	g/m³	0	0	0	0	0	0	0	0	0	0	0	0
Sulfate	g/m³	2100	2050	2030	2050	1970	1900	2000	1990	1710	2100	1710	1978
Chloride	g/m³	3490	3410	3550	3510	3510	3480	3480	3460	3380	3550	3380	3474
T-alkalinity as $CaCO_3$	g/m³	36	36	45	36	36	42	37	38	46[b]	46	36	39
P-alkalinity as $CaCO_3$	g/m³	0	0	0	0	0	0	0	0	0[b]	0	0	0
T-phosphorus as PO_4	g/m³	0.0	0.0	0.0	0.2	0.6	0.7	0.7	0.9	-	0.9	0	0.4
E.F. [TDS(calculated)/cond.]		7.1	7.1	7.3	7.0	7.0	6.9	6.9	6.8	6.8			
E.F. [TDS(evaporated)/cond.]		7.0	7.0	7.1	6.9	6.9	6.9	6.8	6.8	7.3			
Σ Anions	eq/m³	142.8	139.6	143.2	142.2	140.5	138.2	140.4	139.7	130.8			
Σ Cations	eq/m³	139.5	139.8	144.0	138.3	135.3	139.0	141.0	139.0	136.0			
Control value	eq/m³	+1.42	-0.09	-0.34	+1.69	+2.28	-0.36	-0.26	+0.31	-2.44			

[a] Calculated TDS uses only one-half of the bicarbonate.
[b] Analysis conducted at Denver laboratory.
[c] Analysis conducted at LVSTS.
[d] TDS (evaporated at 105 °C).

Table 14 — *Effluent composition, filter 4, lime-softening steady-state operation*

Date		12/10/79	12/17/79	2/4/80	2/11/80	3/25/80	4/1/80	4/7/80	Summary		
Time collected		1115	0820	0815	0755	0750	0750	0850			
Date analyzed at Boulder City		12/13/79	12/21/79	3/13/80	3/13/80	5/6/80[b]	6/3/80	6/5/80	Max	Min	Avg
pH		7.2 (7.1)[c]	7.1 (7.2)[c]	7.7 (7.2)[c]	7.5 (6.8)[c]	7.8 (7.0)[c]	4.6 (7.0)[c]	7.8 (6.8)[c]	7.8 (7.2)[c]	4.6 (6.8)[c]	5.4 (7.0)[c]
TDS (calculated)	g/m^3	8600[a]	8650[a]	8770[a]	8660[a]	8400	8390	8580	8770	8390	8579
TDS (evaporated @ 180°C)	g/m^3	8700	8680	8670	8700	8630[d](8657)[c]	9430	8650	9430	8630	8780
Conductivity @ 25°C	mS/m	1270	1260	1260	1250	1240(1390)[c]	1340	1150	1390	1150	1274
Silica	g/m^3	6.0	12	17	10	10.9	8.9	6.5	17	6	10
Calcium	g/m^3	415	420	425	406	432	432	464	464	406	428
Magnesium	g/m^3	125	143	140	113	117	381	106	381	106	161
Sodium	g/m^3	2300	2420	2420	2270	1940	2380		2420	1940	2290
Potassium	g/m^3	196	215	215	215	217	206	201	217	196	209
Iron, total	g/m^3	0.06	0.08	0.06	0.07	0.23	0.24	0.11	0.24	0.06	0.12
Manganese, total	g/m^3	0.00	0.00	<0.06	<0.06	0.02	0.03	0.01	<0.06	0	-
Strontium	g/m^3	4.9	4.6	7.0	6.3	5.9	5.1	5.4	7.0	4.6	5.6
Bicarbonate	g/m^3	36.6	39.7	45.8	47.0	45.8 (49.0)	1.83(51)[c]	41.5 (46)[c]	5	36.6	45
Carbonate	g/m^3	0.0	0	0	0	0(0)[c]	0	0	0	0	0
Hydroxide	g/m^3	-	-	-	-	-	0[c]	0[c]	-	-	-
Sulfate	g/m^3	2090	2030	2020	1990	1930	1940	1830	2090	1830	1976
Chloride	g/m^3	3440	3510	3500	3480	3380	3480	3550	3550	3380	3477
T-alkalinity as $CaCO_3$	g/m^3	30	33	38	39	(40)[c] 42	38		42	30	37
P-alkalinity as $CaCO_3$	g/m^3	0	0	0	0	-(0)[c]	0	0	0	0	0
T-phosphorus as PO_4	g/m^3	0.5	0.0	1.5	0.8	-	-	-	1.5	0	0.7
E.F. [TDS(calculated)/cond.]		6.9	6.9	7.0	6.9	6.8	6.3	7.5			
E.F. [TDS(evaporated)/cond.]		6.8	6.9	6.9	7.0	7.0(6.2)[c]	7.0	7.5			
Σ Anions	eq/m^3	141.0	141.7	141.2	140.2	136.25	138.53	138.68			
Σ Cations	eq/m^3	136.0	138.3	143.3	140.0	135.35	142.55	139.98			
Control value	eq/m^3	+2.18	+1.48	-0.91	+0.09	+0.41	-1.78	-0.58			

[a] Calculated TDS uses only one-half of the bicarbonate

[b] Analysis conducted at Denver laboratory

[c] Analysis conducted at LVSTS

[d] TDS (evaporated at 105 °C)

Table 15. — *Lime-softening pretreatment optimization test operating conditions, April 21, 1980 through May 19, 1980*

	Test 1	Test 2	Test 3	Test 4	Test 5	Test 6
Dates	4/21/80 to 4/25/80	4/26/80 to 4/27/80	4/27/80 to 4/28/80	4/28/0 to 4/29/80	4/30/80 to 5/1/80	5/1/80 to 5/2/80
Flow rate ($m^3/s \times 10^{-3}$)	0.6	0.6	0.6	0.6	0.6	0.6
Flocculation Tank Control pH	9.8	9.8	9.8	9.8	9.8	9.8
Ferric Sulfate Dosage (g/m^3)	7.5	7.5	7.5	7.5	10.0	6.5
Flocculator Speed (rpm)	140	100	75	100	100	100
Rake Speed (rpm)	0.6	0.6	0.6	0.6	0.6	0.6
Sludge Recycle Rate ($m^3/s \times 10^{-3}$)	0.32	0.32	0.32	0.5	0.5	0.5
Clarifier Sludge Control Level* (height in meters at 0.5% by volume maximum concentration)	0.6	0.6	0.6	0.6	0.6	0.3

	Test 7	Test 8	Test 9	Test 10	Test 11	Test 12
Dates	5/3/80 to 5/5/80	5/5/80 to 5/6/80	5/6/80 to 5/12/80	5/12/80 to 5/12/80	5/14/80 to 5/16/80	5/16/80 to 5/19/80
Flow rate ($m^3/s \times 10^{-3}$)	0.6	0.6	0.6	0.6	0.6	0.6
Flocculation Tank Control pH	9.8	9.8	9.8	9.8	9.8	9.8
Ferric Sulfate Dosage (g/m^3)	-	7.5	7.5	7.5	7.5	7.5
Flocculator Speed (rpm)	100	100	120	120	140	140
Rake Speed (rpm)	0.6	0.6	0.5	0.5	0.5	0.5
Sludge Recycle Rate ($m^3/s \times 10^{-3}$)	0.5	0.5	0.5	0.2	0.2	0.2
Clarifier Sludge Control Level* (height in meters at 0.5% by volume maximum concentration)	0.6	0.6	0.9	0.9	0.9	0.6

*Sludge wasting intermittent as required

Table 16. — *Lime-softening pretreatment optimization test results, task LT-V - April 21, 1980, through May 19, 1980 (Average operating data, lime train)*

		Test (average)											
		1	2	3	4	5	6	7	8	9	10	11	12
Aerated train influent													
Flow[1]	$m^3/s \times 10^{-3}$	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
pH (bench)		6.9	7.4	7.5	7.4	7.5	7.6	7.4	7.5	7.5	7.5	7.5	7.5
CO_2	g/m^3	38	10.1	10.6	17.3	4.0	0.4	0.7	3.1	7.9	1.8	1.5	9.5
Sulfides, total	g/m^3	0.6	0.7	0.8	1.0	1.3	0.9	1.7	2.1	1.3	1.4	1.4	1.1
Calcium	g/m^3	736	720	724	734	629	641	625	614	645	625	632	634
Hardness as $CaCO_3$	g/m^3	2462	2400	2350	2437	2160	2180	2180	2267	2223	2160	2175	2176
T-alk. as $CaCO_3$	g/m^3	880	831	860	868	602	611	595	564	6716	648	615	634
Temperature	C	34.0	34.8	34.0	35.2	33.6	33.2	35.1	34.0	33.6	34.5	33.0	34.0
Flocculation tank													
pH (in-line)		9.9	9.8	9.8	9.8	9.8	9.8	9.8	9.8	9.8	9.8	9.8	9.8
Solids conc	vol pct[2]	4.3	2.8	2.0	5.5	3.4	3.7	2.8	2.8	8.1	7.0	3.3	7.6
Clarifier													
0.3 m level	vol pct[2]	0.6	0.4	0.3	0.8	1.1	1.0	0.7	0.6	2.4	1.2	0.5	1.0
0.6 m level	vol pct[2]	0.5	0.4	0.3	0.7	1.0	0.8	0.5	0.5	0.3	Tr	0	0.2
Sludge recycle													
Flow[1]	$m^3/s \times 10^{-3}$	0.3	0.3	0.3	0.4	0.5	0.5	0.5	0.5	0.5	0.2	0.2	0.2
Sludge recycle solids	vol pct[2]	17	12	12	17	18	10	6	3.6	38	62	36	64
Clarifier effluent													
pH (bench)		9.8	9.7	9.9	9.7	9.7	9.7	9.6	9.7	9.7	9.6	9.6	9.7
Turbidity (in-line)	JTU	-	204	202	188	232	192	214	200	227	210	166	171
Calcium	g/m^3	422	410	417	404	409	422	408	426	413	417	410	409
Hardness as $CaCO_3$	g/m^3	1555	1533	1510	1520	1555	1567	1561	1570	1553	1560	1578	1548
T-alk. as $CaCO_3$	g/m^3	56	55	54	55	52	57	62	51	56	60	57	55
P-alk. as $CaCO_3$	g/m^3	32	29	28	30	29	32	36	32	31	32	29	31
Summary													
Calcium removal[4]	g/m^3	314	310	307	330	220	219	217	188	232	208	222	225
Calcium removal as $CaCO_3$	g/m^3	785	775	768	825	550	548	542	470	580	520	555	562
Hardness removal as $CaCO_3$	g/m^3	907	867	840	917	606	613	619	697	670	600	597	628
T-alk. removal as $CaCO_3$	g/m^3	824	776	806	813	550	554	533	513	615	588	558	579
Excess lime present in clarifier effluent	g/m^3	8	3	2	5	6	7	10	13	6	4	1	7

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[1] Flow rates when in service.

[2] Determined after 15-minute setting time in a graduated cylinder.

[3] For test condition 6, the reported values are for the 0.0-m and 0.3-m levels, respectively.

[4] Calcium removal by lime treatment on aerated water.

Table 17. — *Clarifier effluent composition, lime-softening pretreatment optimization testing, task LT-V*

		Raw water				Acid-stabilized water			
Date		4/28/80	5/5/80	5/12/80	5/19/80	4/28/80	5/5/80	5/12/80	5/19/80
Time Collected		0755	0840	0810	0820	0800	0840	0810	0820
Date analyzed at Yuma		6/5/80	6/9/80	6/8/80	6/23/80	6/5/80	6/9/80	6/8/80	6/23/80
pH[a]		9.5	9.7	9.6	9.7	6.4	6.7	6.2	6.8
TDS (calculated)[b]	g/m^3	8587	8493	8121	8226	8637	8577	8269	8232
TDS (evaporated @ 100 °C)[a]	g/m^3	8439	8528	8450	8510	8551	8524	8535	8572
Conductivity @ 25 °C[a]	mS/m	1280	1320	1310	1340	1290	1340	1340	1360
Silica	g/m^3	18.5	13.4	15	4	18.9	14.4	16	5
Calcium	g/m^3	414	423	413	438	425	427	411	431
Magnesium	g/m^3	142	140	127	118	134	146	127	116
Sodium	g/m^3	2364	2304	2200	2244	2404	2384	2248	2312
Potassium	g/m^3	164	175	164	163	163	169	166	159
Iron, total	g/m^3	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Manganese, total	g/m^3	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30
Strontium	g/m^3	4.70	3.40	4.3	4.2	4.80	3.23	4.3	4.1
Bicarbonate[a]	g/m^3	ND	ND	ND	22	54	26.8	34	29
Carbonate[a]	g/m^3	26	24	24	24	ND	ND	ND	ND
Sulfate	g/m^3	1970	1900	1830	1850	1970	1920	1960	1820
Chloride	g/m^3	3480	3510	3340	3370	3490	3500	3320	3370
T-alkalinity as $CaCO_3$[a]	g/m^3	56	54	50	58	44	22	28	24
P-alkalinity as $CaCO_3$[a]	g/m^3	34	34	30	20	ND	ND	ND	ND
Hydroxide[a]	g/m^3	4.0	3.4	3.4	ND	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		6.7	6.4	6.2	6.1	6.7	6.4	6.2	6.1
E.F. [TDS(evaporated)/cond.][a]		6.6	6.5	6.5	6.4	6.6	6.4	6.4	6.3
Σ Anions	eq/m^3	140.1	139.4	133.3	134.8	140.4	139.2	135.0	133.4
Σ Cations	eq/m^3	138.8	137.4	131.0	133.4	140.7	141.4	133.1	135.7
Control value	eq/m^3	+0.57	+0.88	+1.05	+0.64	-0.13	-0.99	+0.89	-1.09

[a] Analysis conducted at LVSTS.
[b] Calculated TDS uses only one-half of the bicarbonate.

Table 18. — *Lime-softening pretreatment optimization testing summary operating data, task LT-V, filters 3 and 4*

Filter no.		3			4		
From		4-21-80			4-24-80		
To		5-17-80			5-19-80		
Duration (hr), typical		72			72		
Number of filter runs		4			4		
Final wash water turbidity, JTU, low/high		3.0-8.0			2.0-8.2		
		Maximum	Minimum	Average	Maximum	Minimum	Average
Filter influent (pipeline header)							
Flow [1]	$m^3/s \times 10^{-3}$	0.9	0.5	0.6	0.6	0.5	0.6
pH (approx.)		6.7	4.7	5.4	6.6	4.4	5.2
Turbidity	JTU	95	35	72	130	45	73
Calcium	mg/L	489	401	430	465	357	426
Hardness as $CaCO_3$	mg/L	1700	1540	1598	1660	1440	1596
T-alk. as $CaCO_3$	mg/L	44	12	23	40	16	24
P-alk. as $CaCO_3$	mg/L	ND	ND	ND	ND	ND	ND
Filter effluent							
pH		7.2	6.7	6.9	7.0	6.8	6.9
Turbidity	JTU	1.5	0	0.28	1.2	0.07	0.30
Plugging factor	percent	95	11	59	98	15	66
Calcium	mg/L	473	4[illegible]3	434	449	357	428
Hardness as $CaCO_3$	mg/L	1660	1550	1612	1640	1500	1608
T-alk. as $CaCO_3$	mg/L	52	32	42	52	26	40
P-alk. as $CaCO_3$	mg/L	ND	ND	ND	ND	ND	ND
Iron	mg/L	-	-	-	-	-	-
Manganese	mg/L	-	-	-	-	-	-
Free Cl_2 residual	mg/L	3.6	ND	-	1.8	ND	-
Total Cl_2 residual	mg/L	4.0	ND	-	3.6	ND	-
Temperature	°C	33.5	27.0	29.8	35.0	27.0	29.6
Total head loss	m	0.33	0.15	0.22	0.34	0.15	0.24

[1] Flow rates when in service

Table 19. — *Effluent composition, lime-softening pretreatment optimization testing, task LT-V*

		Filter 3	Filter 4		
Date		4/28/80	5/5/80	5/12/80	5/19/80
Time collected		0800	0840	0810	0820
Date analyzed at Yuma		6/5/80	6/9/80	6/8/0	6/23/80
pH[a]		6.6	6.7	6.6	6.7
TDS (calculated)[b]	g/m^3	8542	8604	8343	8213
TDS (evaporated @ 105 °C)[a]	g/m^3	8492	8612	8520	8605
Conductivity @ 25 °C[a]	mS/m	1270	1360	1320	1300
Silica	g/m^3	18.0	12.2	14	5
Calcium	g/m^3	410	427	418	430
Magnesium	g/m^3	154	1[illegible]4	127	116
Sodium	g/m^3	2356	2372	2290	2223
Potassium	g/m^3	163	185	167	159
Iron, total	g/m^3	<0.10	<0.10	<0.10	<0.10
Manganese, total	g/m^3	<0.30	<0.30	<0.30	<0.30
Strontium	g/m^3	4.8	3.40	4.3	4.2
Bicarbonate[a]	g/m^3	51	41.5	46	32
Carbonate[a]	g/m^3	ND	ND	ND	ND
Sulfate	g/m^3	1960	1930	1950	1880
Chloride	g/m^3	3450	3510	3350	3380
T-alkalinity as $CaCO_3$[a]	g/m^3	42	34	38	26
P-alkalinity as $CaCO_3$[a]	g/m^3	ND	ND	ND	ND
Hydroxide[a]	g/m^3	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		6.7	6.3	6.3	6.3
E.F. [TDS(evaporated)/cond.][a]		6.7	6.3	6.5	6.6
Σ Anions	eq/m^3	139.0	139.9	135.9	135.0
Σ Cations	eq/m^3	138.8	141.1	135.3	131.9
Control value	eq/m^3	+0.09	-0.55	+0.27	+1.43

[a] Analysis conducted at LVSTS.

[b] Calculated TDS uses only one-half of the bicarbonate.

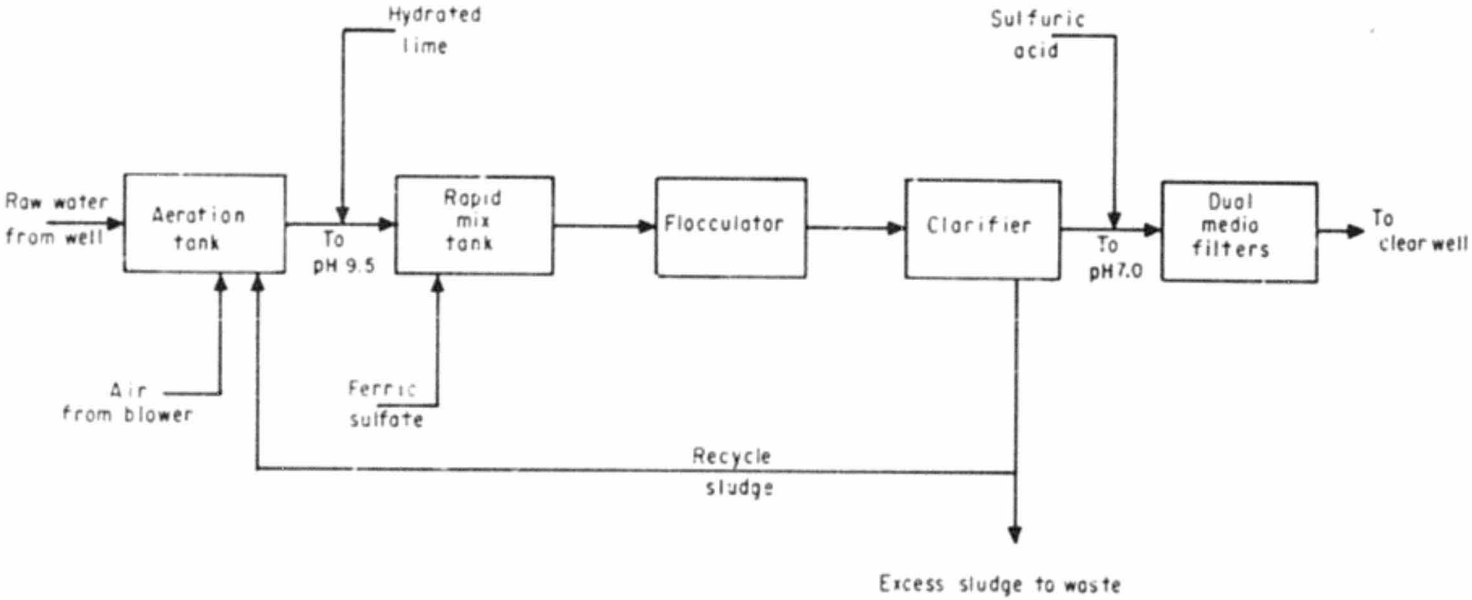
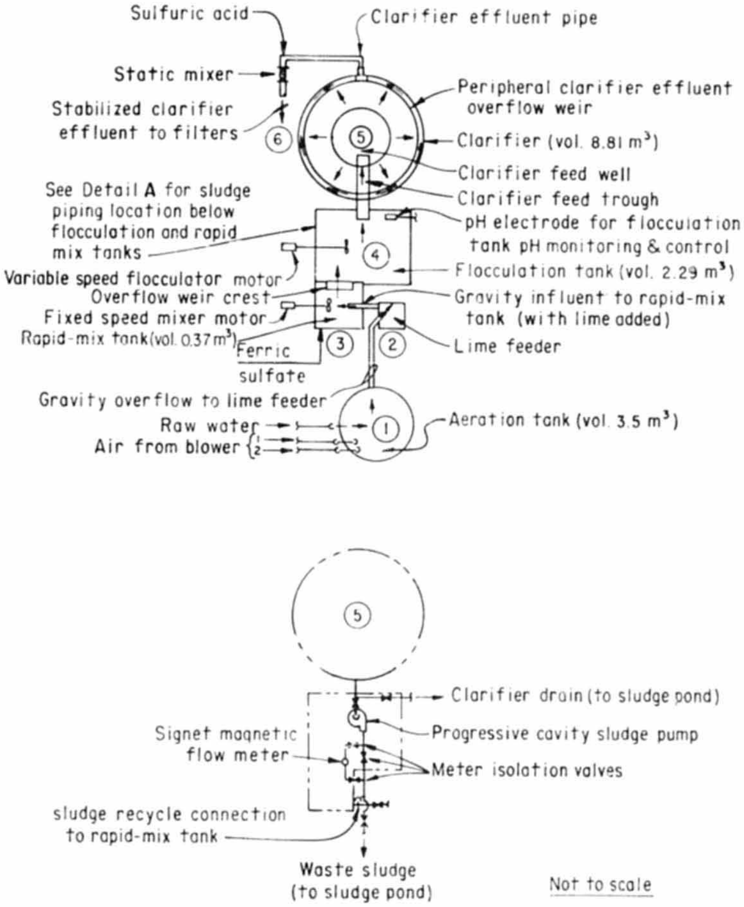


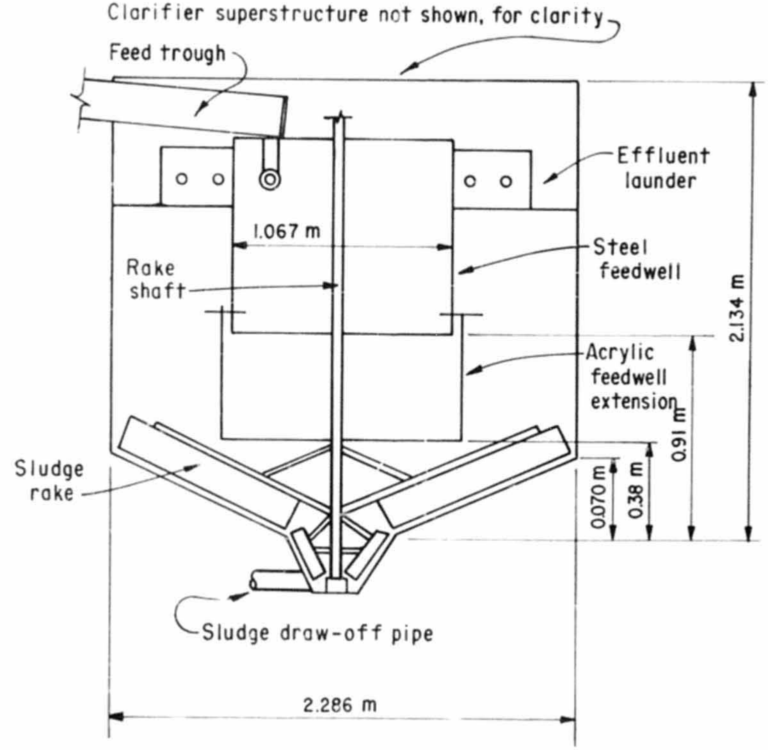
Figure 16.—Lime-softening pretreatment system flow diagram.



Not to scale

DETAIL A

Figure 17.—Lime-softening pretreatment final equipment configuration.



Not to scale

Figure 18.—Clarifier cross section showing feedwell modification.

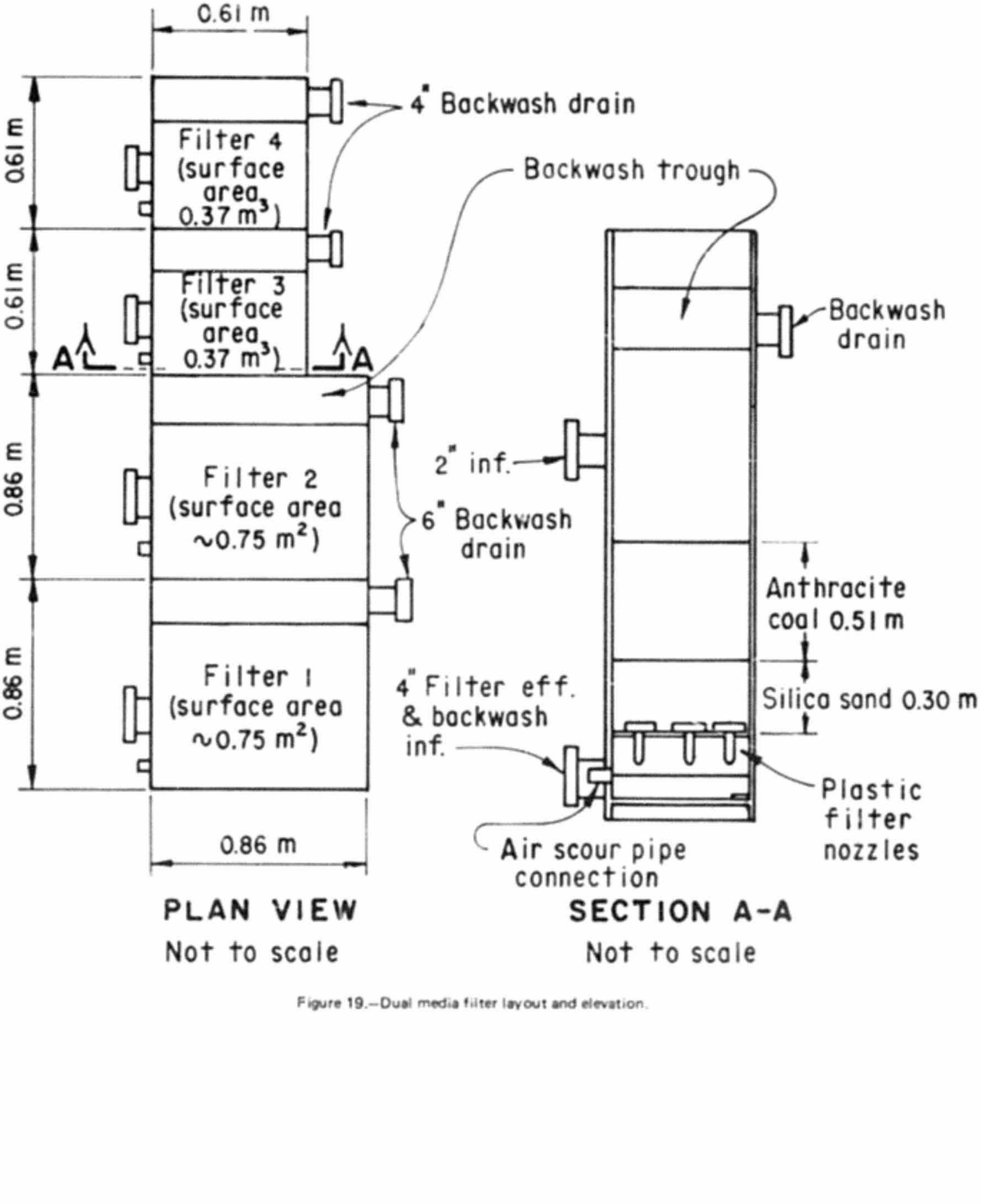


Figure 19.—Dual media filter layout and elevation.

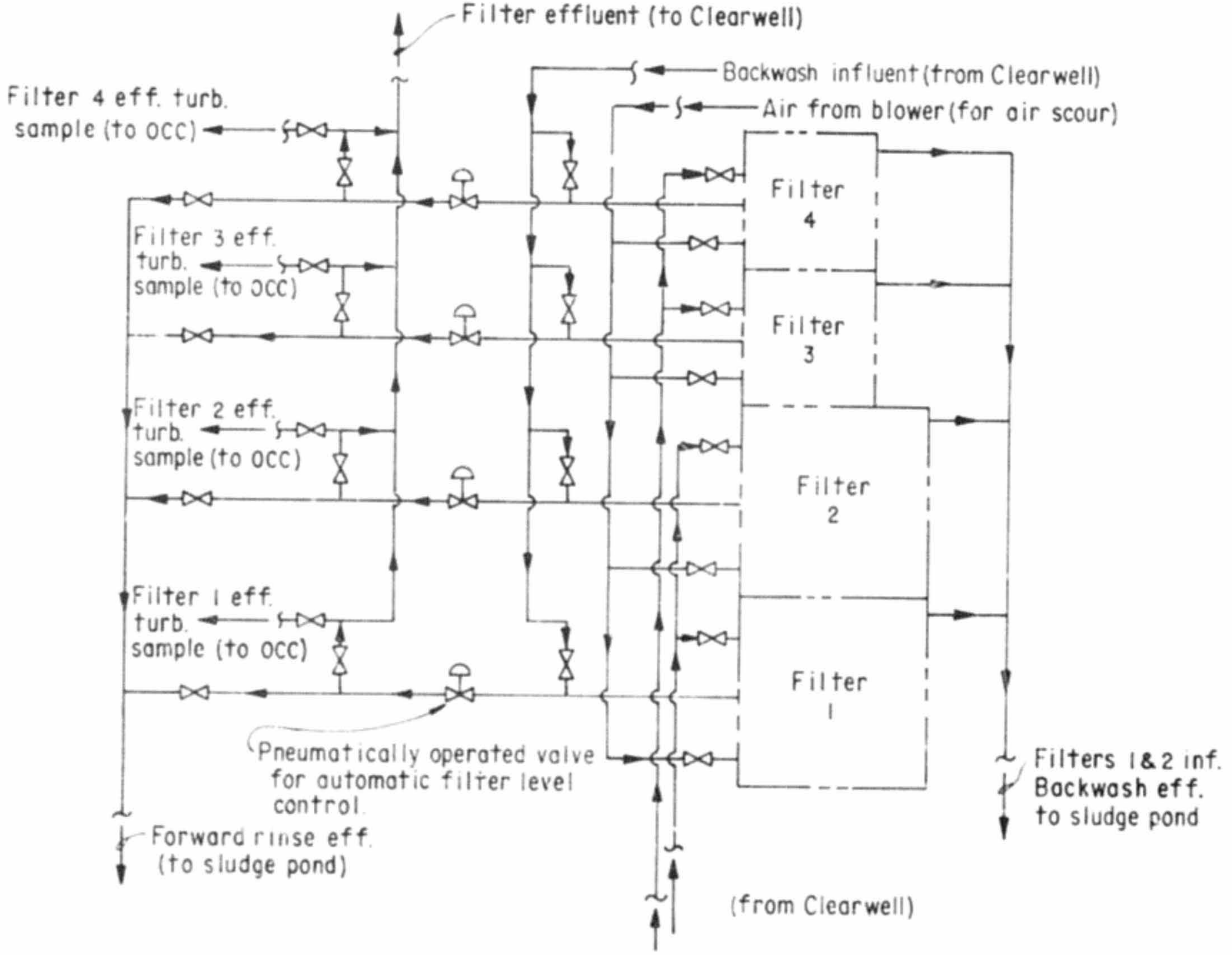


Figure 20.—Filter flow control piping manifold.

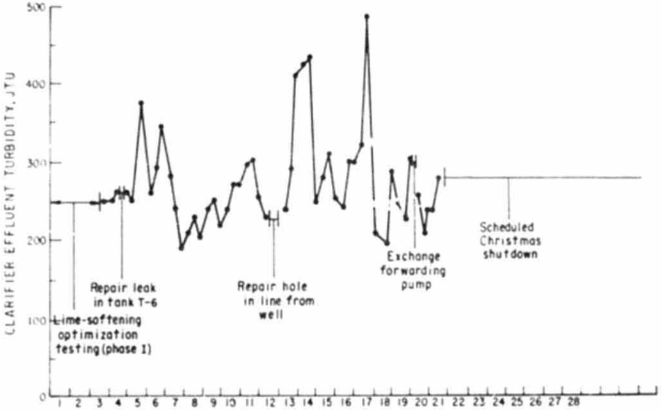


Figure 21a.–Clarifier effluent turbidity, lime-softening steady-state operation (December 1 through December 28, 1979).

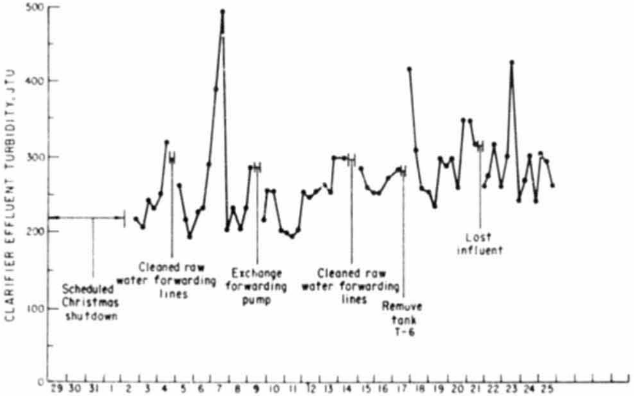


Figure 21b.–Clarifier effluent turbidity, lime-softening steady-state operation (December 29, 1979 through January 25, 1980).

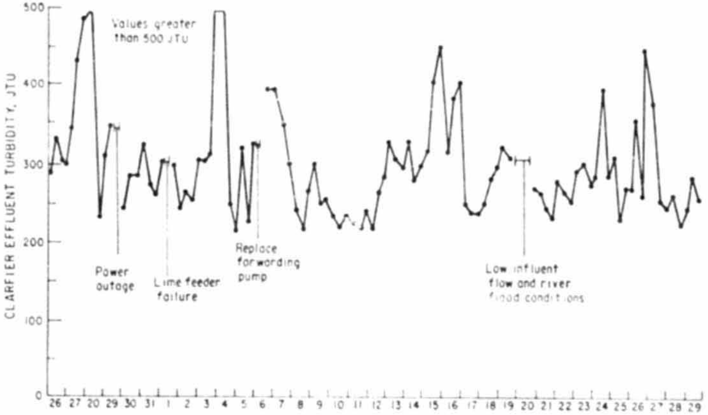


Figure 21c.–Clarifier effluent turbidity, lime-softening steady-state operation (January 26 through February 29, 1980).

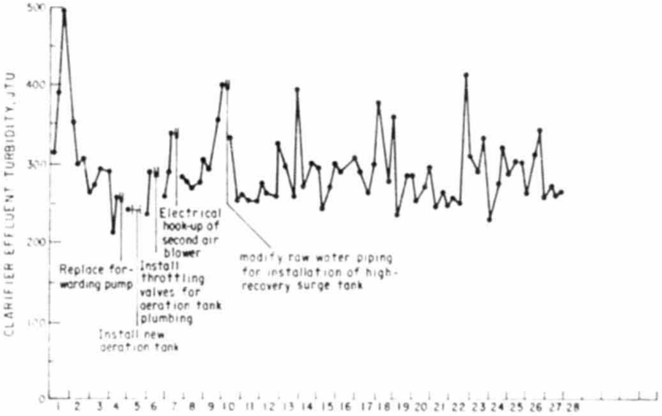


Figure 21d.–Clarifier effluent turbidity, lime-softening steady-state operation (March 1 through March 28, 1980).

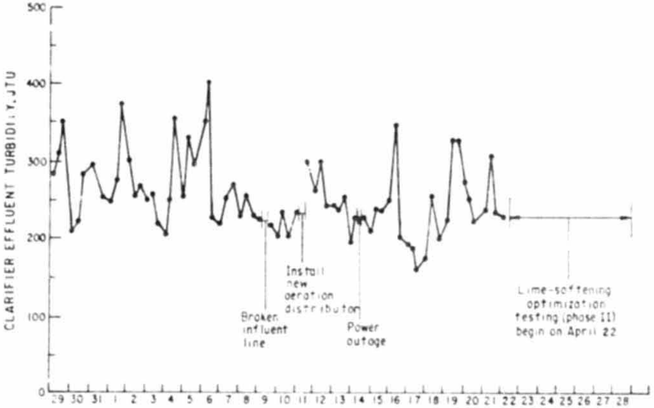


Figure 21e.–Clarifier effluent turbidity, lime-softening steady-state operation (March 29 through April 22, 1980)

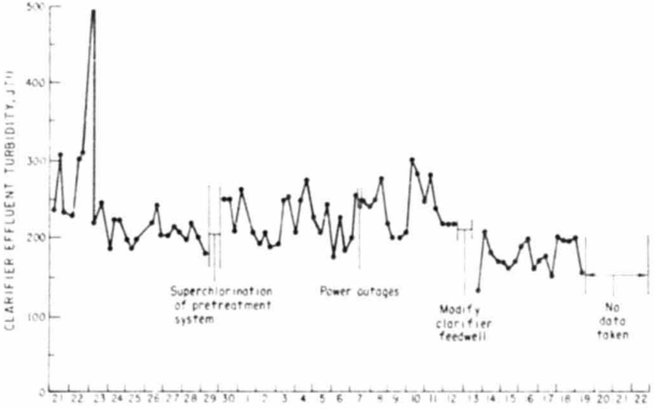


Figure 22.–Clarifier effluent turbidity, lime-softening optimization testing, Task LT-V (April 21 through May 19, 1980).

EXPERIMENTAL LIME-SODA ASH SOFTENING PROCESS

LaVerkin Springs raw water typically contains 820 mg/L of calcium ion and 1266 mg/L of bicarbonate ion. Aeration followed by partial lime treatment eliminated bicarbonate by precipitation of 400 mg/L of calcium as 1000 mg/L of calcium carbonate. The rest of the 420 mg/L of calcium remained in solution as soluble chloride salt. This remaining "permanent hardness" calcium could be precipitated as calcium carbonate by introducing carbonate ion into the water by the addition of soda ash as soluble sodium bicarbonate salt. An A-L-SA (aeration-lime-soda ash) process was evaluated for removal of both "temporary" and "permanent" calcium hardness from LVS water.

A. Equipment Description

Except for additional soda ash feeder equipment, equipment utilized for the A-L-SA process has already been described. This system adds a second volumetric feeder to proportion dry soda ash into a solutioning tank, from which soda ash solution is gravity fed through a flow control valve into the rapid-mix tank. The A-L-SA softening pretreatment system consisted of:

- Aeration system
- Pretreatment system
  - Lime feeder
  - Soda ash feeder
  - Rapid mix tank
  - Flocculation tank
  - Clarifier
- Dual media filters

B. Process Operations

The lime-soda ash softening operation is essentially the operation previously described. Figure 23 presents the flow diagram.

Figure 24 shows the flows through lime-soda ash pretreatment equipment configuration. Clarifier effluent is recycled forward to the soda ash solutioning tank through a rotometer at a flow rate manually preset by a needle valve. Dry powdered soda ash is volumetrically proportional with water in the solutioning tank. Aerated water, powdered lime, ferric sulfate, and soda ash solutions are blended in the rapid-mix tank. Lime addition is again controlled by pH sensors. However, to compensate for the effect of soda-ash on pH, a pH setting near 10.2 is needed to assure a stoichiometric amount of calcium hydroxide (lime) is added to neutralize dissolved bicarbonate ion. Coagulation occurs in the flocculator and precipitation occurs in the clarifier.

1. *Pretreatment system operation.* — Train operating data for this period are presented in table 20. Average calcium reduction in the lime-soda ash softening system, from June 17 through June 30 when the train operated at a desired control pH of 10.3, was approximately 540 g/m³ [1350 g/m³ as ($CaCO_3$)]. Clarifier effluent calcium concentration was 98 g/m³ (245 g/m³ as $CaCO_3$) and clarifier-effluent pH averaged 10.3. The CO_2 concentration in the aerated train influent averaged 43 g/m³. Total alkalinity as $CaCO_3$ was reduced, on the average, from 648 g/m³ in the aerated influent to 72 g/m³ in the clarifier effluent. Phenolphthalein alkalinity was zero in the aerated train influent and averaged 41 g/m³ in the clarifier effluent. Therefore, approximately 10 g/m³ of excess lime were present in the clarifier effluent.

On June 30, the flocculation tank control pH was reduced to 10.2. This was done for two reasons: (1) a 10.3 control pH was difficult to maintain, and (2) based upon observations, treatment efficiency was adequate at a lower pH. The train operated at this set of conditions until August 6. During the period of June 30 through August 6 and August 23 through August 29, the train operated with a desired control pH of 10.2; average calcium reduction was 600 g/m³ (1500 g/m³ as $CaCO_3$). Clarifier effluent calcium concentration averaged 44 g/m³ (110 g/m³ as $CaCO_3$) and clarifier-effluent pH averaged 10.2. The CO_2 concentration in the aerated train influent averaged 36 g/m³. Total alkalinity as $CaCO_3$ was reduced on the average from 658 g/m³ in aerated influent to 104 g/m³ in clarifier effluent. Phenolphthalein alkalinity was zero in aerated influent and averaged 40 g/m³ in clarifier effluent. Therefore, no excess lime was present in clarifier effluent.

On August 6, flow through the train was increased from design basis flow rate 0.63×10^{-3} m³/s to 0.95×10^{-3} m³/s to operate both the reverse osmosis system and the electrodialysis system concurrently. Flow was reduced back to 0.6×10^{-3} m³/s on August 22, when it was no longer necessary to operate the ED system. The train operated at 0.63×10^{-3} m³/s through the end of the LVS test program to August 29. During the period of August 6 through August 22, train influent flowrate was increased from 0.6×10^{-3} m³/s to 0.9×10^{-3} m³/s, and average calcium reduction was 648 g/m³ (1640 g/m³ as $CaCO_3$). Clarifier effluent calcium concentration averaged 37 g/m³ (92 g/m³ as $CaCo_3$) and the clarifier-effluent pH averaged 10.1 (pH 10.2 desired). CO_2 concentration in the aerated train influent averaged 60 g/m³. Total alkalinity as $CaCO_3$ was reduced on the average from 727 g/m³ in the aerated influent to 115 g/m³ in the clarifier effluent. Phenolphthalein alkalinity was zero in the aerated influent and averaged 57 g/m³ in the clarifier effluent.

Figure 25 shows the clarifier effluent turbidity during steady-state operations between May 31 and August 29, 1980. Tables 21a and 21b show clarifier effluent composition and acidified clarifier effluent composition, respectively, for phase II testing.

2. *Dual-media filter operation.* — Alternate operation of filters 3 and 4 continued throughout the period June 12 through August 29 to maintain a sufficient supply of backwash water and process water for support of the reverse osmosis and spiractor testing.

Maximum, minimum, and average values for various chemical constituents and operational parameters for filters 3 and 4 are presented in table 22. The average filter-effluent pH for filters 3 and 4 was 6.8 and 6.7, respectively. The average calcium carbonate saturation pH for filters 3 and 4 was 8.0 and 7.9, respectively. Since the average effluent pH was lower than calculated calcium carbonate saturation pH, the Langelier Saturation Index was negative, indicating that the effluent was undersaturated with respect to calcium carbonate.

Tables 23, 24, and 25 present concentrations of the various constituents in weekly grab samples of raw and aerated influents, clarifier effluent, stabilized clarifier effluent, filter 3 effluent, and filter 4 effluent, respectively.

C. Results and Discussion

During phase II of the LVS site test program, a study was made of the removal of the calcium remaining in solution after raw water was treated by aeration and lime addition. Typically, LVS raw water contained 820 mg/L of calcium. Aeration reduced calcium in solution to 600 mg/L. Subsequent lime treatment coprecipitated all bicarbonate and reduced calcium in solution to 420 mg/L. Task LT-SA was a study of the process for removing most of the calcium left in solution after aeration and lime treatment.

The primary objective of task LT-SA was to demonstrate the feasibility of removing permanent hardness-type calcium from previously aerated and lime-softened water. This was accomplished by two experimental tasks. Task LT-SA I provided for phasing in soda ash addition after demonstrating optimized steady-state aeration and lime-treatment at design basis flow rate of 0.63 x 10^{-3} m^3/s (10 gal/min) for the LTPP. Task LT-SA II optimized lime-soda ash treatment with a parametric study. A second objective was to provide feed water for LVS reverse osmosis membrane testing or task RO. This task LT-SA III called for steady-state operation at best conditions, as determined by task LT-SA II, to provide feed water for the various other systems to be tested as part of the LVS test program.

1. *Task LT-SA I phase in of soda ash addition to the lime process.* — This task calls for the modification of the lime-softening pretreatment system to provide for addition of soda ash for removal of permanent hardness. Modifications were made to the LTPP while the train was shut down during the week of May 12, 1980.

a. Soda ash equipment requirements. — Soda ash treatment of LVS water after aeration-lime softening required the following modifications to the chemical treatment train.

(1) Two 950-L (250-gal) (12-hr) capacity agitated soda ash solutioning tanks were installed with level indication the length of tank shells;

(2) Level gages on the solutioning tanks were calibrated to every 1-inch division as gallons of water. Divisions were marked as pounds of 5 percent soda ash solution by multiplying liters of water x 8.33 x 1.05. The 950-L (230-gal) (1910-1b) "water" level was marked on each tank;

(3) A manually preset metering system was provided for the range of 1.00 to 3.00 pounds of 5 percent soda solution per minute; and

(4) A pipe was provided to permit transfer from one solution tank at a time through a common metering system into the flocculator tank. Soda ash solution should enter to a point below the agitator remote from the pH electrodes.

b. The following steps were taken to change operation of the LTPP from lime treatment (only) to lime-soda ash treatment.

Operation on lime treatment was continued,

Lime treatment was brought to the best condition, determined by task LT-V;

Throughout the lime-soda ash pretreatment study, a constant lime dosage was maintained, at the same level used for task LT-V, and

The soda ash solution was prepared by dissolving 100 pounds of commercial soda ash in 1,910 pounds of aerated water.

c. Shut down. — During the week of May 11, 1980, the LTPP was shut down for the train modifications required to complete task LT-SA I.

2. *Task LT-SA II — Optimization of the lime-soda ash process.* — After completion of the modifications to the LTPP required by task LT-SA I, the pilot plant was put back on lime treatment and brought to optimum steady-state conditions established by task LT-V. Based on a water treatment flow rate of 0.6 x 10^{-3} m^3/s and a calcium content of 420 g/m^3, a stoichiometric addition rate of 5 percent soda ash solution was started. A stoichiometric addition rate of 5 percent soda ash was 14.38 g/s.

a. Variables. — Since addition of soda ash affected flocculation and turbidity, the effects of the following variables on filter influent and effluent were redetermined at stoichiometric addition rate:

Rake speed. — Adjustments to rake speed based on visual observations of effect on sludge blanket;

Solids level. — Adjustments of sludge wasting rate based on effects of solid contents at 1-, 2-, 3-, and 4-foot levels;

Sludge recycle rates. — Effect of 0-, 11-, and 28-L/min sludge recycle rate; and

Flocculator speed. — Minimum r/min needed to maintain suspension.

b. Parameters study. — Task LVS-LT-SA-II optimized lime-soda ash pretreatment with a parametric study. The system was operated at all the best conditions determined above except for variations in the rate of addition of soda ash solution. The LTPP operated for about 3 days to demonstrate equilibrium conditions at each of the following 5 percent soda ash solution feed rates:

Sub-Task	Feedrate (g/s)
A	12.12
B	13.84
C	14.38
D	14.66
E	15.12

The following responses were monitored:

- Residual, calcium, mg/L
- Residual, magnesium, mg/L
- Residual, silica, mg/L
- Residual, iron, mg/L
- Residual, manganese, mg/L
- Stabilized reactor effluent turbidity — JTU, plugging factor — percent
- Filter effluent turbidity — JTU, plugging factor —percent

c. Task procedure. — Lime-soda ash softening optimization testing was initiated on May 19 and continued until June 11, 1980. A total of eight operating conditions were tested. These are summarized in table 26.

d. Task results. — Table 27 presents operating data obtained under various conditions used during this phase of the test program. Based upon a comparison of both calcium and total hardness removal, test condition 7 was the most viable. This was followed by conditions 8, 2, and 6, respectively. The total alkalinity removals indicate that of these four conditions, condition 8 was the most viable, followed by conditions 7, 6, and 2.

Data in table 27 demonstrate test condition 7, as described in table 26, give highest amount of calcium removal, lowest clarifier effluent turbidity, and highest excess lime. Results are summarized as follows:

Conditions	
Water flow rate	0.6 x 10^{-3} m^3/s
Ferric sulfate dosage	7.5 x 10^{-3} m^3/s
Lime addition control pH	9.8
Lime dosage	1140 g/m^3
Flocculation tank control pH	10.2
Soda ash dosage	1200 g/m^3
Clarifier rake speed	0.6 r/min
Flocculator speed	140 r/min
Sludge recycle rate	0.34 x 10^{-3} m^3/s
Sludge control level	0.6 m
Responses	
Calcium removal (A-L-SA)	775 g/m^3
Residual calcium	32 g/m^3
Clarifier effluent turbidity	171 JTU
Excess lime	9 g/m^3

D. Optimum Lime-Soda Ash Treatment Conditions

Optimum steady-state conditions and water quality changes for operation of the aeration-lime-soda ash train are as follows:

Process conditions	
Raw water flow rate	0.63 x 10^{-3} m^3/s
Air rate	47.2 x 10^{-3} m^3/s
Lime addition control pH	9.8
Lime dosage	1140 g/m^3
Flocculation tank control pH	10.3
Ferric sulfate dosage	7.5 g/m^3
Soda ash dosage	1200 g/m^3
Flocculator speed	140 r/min
Rake speed	0.6 r/min
Sludge recycle rate	0.34 x 10^{-3} m^3/s
Clarifier sludge control level height (at 0.5 percent by volume maximum concentration)	0.6 m
Filter-effluent control pH set point	6.9

Water quality changes

	Raw water influent g/m³	Filter effluent g/m³		Raw water influent g/m³	Filter effluent g/m³
TDS	9507	9062	Carbonate	ND	43
Silica	40	2	Sulfate	1860	1970
Calcium	820	60	Chloride	3345	3350
Magnesium	1.0	93	Carbon dioxide	750	ND
Sodium	2220	2690	Boron	5.0	4.2
Potassium	182	188	Strontium	10.2	1.0
Bicarbonate	1266	ND*	Sulfide	4.9	ND

*None detected.

Table 20. — *Lime-soda ash softening steady-state operation operating data, June 17, 1980, through August 29, 1980*

		6/17/80 thru 6/30/80			6/30/80 thru 8/6/80[a,b]			8/6/80 thru 8/22/80[c]			8/22/80 thru 8/29/80[a]			SUMMARY 6/17/80 thru 8/29/80		
Aerated Train Influent	Units	Max	Min	Avg	Max	Min	Avg	Max	Min	Avg	Max	Min	Avg	Max	Min	Avg
Flow[d]	$m^3/s \times 10^{-3}$	0.6	0.6	0.6	0.6	0.6	0.6	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.6	0.8
pH		7.6	7.2	7.4	7.6	6.7	7.3	7.5	6.6	7.2	7.5	7.0	7.3	7.6	6.6	7.3
Calcium	g/m³	721	585	638	761	545	640	834	577	685	720	585	649	834	545	653
Hardness as $CaCO_3$	g/m³	2660	2060	2202	2400	1970	2186	2620	1960	2254	2600	2060	2258	2660	1960	2225
T-alk. as $CaCO_3$	g/m³	780	544	648	920	492	664	824	500	727	700	556	652	920	492	673
P-alk. as $CaCO_3$	g/m³	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Clarifier Effluent (before acid)																
Flow[d]	$m^3/s \times 10^{-3}$	0.6	0.6	0.6	0.6	0.6	0.6	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.6	0.8
pH		10.6	10.1	10.3	10.5	10.0	10.2	10.3	9.8	10.1	10.4	9.9	10.1	10.6	9.8	10.2
Calcium	g/m³	289	40	98	144	24	62	176	16	37	40	8	27	289	8	56
Hardness as $CaCO_3$	g/m³	1020	380	481	680	400	484	760	360	450	520	360	474	1020	360	472
T-alk. as $CaCO_3$	g/m³	100	64	72	120	56	78	188	72	115	176	88	131	188	56	99
P-alk. as $CaCO_3$	g/m³	60	34	41	56	20	41	88	36	57	92	40	65	92	20	51
Turbidity	JTU	260	130	188	310	150	197	320	150	225	350	190	261	350	130	218

[a] These two sets of data were collected under the same operating conditions. The data are presented separately due to variation between the averages for the two periods.

[b] Flow through the aeration tank was increased from 0.6×10^{-3} m³/s to 1.1×10^{-3} m³/s for the period of July 31 through August 2 in order to support Spiractor testing. This flow increase resulted in a decrease in the CO_2 removal, which in turn resulted in the higher concentrations during this period. It should be noted that the additional flow was not put through the pretreatment train.

[c] Train throughput was increased to 0.9×10^{-3} m³/s during this time to ensure that sufficient feedwater was available to support simultaneous operation of both the RO and ED systems.

[d] Flow rate when in service.

Table 21b. — *Lime-soda ash softening steady-state operation stabilized clarifier effluent composition*

Date		6/17/80	6/23/80	6/30/80	7/14/80	7/21/80	7/28/80	8/11/80	8/18/80	8/25/80	Summary		
Time Collected		0745	0815	0740	0830	1045	0700	0655	0710	0815			
Date analyzed	Units	7/15/80	7/15/80	7/31/80	7/29/80	8/21/80	9/5/80	9/5/80	9/16/80	10/21/80	Max	Min	Avg
pH*		5.2	7.3	6.9	6.7	6.7	6.8	6.4	6.5	6.5	7.3	5.2	6.0
TDS (calculated)¹	g/m³	8319	8256	8255	8287	8506	8687	8511	8532	8517	8687	8255	8430
TDS (evaporated @ 105 °C)*	g/m³	8296	8474	8400	8477	8710	8515	8500	8633	8679	8710	8296	8520
Conductivity @ 25 °C*	mS/m	1390	1380	1480	1400	1430	1380	1400	1420	1500	1500	1380	1420
Temperature*	°C	32.5	33.0	35.5	33.3	35.5	35.0	35.5	34.1	33.0	35.5	32.5	34.2
Silica*	g/m³	2.0	1.0	2.0	1.3	1.9	1.5	2.3	2.2	3.1	3.1	1.0	1.9
Calcium	g/m³	171	99	120	100	55.0	48	59.0	31.3	34.8	171	31.3	80
Magnesium	g/m³	19.0	66.1	56.7	68.1	84.0	78.2	64.4	87.7	102	102	19.0	69.6
Sodium	g/m³	2750	2690	2680	2700	2870	2970	2840	2860	2850	2970	2680	2801
Potassium	g/m³	171	177	183	182	167	174	176	173	176	183	167	175
Iron, total	g/m³	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	0.04	<0.04	<0.04	0.04	<0.04	-
Manganese, total	g/m³	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.02	<0.02	<0.02	0.30	<0.02	-
Strontium	g/m³	1.6	0.9	1.1	0.9	0.5	0.6	0.6	0.5	0.5	1.6	0.5	0.8
Bicarbonate*	g/m³	39	44	44	49	54	59	27.0	85	61	85	27	51
Carbonate*	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Sulfate	g/m³	1845	1850	1845	1865	1860	1955	1925	1930	1920	1955	1845	1888
Chloride	g/m³	3340	3350	3345	3345	3440	3430	3430	3405	3400	3440	3340	3387
T-alkalinity as $CaCO_3$*	g/m³	32	36	36	40	44	48	22.0	70	50	70	22	42
P-alkalinity as $CaCO_3$*	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Hydroxide*	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		6.0	6.0	5.6	5.9	5.9	6.3	6.1	6.0	5.7			
E.F. [TDS(evaporated)/cond.]		6.0	6.1	5.7	6.1	6.1	6.2	6.1	6.1	5.8			
Σ Anions	eq/m³	133.29	133.76	133.51	134.01	136.67	138.45	137.30	137.65	136.9			
Σ Cations	eq/m³	134.13	131.89	131.94	132.72	138.78	142.49	136.30	137.62	138.6			
Control value	eq/m³	-0.39	+0.86	+0.72	+0.60	-0.95	-1.79	+0.45	+0.01	-0.77			

*Analysis conducted at LVSTS

¹Calculated TDS uses only one-half of the bicarbonate

Table 22 — *Lime-soda ash softening steady-state operating data, filter 3 and 4*

Filter no		Filter 3						Filter 4					
		Summary 1			Summary 2			Summary 1			Summary 2		
From		6-16-80 to 8-1-80			8-6-80			6-19-80 to 8-6-80			8-9-80		
To		8-22-80 to 8-9-80			8-22-80			8-25-80 to 8-28-80			8-22-80		
Typical duration (hr)		72			72			72			72		
Number of filter runs		10			4			10			4		
Final wash water turbidity, JTU, low/high		0.4-13.0			-			2.4-12.0			1.5		
		Max	Min	Avg	Max	Min	Avg	Max	Min	Avg	Max	Min	Avg
Filter influent (pipeline header)	Units												
Flow *	$m^3/s \times 10^{-3}$	0.6	0.5	0.6	0.9	0.8	0.9	0.6	0.5	0.6	0.9	0.9	0.9
pH (approx.)		7.5	4.7	5.7	6.8	5.8	6.3	6.9	3.6	5.3	6.8	5.9	6.4
Turbidity	JTU	170	42	82	175	55	108	155	25	78	160	35	117
Calcium	mg/L	313	16	72	224	32	59	144	12	70	92	24	43
Hardness as $CaCO^3$	mg/L	1040	380	501	880	400	511	580	460	502	500	400	473
T. alk. as $CaCO^3$	mg/L	88	12	38	88	12	54	84	16	32	84	22	60
P. alk. as $CaCO^3$	mg/L	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Filter effluent													
pH		7.3	5.2	6.7	7.1	6.1	6.7	7.4	6.4	6.9	7.2	6.4	6.8
Turbidity	JTU	1.60	0	0.13	0.30	0	0.11	0.80	0	0.14	0.41	0	0.10
Plugging factor	%	62	1	13	45	1	14	34	1	14	32	1	8.8
Calcium	mg/L	305	24	77	232	32	60	144	24	72	96	24	48
Hardness as $CaCO^3$	mg/L	1000	380	508	880	400	513	580	460	510	520	420	484
T-alk. as $CaCO^3$	mg/L	100	20	48	100	24	62	92	20	44	104	24	67
P-alk. as $CaCO^3$	mg/L	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Iron	mg/L	-	-	-	-	-	-	-	-	-	-	-	-
Manganese	mg/L	-	-	-	-	-	-	-	-	-	-	-	-
Free Cl_2 residual	mg/L	3.3	ND	-	0.9	ND	-	4.8	ND	-	0.9	ND	-
Total Cl_2 residual	mg/L	3.5	0	1.1	4.1	0.1	1.3	5.2	ND	-	3.8	ND	-
Temperature	°C	38.0	30.0	34.1	39.0	31.5	34.6	38.0	30.0	34.1	38.0	31.0	34.2
Total head loss	m	0.28	0.14	0.21	0.36	0.28	0.31	0.23	0.16	0.20	0.32	0.24	0.27

*Flow rates when in service
ND = None detected

Table 23. — *Lime-soda ash softening steady-state operation raw and aerated influent composition, Phase II — June 17, 1980, through August 25, 1980*

		Summary (Raw)			Summary (Aerated)		
		Max	Min	Avg	Max	Min	Avg
pH[a]		6.2	6.0	6.1	7.5	6.8	7.3
TDS (calculated)[b]	g/m^3	9405	9235	9296	9128	8749	8892
TDS (evaporated @ 105 °C)[a]	g/m^3	9672	9334	9557	9356	8338	9052
Conductivity @ 25 °C[a]	mS/m	1540	1410	1483	1500	1400	1432
Temperature[a]	°C	41.0	39.0	40.1	38.5	35.5	37.2
Silica[a]	g/m^3	42.0	40.0	40.9	30.0	29.0	29.9
Calcium	g/m^3	820	770	807	770	590	657
Magnesium	g/m^3	145	140	142	144	141	143
Sodium	g/m^3	2280	2210	2240	2280	2220	2237
Potassium	g/m^3	184	169	176	185	165	175
Iron, total	g/m^3	0.29	<0.04	-	<0.10	<0.04	-
Manganese, total	g/m^3	0.02	<0.02	-	0.02	<0.02	-
Strontium	g/m^3	11.2	10.2	10.7	10.2	6.2	7.8
Bicarbonate[a]	g/m^3	1320	1266	1298	952	600	790
Carbonate[a]	g/m^3	ND	ND	ND	ND	ND	ND
Sulfate	g/m^3	1925	1835	1882	2000	1840	1885
Chloride	g/m^3	3375	3300	3344	3420	3330	3361
T-alkalinity as $CaCO_3$[a]	g/m^3	1082	1038	1064	780	492	648
P-alkalinity as $CaCO_3$[a]	g/m^3	ND	ND	ND	ND	ND	ND
Free CO_2[a]	g/m^3	748	590	678	58	16	31
Hydroxide[a]	g/m^3	ND	ND	ND	ND	ND	ND
Sulfides, total[a]	g/m^3	6.5	3.0	4.9	1.9	0.7	1.1

[a] Analysis conducted at LVSTS

[b] Calculated TDS uses only one-half of the bicarbonate

ND - None detected

Table 24. — *Lime-soda ash softening steady-state operation effluent composition, filter 3*

Date Time Collected Date analyzed at Yuma	Units	8/11/80 0655 9/5/80	8/18/80 0710 9/16/80	Summary Max	 Min	 Avg
pH[a]		6.8	6.7	6.8	6.7	6.8
TDS (calculated)[b]	g/m^3	8478	8509	8509	8478	8494
TDS (evaporated @ 105 °C)[a]	g/m^3	8403	8502	8502	8403	8452
Conductivity @ 25 °C[a]	mS/m	1400	1420	1420	1400	1410
Temperature[a]	°C	34.0	31.8	34.0	31.8	32.9
Silica[a]	g/m^3	2.1	2.2	2.2	2.1	2.15
Calcium	g/m^3	59.0	31.4	59.0	31.4	45.2
Magnesium	g/m^3	64.4	88.7	88.7	64.4	76.6
Sodium	g/m^3	2850	2850	2850	2850	2850
Potassium	g/m^3	177	171	177	171	174
Iron, total	g/m^3	<0.04	<0.04	<0.04	<0.04	-
Manganese, total	g/m^3	<0.02	<0.02	<0.02	<0.02	-
Strontium	g/m^3	0.6	0.5	0.6	0.5	0.55
Bicarbonate[a]	g/m^3	29.0	81	81	29.0	55
Carbonate[a]	g/m^3	ND	ND	ND	ND	ND
Sulfate	g/m^3	1900	1930	1930	1900	1915
Chloride	g/m^3	3410	3395	3410	3395	3402
T-alkalinity as $CaCO_3$[a]	g/m^3	24.0	66	66	24.0	45
P-alkalinity as $CaCO_3$[a]	g/m^3	ND	ND	ND	ND	ND
Hydroxide[a]	g/m^3	ND	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		6.1	6.0			
E.F. [TDS(evaporated)/cond.][a]		6.0	6.0			
Σ Anions	eq/m^3	136.25	137.30			
Σ Cations	eq/m^3	136.76	137.22			
Control value	eq/m^3	-0.23	+0.04			

[a] Analysis conducted at LVSTS.

[b] Calculated TDS uses only one-half of the bicarbonate.

ND - None detected.

Table 25. — *Lime-soda ash softening steady-state operation effluent composition, filter 4*

Date		6/17/80	6/23/80	6/30/80	7/14/80	7/21/80	7/28/80	8/25/80	Summary		
Time Collected		0745	0815	0740	0830	1045	0700	0815			
Date analyzed at Yuma	Units	7/15/80	7/15/80	7/21/80	7/29/80	8/12/80	9/5/80	10/21/80	Max	Min	Avg
pH*		6.5	6.9	6.8	6.6	6.6	6.8	6.8	6.9	6.5	6.7
TDS (calculated)[1]	g/m^3	8298	8241	8229	8240	8543	8648	8572	8648	8228	8396
TDS (evaporated @ 106 °C)*	g/m^3	8433	8514	8467	8470	8654	8532	8548	8654	8433	8517
Conductivity @ 25 °C*	mS/m	1380	1380	1460	1400	1430	1380	1490	1490	1380	1417
Temperature*	°C	33.0	33.0	34.0	30.0	35.0	34	30.8	35.0	30.0	32.8
Silica*	g/m^3	2.0	1.0	2.0	1.3	2.1	1.4	3.1	3.1	1.0	1.8
Calcium	g/m^3	169	97	119	98	51.0	47	34.9	169	34.9	88.0
Magnesium	g/m^3	19.9	67.1	55.9	67.4	89.0	77.3	102	102	19.9	68.4
Sodium	g/m^3	2750	2690	2680	2690	2880	2950	2860	2950	2680	2786
Potassium	g/m^3	173	175	178	180	167	172	173	180	167	174
Iron, total	g/m^3	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10	<0.04	<0.10	<0.04	-
Maganese, total	g/m^3	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.02	<0.30	<0.02	-
Strontium	g/m^3	1.6	1.0	1.1	0.9	0.5	0.6	0.5	1.6	0.5	0.9
Bicarbonate*	g/m^3	44	49	44	44	56	59	76	76	44	53
Carbonate*	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Sulfate	g/m^3	1825	1835	1830	1850	1875	1950	1960	1960	1825	1875
Chloride	g/m^3	3335	3350	3340	3330	3450	3420	3400	3450	3330	3375
T-alkalinity as $CaCO_3$*	g/m^3	36	40	36	36	46	48	62	62	36	43
P—alkalinity as $CaCO_3$*	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Hydroxide*	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		6.0	6.0	5.6	5.9	6.0	6.3	5.8			
E.F. [TDS(evaporated)/cond.]*		6.1	6.2	5.8	6.0	6.1	6.2	5.7			
Σ Anions	eq/m^3	132.82	133.53	133.06	133.20	137.30	138.06	138.0			
Σ Cations	eq/m^3	134.16	131.87	131.69	132.07	139.43	141.44	139.0			
Control value	eq/m^3	-0.62	+0.76	+0.63	+0.52	-0.95	-1.50	-0.44			

* Analysis conducted at LVSTS

[1] Calculated TDS uses only one-half of the bicarbonate

ND - Not detected

Table 26. — *Lime-soda ash optimization test operating condition, May 19, 1980, through June 11, 1980*

Dates	Test 1 5/19/80 to 5/20/80	Test 2 5/20/80 to 5/23/80	Test 3 5/23/80 to 5/28/80	Test 4 5/28/80 to 5/30/80	Test 5 5/31/80 to 6/5/80	Test 6 6/5/80 to 6/9/80	Test 7 6/9/80 to 6/10/80	Test 8 6/10/80 to 6/11/80
Flow rate (m^3/s x 10^{-3})	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
Ferric sulfate dosage (g/m^3)	7.5	7.5	7.5	7.5	7.5	7.5	7.5	7.5
Flocculation tank control pH	9.8	10.0	10.0	10.2	10.2	10.2	10.2	10.3
Soda ash dosage (g/m^3)	1160	1160	960	1100	1140	1160	1200	1200
Clarifier rake speed (rpm)	0.5	0.5	0.5	0.5	0.5	0.6	0.6	0.6
Flocculation speed (rpm)	140	140	140	140	140	140	140	140
Sludge recycle rate (m^3/s x 10^{-3})	0.17	0.17	0.17	0.34	0.34	0.34	0.34	0.34
Sludge control level* (height in meters at 0.5% by volume, maximum concentration)	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6

*Sludge wasting intermittent as required

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Table 27. — *Lime-soda ash pretreatment optimization test results*

		Test 1 5/19/80 - 5/20/80			Test 2 5/20/80 - 5/23/80			Test 3 5/23/80 - 5/28/80		
		Max	Min	Avg	Max	Min	Avg	Max	Min	Avg
Aerated Train Influent	Units									
Flow[a]	$m^3/s \times 10^{-3}$	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
pH (bench)		7.5	7.4	7.4	7.6	7.2	7.4	7.6	7.4	7.5
CO_2	g/m^3	11	1.7	5.6	13	2.0	6.0	50	0.9	8.7
Sulfides, total	g/m^3	1.4	1.0	1.2	1.6	1.0	1.3	2.5	0.5	1.3
Calcium	g/m^3	625	609	617	625	577	608	681	577	621
Hardness as $CaCO_3$	g/m^3	2200	2100	2133	2200	2080	2150	2240	2100	2153
T-alk. as $CaCO_3$	g/m^3	652	596	616	640	580	602	676	608	632
Temperature	°C	36.5	35.0	35.6	38.0	35.0	36.0	36.0	31.0	34.0
NFFlocculation Tank										
pH (in-line)		9.9	9.8	9.8	10.0	9.8	9.9	10.0	9.9	10.0
Solids conc.	vol %[b]	21	12	17	88	6.0	19	47	1.0	16
Clarifier										
0.3 m level	vol %[b]	4.0	2.0	2.7	6.0	0.2	2.5	11	0	3.8
0.6 m level	vol %[b]	0.2	0	0.2	0.2	0	0.1	0.2	0	0.1
Sludge Recycle										
Flow[a]	$m^3/s \times 10^{-3}$	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2
Sludge recycle solids	vol %[b]	96	93	95	98	77	90	97	62	88
Clarifier Effluent										
pH (bench)		9.8	9.7	9.8	10.1	9.4	9.8	10.2	9.9	10.0
Turbidity (in-line)	JTU	210	140	172	260	180	218	260	150	206
Calcium	g/m^3	64	16	40	32	16	24	88	24	69
Hardness as $CaCO_3$	g/m^3	600	540	560	520	460	500	660	420	621
T-alk. as $CaCO_3$	g/m^3	204	68	124	160	80	113	100	60	69
P-alk. as $CaCO_3$	g/m^3	88	40	59	72	40	53	56	28	37.6
Summary										
Calcium removal	g/m^3			577			584			552
Calcium removal as $CaCO_3$	g/m^3			1442			1460			1380
Hardness removal as $CaCO_3$	g/m^3			1573			1650			1532
T-alk. removal as $CaCO_3$	g/m^3			492			489			563
Excess lime present in clarifier effluent	g/m^3			12			None			5

[a] Flow rates when in service

[b] Determined after 15-minute settling time in a graduated cylinder

Table 27. — *Lime-soda ash pretreatment optimization test results — Continued*

		Test 4 5/28/80 - 5/30/80			Test 5 5/31/80 - 6/5/80			Test 6 6/5/80 - 6/9/80		
		Max	Min	Avg	Max	Min	Avg	Max	Min	Avg
Aerated Train Influent	**Units**									
Flow*	$m^3/s \times 10^{-3}$	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6	0.6
pH (bench)		7.6	7.3	7.4	7.6	7.3	7.4	7.5	7.3	7.4
CO_2	g/m^3	9.7	4.4	7.0	13	0.4	5.5	15	2.6	7.6
Sulfides, total	g/m^3	1.6	1.0	1.3	2.2	0.8	1.7	1.2	0.2	0.7
Calcium	g/m^3	641	601	624	640	537	599	625	569	599
Hardness as $CaCO_3$	g/m^3	2260	2100	2183	2160	2040	2109	2240	2080	2145
T-alk. as $CaCO_3$	g/m^3	640	620	629	640	560	597	680	548	596
Temperature	°C	36.0	34.5	35.2	37.0	34.0	34.8	38.0	31.2	35.1
Flocculation Tank										
pH (in-line)		10.2	10.2	10.2	10.3	10.1	10.2	10.2	10.1	10.1
Solids conc	vol %[b]	50	40	45	53	11	36	165	9.5	35.8
Clarifier										
0.3 m level	vol %[b]	79	9.0	46	80	1.0	38	11	0.5	4.7
0.6 m level	vol %[b]	30	0	2.9	21	0	2.4	Tr	0	Tr
Sludge Recycle										
Flow*	$m^3/s \times 10^{-3}$	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
Sludge recycle solids	vol %[b]	95	91	94	97	55	89	95	72	84
Clarifier Effluent										
pH (bench)		10.4	10.1	10.2	10.4	10.1	10.2	10.4	10.1	10.3
Turbidity (in-line)	JTU	285	195	231	255	135	192	250	140	179
Calcium	g/m^3	96	72	83	120	40	67	72	24	55
Hardness as $CaCO_3$	g/m^3	600	500	553	560	400	477	460	400	425
T-alk. as $CaCO_3$	g/m^3	76	60	65	80	64	71	100	72	80
P-alk. as $CaCO_3$	g/m^3	36	28	34	48	28	39	60	36	42
Summary										
Calcium removal	g/m^3			541			532			544
Calcium removal as $CaCO_3$	g/m^3			1352			1330			1360
Hardness removal as $CaCO_3$	g/m^3			1630			1632			1720
T-alk. removal as $CaCO_3$	g/m^3			564			526			516
Excess lime present in clarifier effluent	g/m^3			3			7			4

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*Flow rates when in service.

[b] Determined after 15-minute settling time in a graduated cylinder. Tr = trace amounts.

Table 27. — *Lime-soda ash pretreatment optimization test results — Continued*

		Test 7 6/9/80 - 6/10/80			Test 8 6/10/80 - 6/11/80		
		Max	Min	Avg	Max	Min	Avg
Aerated Train Influent	**Units**						
Flow[a]	$m^3/s \times 10^{-3}$	0.6	0.6	0.6	0.6	0.6	0.6
pH (bench)		7.5	7.4	7.4	7.4	7.4	7.4
CO_2	g/m^3	18	10	14	19	12	16
Sulfides, total	g/m^3	1.5	0.6	0.9	1.0	0.5	0.8
Calcium	g/m^3	673	641	652	681	625	646
Hardness as $CaCO_3$	g/m^3	2220	2200	2207	2300	2180	2240
T-alk. as $CaCO_3$	g/m^3	720	600	667	700	648	672
Temperature	°C	38.0	33.5	35.7	38.5	35.0	36.3
Flocculation Tank							
pH (in-line)		10.2	10.1	10.2	10.3	10.2	10.3
Solids conc.	vol %[b]	24	5.0	13	22	2.5	13
Clarifier							
0.3 m level	vol %[b]	4.0	0.5	1.7	2.0	0.5	1.2
0.6 m level	vol %[b]	0	0	0	0	0	0
Sludge Recycle							
Flow[a]	$m^3/s \times 10^{-3}$	0.3	0.3	0.3	0.3	0.3	0.3
Sludge recycle solids	vol %[b]	85	71	76	80	52	67
Clarifier Effluent							
pH (bench)		10.2	10.1	10.2	10.3	10.2	10.3
Turbidity (in-line)	JTU	180	150	171	210	150	178
Calcium	g/m^3	32	24	27	48	40	43
Hardness as $CaCO_3$	g/m^3	480	420	453	460	440	447
T-alk. as $CaCO_3$	g/m^3	104	76	87	80	72	76
P-alk. as $CaCO_3$	g/m^3	56	40	48	44	40	41
Summary							
Calcium removal	g/m^3			625			603
Calcium removal as $CaCO_3$	g/m^3			1562			1508
Hardness removal as $CaCO_3$	g/m^3			1754			1793
T-alk. removal as $CaCO_3$	g/m^3			580			596
Excess lime present in clarifier effluent	g/m^3			9			6

[a] Flow rates when in service

[b] Determined after 15-minute settling time in a graduated cyclinder

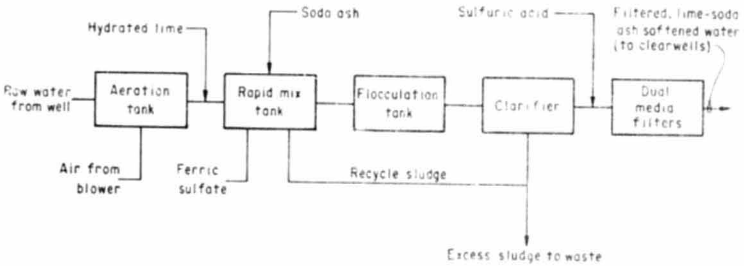
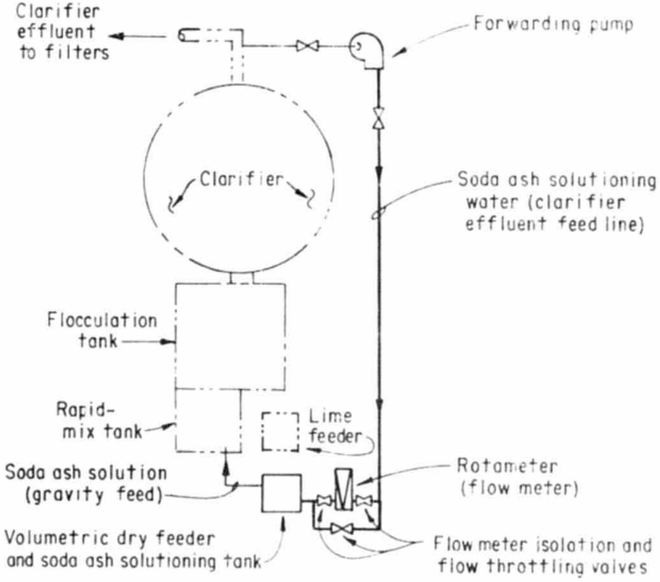


Figure 23.—Lime–soda ash softening pretreatment system flow diagram



Not to scale

Figure 24.—Lime–soda ash softening pretreatment equipment configuration

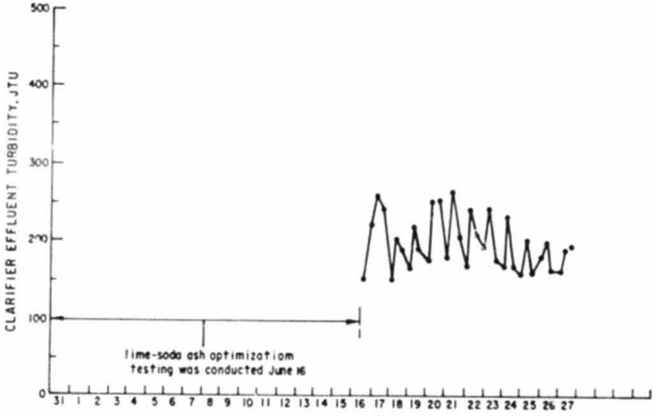


Figure 25a.—Clarifier effluent turbidity, train L-I, lime–soda ash softening steady-state operation (May 31 through June 27, 1980).

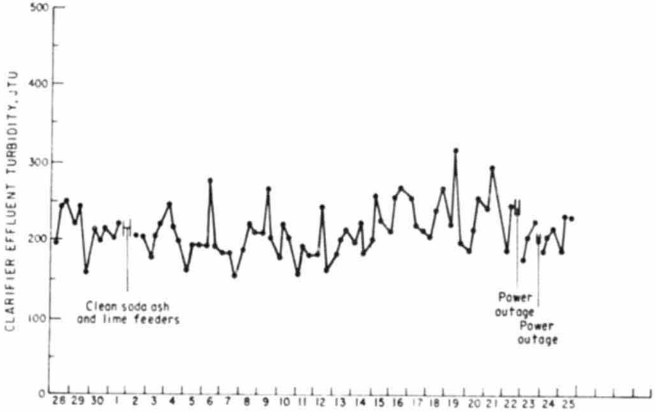


Figure 25b.—Clarifier effluent turbidity, train L-I, lime–soda ash softening steady-state operation (June 28 through July 25, 1980).

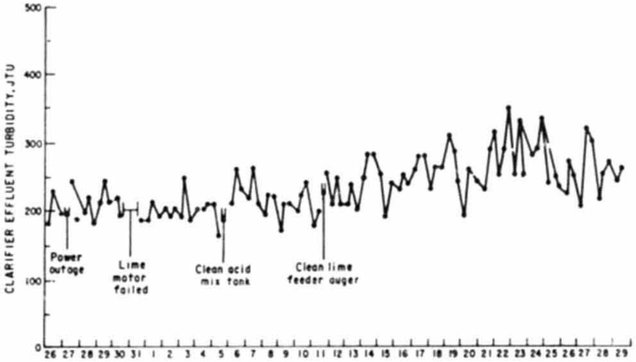


Figure 25c.—Clarifier effluent turbidity, train L-I, lime—soda ash softening steady-state operation (July 26 through August 29, 1980).

EXPERIMENTAL LIME-ION EXCHANGE PROCESS STUDY

In this section, results of the A-L-IX (aeration-lime softening-ion exchange softening) process study are presented. Equipment, process operations, results, and optimum conditions are discussed. A separate report titled "Cation-Exchange Pretreatment Studies for LaVerkin Springs" has been written [4].

Aeration and lime softening are discussed in earlier sections. Aeration of LVS raw water reduces dissolved calcium from 820 mg/L to 600 mg/L and bicarbonate from 1266 mg/L to 795 mg/L. Subsequent additions of lime (calcium hydroxide) to increase water pH from 7.4 to 9.5 neutralizes the remaining bicarbonate and removes 180 mg/L of calcium. Consequently, 450 mg/L of calcium carbonate precipitates and 420 mg/L of calcium remains in solution after A-L treatment. The calcium remaining in solution as highly soluble chloride salt can be effectively removed from solution by cation-ion exchange.

A. Equipment Description

The A-L-IX water softening system consisted of:

- Aeration unit
- Lime treatment train
  - Lime feeder
  - Rapid mix tank
  - Flocculation tank
  - Clarifier
  - Dual Media Filters
  - Clearwell
- Ion exchange unit

The aeration unit and the lime treatment train have been discussed. The IX unit equipment was developed, specified, designed, constructed, and operated during the LVS site test program.

1. *Preliminary design.* — Preliminary design of the mobile IXPP (ion exchange pilot plant) was done by the USBR (Bureau of Reclamation), E&R (Engineering and Research) Center, Division of Design, S&A (Structural and Architectural) personnel. A facsimile of S&A diagram and details for design of the mobile IXPP are given as figures G.10 and G.11.

2. *Intermediate E&R Center design and construction.* — Modifications of the mobile IXPP design was done by USBR Division of Research chemical engineering personnel. A facsimile of the EP&ID (equipment, piping, and instrument diagrams) for mobile IXPP-I is attached as figure G.12. Interconnecting process units of the IXPP mobile IXPP were constructed by Division of Research chemical engineers at the USBR's desalination test station at Longmont, Colorado (fig. G.13). Construction was completed on November 23, 1977. At that time, a decision was made to defer the LVS site test until mid-1979. In the interim, the IXPP was utilized by Kaakinen and Laverty [9] for the YHR (Yuma High Recovery) test program at the YDTF (Yuma Desalting Test Facility) in Yuma, Arizona. For the YHR test, Denver location chemical engineers developed a layout drawing for the IXPP total system to be set up at YDTF. A drawing "Suggested YHR Recovery Layouts and Interconnecting Pipe at the YDTF," is included as figure G.14.

3. *Final YDTF design and modifications.* — The mobile IXPP and associated equipment shown on figure G.12 were trucked to YDTF in December 1977. The IX equipment was set up for the YHR test program according to G.15 and G.16. Initial operations during the spring of 1978 revealed a major discrepancy in the PVC solenoid valves that controlled waterflow patterns for dual IX columns operations. Normally closed plastic flow directional control valves were opened by a microprocessor control system energizing a 40-watt solenoid. When valves were energized for longer than 15 minutes, the heat generated by the solenoid coil would conduct down to the PVC valve body. After repeated cycles, the valve bodies would distort and no longer seat properly. Another fault of the plastic solenoid valve was a water hammer produced in associated pipe when de-energized. Vibrations in the pipe system caused leaks to develop at threaded pipe joints. Both problems were eliminated by replacing most of the solenoid valves with motorized ball valves and cementing PVC pipe joints. Purchase cost of the motorized ball valves was competitive with purchase cost of solenoid valves. Motorized ball valves are also much more energy efficient than solenoid valves. When energized to open or to close positions, full flow ball valves are given a one-quarter turn, and then the motor shuts itself off. Changes made in the flow control valves mode at the YDTF are shown in figure G.4.

4. *Test site IXPP layout.* — The IXPP and associated equipment were shipped from the YDTF, Yuma, Arizona, to the LVS test site, Hurricane, Utah, during October 1979. The equipment was installed according to figures G.15 and G.16.

5. *Specific LVS test site equipment description and modification.* — The IXPP consisted of two IX columns with a nominal capacity of 3.8×10^{-4} m^3/s each, along with storage tanks, pumps, and interconnecting valves and piping. Reject brine was supplied by an Ionics Aquamite V ED (electrodialysis) unit. The ED, in turn, was supplied by the IX

softened water. A microprocessor was used to control operations. Figures 26 and 27 present a process flow diagram and piping diagram, respectively. Each of the two 0.34-meter-diameter columns were filled to a depth of approximately 1.12 meters with Dowex HCR-W-2 cation exchange resin.

Both columns 1 and 2 were modified from the configuration utilized during the YHR program by the addition of a midport distributor plate between the upper and lower sections of each column. This was used as an effluent point during packed-bed regeneration (i.e., resin hold-down) and during concurrent upflow for down-rinse. The midport distributor was fabricated according to figure 28 from a 1-inch-thick acrylic plate with a ½-inch-diameter, PVC-perforated pipe cross. This was connected to the acrylic flange by four unions that allowed assembly and disassembly of the cross. The cross was covered with a fine-mesh, plastic screen to prevent the resin from being washed out of the column. The flange had four exit points (one from each end of the cross) that were connected into the IX system discharge piping.

A heat exchanger was used to control the temperature of the regenerants during task IX-II testing. Bypass raw water was used in the jacket side of the heat exchanger as temperature-control medium. Only column 1 was used with the heat exchanger. The heat exchanger had to be located close to the column to minimize heat loss from the process piping.

B. Process Operations

1. *Proof of concept test.* — The general concept of using concentrated reject brine from a desalting process to regenerate exhausted IX resin has been claimed by Bresler [10] and demonstrated by Haugseth and Beitelshees [11]. They demonstrated the feasibility of softening of brackish well water with a sodium ion-substituted strong acid IX resin. They showed reject brine must contain at least 9000 mg/L of sodium and have a volume of approximately twice the resin volume for effective regeneration. This test is the first application of this concept by the Bureau and, thus, is called the "proof of concept test." Raw water with a sodium to total hardness weight ratio greater than 2.0 was favorable for producing a good reject brine regenerant. Regeneration by a reject brine with 9000 mg/L of sodium recovered about 50 percent of the calcium capacity of the resin, while 20 000 mg/L or a stoichiometric amount of sodium regenerated 85 percent of exchange sites. A 90 percent calcium loading was demonstrated after regeneration with a reject brine containing 35 000 mg/L of sodium ion.

The nature and amounts of dissolved salts in LVS water are favorable for pretreatment by IX softening and regeneration with the reject brine. Sodium total hardness ratio is 2.6, and since sodium content of the raw water is 2530 mg/L, reject brine from membrane desalting at 80 percent water recovery would contain 12 650 mg/L. Reject brine from 90 percent water recovery could contain 25 300 mg/L; consequently, the high sodium content of reject brines from LVS water would be viable for a pretreatment process. At greater than 80 percent water recovery, the volume of reject brine would be too low for one-pass regeneration. Additional process steps can be taken to compensate for low reject brine volume. High sodium content should permit dilution with raw water. Another process optimization would be to accumulate spent regenerant from subsequent IX cycles removing calcium sulfate precipitate and recycling as a regenerant.

2. *Preliminary offsite process demonstration.* — The concept of removal of calcium ions by IX with subsequent regeneration of resin with waste brine recovered from water desalting was first tested on actual LVS water by Eisenhauer and Schiller [12]. IX softening as a practical pretreatment method for membrane desalting of water was conducted at the Bureau Testing Desalinization Test Station, Longmont, Colorado, from July 1976 to October 1976. Water was trucked in from LaVerkin Springs, Utah, for use at the test station. The concept demonstrated was one of reducing calcium in raw LVS water by IX below sulfate fouling concentration levels, desalting by RO, and using the recovered reject brine to regenerate exhausted IX resin. Test results demonstrated IX softening of LVS water can be operated as a closed-loop exhaustion-regeneration cycle. Only reject brine from RO needs to be used to regenerate the IX resin. As a result of this study, a process development test program was recommended for the LVS site to determine optimum process conditions, cost, and design criteria for IX pretreatment.

3. *The YHR (Yuma high recovery) process demonstration.* — The concept of removal of calcium ion by IX with subsequent resin regeneration by waste desalting brine was also evaluated at a Bureau test site by Kaakinen and Laverty [9]. Their pilot plant study of cation exchange softening was carried out at the YDTF (Yuma Desalting Test Facility) in 1978 and 1979 as a YHR project in support of the YDP (Yuma Desalting Plant). The YDP, under design by the Bureau, will produce 95 Mgal/d of low-salt water by 70 percent water recovery RO desalting. At 70 percent water recovery, 40 Mgal/d of waste (reject) brine would be produced. The disposal of the reject brine, usually in evaporation ponds or by deep well injection, can be one of the most expensive features of a desalting project. One method under consideration for reducing the amount of reject brine requiring disposal is to further desalt the brine produced by 70 percent water recovery desalting by electrodialysis. Since the reject brine is saturated with calcium sulfate, calcium must be removed prior to feeding the brine to electrodialysis. The YHR test demonstrated IX with resin regeneration by reject brine is a viable softening method.

4. *Test site process studies.* — The IX system operation at the LVS site was initiated on January 7, 1980. This initial operation, using sodium chloride brine, was to provide feed water for electrodialysis system startup and checkout. Upon completion of that task, the IX test program was begun.

During the course of the testing, several runs were found to be nonviable; all 90 percent recovery runs attempted showed a tendency to scale during regeneration. This scaling varied from light to extreme; several runs were discarded because severe scaling prevented their completion.

In an attempt to overcome the scaling problems encountered in the testing, 10 additional runs were added to the original set. Several parameters were changed or added for these additional runs. The maximum flow rate was increased to 36 L/min from 24 L/min. The volume of RR (recycle regenerant) was decreased to 800 liters for several of the runs. Runs L 23.00 and L 24.00 were conducted with SHMP (sodium hexametaphosphate) added to the regenerant to produce a final residual concentration of 100 g/m^3.

Table 28 lists the original test program runs (as revised on February 15, 1980), plus the additional 10 runs (runs L 17.00 through L 26.00) that were added by USBR, effective March 15, 1980. A total of 18 runs was attempted; only 16 were completed. Runs L 11.00 and L 21.00 could not be completed because of severe scaling of the midport distributor. Table 29 lists the completed runs with the data of completion. The LVS-IX program was completed on April 21, 1980.

5. *The IX operation in support to other testing.* —Upon completion of the LVS IX Test Program, the IX system was operated at various times throughout the remainder of the LVS site operations in support of various other tests. These included the special electrodialysis evaluation testing, reverse osmosis testing, and spiractor testing. No IX data cycles were conducted during any of this testing. Only control sampling and analyses were made of the IX operation during these tests.

C. Results and Discussion

The ion exchange operation had the following objectives:

Demonstrate that IX is possible for removal of residual calcium in lime-treated water to meet feed water specifications for membrane desalting;

Optimize the IX process by performing a predesigned statistical factorial experiment;

Demonstrate the viability of regeneration of IX resin with brine from 90 percent water recovery membrane desalting; and

Operate for 3 months on different cycle conditions to obtain data for plant design and for calculations of O&M and capital cost.

The following experimental IX tasks were evaluated:

Task IX-I. — installation and shakedown of the IXPP system;

Task IX-II. — statistical factorial designed experiment to optimize the IX process;

Task IX-III. — comparison of process performance of a Dow gel resin and a Rohn and Haas Amberlite resin (omitted because of insufficient test time), and

Task IX-IV. — operation of IXPP at the best condition determined by IX-II to provide feed water for the desupersaturation test.

1. *Task IX-I — Installation and startup of the IXPP system.* — Installation of the IX system began on November 26, 1979, and continued through the week of December 17, 1979.

a. The IXPP system was reinstalled at the LVS site according to figures G 15 and G 16, with the following changes:

- A 1-inch-thick, 14-inch-diameter open section flange was installed below the bottom flow distribution plate of the dual columns to improve the flow distributions.
- Each column was charged with 100 liters of Dowex gel resin HCR-W2, instead of the Rohm and Haas Amberlite 200 which was used at YDTF.
- A resin holddown purge flange was installed between the upper and lower sections of the dual columns (fig. 28).

b. The system was hydraulically checked and leaks eliminated using only city water and by operating through the following cycle:

Mode	Input (city water)	Output	Duration (min)	Avg flow (L/min)
Regen 1	Re regen	Waste	10	24
Regen 2	Re regen	Sp. regen	10	24
Regen 3	Fr regen	Sp. regen	10	24
Drain 1	(Vent)	Waste	3	20
Rinse	Feed	Waste	10	12
Service	Feed	Product	10	30
Drain 2	(Vent)	Waste	2	20

c. The system was operated through the following cycle during the initial startup and parametric testing of the ED. These cycles were continued to make an initial inventory of 1600 liters of 80 percent recovery fresh brine.

Mode	Input	Output	Duration (min)	Avg flow (L/min)
Backwash	Lime-treated water	Waste	10	[2]
Regen 3	10 percent sodium chloride	Waste	[1]	24
Drain 1	(Vent)	Waste	3	-
Rinse	Lime-treated water	Waste	10	15
Service	Lime-treated water	Product	120[3]	30
Drain 2	(Vent)	Waste	2	-

---

[1] Used established makeup volume of 10 percent sodium chloride solution.

[2] Used flow to maintain 50 percent bed expansion.

[3] Break through 40 mg/L of calcium (depending on pH control point for lime treatment, calcium in feed averaged 400 to 500 mg/L).

2. *Task IX-II — Study of the significance of variables on run length.* — Task IX-II was a research-level statistically designed experiment (table 30) to determine the significant effects of all the variable IX process conditions on the optimum performance of the process and on the amount of product water produced each cycle. Because of operational difficulties and time constraints, this experiment was modified during the course of the testing. The final experiment consisted of six of the originally designed runs plus the 10 additional runs summarized in table 28. The best process conditions were used to estimate optimum process conditions, equipment capital cost, and O&M cost.

a. Best run condition. — From the standpoint of trouble-free operation and maximum product water throughput, run condition L25 was the best. Test conditions and results from run condition L25 and cycle L25.22 are summarized in table 31 through table 37:

- Table 31 — The IX-run L.25.00 conditions
- Table 32 — Chemical compositions of tank water before cycle 25.22
- Table 33 — Operating conditions for cycle 25.22
- Table 34 — Fresh regenerant volume balance (ratio)
- Table 35 — Influent and effluent compositions
- Table 36 — Major cation concentration
- Table 37 — Service performance summary

Major cation breakthrough concentrations for regeneration effluents and service effluents are shown in figure 29.

3. *Task IX-III — Testing of Rohm and Haas Amberlite 200.* — Initially, tasks IX-II and IX-III were proposed to perform the 18 run conditions and cycle order shown in table 38. Cycles No. 1, 2, 5, and 6 were completed in orders No. 1, 2, 3, and 4 for x_2 values of 80 percent water recovery brine. Then, cycle condition 11 was tested. Heavy calcium sulfate scaling occurred during cycle 11 regeneration at the x_2 value of 90 percent later recovery brine. Consequently, a decision was made to complete the IX testing according to the cycle conditions and cycle order shown in table 30. This decision was made for the following reasons:

Not enough time was left to complete the initially scheduled cycles;

For a desalting plant to be feasible for the LVS site, it is necessary to demonstrate 90 percent water recovery;

The feasibility of desalting at 80 percent water recovery was demonstrated by the first four cycle conditions tested, and there would be little to gain by completing the other four 80 percent cycles; and

Testing the old Rohm and Haas Amberlite 200 resin against the new Dowex gel HCR-W2 was no longer necessary. When Amberlite 200 was used in the YHR test, it became tightly cemented to the resin by calcium sulfate and took a long time to rinse free. It was observed during cycle condition 11 at LVS that calcium sulfate scale did not cement to the Dowex gel resin.

4. *Operation of IX at best conditions.* — Task IX-IV called for "calcium polishing" of lime-soda ash softened water to provide low calcium feed water for RO. For task IX-IV, the IXPP was operated at the best conditions determined by task IX-II data cycle L25.22 (tables 31, 32, 33, 34, 35, 36, and 37). The product water quality is summarized in column (9) of table 3 (earlier in report). The IX polishing reduced calcium of the A-L-SA effluent from 60 to 6 mg/L and magnesium from 93 to 46 mg/L.

Table 28. — *Ion exchange test program parameters versus breakthrough bed volumes*

Run Number[b]	Fresh Regenerant		Recycle Regenerant[a]		Resin	Heat	BV[i]
	Nominal Recovery (%)	Flow Rate ($m^3/s \times 10^{-6}$)	Flow Rate ($m^3/s \times 10^{-6}$)	Volume (m^3)	Bed Hold Down	Exchanger	Calcium 6 meq/L
L.01.00[c]	80	200	0	0	NO	YES	30
L.02.00[c]	80	200	400	1.6	NO	YES	36
L.03.00[c]	80	200	0	0	YES	YES	34
L.04.00[c]	80	200	400	1.6	YES	YES	41
L.05.00[c]	90	400	0	0	NO	YES	21
L.06.00	90	200	0	0	NO	YES	*h*
L.07.00	90	200	0	0	YES	YES	*h*
L.08.00	90	200	400	1.6	NO	YES	*h*
L.09.00	90	200	400	1.6	YES	YES	*h*
L.10.00[c]	90	400	400	1.6	NO	YES	48
L.11.00[d]	90	400	0	0	YES	YES	*h*
L.12.00[c]	90	400	400	1.6	YES	YES	50
L.13.00	80	400	0	0	NO	YES	*h*
L.14.00	80	400	400	1.6	NO	YES	*h*
L.15.00	80	400	0	0	YES	YES	*h*
L.16.00	80	400	400	1.6	YES	YES	*h*
L.17.00[c]	90	600	600	1.6	YES	YES	42
L.18.00[c]	90	400	400	1.6	YES	NO	51
L.19.00[c]	90	600	600	1.6	YES	NO	48
L.20.00[c]	90	400	400	1.6	NO	NO	50
L.21.00[d]	90	400	0	0	YES	NO	*h*
L.22.00[c]	90	600	600	0.8	YES	NO	47
L.23.00[c,e] (SHMP)	90	400	0	0	YES	NO	46
L.24.00[c,e] (SHMP)	90	400	0	0	YES	YES	39
L.25.00[c,f]	90	600	600	0.8	YES	NC	53
L.26.00[c,g]	90	600	600	0.8	YES	NO	53

Table 28 Footnotes

[a] A backwash mode at 400 m^3/s for 10 minutes for all runs was utilized as follows
1) A feedwater backwash was used for those runs which did not utilize a recycle regeneration mode
2) A recycle regenerant backwash was used in conjunction with all runs which did utilize a recycle regeneration

[b] The calcium breakthrough concentration was as follows
1) 12.0 eq/m^3 for 80% Recovery and
2) 6.0 eq/m^3 for 90% Recovery

[c] These sixteen runs were completed. All othar run conditions shown on this table were not completed

[d] This run condition could not be tested to completion because of severe scaling of the midport distributor

[e] SHMP (sodium hexametaphosphate) was injected into the fresh regenerant so as to have a final concentration of 100 g/m^3

[f] This run utilized an up-flow rinse rather than the normal down-flow rinse. The rinse effluent was removed through the midport distributor

[g] This run utilized a concurrent up-flow/down-flow rinse. The effluent from this mode was removed through the midport distributor

[h] Data run cycle could not be achieved due to flow through column being inhibited by calcium sulfate scale formation

[i] One BV (bed volume) = 0.1 m^3 Dow Gel Resin HCR-W2

Table 29. — *Completed ion-exchange test runs*

Run Order	Run Number of Data Cycle	Date Completed
1	L.01.97[a]	1/29/80
2	L.02.12[a]	2/1/80
3	L.03.35[a]	2/14/80
4	L.04.11[a]	2/17/80
5	L.05.54	2/20/80
6	L.10.14	3/ 8/80
7	L.12.22	3/14/80
8	L.17.17	3/18/80
9	L.18.13	3/21/80
10	L.19.27	3/27/80
11	L.20.09	3/29/80
12	L.22.15	4/ 4/80
13	L.25.22[b]	4/ 9/80
14	L.26.14[c]	4/12/80
15	L.23.19[d]	4/16/80
16	L.24.11[d]	4/18/80

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[a] This run was conducted at a nominal recovery of 80%. All other runs were conducted at 90% recovery

[b] This run utilized an up-flow rinse followed by a down-flow rinse

[c] This run utilized a concurrent up-flow/down-flow rinse

[d] SHMP was injected into the fresh regenerant so as to have a final concentration of 100 g/m^3

Table 30. — *Operation of IX-PP on a 2^4 factorial designed experiment as task LVS-IX-II*

Constants:

C_1 = Backwash (feed water): Omitted when X_1 was at (+) level
= When X_2 was at (-) level, 24 L/min for 10 min
C_2 = Regenerant 2 (spent brine)[1]: 24 L/min
C_3 = Regenerant 3 (fresh brine): (80 percent recovery) (f)
: (90 percent recovery
C_4 = Drain 1: 3 minutes
C_5 = Rinse (feed water): 15 L/min for 2 min
C_6 = Drain 2: 2 minutes
C_7 = Recycle brine temperature: 27 °C
C_8 = Breakthrough: calcium (meq/L) for 80 percent = 12; for 90 percent = 6
C_9 = Feed flow rate: 30 L/min

Variables	Low (-)	High(+)	Responses
X_1 = Regenerant 2 recycle volume, L	0	1600	Calcium
X_2 = Brine regenerant concentration, percent water recovery	80	90	Magnesium Service time
X_3 = Resin hold down	No	Yes	
X_4 = Fresh regenerant flow rate, L/min	12	24	

Run order	PRC data cycle No.	Variables x1 Recycle reg. (L)	x1 Recycle reg. (L/min)	x2 Recovery (%)	x3 Resin hold down	x4 Fresh reg. (L/min)
1	L01.97 (a)	0	0	80	No	12
2	L02.12 (a)	1600	24	80	No	12
3	L03.35 (b)	0	0	80	Yes	12
4	L04.11 (b)	1600	24	80	Yes	12
5	L05.54 (b)	0	0	90	No	24
6	L11.19 (b)	0	0	90	Yes	24
7	L10.14 (b)	1600	24	90	No	24
8	L12.22 (b)	1600	24	90	Yes	24
9	L17.17 (b)	1600	33	90	Yes	33
10	L18.13 (a)	1600	24	90	Yes	24
11	L19.27 (a)	1600	33	90	Yes	33
12	L20.09 (a)	1600	24	90	No	24
13	L21.10 (a)	0	0	90	Yes	24
14	L22.15 (a)	800	33	90	Yes	33
15	L25.22 (a) (c)	800	33	90	Yes	33
16	L26.14 (a) (e)	800	33	90	Yes	33
17	L23.19 (a) (d)	0	0	90	Yes	24
18	L24.11 (b) (d)	0	0	90	Yes	24

Note
(a) Regenerant at ambient temperature
(b) Regenerant heated to approximately 27 °C
(c) After regeneration backwash at 33 L/min for 7 minutes
(d) 100 mg/L of SHMP added to regenerant
(e) Same as (c) with down rinse at the same time
(f) FRV (fresh regenerant volume) = [SER V (service volume)] [1 - REC (recovery)]

Table 31. — *Ion-exchange, run L.25.00*

Date: 4/9/80
Cycle: L.25.22
Conditions: Feedwater: LaVerkin Springs lime-softened and pretreated
Regenerants: recycled regenerant and fresh ED brine

		Target	Actual
Control variables:	Fresh regeneration conc. (mg/L TDS-calc.)	—	95 160
	Fresh regeneration conc. (mg/L TDS-evap.)	—	85 900
	Fresh regeneration flow rate (L/min)	33.0	30.6
	Recycled regenerant flow rate (L/min)	33.0	33.4
	Recycled regenerant volume (L)	800	791
	Service termination point (meg/L Ca++)	6.0	7.0
	Heat exchanger used	No	No
	Packed bed regeneration mode used (i.e., resin hold-down)	Yes	Yes

Standard resin bed: Height* = 1 410mm
Volume = 115.0 L

*The resin bed height at the end of drain-down after Regen 1 was used as the standard resin height in calculating bed expansion.

Table 32. — *Chemical compositions of tank waters (prior to cycle) L.25.22*

Tank	pH	Conductivity (mS/m)	Ca^{++} (meq/L)	Mg^{++} (meq/L)	TH (meq/L)
Recycle regenerant (T-5)	7.3	9080	41.0	94.0	135.0
Spent regenerant (T-6)	7.2	9010	42.0	89.0	131.0
Lime-softened feed (T-9)	7.1	1370	21.6	9.8	31.4
Lime-softened feed (T-10)	7.2	1380	21.6	9.8	31.4
Fresh ED brine (T-28)	6.2	9800	13.2	41.2	54.4
IX product/ED feed (T-33)	6.9	1490	1.6	4.4	6.0
Lime-softened feed (clearwell)	7.2	1370	21.4	10.2	31.6

Table 33. *Operating conditions, cycle L.25.22*

Mode	Input	Output	Duration (min)	Throughput (L)	Volume (BV)	Avg Flow Rate (L/min)	Avg Flow Rate BV/min	Bed expansion (%)	Temperature (°C)
Regen 1	Re regen	Waste	10	247	2.15	24.7	0.215	33.	19.1
Drain 1	(Vent)	Waste	2	58	0.50	29.0	.252	0.0	—
Regen 2	Re regen	Sp regen	24	791	6.88	33.1	.288	0.4	17.4
Regen 3	Fr regen	Sp regan	19	575	5.00	30.6	.266	4.6	19.4
Up rinse	Feed	Waste	12	385	3.35	32.1	.279	0.0	—
Dwn rinse	Feed	Waste	20	300	2.61	15.0	.130	0.0	—
Service	Feed	Product	193	5800	50.4	30.1	.261	0.0	—
Drain 2	(Vent)	Waste	2	30	0.26	15.0	.130	0.0	—

Table 34. — *Fresh regenerant volume balance, cycle L.25.22*

Run L.25.00 cycle no.	Date	Fresh regenerant volume (V_3) (L)	Fresh regenerant TDS (mg/L)	Service volume (V_s) (L)	ED feed TDS (mg/L)	R (%)	$V^3/(1-R)V_s$
12	4/7/80	495	—	5310	—	90	0.93
13	4/7/80	500	—	5990	—	90	0.83
14	4/7/80	497	—	5530	—	90	0.90
15	4/8/80	499	—	5740	—	90	0.87
16	4/8/80	570	—	5790	—	90	0.98
17	4/8/80	570	—	5630	—	90	1.01
18	4/8/80	117	—	4980	—	90	0.23
19	4/8/80	573	—	5580	—	90	1.03
20	4/9/80	573	—	5430	—	90	1.06
21	4/9/80	570	—	5670	—	90	1.01
22	4/9/80	575	—	5800	—	90	0.99

Table 35. — *Influent and effluent compositions during IX regenerations and service, cycle L.25.22*
(All samples composites except regenerations influent)
(Analysis conducted at Denver on 6/20/80)

	Units	Regen 1, 2 influent	Regen 1 effluent	Regen 2 effluent	Regen 3 Influent	Regen 3 Effluent	Rinse & Service influent	Service effluent
pH		—	—	—	7.1	—	7.5	7.3
TDS (calculated)	mg/L	74 420	38 110	73 700	96 160	94 660	8 530	8 900
TDS (evaporated @ 105°C)	mg/L	77 400	47 100	75 500	85 900	88 000	8 630	8 440
Conductivity @ 25°C	mS/m	—	—	—	9 220	—	1 270	1 350
Silica	mg/L	0.5	0.5	0.5	6.3	0.5	10.7	11.5
Calcium	mg/L	800	3 070	2 530	800	1 120	432	48.0
Magnesium	mg/L	1 200	1 170	1 390	1 320*	781	129	56.6
Sodium	mg/L	23 800	8 860	22 300	30 600	30 100	2 270	2 940
Potassium	mg/L	1 810	778	1 570	1 990	2 150	210	271
Iron, total	mg/L	ND	ND	ND	ND	ND	ND	ND
Manganese, total	mg/L	ND	ND	ND	ND	ND	ND	ND
Strontium	mg/L	27.3	37.8	38.0	—	25.7	4.1	0.3
Bicarbonate	mg/L	11.0	11.0	11.0	42.1	11.0	34.8	34.8
Carbonate	mg/L	0	0	0	0	0	0	0
Hydroxide	mg/L	—	—	—	—	—	—	—
Sulfate	mg/L	13 000	7 820	12 800	27 200	27 100	1 820	1 920
Chloride	mg/L	33 800	16 400	33 100	33 200	33 400	3 620	3 620
T-alkalinity as $CaCO_3$	mg/L	—	—	—	—	—	—	—
P-alkalinity as $CaCO_3$	mg/L	—	—	—	—	—	—	—
T-phosphorus as PO_4	mg/L	—	—	—	—	—	—	—
E.F. [TDS(calculated)/cond.]		—	—	—	10.3	—	6.7	6.6
E.F. [TDS(evaporated)/cond.]		—	—	—	9.3	—	6.8	6.3
Σ Anions	meq/L	1 222.18	625.18	1 200.18	1 502.69	1 507.18	140.47	142.57
Σ Cations	meq/L	1 214.70	654.90	1 248.30	1 528.90	1 485.10	136.17	141.98
Control value	meq/L	+0.39	-3.03	-2.57	-1.12	+0.94	+1.88	+0.25

*Questionable chemical analysis result

Table 36. — *Major cation concentrations of samples analyzed by atomic absorption, cycle L.25.22*

Mode	Process stream	Throughput (BV)	Ca (meq/L)	Mg (meq/L)	Th (meq/L)	Na (meq/L)
Regen 1	Effluent[a]	0.00	37.92	19.75	57.68	300.13
Regen 1	Effluent[a]	1.29	234.53	156.38	390.91	1000.43
Regen 2	Effluent[a]	2.15	2[illegible]4.57	172.84	387.41	1100.48
Regen 2	Influent[a]	3.87	[illegible]9.92	98.77	138.69	1035.23
Regen 2	Effluent[a]	5.60	99.80	98.77	198.57	1300.57
Regen 3	Effluent[a]	9.05	69.86	90.53	160.40	1300.57
Regen 3	Influent	9.59	39.92	108.64	148.56	1331.01
Regen 3	Effluent[a]	10.12	69.86	69.96	139.82	1500.65
Regen 3	Effluent[a]	11.18	49.90	57.61	107.51	1500.65
Regen 3	Effluent[a]	12.25	43.91	55.97	99.88	1500.65
Regen 3	Effluent[a]	14.11	41.92	50.21	92.12	1500.65
UF Rinse[c]	Effluent	1.67[b]	3.39	4.20	7.59	190.08
UF Rinse[c]	Effluent	3.35	2.20	.60	2.80	160.07
DF Rinse[d]	Effluent	3.35	14.47	9.05	23.52	140.06
DF Rinse[d]	Effluent	4.65	2.20	2.39	4.58	160.07
DF Rinse[d]	Effluent	5.96	2.40	1.23	3.63	140.06
Service	Effluent	5.96	1.80	1.81	3.61	150.07
Service	Influent	9.09	21.46	12.35	33.80	120.05
Service	Effluent	11.71	1.60	1.56	3.16	160.07
Service	Influent	28.43	21.96	9.05	31.01	130.06
Service	Effluent	29.74	1.40	1.23	2.63	150.07
Service	Effluent	41.50	1.40	1.23	2.63	150.07
Service	Effluent	47.51	2.79	9.88	12.67	140.06
Service	Effluent	53.52	4.99	20.58	25.57	130.06
Service	Influent	56.39	21.96	11.52	33.48	120.05
Service	Effluent	56.39	6.99	20.58	27.56	130.06

[a] Sample was diluted in the field (1/10). Values reflect undiluted concentrations.

[b] No sample of the rinse effluent was collected at the beginning of the mode, therefore the first reported throughput BV is for the sample collected at the midway point of the rinse mode.

[c] Sample collected during upflow rinse.

[d] Sample collected during downflow rinse.

Table 37. — *Service performance summary, cycle L.25.22*

	Average concentrations, meq/L			Removal	Resin capacity	Time-weighted
	Influent	Effluent	Difference	(%)	(eq/L)	resin capacity (eq/L)
Ca	21.79	2.01	19.78	91	0.998	0.679
Mg	10.97	4.40	6.57	60	.331	.225
TH	32.76	6.41	26.36	80	1.329	.904
Na	123.39	141.58	-18.19			

Table 38. — *Operation of IX-PP on a 2^4* factorial designed experimant as task LVS-IX-II and -III

Constants:

C_1 = Backwash (feed water): Omitted when X_1 is at (+) level
: When X_2 is at (-) level, 24 L/min for 10 min
C_2 = Regenerant 2 (spent brine): 24 L/min
C_3 = Regenerant 3 (fresh brine): [1](80 percent recovery)
: (90 percent recovery)
C_4 = Drain 1: 3 minutes
C_5 = Rinse (feed water): 15 L/min for 2 min
C_6 = Drain 2: 2 minutes
C_7 = Recycle brine temperature: 27 °C
C_8 = Breakthrough: calcium (meq/L) for 80 percent = 12; for 90 percent = 6

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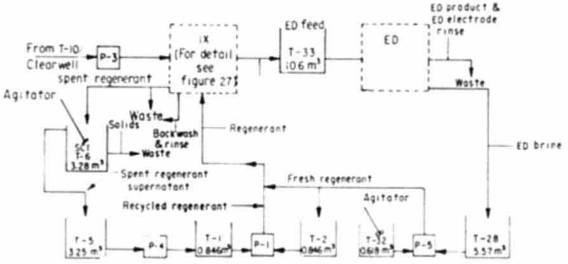
[1] FRV (fresh regenerant volume) = [SER V (service volume)]
[1 - REC (recovery)]

Variables	Low (-)	High (=)	Responses
x_1 = Regenerant 2 recycle volume, L	0	1600	Calcium
x_2 = Brine regenerant concentration, percent water recovery	80	90	Magnesium Service time
x_3 = Resin hold down	No	Yes	
x_4 = Fresh regenerant flow rate, L/min	12	24	

Number cycles	Cycle order	Cycle order IX-III	x1	x2	x3	x4
1	1	3	-	-	-	-
2	2		+	-	-	-
3	6	1	-	+	-	-
4	8		+	+	-	-
5	3		-	-	+	-
6	4		+	-	+	-
7	7		-	+	+	-
8	14		+	+	+	-
9	13		-	-	-	+
10	14		+	-	-	+
11	5		-	+	-	+
12	10		+	+	-	+
13	15	4	-	-	+	+
14	16		+	-	+	+
15	11		-	+	+	+
16	12	2	+	+	+	+



PUMPS

P-1	Regenerant
P-3	IX feed
P-4	Recycled regenerant transfer
P-5	Fresh regenerant transfer

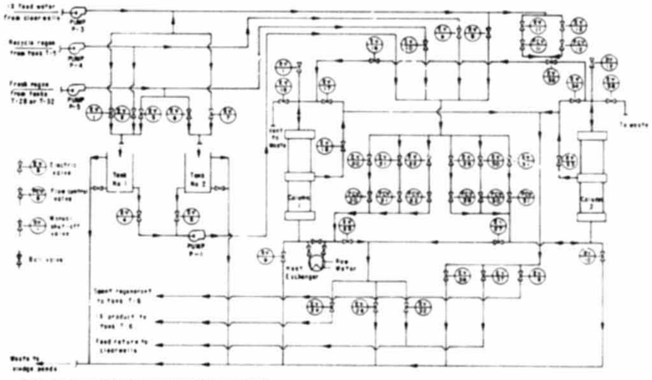
PROCESSES

IX	Ion exchanger
ED	Electrodialyzer
SCT	Solids contact tank (T-6) (Calcium sulfate precipitator)

TANKS

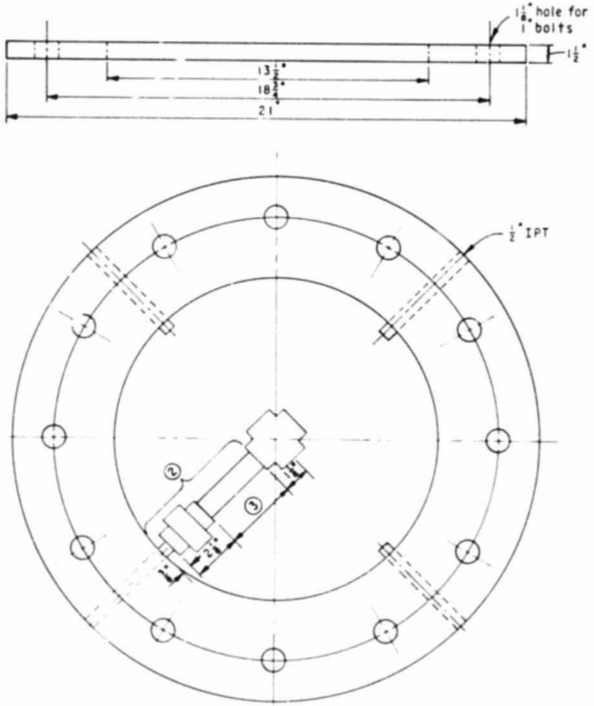
T-1	Recycled regenerant (metering)
T-2	Fresh regenerant (metering)
T-5	Spent regenerant supernatant
T-10/	IX feed
T-28	ED brine
T-32	Sodium chloride solutioning
T-33	IX product/ED feed

Figure 26.—Flow diagram of the ion-exchange: electrodialysis pilot plant.



Note: See figure 26 for identification of pumps and tanks

Figure 27.—Ion-exchange piping diagram.



NOTES

(1) Use schedule 40 PVC threaded fittings
(2) Typical, one of four
(3) Determine nipple length during fabrication
(4) Drill 4-$\frac{1}{2}$ inch NPT taps from both ends of the holes
(5) Drill $\frac{1}{2}$" nipples with $\frac{1}{4}$" drill through 90° axis on $\frac{1}{2}$" centers
(6) Wrap around and seal $\frac{1}{2}$" nipples with 53 mesh polyester screen

Figure 28.—The IX column resin hold-down flange.

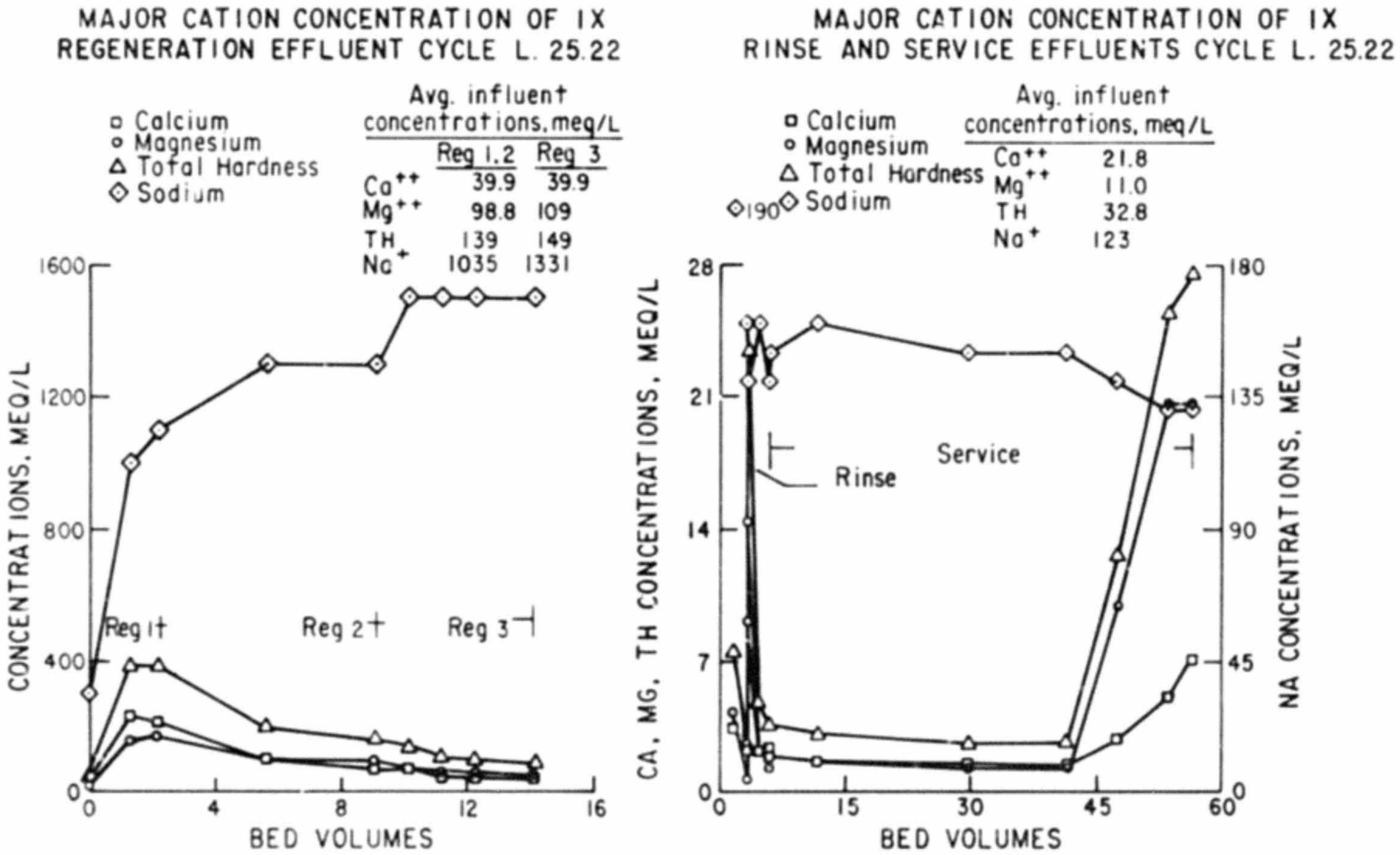


Figure 29.—Major effluent cation concentration versus bed volumes of throughput (sodium, calcium, and magnesium concentrations measured by atomic absorption; total hardness calculated by summing calcium and magnesium concentrations).

EXPERIMENTAL SPIRACTOR PROCESS STUDIES

The Permutit Spiractor [13] is a high-rate, cold lime-soda water softener embodying a unique softening principle — catalytic precipitation. Besides hastening the softening process, which takes less than 8 minutes, a charge of catalyst granules serves as nuclei around which the precipitated hardness compounds build up. These enlarged granules are dense crystalline pebbles, much like coarse sand.

Thus, the voluminous wet sludge (containing 85 to 95 percent water by weight), characteristic of all other lime-water softening processes, is eliminated. Such wet sludge has presented problems in disposal because it usually must first be lagooned or otherwise dried. The spiractor produces a waste product consisting of enlarged catalyst granules that, after drainage, has a moisture content similar to damp sand. The sludge dewatering problem is solved and the use of large land areas for lagooning is avoided.

The spiractor's conical steel shell is partially filled with a granular catalyst of proper grain size. Raw water and the usual chemicals are introduced at the bottom of the cone through separate inlets. The tangential direction of these inlets, plus the conical shape of the shell, imparts a turbulent spiral motion to the incoming water and chemicals; thus, intimate mixing of the water and chemicals with the granules of catalyst is assured. As the water rises in the spiractor cone, its velocity is gradually reduced by the increasing cross-sectional area. Since suspension of the granules of catalyst requires a certain upward velocity of flow, a point is reached where the velocity is no longer sufficient to suspend the catalyst. By design, when this point of low velocity is reached, the time of contact between the water, chemicals, and catalyst has been sufficient to carry the softening reactions to completion. Also, the water has lost almost all of its turbulent motion and rises to the outlet soft and clarified.

As the spiractor continues in operation, the hardness compounds, precipitated by the chemicals and build up on the catalyst granules as hard crystalline shells and enlarge them. These larger, heavier particles gravitate toward the bottom of the spiractor. Periodically, the enlarged granules are drawn off by opening a drain valve at the bottom of the cone. Fresh catalyst is introduced into the spiractor at the top.

A. Equipment Description

1. *Description of equipment set up at test site.* —Figure 30 is an EP&ID (equipment piping and instrument diagram) for the spiractor installation. Figure 31 is a simplified representation of all the spiractor influent and effluent points. The basic piece of equipment utilized in this phase of testing was a 3.81-meter-tall Permutit spiractor. The top 1.83 meters section was 0.30 meter in diameter, and from there the diameter tapered to near the bottom to approximately 0.08 meter.

A Kates flow control valve and a rotometer (flow rate meter) were used at various stages to measure and control raw water flow rate. The chemical feed system consisted of two 0.19-m^3 uncoated steel feed tanks with attached steel chemical feed pumps. Steel was not the preferred construction material, but had to be used to keep onsite installations on schedule. Pump discharges were connected to the spiractor influent line through shut-off valves. The output of these pumps was manually controlled; the pumping rates were volumetrically measured. The spiractor overflowed into a third 0.19-m^3 uncoated steel tank, and then gravity flowed to waste. Catalyst was manually loaded, as needed, into the top. Catalyst bed level was monitored by collecting samples from the appropriate sample valves. A total of 12 sample valves were located at 0.30 meter intervals along the entire length.

2. The equipment description and operating instructions provided to the operating contractor by the "Spiractor" vendor are attached as appendix H. Appendix H contains the following information:

- Specifications and data
- General description
- Principle of operation
- Flow control
- Sampling connections
- Placing in operation
- Control of catalyst bed
- Anticipated operating results
- Chemical control
- Eductor
- Trisodium phosphate feed
- Effluent control
- Spiractor assembly
- Bill of materials

B. Process Operations

1. *Operations of Spiractor equipment.* The spiractor was operated according to the flow pattern shown in figure 32 and the following procedures:

The raw water flow into the base of spiractor was started through valve V-9;

Valve V-8 was opened to permit overflow from spiractor into the temporary holding tank;

The raw water flow rate was adjusted with a Kates valve FCV-1 to 26.5 L/min (7 gal/min);

Sand (AGSCO No. 7 silica) was added by bucket into the top opening of spiractor;

Each bucket of sand was allowed to equilibrate for 10 minutes;

After each charge of sand, the process of the rise was followed in the sand-water interface up the spiractor by catching a sample in a beaker and making visual observation, starting at the lowest sample valve SV-1 and working up. Fresh sand will probably have ultrafines that will make sample of water smokey. Therefore, look for sand particles. Dump all samples back in top of spiractor;

The sand charging and sampling procedure was repeated until sand was detected at sample point SV-9; and then

The charging of sand was discontinued, the operation of system proceeded.

2. *Lime-soda ash softening procedure.* — Operation of the spiractor for lime-soda ash softening went as follows:

a. To prepare a 10 percent lime milk, charge lime, and clearwell water to makeup tank, and agitate.

b. To prepare a 10 percent soda ash solution, continue agitation in the makeup tank and charge with soda ash and clearwell water.

c. Based on the operation of the lime treatment train, the lime required to neutralize all temporary hardness in the raw water flowing at 7 gal/min would be 4.0 lb/h (dry basis). Therefore, start pump P-2 and adjust the flow rate of the 10 percent lime slurry to 40 lb/h;

d. Based on the operation of the lime-soda ash train, 4.2 lb/h (dry basis) soda ash is required to reduce calcium to less than 50 mg/L. Therefore, start pump P-3 and adjust the flow rate of the 10 percent soda ash solution to 42 lb/h; and

e. As calcium carbonate forms in the spiractor, it will nucleate and coat on the sand particles. This might result in a change in density of the particles, and the sand interface could move up above the SV-9 sample point. If this occurs, reduce the flow to 24.6 L/min (6.5 gal/min), and proportionately reduce the lime and soda ash feeds. If necessary, reduce flows to 22.7 L/min (6.0 gal/min) so the interface drops back to SV-9. Do not go any lower than 22.7 L/min (6.0 gal/min). If necessary, the next corrective step would be to operate with the interface at SV-10.

C. Results and Discussion

The spiractor operation had the following objectives:

- Lime-soda ash treatment of LVS water with nucleation and densification of precipitated calcium carbonate on sand particles.
- Calcium sulfate desupersaturation of ED- (electrodialysis) produced reject brine for IX (ion exchange) regenerant by nucleation on calcium sulfate pellets.

The following experimental spiractor tasks were performed to evaluate the spiractor.

- Task SP-I, installation and hydraulic testing of SPPP (spiractor pilot plant)
- Task SP-II, lime-soda ash treatment with the SPPP
- Task SP-III, modification of SPPP installation for desupersaturation test
- Task SP-IV, desupersaturation IX recycle regenerant with the SPPP

1. *Installation and hydraulic testing.* — Task SP-I included receiving, setting and assembling components for the SPPP and hydraulically testing of the spiractor system. Hydraulic tests and correction of leaks and malfunctions were accomplished on July 1, 1980. The assembly drawing of the spiractor is attached in appendix H. The spiractor was assembled in the total system according to figure 32. The spiractor was set up on a concrete pad as close as possible to the recycle regenerant tank No. T-5 for the desupersaturation experiment. Since the spiractor was first tested on lime-soda ash treatment, raw water was transferred from where it enters the lime treatment train to the spiractor through 1½-inch and 2-inch pipe nozzles.

2. *Demonstration of chemical softening with the spiractor.* — This task called for initiation of operation of the spiractor with the following chemical dosages:

Lime	1140 g/m^3
Soda ash	1200 g/m^3

The initial flow rate of feed water was to be 0.44×10^{-3} m^3/s. This flow rate was to be adjusted as required to maintain the catalyst level at the required control point. The minimum allowable for equilibrium conditions flow was 0.38 x 10^{-3} m^3/s; the recommended catalyst level was at level 9 (SV-9), figure 31. Catalyst used for this phase of the testing was silica sand; the proper level was to be maintained by recharging with additional catalyst as required. Once stable operation was attained, data were to be collected to allow evaluation of the system.

3. *Operating conditions.* — Lime and lime-soda ash treatment in the spiractor was done at the 16 different conditions summarized in table 39.

4. *Summary of results.* — Table 40 presents the operating data obtained for the various conditions tested during task SP-II. No data are presented for tests 1 through 8, since at these conditions, the efficiency of operation ranged from no softening to very slight softening. Test 9 was the first condition that showed any appreciable degree of softening.

Analysis of the data indicates that the spiractor will soften LVS water effectively either before or after carbon dioxide has been removed in the aeration system. In tests using only soda ash (tests 10, 11, 12, and 13), as the soda ash dosage increases, calcium hardness and total hardness removals also increase. Residual total alkalinities also increase proportionately as the soda ash dosage increases.

Tests 9 and 15 utilized both lime and caustic soda. During test 9, caustic soda was injected only for short periods because of a limited supply on hand. The primary purpose of the caustic soda was to aid in initiating the scaling of the silica sand catalyst; the use of caustic soda throughout the test was not planned. Operating data obtained at this set of conditions indicates partial treatment occurred.

One goal of parametric testing was to establish and maintain the effluent at high pH to promote a complete reaction. High pH was not attained until caustic soda was injected into the feed water (test 15, table 42). After 1.3 hours of operation, lime addition was initiated. Average calcium reduction was 763 g/m^3 (1908 g/m^3 as $CaCO_3$). Spiractor effluent calcium concentration averaged approximately 47 g/m^3 (188 g/m^3 as $CaCO_3$); effluent pH averaged 10.3. Total alkalinity as $CaCO_3$ was reduced from 1081 to 848 g/m^3. Phenolphthalein alkalinity was zero in influent and averaged 397 g/m^3 in effluent.

On July 31, test 16, the final softening test, was initiated. This was the only test to use aerated raw water as a feed source. Both lime and soda ash were used. Treatment efficiency was excellent (table 41). Average calcium reduction in the spiractor was 654 g/m^3 (1635 g/m^3 as $CaCO_3$). Effluent calcium concentration averaged approximately 71 g/m^3 (178 g/m^3 as $CaCO_3$); effluent pH averaged 10.3. Total alkalinity was reduced, on the average, from 852 to 313 g/m^3. Phenolphthalein alkalinity was zero in influent and averaged 124 g/m^3 in effluent.

Laboratory data presented in table 41 for August 3 indicate that the treatment at the time the sample was collected was better than the average treatment, as indicated by the operating data. With the exception of the extremely high carryover during test condition 16, the quality of treatment was generally as good as, if not better than, the treatment attained with the in-line softening system.

5. *Conclusions and recommendations.* — The following conclusions can be drawn from task SP-II testing:

a. The spiractor can be used in a lime-soda ash softening system in treating the LVS water. Calcium can be reduced to as low as 10 g/m^3;

b. Catalyst handling and disposal present a totally different problem than does normal softening sludge. However, the spiractor produces a much more compact sludge; and

c. Carryover problems that occurred when lime was used in the system requires further investigation. Spiractor effluent might require a clarification step to reduce the load on the filters.

6. *Optimum steady-state conditions.* — Optimum steady-state conditions for lime-soda ash softening of aerated LVS water with a 12-inch-diameter spiractor were as follows:

Process conditions

Water feed rate	0.3 x 10^{-3} m^3/s
Lime dosage	830 g/m^3
Soda ash dosage	1200 g/m^3
Catalyst control level	SV-9

Water quality changes

	Aerated influent (g/m^3)	Spiractor effluent (g/m^3)
pH	7.0	10.3
TDS	9166	8456
Silica	30	2
Calcium	770	10
Magnesium	144	677
Sodium	2220	2910
Strontium	10.4	0.3
Bicarbonate	932	ND*
Carbonate	ND	80

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*Filtered samples. ND: not detected.

7. *Task SP-III, conversion of spiractor to desupersaturation mode.* — After completion of the 16 lime-soda ash addition tests, the spiractor was re-piped according to figures 30 and 32. Conversion of the spiractor for testing supersaturation of IX brine regenerant was completed on August 10, 1980.

8. *Task SP-IV, desupersaturation testing.* — This task called for the desupersaturation of the IX regeneration effluents by diverting both the recycle (Regen 2) and fresh (Regen 3) regenerants through the spiractor and then into tank T-5. Upon completion of the transfer of these spent regenerants to tank T-5, the tank contents were to be continually cycled through the spiractor in an attempt to further reduce the possibility of calcium sulfate scaling.

The catalyst used for task SP-IV of the operation was insoluble calcium sulfate pellets; no other chemicals were used in this task. Catalyst was maintained at a specified level by recharging with additional catalyst after blowing down the catalyst as required to maintain proper system operation.

a. Equipment description. — During the desupersaturation testing, IX regenerants were used as the feed water for the spiractor. Figure 33 presents the process flow diagram for this phase of testing. When the IX column was being regenerated, the spent regenerant was pumped to tank T-5, the IX recycle regenerant tank. Upon completion of the regeneration mode valves on the IX column were set to allow the tank T-5 contents to be continually recirculated through the spiractor via the recycle regenerant pump P-5. During the regeneration mode, this pump was used to supply the regenerant to the IX column.

b. Operating procedure. — During the portion of the operation when the recycle regenerant in tank T-5 was being constantly recirculated through the spiractor, the flow rate was maintained at approximately 0.44 x 10^{-3} m^3/s. The catalyst bed was maintained at level SV-9. The flow rate was the same as the regenerant flow rate used in the IX regenerations. The operating conditions for the IX during the three-step regenerations were the same as run L 20.00 (table 42) where the flow rate for all three regeneration steps was 0.4 x 10^{-3} m^3/s. The volume of the recycle regenerant column (Regen 2) was 1.5 m^3, the volume of fresh regenerant used was 0.50 m^3. The final cycle, conducted on August 22, utilized a higher flow rate than the other cycles. The flow rate was increased from 0.4 x 10^{-3} m^3/s to 0.53 x 10^{-3} m^3/s. The remainder of the operating conditions remained unchanged.

c. Summary of results. — Samples of spiractor influent (IX spent regenerant) and effluent were collected for laboratory analysis on August 19, 21, and 22. Table 42 presents the results. Samples were collected during regeneration steps using both recycled regenerant (Regen mode II) and fresh regenerant (Regen mode III). The data were analyzed using the Yeats ED method[7] (table 43). These data indicate the spiractor influent, during the IX Regen III mode on August 19, was supersaturated and after passing through the spiractor became more supersaturated. Influents, during IX Regen III modes of August 21 and 22, were not supersaturated with respect to gypsum and, after passing through the calcium sulfate catalyst bed in the spiractor, became supersaturated.

Five brine cycles, utilizing column 1, were conducted from August 18 through 22. Samples for laboratory analysis were collected during only three of these five cycles. Regeneration modes 1 and 2 used spent regenerant that had been recirculated through the spiractor in an attempt to desupersaturate the calcium sulfate in the spent regenerant. Regeneration mode III of these cycles utilized 90 percent fresh ED brine as the regenerant. During both regeneration modes II and III of these cycles, the column effluent was sent through the spiractor and into tank T-5. The contents of tank T-5 were then recirculated during the rinse and service modes of these cycles.

During Regen I, the resin bed expanded to the top of the column. At the beginning of Regen II, approximately 10 to 15 cm of the bed remained at the top of the column. This top portion of the resin bed appeared to be slightly cemented with calcium sulfate during Regens II and III. The remainder of the bed remained fluidized. During Regen II of the third and fourth cycles, a gap approximately 3- to 5-cm wide was observed between the cemented and fluidized portions of the resin bed.

d. Conclusions. — The following conclusions can be drawn from this phase of the testing:

(1) Data collected during this phase of the testing indicates the spiractor operation was only partially successful;

(2) According to table 43, three of the regenerant feed waters increased in calcium concentration and the other three regenerant feed waters decreased in calcium concentration after contact with suspended calcium sulfate pallets in the spiractor; and

(3) One of the objectives of this test was to visually compare the results of the IX system

operation using the spiractor with the results obtained during cycle L.10.00 (table 44) of the IX program. Cycle L.10.00 was performed in phase I of the LVS site test program with lime-treated water. The desupersaturation tests were run during phase II with lime-soda ash treated water. The test schedule did not permit using lime-treated water for the desupersaturation test, and a visual comparison was not conclusive.

Table 39. — *Operating conditions, spiractor softening testing*

Date	Test 1[b] 7/8/80 to 7/9/80	Test 2 7/9/80 to 7/11/80	Test 3 7/11/80 to 7/12/80	Test 4 7/12/80 to 7/13/80	Test 5 7/13/80 to 7/14/80	Test 6 7/14/80 to 7/15/80	Test 7 7/15/80 to 7/16/80	Test 8 7/16/80 to 7/17/80
Flow rate ($m^3/s \times 10^{-3}$)	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.31
Feedwater	Raw	Raw	Raw	Raw	Raw	Raw	Raw	Raw
Lime dosage (g/m^3)	1140	1200	1200	1200	1200	1200	1600	1200
Soda ash dosage (g/m^3)	1200	0	0	0	0	0	0	0
Caustic soda dosage (g/m^3)	0	0	0	0	0	0	0	0
Trisodium phosphate dosage (g/m^3)	0	0	10	30[c]	40	50	60	70
Catalyst control level[a]	SV-9	SV-9	SV-9	SV-9	SV-9	SV-9	SV-9	SV-9
Date	**Test 9 7/18/80 to 7/21/80**	**Test 10 7/22/80 to 7/23/80**	**Test 11 7/23/80 to 7/24/80**	**Test 12 7/24/80 to 7/25/80**	**Test 13 7/25/80 to 7/26/80**	**Test 14 7/28/80 to 7/28/80**	**Test 15[f] 7/29/80 to 7/30/80**	**Test 16 7/31/80 to 8/ 3/80**
Flow rate ($m^3/s \times 10^{-3}$)	0.31	0.31	0.31	0.31	0.31	0.19	0.31	0.31
Feedwater	Raw	Raw	Raw	Raw	Raw	Raw	Raw	Aerated
Lime dosage (g/m^3)	1200	0	0	0	0	1180[e]	1200	830
Soda ash dosage (g/m^3)	0	1200	2400	3600	3600	0	0	1200
Caustic soda dosage (g/m^3)	650[d]	0	0	0	0	0	1590	0
Trisodium phosphate dosage (g/m^3)	0	0	0	0	0	0	0	0
Catalyst control level[a]	SV-9	SV-9	SV-9	SV-9	SV-10	SV-10	SV-9	SV-9

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[a] Catalyst blowdown was as required. Catalyst was added, as required after blowdowns, to maintain control levels.

[b] No softening occurred at this condition, and Permutit was consulted. The test conditions which followed were the result of the recommendations received from Permutit and the USBR Project Manager.

[c] This condition was initiated at a TSP dosage of 20 g/m^3, but was subsequently increased to 30 g/m^3.

[d] Lime addition was initiated after 4.0 hours of operation. The caustic soda feed terminated 0.8 hour later because the caustic soda supply was depleted. The remaining operation at this condition was with lime addition only.

[e] The lime solution feed source for this test condition was tank T-5. A total of 3.380 m^3 of 10-percent solution was used. This was injected at a rate of 0.19×10^{-3} m^3/s to match the raw water flow rate of 0.19×10^{-3} m^3/s. This test continued until tank T-5 was drained.

[f] This test condition was initiated with an initial caustic soda dosage of 1270 g/m^3. After approximately 0.7 hour, the caustic soda dosage was increased to 1 590 g/m^3. After 0.6 hour more, the lime injection at 1200 g/m^3 was initiated. The caustic soda supply was depleted shortly thereafter and the system continued to operate with lime addition only.

Table 40. — *Average operating data spiractor softening tests*

Time period, 1980 Test condition No.[1]		July 18 - July 21 9	July 22 - July 23 10	July 23 - July 24 11	July 24 - July 25 12	July 25 - July 26 13	July 28 14	July 29 - July 30 15	July 31 - August 3 16
SPIRACTOR INFLUENT									
Flow[2]	$m^3/s \times 10^{-3}$	0.3	0.3	0.3	0.3	0.3	0.2	0.3	0.3
pH (bench)		6.2	6.1	6.2	6.2	6.2	6.2	6.2	7.4
CO_2	g/m^3	624	687	705	666	666	634	623	47
Sulfides, total	g/m^3	5.0	4.5	5.4	5.0	4.8	6.5	5.2	1.7
Calcium	g/m^3	817	801	794	788	820	822	810	725
Hardness as CaCO	g/m^3	2643	2627	2620	2540	2580	2620	2610	2280
Turbidity (bench)	JTU	3.8	2.2	4.9	1.6	4.4	2.1	3.6	43
T-alk as $CaCO_3$	g/m^3	1063	1123	1146	1127	1052	1080	1081	852
Temperature	°C	40.7	40.0	40.5	41.0	40.3	40.6	41.1	40.4
SPIRACTOR EFFLUENT									
pH (bench)		9.0	6.6	6.9	7.0	7.6	7.2	10.3	10.3
Calcium	g/m^3	465	625	284	184	40	491	47	71
Hardness as CaCO	g/m^3	1747	2090	1300	1065	640	1710	325	482
T-alk as $CaCO^3$	g/m^3	659	1843	2034	2278	2780	730	848	313
P-alk as $CaCO_3$	g/m^3	142	ND	ND	ND	ND	ND	397	124
Turbidity (bench)	JTU	>500	-	39	92	272	>500	-	>500
Temperature	°C	40.4	40.0	40.0	40.0	39.3	38.6	40.4	39.8
CO_2	g/m^3	-	-	230	204	56	66	-	ND
SUMMARY									
Calcium removal	g/m^3	352	176	510	604	780	331	763	654
Calcium removal as $CaCO_3$	g/m^3	880	440	1275	1510	1950	828	1908	1635
Hardness removal as $CaCO_3$	g/m^3	896	537	586	262	-	910	2285	1798
Average effective turbidity	JTU	>500	-	39	92	272	>500	-	>500
Lime feed	g/m^3	2976	-	-	-	-	1199	1144	-

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[1] No data are presented for tests 1 through 8 since no apparent softening took palce.

[2] Flow rates when in service.

ND — None detected.

Table 41. — *Influent and effluent composition, spiractor softening testing*

Date Time Collected Date analyzed at Yuma	Units	Raw water influent 7/70/80 0750 9/5/80	Effluent^{c} 7/30/80 0750 9/5/80	Aerated Influent 8/3/80 0000 9/5/80	Effluent^{c} 8/3/80 0000 9/5/80
pH^{a}		6.2	10.4	7.0	10.3
TDS (calculated)^{b}	g/m^3	9330	8978	9107	8532
TDS (evaporated @ 105 °C)^{a}	g/m^3	9654	8921^{d}	9166	8456^{d}
Conductivity @ 25 °C^{a}	mS/m	1500	1500^{d}	1380	1400^{d}
Temperature^{a}	°C	41.0	40.0	—	—
Silica	g/m^3	39.0	3.4	30.0^{a}	2.4
Calcium	g/m^3	830	7.1	770	10.4
Magnesium	g/m^3	142	29.2	144	66.7
Sodium	g/m^3	2220	3180	2220	2910
Potassium	g/m^3	168	168	172	173
Iron, total	g/m^3	<0.10	<0.10	0.18	<0.10
Manganese, total	g/m^3	<0.30	<0.30	<0.30	<0.30
Strontium	g/m^3	10.6	0.30	10.4	0.3
Bicarbonate	g/m^3	1281^{a}	64.9	932^{a}	ND
Carbonate	g/m^3	ND^{a}	263	ND^{a}	80.2
Sulfate	g/m^3	1905	1900	1925	1895
Chloride	g/m^3	3375	3395	3370	3390
T-alkalinity as $CaCO_3$	g/m^3	1050^{a}	491	764^{a}	144
P-alkalinity as $CaCO_3$	g/m^3	ND^{a}	219	ND^{a}	77.2
Hydroxide	g/m^3	ND^{a}	ND	ND^{a}	3.5
E.F. [TDS(calculated)/cond.]		6.2	6.0^{d}	6.6	6.1
E.F. [TDS(evaporated)/cond.]		6.4	5.9	6.6	6.0^{d}
Σ Anions	eq/m^3	155.89	145.17	150.44	137.83
Σ Cations	eq/m^3	154.21	145.39	151.47	137.02
Control value	eq/m^3	+0.67	-0.09	-0.42	+0.36

[a] Analysis conducted at LVSTS.
[b] Calculated TDS uses only one-half of the bicarbonate.
[c] The data presented for the Spiractor product was obtained from a filtered sample.
[d] The conductivity and evaporated TDS values were obtained from an unfiltered sample. The E.F.'s were calculated from this conductivity value.
ND — None detected.

Table 42a. — *Ion-exchange, run L.20.00*

Date: 3/29/80
Cycle: L.20.09
Conditions: Feedwater — LaVerkin Springs — lime-softened, pretreated
Regenerants — recycled regenerant and fresh ED brine

		Target	Actual
Control variables:	Fresh regeneration conc. (mg/L TDS-calc.)	—	87 980
	Fresh regeneration conc. (mg/L TDS-evap.)	—	91 900
	Fresh regeneration flow rate (L/min)	24.0	23.9
	Recycled regenerant flow rate (L/min)	24.0	23.6
	Recycled regenerant volume (L)	1600	1603
	Service termination point (meq/L Ca^{++})	6.0	6.6
	Heat exchanger used	No	No
	Packed bed regeneration mode used (i.e. resin hold-down)	No	No
Standard resin bed:	Volume = 115 L		

Table 42b. — *Chemical compositions of tank waters prior to cycle L.20.09*

Tank	pH	Conductivity (mS/m)	Ca^{++} (meq/L)	Mg^{++} (meq/L)	TH (meq/L)
Recycle regenerant (T-5)	7.3	9 550	43.0	85.0	128.0
Spent regenerant (T-6)	7.2	9 900	37.0	90.0	127.0
Lime-softened feed (T-9)	7.1	1 300	21.6	10.0	31.6
Lime-softened feed (T-10)	7.2	1 380	22.0	9.6	31.6
Fresh ED brine (T-28)	6.7	>10 000	26.0	94.0	120.0
IX product/ED feed (T-33)	6.9	1 300	1.1	4.5	5.6
Lime-softened feed (clearwell)	7.3	1 410	21.8	10.0	31.8

Table 42c. — *Operating conditions, cycle L.20.09*

Mode	Input	Output	Duration (min)	Throughput (L)	Volume (BV)	Avg flow rate (L/min)	Avg flow rate (RV/min)	Bed Expansion (%)	Temperature (°C)
Regen 1	Re regen	Waste	10	240	2.09	24.0	0.209	—	15.0
Regen 2	Re regen	Sp regen	68	1603	13.9	23.6	.205	—	11.9
Regen 3	Re regen	Sp regen	25	598	5.20	23.7	.206	—	13.6
Drain 1	(Vent)	Waste	2	30	0.26	15.0	.130	0.0	—
Rinse	Feed	Waste	10	150	1.30	15.0	.130	0.0	—
Service	Feed	Product	198	5940	51.7	30.0	.261	0.0	—
Drain 2	(Vent)	Waste	3	41	0.36	13.7	.119	0.0	—

Table 42d. — *Spiractor influent and effluent composition*

	Units	Influent during Regen 2	Effluent during Regen 2	Influent during Regen 3	Effluent during Regen 3
Date		8/19/80	8/19/80	8/19/80	8/19/80
Time collected		0850	0850	0942	0942
Date analyzed at Yuma		10/17/80	10/17/80	10/17/80	10/17/80
pH[a]		8.0	8.0	7.5	7.8
TDS (calculated)[b]	g/m^3	92 535	91 977	87 935	89 868
TDS (evaporated @ 105 °C)[a]	g/m^3	96 040	94 624	89 625	90 861
Conductivity @ 25 °C[a]	mS/m	9 270	9 120	9 300	9 200
Temperature[a]	°C	24.8	26.0	26.2	25.0
Silica	g/m^3	1.20	1.00	2.20	2.00
Calcium	g/m^3	790	680	620	650
Magnesium	g/m^3	1 092	1 179	610	796
Sodium	g/m^3	30 700	30 500	29 700	30 000
Potassium	g/m^3	1 680	1 650	1 580	1 590
Iron, total	g/m^3	<0.04	<0.04	<0.04	<0.04
Manganese, total	g/m^3	0.30	0.30	0.10	0.20
Strontium	g/m^3	12.0	12.0	8.0	10.0
Bicarbonate	g/m^3	220	210	229	239
Carbonate	g/m^3	ND	ND	ND	ND
Sulfate	g/m^3	22 100	21 800	21 850	21 750
Chloride	g/m^3	36 050	36 050	33 450	34 950
T-alkalinity as $CaCO_3$	g/m^3	180	172	188	196
P-alkalinity as $CACO_3$	g/m^3	ND	ND	ND	ND
Hydroxide	g/m^3	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		10.0	10.1	9.5	9.8
E.F. [TDS(evaporated)/cond.][a]		10.4	10.4	9.6	9.9
Σ Anions	eq/m^3	1 481	1 475	1 402	1 443
Σ Cations	eq/m^3	1 508	1 500	1 414	1 444
Control value	eq/m^3	-1.17	-1.12	-0.51	-0.04

[a] Analysis conducted at LVSTS.

[b] YDTF used only one-half the bicarbonate in calculating the TDS.

Table 42d. — *Spiractor influent and effluent composition — Continued*

	Units	Influent during Regen 2	Effluent during Regen 2	Influent during Regen 3	Effluent during Regen 3
Date		8/19/80	8/19/80	8/19/80	8/19/80
Time collected		0850	0850	0942	0942
Date analyzed at Yuma		10/17/80	10/17/80	10/17/80	10/17/80
pH[a]		8.0	8.0	7.4	7.7
TDS (calculated)[b,c]	g/m^3	88 674	88 258	85 960	85 689
TDS (evaporated @ 105 °C)[a,c]	g/m^3	95 764	92 606	88 678	92 332
Conductivity @ 25 °C[a]	mS/m	9 380	9 350	9 030	8 960
Temperature[a]	°C	25.2	25.0	26.5	25.7
Silica	g/m^3	0.80	0.60	1.80	1.40
Calcium	g/m^3	710	630	363	566
Magnesium	g/m^3	1 270	1 360	622	835
Sodium	g/m^3	28 600	28 500	29 000	28 400
Potassium	g/m^3	1 550	1 520	1 520	1 530
Iron, total	g/m^3	<0.04	<0.04	<0.04	<0.04
Manganese, total	g/m^3	0.30	0.30	0.10	0.30
Strontium	g/m^3	9.00	11.0	7.00	8.00
Bicarbonatre	g/m^3	268	273	293	298
Carbonate	g/m^3	ND	ND	ND	ND
Sulfate	g/m^3	21 000	20 700	21 600	21 600
Chloride	g/m^3	35 400	35 400	32 700	32 600
T-alkalinity as $CaCO_3$	g/m^3	220	224	240	244
P-alkalinity as $CaCO_3$	g/m^3	ND	ND	ND	ND
Hydroxide	g/m^3	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		9.4	9.4	9.5	9.6
E.F. [TDS(evaporated)/cond.][a]		10.2	9.9	9.8	10.3
Σ Anions	eq/m^3	1 440	1 434	1 377	1 374
Σ Cations	eq/m^3	1 424	1 422	1 370	1 372
Control value	eq/m^3	+0.74	+0.54	+0.34	+0.13

[a] Analysis conducted at LVSTS.

[b] YDTF used only one-half the bicarbonate in calculating the TDS.

[c] The discrepancy between the calculated and evaporated TDS is attributed to the unstable nature of the sample at time of initial dilution. Samples were diluted in the field (1/10) and filtered.

Table 42d. — Spiractor influent and effluent composition — Continued

	Units	Influent during Regen 2	Effluent during Regen 2	Influent during Regen 3	Effluent during Regen 3
Date		8/22/80	8/22/80	8/22/80	8/22/80
Time collected		0740	0740	0810	0810
Date analyzed at Yuma		10/21/80	10/21/80	10/21/80	10/21/80
pH[a]		8.0	8.1	7.6	7.8
TDS (calculated)[b]	g/m^3	89 752	89 191	84 301	84 292
TDS (evaporated @ 105 °C)[a,c]	g/m^3	93 163	92 932	122 503	125 265
Conductivity @ 25 °C[a]	mS/m	9 410	9 480	9 100	9 110
Temperature[a]	°C	25.3	25.3	27.6	26.8
Silica	g/m^3	0.40	0.60	1.60	1.40
Calcium	g/m^3	700	640	350	640
Magnesium	g/m^3	1 350	1 360	624	877
Sodium	g/m^3	29 200	28 900	28 400	27 800
Potassium	g/m^3	1 570	1 550	1 590	1 530
Iron, total	g/m^3	<0.04	<0.04	<0.04	<0.04
Manganese, total	g/m^3	0.30	0.30	0.20	0.30
Strontium	g/m^3	10.0	11.0	8.00	10.0
Bicarbonatre	g/m^3	244	259	254	268
Carbonate	g/m^3	ND	ND	ND	ND
Sulfate	g/m^3	21 000	20 800	20 600	20 700
Chloride	g/m^3	35 800	35 800	32 600	32 600
T-alkalinity as $CaCO_3$	g/m^3	200	212	208	220
P-alkalinity as $CaCO_3$	g/m^3	ND	ND	ND	ND
Hydroxide	g/m^3	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		9.5	9.4	9.3	9.3
E.F. [TDS(evaporated)/cond.][d]		9.9	9.8	13.5	13.8
Σ Anions	eq/m^3	1 451	1 447	1 353	1 355
Σ Cations	eq/m^3	1 457	1 441	1 345	1 353
Control value	eq/m^3	-0.23	+0.29	+0.38	+0.12

[a] Analysis conducted at LVSTS.

[b] YDTF used only one-half the bicarbonate in calculating the TDS.

[c] The discrepancy between the calculated and evaporated TDS is attributed to the unstable nature of the sample at the time of initial dilution. Samples were diluted in the field (1/10) and filtered.

Table 43. — *Analysis of gypsum solubility (based on the Marshall/Slusher Program)*[7]

	8/19/80	8/19/80	8/21/80	8/21/80	8/22/80	8/22/80
Test Date Spiractor feed source	Regen II	Regen III	Regen II	Regen III	Regen II	Regen III
Spiractor influent						
Calcium conc. (g/m^3)	790	620	710	363	700	350
Initial molal ionic strength (I_o)	1.789	1.676	1.721	1.633	1.746	1.598
Saturated molal ionic strength (I_s)	1.271	1.390	1.359	1.976	1.406	2.037
Concentration factor (CF) = $I_s \div I_o$	0.710	0.829	0.789	1.210	0.805	1.275
Spiractor effluent						
Calcium conc. (g/m^3)	680	650	630	566	640	640
Initial molal ionic strength (I_o)	1.780	1.719	1.715	1.646	1.733	1.622
Saturated molal ionic strength (I_s)	1.436	1.397	1.496	1.464	1.496	1.356
Concentration factor (CF) = $I_s \div I_o$	0.807	0.813	0.872	0.889	0.864	0.836
Summary						
Calcium reduction (g/m^3)	(-)110	(=) 30	(-) 80	(=)203	(-) 60	(=)290
Concentration factor differential ($CF_{eff}-CF_{inf}$)	(+) 0.097	(-) 0.016	(+) 0.083	(-) 0.321	(+) 0.059	(-) 0.439

Table 44a — *Ion-exchange, run L 10 00*

Date	3/8/80		
Cycle	L 10 14		
Conditions	Feedwater — LaVerkin Springs — lime-softened, pretreated		
	Regenerants — recycled regenerant and fresh ED brine		
Control variables	Fresh regeneration conc (mg/L TDS-calc)	—	94 000
	Fresh regeneration conc (mg/L TDS-evap)	—	93 400
	Fresh regeneration flow rate (L/min)	24 0	23 7
	Recycled regenerant flow rate (L/min)	24 0	23 2
	Recycled regenerant volume (L)	1 600	1 594
	Service termination point (meq/L Ca++	6 0	8 2
	Heat exchanger used	Yes	Yes
	Packed bed regeneration mode used (i.e., resin hold-down)	No	No
Standard resin bed	Height[a] = 1 320 mm		
	Volume = 115 0 L		

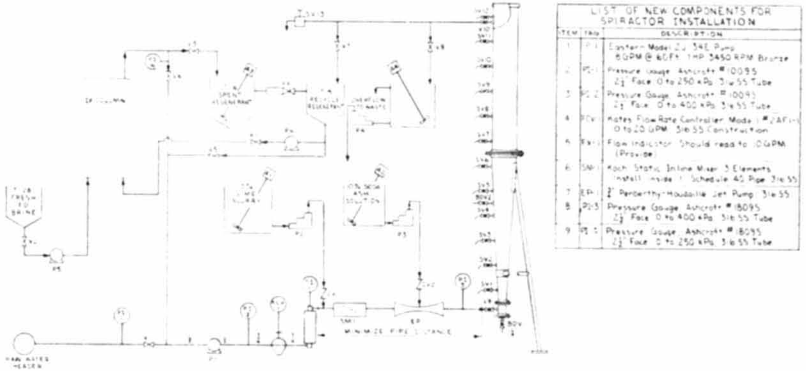
[a] The resin bed height at the end of drain-down after regeneration with fresh regenerant was used as the standard resin bed height in calculating bed expansion

Table 44b — *Chemical compositions of tank waters prior to cycle L 10 14*

Tank	pH	Conductivity (mS/m)	Ca++ (meq/L)	Mg++ (meq/L)	TH (meq/L)
Recycle regenerant (T-5)	7 6	8100	45 0	93 0	138 0
Spent regenerant (T-6)	7 5	8210	44 0	91 0	135 0
Lime-softened feed (T-9)	7 3	1250	21 6	8 2	29 8
Lime-softened feed (T-10)	7 4	1290	23 8	9 2	33 0
Fresh ED brine (T-28)	6 5	9910	23 2	29 6	52 8
IX product/ED feed (T-33)	6 0	1250	1 3	2 5	3 8
Lime-softened feed (clearwell)	7 4	1220	21 0	9 4	30 4

Table 44c — *Operating conditions, cycle L 10 14*

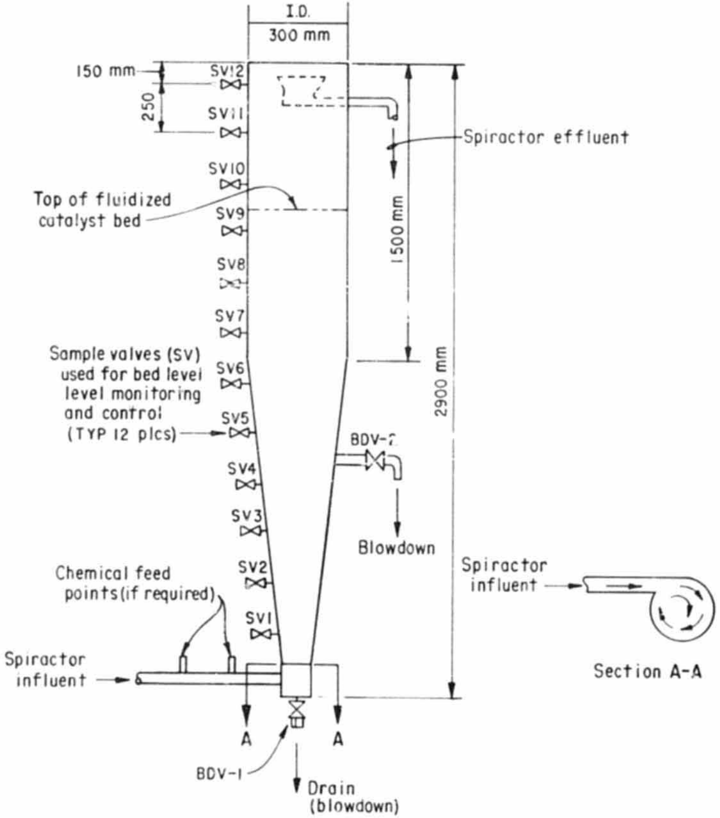
Mode	Input	Output	Duration (min)	Throughput (L)	Volume (BV)	Avg flow rate (L/min)	Avg flow rate (BV/min)	Bed expansion (%)	Temp (°C)
Regen 1	Re regen	Waste	10	250	2 17	25 0	0 217	42	19 5
Regen 2	Re regen	Sp regen	69	1594	13 9	23 1	201	42 0	26 6
Regen 3	Re regen	Sp regen	22	522	4 54	23 6	205	42	26 4
Drain 1	(Vent)	Waste	3	62	0 54	20 7	180	0 0	
Rinse	Feed	Waste	15	220	1 91	14 7	128	0 0	
Service	Feed	Product	187	5610	48 8	30 0	261	0 0	
Drain 2	(Vent)	Waste	2	41	0 36	20 7	180	0 0	



LIST OF NEW COMPONENTS FOR SPIRACTOR INSTALLATION

ITEM	TAG	DESCRIPTION
1	P-1	Eastern Model 2U 34E Pump 8GPM @ 60Ft 1HP 3450 RPM Bronze
2	PI-1	Pressure Gauge Ashcroft #10095 2½" Face 0 to 250 kPa 316 SS Tube
3	PI-2	Pressure Gauge Ashcroft #10095 2½" Face 0 to 400 kPa 316 SS Tube
4	FCV-1	Kates Flow Rate Controller Model #2AF-1 0 to 20 GPM 316 SS Construction
5	FI-1	Flow Indicator Should read to 10 GPM (Provide)
6	SM-1	Koch Static Inline Mixer 3 Elements Install inside 1" Schedule 40 Pipe 316 SS
7	EP-1	¾" Penberthy-Houdaille Jet Pump 316 SS
8	PI-3	Pressure Gauge Ashcroft #18095 2½" Face 0 to 400 kPa 316 SS Tube
9	PI-1	Pressure Gauge Ashcroft #18095 2½" Face 0 to 250 kPa 316 SS Tube

Figure 30 – The EP & ID for spiractor installation



For clarity, support legs are not shown.
All dimensions in millimeters.

Figure 31.—Spiractor elevation.

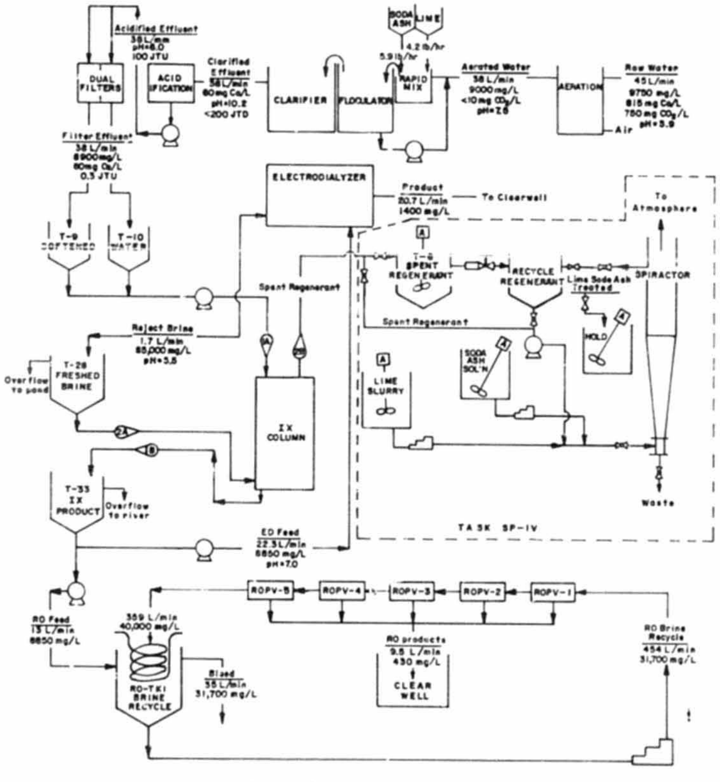


Figure 32.—Phase II total system flow sheet.

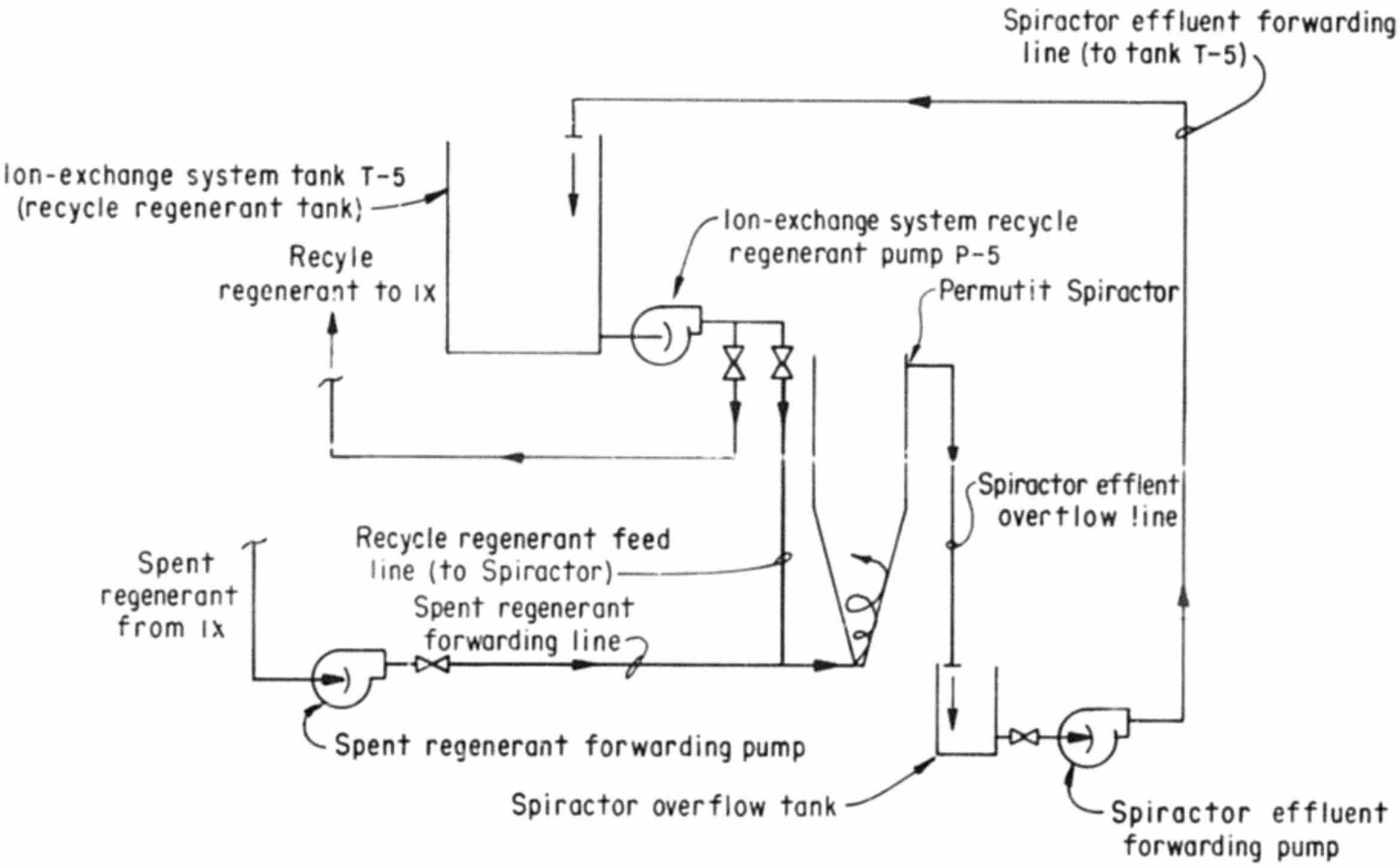


Figure 33.–Spiractor process flow diagram, regenerant desupersaturation testing.

EXPERIMENTAL ELECTRODIALYSIS PROCESS STUDY

Electrodialysis is an electrochemical separation technology [14] with the capability to reduce or concentrate ionized minerals or salts in water solution. The type of energy used to perform the separation process is direct current electricity. Separation media are two types of plastic sheets called membranes, which have the capability of selectively transferring either positively or negatively charged particles, called ions, in solution. Ions are electrically charged atoms or chemical groups.

Commercial units consist of at least one pair of electrodes and several hundred "cell pairs" of membranes and water flow spacers to achieve the required water production rate of the system. A single cell pair consists of four elements as shown in figure 34.

- Cation selective membrane
- Ion depleting cell
- Anion selective membrane
- Ion concentrating cell

With multiple membranes and water passages located between a pair of electrodes, the alternating passages contain ion concentrating or ion depleting solution. The flow paths are isolated, and the ED (electrodialysis) process operates with a continuous flow of water through each cell. Reduction in concentration of ions in the depletion passage is a direct function of the amount of electric current applied and the flow rate of water through a cell.

Membranes are designated as anion permeable (negative ion transfer), or cation permeable (positive ion transfer). Typical anions found in brine water are: chloride, sulfate, nitrate, and bicarbonate. Cations include: sodium, calcium, and magnesium. An anion permeable membrane is an inert organic molecular structure having fixed positively charged groups in place in the membrane. Normally, this is a quaternary amine type of group. Structure of the cation permeable membrane is similar except that a fixed negative charge is in place such as a sulfonate group. Under the driving force of an impressed direct current electrical field, positively charged cations will move in direction of the negative electrode (cathode) until they are repelled by an anion membrane. Negatively charged anions will move toward the positive electrode (anode) until repelled by a cation membrane. Ions in solution of positive charge will electrically transfer through the cation membrane by means of fixed negative charge groups. In like manner, negative ions will transfer through the anion membrane by means of fixed positive charge groups. Negative ions in solution will be repelled by negative fixed charges of the cation membrane with the opposite case being true for the anion membrane.

An electrochemical membrane cell can be constructed to remove and concentrate ions in water passages between membranes utilizing the selective transfer and rejection functions of the membranes along with a direct electric current as shown in figure 35. Water will exit from an ED cell membrane stack as two streams: a purified stream and a concentrated stream. In water desalting applications, the purified stream is the product while the concentrate goes to waste. The ratio of product to waste is normally controlled through concentrate recycle.

Removal of minerals from water by the ED process is a function of the amount of electric current applied. The process becomes more efficient with increasing temperature, up to a limit of 110 °F to 160 °F. The process efficiency is enhanced by hydraulic turbulence to aid in the transfer of ions. Process inefficiencies can be caused by scaling or fouling of membrane surfaces. Membrane efficiency also depends on the membrane materials and construction used since there are many different types of membranes for differing applications.

Addition of chemicals, such as acid, to the membrane stack may be required to avoid fouling of the membrane surfaces. Chemicals may also be added to improve stack operation and efficiency by increasing the electrical conductivity of water, thus reducing power requirements.

A special technique called polarity reversal has been developed to overcome membrane scaling and fouling problems without the addition of acids or other chemicals. With this process, the electrical polarity of electrodes is reversed three or four times per hour. At the same time, automatic valves reverse stream flow so depleting and concentrating flow paths are interchanged. This reversal causes ions to flow in opposite direction, breaking up concentrated scale-forming layers at the membrane surfaces. Colloids carrying electrical charges are also removed from the membrane and carried off in the waste stream.

Certain materials, when present in the feed water, should be removed by pretreatment. Iron or hydrogen sulfide concentrations above 0.3 mg/L, manganese concentrations above 0.1 mg/L, and free chlorine must be avoided. Since most natural brines contain sulfate ion, lime, and possibly soda ash, pretreatment is required to reduce the calcium ion concentration. The soluble calcium content of the feed water needs to be reduced to a level which prevents membrane fouling by precipitation of calcium sulfate in the brine. These constituents, along with turbidity in excess of 2 JTU (Jackson Turbidity Units), are normally controlled by means of standard aeration chemical treatment, filtration, or ion exchange techniques. These limitations apply to water desalting, and their control is essential to maintain optimum operation and cost performance.

A. Equipment Description

The same Ionics, Inc., AQ-V (Aquamite V) used at YDTF for Yuma High Recovery studies [9] was used at the LVS test site. The Aquamite V flow diagram, equipment layout, and service connection facsimile drawings are given in appendix G as figures G.18, G.19, and G.20.

The equipment utilized for LVS testing consisted of the Ionics AQ-V ED desalting unit opewrated as shown in figure 36.

1. *Stack configuration.* — Figure 37 presents the ED stack configuration and stack flow diagram. This equipment utilized a two-stage electrical, six-stage hydraulic stack configuration with 260 cell pairs. Each cell pair consisted of an anion and cation membrane with spacer placed between each individual membrane. Special heavy cation membranes were used to separate the various hydraulic stages. Membranes used were AR204 SXZL-183 anion membrane, CR61 CZL-183 cation membrane, and CR61 AZL-065 heavy cation membrane. Ionics Inc., membrane specifications are given in appendix I.

2. *Design conditions.* The Aquamite V demineralizer was based on the following design condition:

Feed water salinity	8440 mg/L	0.136 normality
Product water salinity	1800 mg/L	0.029 normality
Feed water temperature	93 °F	33.9 °C
Production per 24-hour-day (corrected for water transfer)	7350 gal (US)/day	27.8 m^3/day
Total waste per 24-hour-day[1]	605 gal (US)/day	2.3 m^3/day
Brine makeup per 24-hour-day	132 gal (US)/day	0.5 m^3/day
Electrical energy consumption per 1,000 gallons product	16.4 kWh/1000 gal	
Power characteristics	460 V	3-phase 60 cycle
Input power requirements	35 kVA	

[1] Assuming electrode stream reuse

3. *Proposed operating conditions.* — The Aquamite V demineralizer was to operate at the following conditions:

Temperature	93 °F
Dilute flow	5.85 gal/min
Brine make-up flow	0.09 gal/min (for maximum brine salinity)
Stack pressure drop (stack inlet pressure minus stack outlet pressure)	42 lb/in²
Feed water TDS	8440 ms/L
Product water TDS	1800 mg/L
Stage 1 — Volts	98
Amps	25.7
Stage 2 — Volts	91
Amps	17.3
Product conductivity	2800 micromhos (approximate)
Input power requirement	460 V, 3ϕ, 60 Hz

4. *Stack components.* — The electrodialysis stack used at the LVS test site was made up of the following components:

Stack components	Number (per stack)	Type
Anion membranes	260	AR204 SXZL-183
Cation membranes	254	CR61 CZL-183
Heavy cation membranes	8	CR61 AZL-065
Spacers	520	MKII-4
Special electode spacer	4	MKII-4
Electrodes	4	Platinized titanium or niobium

B. Process Operations

When all the stack components are alined and compressed together, the manifold holes form flow paths to carry dilute and brine streams to and from the spacers. These manifolds are blocked in certain places to force the water through the spacers. For instance, the dilute inlet stream will enter one of the manifolds through the top of the stack. The heavy cation membrane which separates the first and second hydraulic stages is positioned to block inlet manifold holes at the bottom of the first stage. This blockage creates the back pressure needed to drive the dilute stream through spacers which have a flow patch through the manifold. The stream will exit the spacers through the other manifold to which the flow path is connected. This manifold will then be the inlet to the next hydraulic stage. By alternating the position of heavy cation membranes down through the stack (and therefore alternating those manifolds which are blocked), the flow paths make a number of passes across the stack.

Under most operating conditions, the Ionics AQ-V would use ED feed water for electrode rinsing. This operation was changed for the LVS test site operation by installing an electrode rinse pump which uses IX feed water from the clearwell. This change was done to reduce the amount of feed water utilized and increase overall system on-line time.

The ED unit was operated intermittently at LVS from January 10 through August 19, 1980. The opration was conducted in this manner because the primary objective was to produce brine in support of the IX test program, and insufficient personnel were available to monitor and operate both IX and ED on a continuous basis. No long-term, steady-state operating data were obtained.

C. Discussion of Experimental Task

The ED operation had the following objectives:

Determine the operating performance of the ED system while desalting LVS water at different water recovery levels, and

Provide regenerant brine concentrations as required for IX tasks II, III, and IV.

The following experimental ED tasks were performed:

- Task ED-I, stack assembly and hydraulic testing
- Task ED-II, startup and ED system performance test
- Task ED-III, operation of ED system to provide the regenerant brine concentrates required for IX tasks II, III, and IV
- Task ED-IV, additional ED cell performance studies
- Task ED-V, membrane stack inspection

1. *Task ED-I, stack assembly and hydraulic testing.* — Task ED-I assembled the Aquamite V membrane cell stack and hydraulically test the electrodialysis pilot plant by January 1, 1980.

- The membrane cell stack was assembled with six hydraulic stages and two electrical stages. An electrode feed pump for 3 L/min of flow at 275 kPa was installed.

- After completion of assembly of the membrane stack, the ED system was hydraulically tested to correct leaks, condition equipment, and evaluate and adjust operations of pumps and flow control devices.

2. *Task ED-II, startup and ED system performance test.* — Task ED-II required an Ionics field engineer at the LVS site to direct the electrical modifications of the AQ-V system and to conduct an ED performance test.

The Ionics field engineer established electrical power control settings. He optimized flow rates, minimize cell voltages, and generally tuned the operation of the system.

After the electrical operation was established, the ED system was operated through polarity reversals at different water recoveries to obtain the data summarized in table 45. Data were collected just prior to the end of each 30-minute polarity reversal when steady-state conditions have been achieved.

a. Site operations for performance test. — Performance tests wre comducted by the Ionics field engineer, who arrived on January 8, 1980. Startup of the ED was attempted on January 8; failure of the brine pump delayed testing until January 10. At that time, it was discovered that two improperly sized electrode spacers had been installed. The stack was disassembled and the electrode spacers were replaced. On January 11 and 12, the ED unit was operated by the Ionics' engineer at the conditions shown in table 46. Test 1 was conducted with a 15-minute cycle duration at a recovery of approximately 90 percent to provide data at the normal cycle duration. All other tests were conducted with a 30-minute cycle duration.

b. Performance at various water recoveries. — Task ED-II was conducted with the Aquamite V ED cell system to demonstrate performance at various water recovery levels. The data obtained are summarized intable 45 and plotted as figure 38. The feed water was controlled as shown. Feed TDS was held constant at 8800 mg/L for water recoveries from 72.8 through 92.8 percent. The ED system had no temperature control capability, and temperature variations shown in figure 38 were due to fluctuations in ambient air temperature at the time of specific water recovery test. Target feed control rate was 22.5 L/min. At higher water recoveries, feed flow rate was difficult to hold constant, and it varied between 21.5 and 22.5 L/min. Holding feed TDS and flow rate constant, product TDS rose from 1160 mg/L to 1268 mg at 72.8 and 92.8 percent water recoveries.

Stack responses are plotted at the bottom of figure 38. Stack emf (electromotive force, or voltage) remained constant between 72.8 and 88.5 percent water recoveries. The emf for stage 1 was 109 V and for stage 2 was 105 V. Both stacks 1 and 2 dropped 1 V between 88.5 and 92.1 percent recovery and another 1 V when water recovery was increased to 92.8 percent. The current in stack 1 raised gradually from 22.8 A to 25.4 A and in stack 2, from 15.0 A to 16.5 A between 72.8 and 88.5 percent water recoveries. Power consumption rose gradually

from 4.1 to 4.4 kW between 72.8 and 88.5 percent water recoveries. Power consumption then rose at a more rapid rate to 4.9 kW at 92.8 percent water recovery.

3. *Task ED-III, ED operation in support of the IX test program.* — Operation of the ED in support of IX test program began on January 12 and continued until April 18. The system was operated initially at 80 percent recovery. On February 15, the 80 percent recovery IX cycles were terminated, and the ED desalting recovery was increased to 90 percent. It operated at this value until just prior to the end of the IX test program on April 21.

Site operations and ED desalting performance. — The AQ-V ED unit performance data are summarized in table 47. Analyses of the feed, product, and brine streams are summarized in table 48.

4. *Task ED-IV, additional ED cell performance studies.* — Special ED testing was requested to ensure that sufficient data would be available to the Bureau designers to allow them to properly evaluate the ED desalting process and its potential application to the treatment of the LVS and similar waters. The data requested included normal operating data and laboratory samples at 80 and 90 percent recoveries. Data and samples were collected during both the positive and negative polarities. In addition, midstack samples were collected.

a. Site operations and ED desalting performance. — Beginning on March 26, while the IX test program was still underway, the first 90 percent data were collected on March 28, April 2, and April 3. The first 80 percent recovery data were not collected untril April 19, just prior to the end of the IX test program. The change to 80 percent recovery was possible at this time because sufficient 90 percent brine inventory was on hand to complete the testing. The last three 80 percent recovery runs were all conducted on April 21. The 80 percent recovery data are presented in tables 49 and 50; the 90 percent recovery data are presented in tables 51 and 52.

An analysis of the operating data obtained at 80 percent recovery indicates excellent correlation among all four sets of data. A comparison of current efficiencies for these four runs indicates a fairly wide range. The average current efficiency is 81.6 percent, but the standard deviation is 10.6. Current efficiency was derived from the stack currents, voltages, and feed product cationic concentrations in meq (milliequivalents). The evaporative feed TDS values averaged 8800 g/m^3 with a standard deviation of only 36. The calculated feed TDS values (based upon a summation of ions) averaged 8935 g/m^3, but with a standard deviation 1036. It was concluded that the data presented in table 49, check no. 3, represent the only run with proper correlation of the evaporative and calculated TDS values. The other ED process stream samples among the 80 percent recovery data with poor correlation of the TDS values are all midstage samples. All other samples have excellent correlation.

Analysis of the 90 percent recovery data indicates excellent correlation of the operating data among the four sets. A comparison of the current efficiencies for these sets of data indicates, once again, the laboratory analysis results are suspect. Average current efficiency was 75.1 percent with a standard deviation of 8.6. Special checks 1, 2, and 4 compare very favorably with an average current efficiency of 71.0 percent and a standard deviation of only 2.8. Special check 3 (table 51) had a current efficiency of 87.5 percent.

A comparison of the TDS values indicates two different occurrences that influence the data. The first is that the feed TDS (evaporative) was significantly higher for special checks 3 and 4 than for the other two runs. Evaporative TDS for special checks 3 and 4 averaged 9350 g/m^3 with a standard deviation of 127. The feed TDS of checks 1 and 2 compare very favorably with the TDS (evaporative) of the feed water of the 80 percent checks.

The reason for the increased feed TDS is unknown. However, during this time, severe scaling problems were being encountered during the regeneration of the IX. This might account for the poorer quality of the ED feed.

The other factor causing the high standard deviation in the calculated current efficiency of special check 3 (table 51) is deviation between individual evaporative and calculated TDS values. This difference is approximately 1.1, 8.0, 21.6, and 0.2 percent for special checks 1, 2, 3, and 4 (table 51), respectively.

Based upon the foregoing data, it can be seen that the data from special checks 1 and 2 are more representative than from checks 3 and 4. The data for the later two special checks should probably not be used for evaluating the system.

b. Performance at 80 and 90 percent water recovery levels. — Task ED-IV was an analytical and electrical data acquisition study of the electrodialysis pilot plant operation at 80 and 90 percent water recovery levels. A summary of the average mass flow and electrical data obtained are given as table 53. Data were obtained at constant feed water conditions of 375 x 10^{-6} m^3/s and 8800 g/m^3 and at voltage and product flow balance conditions required for 80 and 90 percent water recoveries.

Cell pair resistivities for the period January 28, 1980 through March 17, 1980, are summarized in table 54 and plotted as figure 39. Resistivities remained constant during opreation at 80 and 90 percent water recovery levels. The first electrical stage averaged 0.00475 and 0.00385 ohm-m^2 at 80 and 90 percent recoveries, respectively. Normally, cell resistivities rise with operating time due to gradual fouling of the membranes. The constant resistivities experienced at LVS reflect on the high quality of the feed water used for electrodialysis. The feed water for ED went through a three-step pretreatment of aeration, partial lime treatment, and ion exchange. Table 3 (earlier in report) showed that the ED feed water had a desalting percent water recovery based on calcium sulfate solubility of 92.0 percent at 20 °C, i.e., the ED cell system could be operated up to 92 percent water recovery before any calcium sulfate membrane fouling would occur.

5. *Task ED-V, membrane stack inspection.* —Task ED-V involved opening the membrane stack, individually inspecting and logging the condition of each component, sampling deposits of scale and slime, hand cleaning each component, and reassembling the membrane stack. The ED system was brought back to first-class operating condition by draining and flushing all tanks, pumps, and pipes in the hydraulic system.

On April 29, inspection started at the top and removed one component at a time. The condition of each component giving the size and distribution of any damage, scale, and slime are summarized in table 55. The quantities of deposits found in the stack were very small, and the number of deposits, as indicated in the table, were very few. The overall condition of the ED stack was excellent.

The ED stack, system piping, valves, and pumps were thoroughly flushed with IX product water. On April 30, the stack was reassembled, and the system returned to operating condition.

D. Optimum Steady-State Conditions

The optimum steady-state conditions for the AQ-V at 90 percent water recovery of A-L-IX pretreated LVS water were as follows:

	Units	Process conditions
Brine concentration, TDS	g/m^3	93 800
Temperature	°C	24.1
Feed flow rate	L/min	21.8
Brine makeup flow rate	L/min	0.2
Product flow rate	L/min	20.5
Reject flow rate	L/min	1.5
Reject pH		5.8
Electrical stage 1		
Current	A	31.0
emf	V	105
Specific cell pair resistance	cm^2	39.1
Electrical stage 2		
Current	A	18.1
emf	V	98
Special cell pair resistance	cm^2	63.7
Feed cation concentration	meq/L	143.2
Product cation concentration	meq/L	17.6
Current efficiency	%	70.9

Water quality changes

	Units	ED streams		
		Feed	Dilute	Concentrate
pH		7.7	7.1	6.9
TDS (evaporated at 105 °C)	g/m^3	8660	1060	93 800
Conductivity at 25 °C	mS/m	1340	206	7 540
Silica	g/m^3	10	9	14
Calcium	g/m^3	32	2	320
Magnesium	g/m^3	78	2	439
Sodium	g/m^3	2940	406	32 400
Potassium	g/m^3	188	16	2 480
Iron	g/m^3	8.16	0.04	2.1
Manganese	g/m^3	0.01	ND	1.1
Bicarbonate	g/m^3	43	14	96
Carbonate	g/m^3	0	0	0
Sulfate	g/m^3	1520	197	21 900
Chloride	g/m^3	4440	490	35 200
Anions	meq/L	157.4	18.1	1450
Cations	meq/L	143.3	18.4	1525

Table 45. – *Performance of Aquamite V ED cell at various water recovery levels*

		Parameter	Units	Percent water recovery 1	 2	 3	 4	 5	 6	 7
				72.8	79.3	87.8	88.5	90.0	92.1	92.8
Feed	1	TDS	g/m³	8851	8851	8851	8851	8757	8851	8851
	2	Conductivity	mS/m	140	140	140	140	140	140	140
	3	Temperature	°C	20.5	20.0	20.5	17.3	20.0	20.0	18.0
	4	Flow	L/min	22.5	22.5	22.1	22.2	21.5	22.5	22.2
	5	Cycle duration	min	30	30	30	30	15	30	30
	6	Volume	L	675	675	663	666	323	675	666
	7	Quantity of salt	g	5974	5974	5868	5895	2824	5974	5895
Product	8	TDS	g/m³	1160	1183	1011	1450	1200	1268	1268
	9	Conductivity	mS/m	250	250	240	270	260	260	280
	10	Flow	L/min	16.5	19.9	19.5	20.2	19.5	20.4	21.0
	11	Cycle duration	min	30	30	30	30	15	30	30
	12	Volume	L	495	597	585	606	293	612	630
	13	Quantity of salt	g	574	706	591	879	351	776	799
	14	TDS project	g/m³	29 426	38 278	65 241	65 802	54 718	92 398	106 876
	15	Conductivity	mS/m	3700	4800	6250	7200	8800	9700	10 000
	16	Flow	L/min	5.0	5.3	2.9	2.55	1.5	1.5	1.03
	17	Cycle duration	min	30	30	30	30	15	30	30
	18	Volume	L	150	159	87	77	23	45	31
	19	Quantity of salt	g	4414	6086	5676	5034	1231	4158	3302
Stack	20	Pressures, in	kPa	305	304	310	312	302	310	313
	21	Pressures, out	kPa	292	281	296	293	285	308	299
	22	Differential pressure	kPa	12.8	23.5	13.7	18.6	16.7	2.0	13.7
	21	emf, stage 1	V	109	109	109	108	108	109	107
	22	emf, stage 2	V	105	105	105	104	104	104	103
	23	Current stage 1	A	22.8	23.4	25.4	25.2	28.0	25.8	28.4
	24	Current stage 2	A	15.0	15.2	15.9	16.5	17.0	17.7	18.4
	25	Power stage 1	W	2485	2551	2769	2722	3024	2812	3039
	26	Power stage 2	W	1575	1596	1670	1716	1768	1841	1895
	27	Power stages 1 and 2	W	4060	4147	4439	4438	4792	4653	4934

Table 47. — *Electrodialysis unit performance, Ionics Aquamite V*

	Units	1/28/80	2/11/80	2/19/80[b]	2/25/80
Operating time	hr	9 693.0	9 786.5	9 907.4	9 988.8
Brine concentration	g/m^3	41 100	41 600	96 800	99 200
Recovery	%	79.6	80.2	91.6	90.7
Feed temperature	°C	23.8	23.8	22.0	23.8
Dilute flow rate	$m^3/s \times 10^{-6}$	366.7	366.7	366.7	366.7
Brine make-up flow rate	$m^3/s \times 10^{-6}$	63.3	61.7	3.3	0[d]
Product flow rate	$m^3/s \times 10^{-6}$	——[a]	333.3	343.3	346.7
Brine flow rate	$m^3/s \times 10^{-6}$	80.0	71.7	25.0	30.0
Brine pH		6.0	6.3	——[c]	5.6
Electrical stage 1					
Current	amperes	26.6	26.2	29.7	28.9
Voltage	volts	109	109	109	97[e]
Specific cell pair resistance	$\Omega \cdot m^2 \times 10^{-3}$	4.71	4.78	4.06	3.85
Electrical stage 2					
Current	amperes	15.5	15.6	17.7	19.1
Voltage	volts	104	104	104	97[e]
Specific cell pair resistance	$\Omega \cdot m^2 \times 10^{-3}$	7.70	7.65	6.50	5.82
Feed cationic concentration	eq/m^3	142.6	147.2	145.2	146.7
Product cationic concentration	eq/m^3	14.7	18.2	12.4	21.6
Current efficiency	percent	82.6	84.0	76.3	71.0

[a] No product flow rate presented because of erroneous data recorded.

[b] Because of operator error, ED conditions were not recorded at the time samples were collected. Data from conditions taken on February 20, 1980, are presented as being representative of the conditions which existed when the samples were collected on February 19, 1980.

[c] Brine pH was not recorded when unit conditions were taken.

[d] This zero brine make-up is apparently due to operator error during the data collection.

[e] The stage voltages were reduced prior to collecting this data because of low product conductivity readings were being encountered prior to the change.

Table 47. — *Electrodialysis unit performance, Ionics Aquamite V — Continued*

	Units	3/3/80	3/10/80	3/17/80
Operating time	hr	0096.0	0210.7	0351.5
Brine concentration	g/m^3	96 500	96 500	95 940
Recovery	%	90.3	90.6	91.6
Feed temperature	°C	26	25.1	23.1
Dilute flow rate	$m^3/s \times 10^{-6}$	336.7	365.0	366.7
Brine make-up flow rate	$m^3/s \times 10^{-6}$	2.3	3.3	2.3
Product flow rate	$m^3/s \times 10^{-6}$	248.3	345.0	350.0
Brine flow rate	$m^3/s \times 10^{-6}$	26.7	26.7	20.0
Brine pH		5.7	5.5	6.2
Electrical stage 1				
Current	amperes	29.9	31.9	30.4
Voltage	volts	97	104	105
Specific cell pair resistance	$\Omega \cdot m^2 \times 10^{-3}$	3.90	3.85	3.91
Electrical stage 2				
Current	amperes	17.9	18.1	18.3
Voltage	volts	92	98	100
Specific cell pair resistance	$\Omega \cdot m^2 \times 10^{-3}$	6.17	6.38	6.18
Feed cationic concentration	eq/m^3	150.8	148.9	140.5
Product cationic concentration	eq/m^3	23.1	20.5	18.5
Current efficiency	percent	72.7	69.5	68.1

Table 48. — *Chemical analysis, Ionics Aquamite V unit*

Date		1/28/80			2/11/80			2/19/80			2/25/80		
Date analyzed at Boulder City		3/11/80			3/17/80			3/17/80			3/18/80		
		Feed	Product	Brine	Feed	Product	Brine	Feed	Product	Brine	Feed	Product	Brine
pH	Units	7.8	7.4	7.3	7.9	7.5	7.4	7.8	7.5	6.6	7.7	6.9	7.7
TDS (calculated)	g/m^3	9 110	909	41 100	9 140	1 150	41 600	9 170	775	96 800	9 310	1 360	99 200
TDS (evaporated @ 180 °C)	g/m^3	9 180	905	41 300	8 880	1 130	41 300	8 950	752	98 000	9 280	1 370	99 700
Conductivity (@ 25 °C)	mS/m	1 350	165	5 050	1 360	205	5 090	1 360	144	9 850	1 390	245	9 830
Silica	g/m^3	13	12	15	12	12	17	14	13	22	18	16	25
Calcium	g/m^3	95	5.6	420	60	2.4	310	50.0	0	430	50.0	0	490
Magnesium	g/m^3	64.1	4.4	287	54.9	3.9	275	36.6	2.0	348	36.6	4.3	415
Sodium	g/m^3	2 880	315	13 500	3 110	402	13 900	3 110	274	33 400	3 110	475	32 800
Potassium	g/m^3	284	13.3	684	176	12.5	860	176	10.9	1 740	235	21.5	2 420
Iron, total	g/m^3	0.04	0	0.29	0.06	0	0.42	0.07	0	1.0	0.06	0	0.72
Manganese, total	g/m^3	<0.05	<0.05	0.11	<0.05	<0.05	0.13	<0.05	<0.05	0.32	<0.05	<0.05	0.22
Strontium	g/m^3	0	0.1	6.0	1.5	0	5.5	1.0	0.1	9.0	0.5	0	5.3
Bicarbonate	g/m^3	45.1	20.1	70.2	46.4	22.0	82.4	43.9	12.2	107	45.1	23.2	54.9
Carbonate	g/m^3	0	0	0	0	0	0	0	0	0	0	0	0
Sulfate	g/m^3	2 120	133	9 580	2 040	196	9 620	2 100	90.2	22 800	2 210	197	25 500
Chloride	g/m^3	3 630	415	16 600	3 660	513	16 600	3 660	378	38 000	3 630	636	37 500
T-alkalinity as CaCO$_3$	g/m^3	37	17	58	38	18	68	36	10	88	37	19	45
P-alkalinity as CaCO$_3$	g/m^3	0	0	0	0	0	0	0	0	0	0	0	0
T-phosphorus as PO$_4$	g/m^3	0.9	0.7	0.7	0.5	0.4	0.4	0.5	1.4	1.6	1.2	1.2	0.6
E.F. [TDS(calculated)/cond.]		6.7	5.5	8.1	6.7	5.6	8.2	6.7	5.4	9.8	6.7	5.6	10.1
E.F. [TDS(evaporated)/cond.]		6.8	5.5	8.2	6.5	5.5	8.1	6.6	5.2	9.9	6.7	5.6	10.1
Σ Anions	eq/m^3	147.3	14.79	667.4	146.4	18.88	668.6	147.6	12.74	1 549.0	149.0	22.40	1 589.0
Σ Cations	eq/m^3	142.3	14.66	648.0	147.0	18.24	666.0	145.0	12.36	1 544.0	146.5	21.55	1 549.0
Control value	eq/m^3	+2.09	+0.39	+1.86	-0.25	+1.60	+0.25	+1.09	+1.25	+0.21	+1.03	+1.87	+1.62

Table 48. — *Chemical analysis, Ionics Aquamite V unit — Continued*

Date		3/3/80			3/10/80			3/17/80		
Date analyzed at Boulder City		3/21/80			3/25/80			5/1/80		
		Feed	Product	Brine	Feed	Product	Brine	Feed	Product	Brine
pH		6.9	6.7	6.3	7.3	7.4	6.8	7.5	7.1	7.0
TDS (calculated)	g/m^3	9 380	1 420	96 500	9 180	1 280	96 500	9 330	1 110	95 940
TDS (evaporated @ 180 °C)	g/m^3	9 160	1 490	94 900	9 080	1 250	96 100	9 060	1 150	95 600
Conductivity @ 25 °C	mS/m	1 390	272	9 610	1 370	225	9 670	1 350	282	8 450
Silica	g/m^3	11	10	20	16	19	26	8.0	8.0	7.5
Calcium	g/m^3	45	1.8	390	35	5.0	310	32.0[b]	3.20	192
Magnesium	g/m^3	39.7	1.6	451	51.9	3.1	506	48.8	1.95	76.1
Sodium	g/m^3	3 220	511	32 800	3 170	449	33 000	2 940	406	34 000
Potassium	g/m^3	205	23.5	1 990	196	19.6	1 960	275	21.9	2 970
Iron, total	g/m^3	0.05	0	0.77	0.04	0	0.72	0.04	0.02	0.35
Manganese, total	g/m^3	<0.05	<0.05	0.24	<0.05	<0.05	0.30	0.02	0.01	0.09
Strontium	g/m^3	0.3	0	2.8	0.3	0	2.3	0.07	ND	1.75
Bicarbonate	g/m^3	44.5	20.1	85.4	40.3	26.2	68.9	44.5	15.9	93.9
Carbonate	g/m^3	0	0	0	0	0	0	0	0	0
Sulfate	g/m^3	2 150	181	23 900	2 040	190	22 900	2 100	76.8	23 000
T-alkalinity as CaCO$_3$	g/m^3	37	17	70	33	22	57	42[a]	20[a]	90[a]
P-alkalinity as CaCO$_3$	g/m^3	0	0	0	0	0	0	0[a]	0[a]	0[a]
T-phosphorus as PO$_4$	g/m^3	0.6	0.5	0.2	1.1	0.8	0.2	-	-	-
E.F. [TDS (calculated)/cond.]		6.7	5.2	10.0	6.7	5.7	10.0	6.9	3.9	11.4
E.F. [TDS (evaporated)/cond.]		6.6	5.5	9.9	6.6	5.6	9.9	6.7	4.1	11.3
Σ Anions	eq/m^3	149.5	23.14	1 539.0	145.9	20.72	1 545.0	146.73	18.06	1 480.5
Σ Cations	eq/m^3	150.8	23.02	1 533.0	149.0	20.50	1 542.0	140.63	18.58	1 571.9
Control value	eq/m^3	-0.54	+0.26	+0.25	-1.30	+0.51	+0.12	+2.56	-1.35	-3.96

[a] Analysis conducted at LVSTS

[b] This concentration was originally reported as 320 g/m^3 but was changed

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Table 49. — *Electrodialysis unit performance at 80% recovery*

Date		4/19/80	4/21/80	4/21/80	4/21/80
Special check number		1	2	3	4
Operating time	hr	1023.7	1059.8	1061.8	1064.8
Brine concentration (TDS)	g/m^3	39 840	39 700	40 700	40 500
Feed temperature	°C	32.7	31.0	30.9	31.1
Dilute flow rate	$m^3/s \times 10^{-6}$	366.7	366.7	366.7	366.7
Brine make-up flow rate	$m^3/s \times 10^{-5}$	60.0	58.3	56.7	56.7
Product flow rate	$m^3/s \times 10^{-6}$	345.0	350.0	350.0	348.3
Brine flow rate	$m^3/s \times 10^{-6}$	86.0	85.0	82.5	81.7
Brine pH		6.3	6.1	6.2	6.2
Electrical stage 1					
Current	amperes	27.1	26.2	26.3	26.2
Voltage	volts	94	94	95	93
Specific cell pair resistance	$\Omega \cdot m^2 \times 10^{-3}$	4.88	4.84	4.86	4.80
Electrical stage 2					
Current	amperes	15.0	15.3	15.4	15.7
Voltage	volts	88	88	88	88
Specific cell pair resistance	$\Omega \cdot m^2 \times 10^{-3}$	8.24	7.74	7.68	7.57
Feed cationic concentration	eq/m^3	124.5	164.3	139.7	147.6
Product cationic concentration	eq/m^3	14.6	17.9	20.7	21.4
Current efficiency	percent	71.0	96.0	77.6	82.0

Table 50. — *Chemical analysis for ED performance at 80 percent recovery*

		Feed	Positive Dilute	Positive Mid-Stage Dilute	Positive Concentr.	Negative Dilute	Negative Mid-Stage Dilute	Negative Concentr.	Feed	Positive Dilute	Positive Mid-Stage Dilute	Positive Concentr.	Negative Concentr.	Negative Mid-Stage Concentr.	Negative Dilute
Date		4/19/80	4/19/80	4/19/80	4/19/80	4/19/80	4/19/80	4/19/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80
Time collected		1832	1832	1832	1832	1902	1902	1902	0932	0932	0932	0932	1003	1003	1003
Date analyzed at Denver		6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80
pH		7.8	7.3	7.7	7.1	7.4	7.3	7.4	7.8	7.0	7.5	7.6	7.6	7.6	7.2
TDS (calculated)	g/m³	7 660	890	4 640	39 720	1 060	38 560	40 620	10 060	1 190	5 140	40 920	40 890	40 320	1 210
TDS (evaporated @ 105 °C)	g/m³	8 760	880	3 560	39 840	998	37 600	39 500	8 780	1 010	3 900	39 700	39 400	37 600	1 030
Conductivity @ 25 °C	mS/m	1 340	168	595	4 870	191	5 100	4 660	1 360	188	696	4 660	4 870	4 460	199
Silica	g/m³	6.6	6.8	6.7	5.0	6.1	5.[illegible]	5.0	10.2	10.3	9.6	7.7	-	7.2	9.4
Calcium	g/m³	28.8	1.60	3.20	192	1.60	16[illegible]	181	32.0	1.60	9.60	208	208	208	1.60
Magnesium	g/m³	42.0	2.93	9.76	276	5.37	2[illegible]4	222	46.8	3.66	781	234	215	234	2.93
Sodium	g/m³	2 650	324	1 620	13 100	372	12 900	14 100	3 560	398	1 860	14 100	14 300	13 300	415
Potassium	g/m³	170	8.99	46.9	832	9.78	851	904	161	9.78	50.0	814	825	778	8.99
Iron, total	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Manganese, total	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Strontium	g/m³	0.1	0.2	0.2	0.7	ND	0.7	0.8	0.2	ND	0.3	1.1	ND	1.3	ND
Bicarbonate	g/m³	40.9	17.1	31.1	53.7	17.1	65.3	59.2	37.8	15.3	28.7	56.1	42.7	51.2	15.3
Carbonate	g/m³	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sulfate	g/m³	842	93.1	1 430	9 920	181	9 540	9 850	2 730	357	1 570	10 500	10 000	9 340	290
Chloride	g/m³	3 880	433	1 490	16 340	469	14 800	15 300	3 480	398	1 600	15 000	15 300	16 400	469
T-alkalinity as $CaCO_3$*	g/m³	33.5	14.0	22.5	44.0	14.0	53.6	48.5	31.0	12.6	23.5	46.0	35.0	42.0	12.6
P-alkalinity as $CaCO_3$*	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS (calculated)/cond.]		5.7	5.3	7.8	8.2	5.5	7.6	8.7	7.4	6.3	7.4	8.8	8.4	9.0	6.1
E.F. [TDS (evaporated)/cond.]		6.5	5.2	6.0	8.2	5.2	7.4	8.5	6.5	5.4	5.6	8.5	8.1	8.4	5.2
Σ Anions	eq/m³	127.17	14.42	72.31	639.88	17.25	616.07	637.97	155.52	18.88	78.37	642.92	641.70	657.84	19.50
Σ Cations	eq/m³	124.23	14.65	72.66	621.50	16.97	608.40	663.34	164.56	17.93	83.40	663.40	671.10	626.50	18.55
Control value	eq/m³	+1.42	-0.70	-0.29	+1.83	+0.75	+0.79	-2.54	-3.59	+2.38	-3.81	-2.03	-2.92	+3.04	+2.32

*Alkalinities were not reported by the Denver lab. Values have been calculated from the bicarbonate concentration.

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Table 50. — *Chemical analysis for ED performance at 80 percent recovery — Continued*

		Feed	Positive Dilute	Positive Mid-Stage Dilute	Positive Concentr.	Negative Concentr.	Negative Mid-Stage Concentr.	Negative Dilute	Feed	Positive Dilute	Positive Mid-Stage Dilute	Positive Concentr.	Negative Concentr.	Negative Mid-Stage Concentr.	Negative Dilute
Date		4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80	4/21/80
Time collected		1132	1132	1132	1132	1203	1203	1203	1432	1432	1432	1432	1503	1503	1503
Date analyzed at Denver		7/11/80	7/11/80	7/11/80	7/11/80	7/11/80	7/11/80	7/11/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80	6/19/80
pH		7.9	7.4	7.7	7.8	7.5	7.8	7.4	7.6	6.9	7.4	7.0	6.8	7.1	7.2
TDS (calculated)	g/m^3	8 610	1 240	3 990	38 380	38 440	36 940	1 130	9 410	1 360	4 520	40 590	40 080	38 190	1 270
TDS (evaporated @ 105 °C)	g/m^3	8 820	1 130	3 940	40 700	39 900	38 200	1 080	8 840	1 070	3 970	40 500	40 700	38 700	1 130
Conductivity @ 25 °C	mS/m	1 350	195	614	3 170	4 830	4 610	253	1 470	207	653	4 870	4 870	5 100	221
Silica	g/m^3	9.5	9.6	9.4	6.9	9.6	6.8	9.6	9.6	11.1	10.4	8.5	6.7	8.4	10.3
Calcium	g/m^3	48.0	1.20	16.0	240	224	224	16.0	60.8	1.60	6.40	256	256	256	1.60
Magnesium	g/m^3	58.6	18.8	39.0	224	283	254	24.4	62.5	5.37	16.1	312	312	293	5.86
Sodium	g/m^3	2 940	433	1 340	13 000	12 900	12 600	369	3 090	472	1 510	13 200	12 900	12 300	433
Potassium	g/m^3	181	10.2	54.3	843	796	796	10.2	198	12.1	61.4	977	977	951	12.5
Iron, total	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Manganese, total	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Strontium	g/m^3	0.3	ND	ND	1.2	1.3	1.2	ND	0.2	0.1	0.5	1.4	1.5	1.3	0.3
Bicarbonate	g/m^3	36.0	15.3	26.8	37.8	32.9	45.8	15.3	35.4	14.6	26.8	37.8	32.3	44.5	15.3
Carbonate	g/m^3	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sulfate	g/m^3	1 960	211	847	9 630	9 600	9 110	214	2 680	406	1 360	10 500	10 400	9 740	354
Chloride	g/m^3	3 380	540	1 660	14 400	14 600	13 900	469	3 270	433	1 530	15 300	15 200	14 600	433
T-alkalinity as $CaCO_3$*	g/m^3	29.5	12.6	22.0	31.0	27.0	37.6	12.6	29.0	12.0	22.0	31.0	26.5	36.5	12.6
P-alkalinity as $CaCO_3$*	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS (calculated)/cond.]		6.4	6.4	6.5	12.1	8.0	8.0	4.5	6.4	6.6	6.9	8.3	8.2	7.5	5.8
E.F. [TDS (evaporated)/cond.]		6.5	5.8	6.4	12.8	8.3	8.3	4.3	6.0	5.2	6.1	8.3	8.4	7.6	5.1
Σ Anions	eq/m^3	136.59	19.85	64.94	607.62	612.54	582.75	17.91	148.38	20.89	72.04	650.62	645.53	615.73	19.82
Σ Cations	eq/m^3	139.83	20.66	63.79	619.00	614.80	602.40	19.06	147.23	21.33	69.01	637.40	623.40	595.10	19.68
Control value	eq/m^3	-1.46	-1.96	+1.03	-1.19	-0.24	-2.15	-2.99	+0.48	-1.02	+2.48	+1.30	+2.19	+2.14	+0.34

*Alkalinities were not reported by the Denver lab. Values have been calculated from the bicarbonate concentration.

Table 51. — *Electrodialysis unit performance at 90 percent recovery*

Special check number Date		1 3/26/80	2 3/28/80	3 4/2/80	4 4/3/80
Operating time	hr	0546.1	0587.0	0670.9	0694.9
Brine concentration (TDS)	g/m^3	98 300	94 900	94 800	94 800
Feed temperature	°C	24.8	24.1	21.3	26.0
Dilute flow rate	$m^3/s \times 10^{-6}$	366.7	366.7	366.7	366.7
Brine make-up flow rate	$m^3/s \times 10^{-6}$	3.3	1.7	3.3	3.3
Product flow rate	$m^3/s \times 10^{-6}$	341.7	338.3	333.3	336.7
Brine flow rate	$m^3/s \times 10^{-6}$	23.3	25.0	23.3	28.3
Brine pH		5.8	5.8	5.8	5.4
Electrical stage 1					
Current	amperes	31.0	30.7	29.8	32.3
Voltage	volts	105	104	106	105
Specific cell pair resistance	$\Omega \cdot m^2 \times 10^{-3}$	3.97	3.91	3.88	3.91
Electrical stage 2					
Current	amperes	18.0	17.5	18.5	18.5
Voltage	volts	99	98	98	98
Specific cell pair resistance	$\Omega \cdot m^2 \times 10^{-3}$	6.44	6.46	5.95	6.37
Feed cationic concentration	eq/m^3	142.4	143.2	182.6	156.6
Product cationic concentration	eq/m^3	19.6	17.6	27.2	18.8
Current efficiency	percent	68.2	70.9	87.5	73.8

Table 52. — *Chemical analysis for ED performance at 90 percent recovery*

		Positive Dilute	Positive Mid-Stage Dilute	Positive Concentr.	Feed	Negative Dilute	Negative Mid-Stage Concentr.	Negative Concentr.	Feed	Negative Dilute	Negative Mid-Stage Concentr.	Negative Concentr.	Positive Dilute	Positive Mid-Stage Dilute	Positive Concentr.
Date		3/26/80	3/26/80	3/26/80	3/26/80	3/26/80	3/26/80	3/26/80	3/28/80	3/28/80	3/28/80	3/28/80	3/28/80	3/28/80	3/28/80
Time collected		1335	1335	1335	1335	1405	1405	1405	1225	1225	1225	1225	1255	1255	1255
Date analyzed at Denver		5/16/80	5/16/80	5/16/80	5/16/80	5/16/80	5/16/80	5/16/80	5/16/80	5/16/80	5/16/80	5/16/80	5/16/80	5/16/80	5/16/80
pH		7.4	7.4	7.5	7.7	7.4	7.2	7.0	7.7	7.1	7.8	6.9	7.5	7.6	7.2
TDS (calculated)	g/m^3	1 220	3 800	98 660	8 740	1 150	90 070	90 410	9 350	1 140	86 100	92 850	1 100	3 860	91 540
TDS (evaporated @ 105 °C)	g/m^3	1 160	3 740	98 300	8 840	1 080	95 200	95 100	8 660	1 060	93 900	93 800	1 100	4 000	94 900
Conductivity @ 25 °C	mS/m	220	660	9 600	1 320	189	9 600	6 120	1 340	206	8 800	7 540	200	580	8 800
Silica	g/m^3	10.1	9.8	12.1	10.4	9.1	15.1	14.7	10.1	8.9	14.6	14.0	9.5	8.9	9.5
Calcium	g/m^3	1.60	5.40	416	32.0	1.20	240	320	32.0	1.60	320	320	1.60	4.80	240
Magnesium	g/m^3	2.20	12.3	869	68.3	2.32	744	683	78.1	1.95	615	439	1.46	9.52	493
Sodium	g/m^3	430	1 270	33 200	2 940	406	29 400	29 400	2 940	406	27 600	32 400	389	1 330	32 100
Potassium	g/m^3	23.5	94.2	3 170	286	21.5	3 150	3 110	286	16.4	2 350	2 480	17.2	70.4	2 500
Iron, total	g/m^3	0.05	0.09	2.00	0.18	0.60	1.90	1.90	0.16	0.04	2.0	2.10	0.41	0.07	2.5
Manganese, total	g/m^3	ND	ND	0.28	0.01	0.03	0.26	0.27	0.01	ND	0.28	0.27	ND	ND	0.28
Strontium	g/m^3	ND	ND	1.70	0.30	ND	1.70	1.60	0.2	ND	1.0	1.1	ND	ND	1.0
Bicarbonate	g/m^3	17.7	29.3	93.9	42.7	17.7	18.9	87.2	43.3	14.0	105	96.4	18.9	31.1	88.5
Carbonate	g/m^3	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sulfate	g/m^3	197	850	23 600	2 090	176	22 300	22 600	1 520	197	22 400	21 900	123	672	2 110
Chloride	g/m^3	540	1 530	37 300	3 270	518	34 200	34 200	4 440	490	32 700	35 200	540	1 530	54 000
T-alkalinity as $CaCO_3$*	g/m^3	14.5	24.0	77.0	35.0	14.5	15.5	71.5	35.5	11.5	86.1	79.1	15.5	25.5	72.6
P-alkalinity as $CaCO_3$*	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS (calculated)/cond.]		5.5	5.8	10.3	6.6	6.1	9.4	11.1	7.0	5.5	9.8	12.3	5.5	6.7	10.4
E.F. [TDS (evaporated)/cond.]		5.3	5.7	10.2	6.7	5.7	9.9	11.7	6.5	5.1	10.7	12.4	5.5	6.9	10.8
Σ Anions	eq/m^3	19.60	61.38	1 543.54	136.20	18.55	1 429.31	1 434.43	157.41	18.14	1 389.72	1 449.58	18.08	61.91	1 565.40
Σ Cations	eq/m^3	19.56	58.79	1 613.00	142.51	18.50	1 433.50	1 431.60	143.31	18.36	1 326.60	1 525.40	17.54	60.52	1 516.30
Control value	eq/m^3	+0.10	+2.45	-2.89	2.35	+0.13	-0.19	+0.13	+5.54	-0.57	+2.92	-3.36	+1.40	+1.30	+2.01

*Alkalinities were not reported by the Denver lab. Values have been calculated form the bicarbonate concentration.

Table 52. — *Chemical analysis for ED performance at 90 percent recovery — Continued*

		Feed	Positive Dilute	Positive Mid-Stage Dilute	Positive Concentr.	Negative Concentr.	Negative Mid-Stage Concentr.	Negative Dilute	Feed	Negative Dilute	Negative Mid-Stage Dilute	Negative Concentr.	Positive Dilute	Positive Mid-Stage Dilute	Positive Concentr.
Date		4/2/80	4/2/80	4/2/80	4/2/80	4/2/80	4/2/80	4/2/80	4/3/80	4/3/80	4/3/80	4/3/80	4/3/80	4/3/80	4/3/80
Time collected		1350	1350	1350	1350	1420	1420	1420	-	-	-	-	-	-	-
Date analyzed at Denver		6/5/80	6/5/80	6/5/80	6/5/80	6/5/80	6/5/80	6/5/80	6/5/80	6/5/80	6/5/80	6/5/80	6/5/80	6/5/80	6/5/80
pH		7.4	7.5	7.6	6.7	7.7	7.9	7.4	7.3	6.8	6.5	6.2	6.6	7.5	6.2
TDS (calculated)	g/m^3	11 050	1 660	4 940	93 950	94 360	91 910	1 370	9 630	1 080	91 060	92 650	1 100	3 790	93 510
TDS (evaporated @ 105 °C)	g/m^3	9 090	1 690	4 120	94 800	94 400	94 900	1 370	9 610	1 020	92 600	93 000	1 120	3 780	94 800
Conductivity @ 25 °C	mS/m	1 260	232	644	7 930	8 590	8 590	264	1 290	210	8 590	8 590	191	573	8 590
Silica	g/m^3	5.7	5.4	5.1	8.5	8.1	8.0	5.7	5.2	4.7	8.2	7.3	5.1	5.4	8.2
Calcium	g/m^3	32.0	1.20	12.8	370	366	379	6.40	32.0	12.8	307	336	12.8	16.0	422
Magnesium	g/m^3	45.9	28.5	21.5	585	557	565	19.5	127	11.7	681	634	11.7	24.4	709
Sodium	g/m^3	3 940	556	1630	32 400	32 100	30 100	460	3 160	372	31 300	31 600	381	1 290	30 900
Potassium	g/m^3	230	23.5	82.1	2 550	2 650	2 560	22.7	280	21.5	2 920	3 040	23.1	88.3	3 040
Iron, total	g/m^3	0.05	0.01	0.02	0.80	0.90	0.90	0.01	0.06	0.03	0.72	0.73	0.03	0.05	0.74
Manganese, total	g/m^3	ND	ND	ND	0.34	0.34	0.34	ND	0.05	ND	0.34	0.34	ND	ND	0.34
Strontium	g/m^3	ND	ND	ND	3.0	4.0	3.0	ND	ND	ND	3.0	3.3	ND	ND	3.2
Bicarbonate	g/m^3	43.9	20.7	31.7	32.4	83.0	97.6	20.1	33.6	15.3	48.8	34.2	12.8	26.2	29.9
Carbonate	g/m^3	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sulfate	g/m^3	2 060	170	812	22 600	22 200	22 300	181	1 940	112	22 500	22 400	107	713	22 900
Chloride	g/m^3	4 690	852	2 340	35 400	36 400	35 900	653	4 050	533	33 300	34 600	547	1 630	35 500
T-alkalinity as $CaCO_3$*	g/m^3	36.0	17.0	26.0	26.6	68.1	80.0	16.5	27.6	12.6	40.0	28.0	10.5	21.5	24.5
P-alkalinity as $CaCO_3$*	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS (calculated)/cond.]		8.8	7.2	7.7	11.8	11.0	10.7	5.2	7.5	5.1	10.6	10.8	5.8	6.6	10.9
E.F. [TDS (evaporated)/cond.]		7.2	7.3	6.4	12.0	11.0	11.0	5.2	7.4	4.9	10.8	10.8	5.9	6.6	11.0
Σ Anions	eq/m^3	175.72	27.88	83.42	1 469.35	1 494.36	1 475.60	22.50	154.95	17.59	1 421.80	1 442.56	17.84	61.33	1 476.49
Σ Cations	eq/m^3	182.16	27.20	75.20	1 541.70	1 526.80	1 440.80	22.50	157.15	18.35	1 506.00	1 526.60	18.79	61.07	1 496.50
Control value	eq/m^3	-2.23	+1.26	+5.87	-3.16	-1.39	+1.51	0.00	-0.88	-2.00	-3.80	-3.74	-.248	+0.25	-0.87

*Alkalinities were not reported by the Denver lab. Values have been calculated from the bicarbonate concentration.

Table 53. — *Average performance data for Aquarnite V at constant feed conditions*

	Percent water recovery levels	
	[1]80	[2]90
Feed:		
Rate, $m^3/s \times 10^{-6}$	375	375
TDS, g/m^3	8800	8800
Brine:		
Rate, $m^3/s \times 10^{-6}$	75	38
TDS, g/m^3	40 700	96 500
Product:		
Rate, $m^3/s \times 10^{-5}$	300	338
	1350	1350
emf, V		
Stage 1	104	102
Stage 2	98	97
Current, A		
Stage 1	26.5	30.2
Stage 2	15.4	18.2
Total cell power consumption, W	4200	4700
Total cell current efficiency, %	82	73

[1] Average of 6 to 8 hours steady-state data acquisitions
[2] Average of 11 to 8 hours steady-state data acquisitions
[3] TDS — total dissolved solids

Table 54. – *The SCPR[1] for Ionics Aquamite V and LVS water*

	Date 1980						
	1-28	2-11	2-19	2-25	3-3	3-10	3-17
Operating time, hrs.	312	406	526	608	715	829	971
Water recovery, %	79.6	80.2	91.6	90.7	90.3	90.6	91.6
[1]SCPR, $m^2 \times 10^{-3}$							
Stage 1	4.71	4.78	4.06	3.85	3.90	3.85	3.91
Stage 2	7.70	7.65	6.50	5.82	6.17	6.38	6.18

[1] SCPR — specific cell pair resistance

Table 55 — *Electrodialysis stack inspection results, April 29, 1980*

The following observations were made during the ED stack inspection

1. Electrical Stage 1
    - a. Greenish-white precipitate sample taken from the edge of the "X" stream of the first heavy cation membrane; also slight rippling and discoloration of the edges of the membrane.
    - b. Greenish-white precipitate sample collected from between the anion and cation membranes of the first cell pair.
    - c. Slight discoloration of edges, slight crusty scaling near the stream inlets and outlets, and minimal reddish-brown scale buildup on cation and anion membrane surfaces. Scale easily rubs off the anion membranes, but is more permanent of the cation membranes of the first hydraulic stage.
    - d. Slight greenish-white scale buildup in center bend between inlet and outlet streams of the cation membrane of the fourth cell pair.
    - e. Minimal greenish-white scale of the "X" stream inlet to the first hydraulic stage.
    - f. Reddish-brown precipitate on cell pair 16, uniformly distributed on anion membrane.
    - g. Small black object (apparently electrode backing tape) found in cell pair 21.
    - h. Brown precipitate found in cell pair 26.
    - i. Turquoise-colored precipitate in cell pairs 29 and 34.
    - j. Precipitates of various colors (brown, greenish-white, turquoise) on intermembrane spacer between first and second hydraulic stages.
    - k. Reddish-brown film that easily rubs off of anion membranes. Film lessens in intensity throughout second hydraulic stage.
    - l. Turquoise precipitate on bottom "Y" stream inlet to third hydraulic stage of the cation membrane cell pair 130.
2. Electrical Stage 2
    - a. Turquoise precipitate on PVC block for electrode three, scale around inlet and outlet port of electrode rinse ports.
    - b. Very rippled heavy cation membrane of electrode three.
    - c. No film deposit seen on anion membranes of fourth hydraulic stage.
    - d. Small flakes of the membrane spacer found adhering to the membrane on cell pair 160, when the spacer was separated from the anion membrane.
    - e. Greenish-white precipitate found on the bottom of the "X" stream inlet to fourth hydraulic stage.
    - f. Turquoise precipitate on "Y" stream inlet to sixth hydraulic stage of cell pairs 259 and 260.
    - g. Slight scaling of heavy cation membrane on bottom of electrical stage two.
    - h. Slight scaling of "X" stream outlet from sixth hydraulic stage bottom of electrical stage two.

All other cell pairs appeared normal. The stack as a whole was in excellent condition.

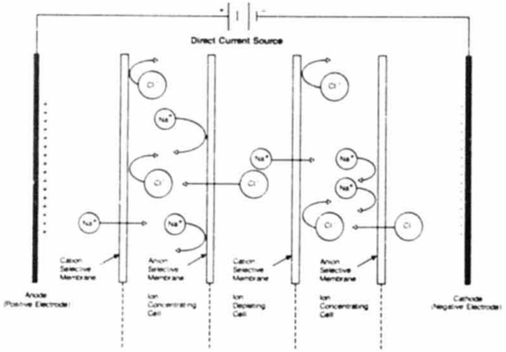


Figure 34 — Basic electrolysis system [14]

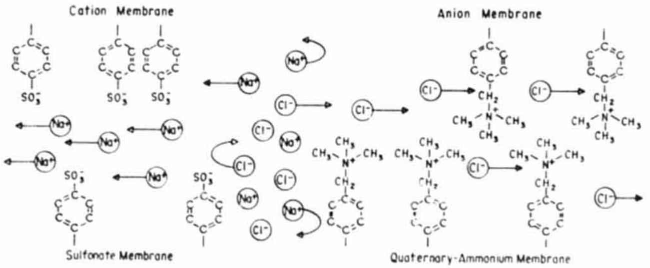


Figure 35 — Selective membrane structure [14]

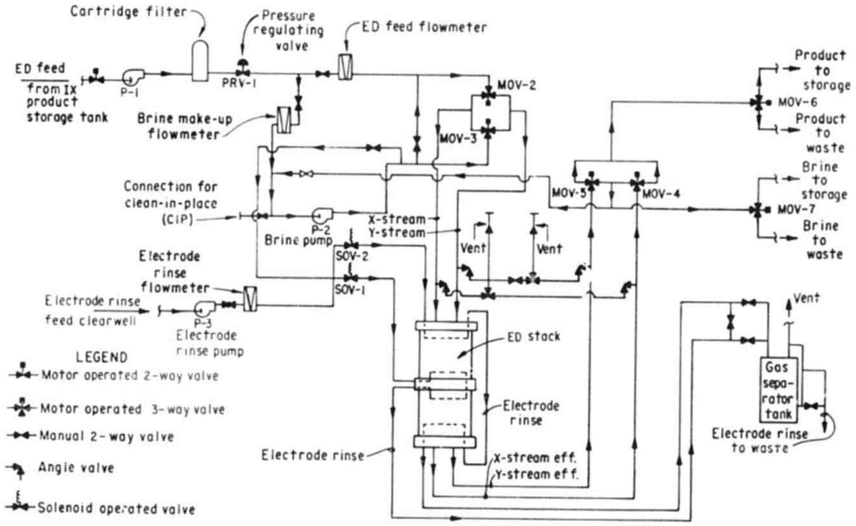
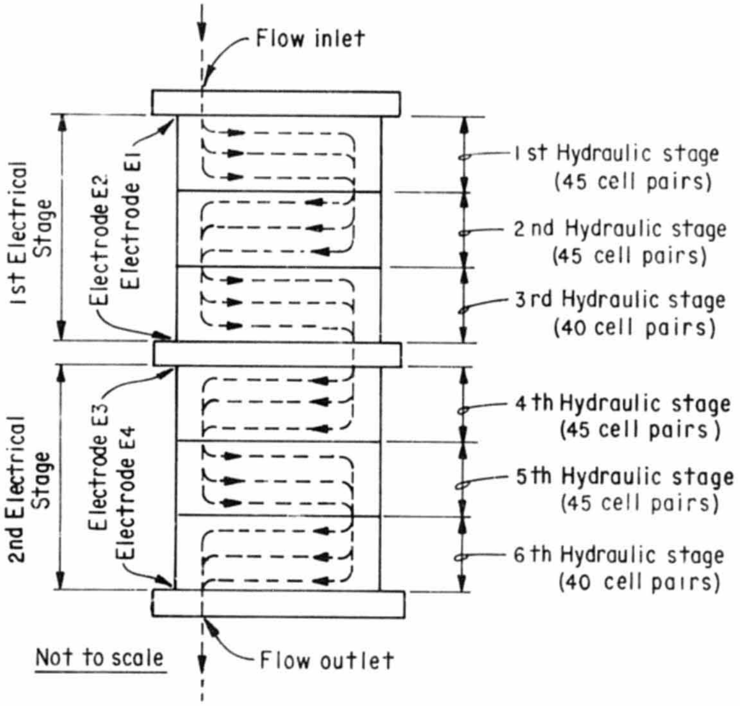


Figure 36 — Ionics Aquamite V electrodialysis demineralizer, simplified flow diagram



Note: One cell pair is one anion, one cation, and two spacers.

Figure 37 — The ED Stack configuration and flow path

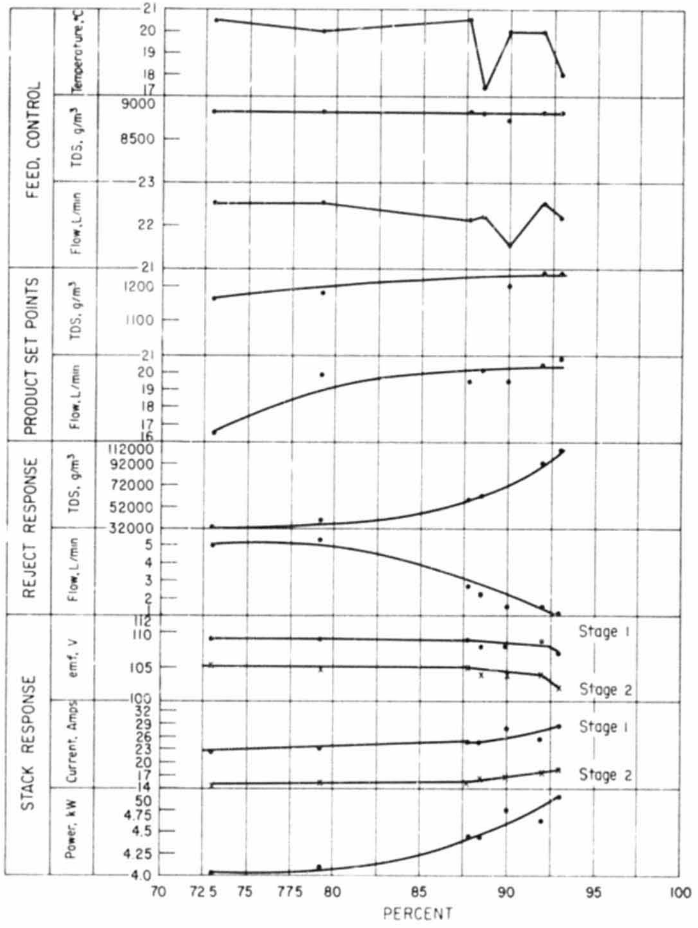


Figure 38 — Electrodialysis performance at various water recovery levels

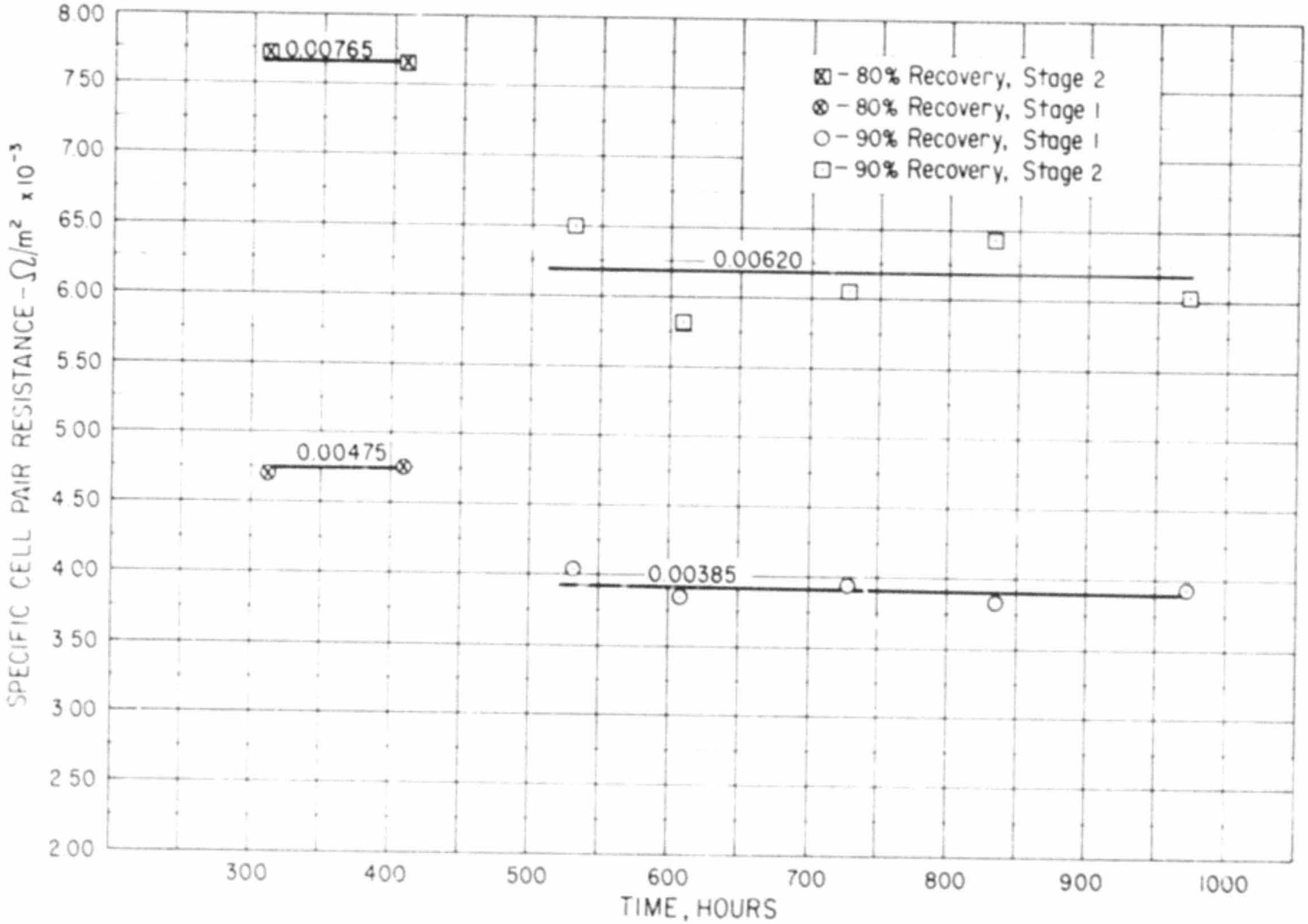


Figure 39 — Specific cell pair resistance

EXPERIMENTAL REVERSE OSMOSIS MEMBRANE EVALUATION

A general theory and description of RO (reverse osmosis) has been given by Buckley [15]. Osmosis (fig. 40) is the spontaneous passage of a liquid from a dilute to a more concentrated solution across an ideal semipermeable membrane that allows passage of the liquid but not of dissolved solids (fig. 39). Obviously, RO is a process in which the natural osmotic flow is reversed. Reversal (fig. 41) is effected by application of pressure to the concentrated solution sufficient to overcome the natural osmotic pressure of less-concentrated (dilute) solution. When the amount of water passing in either direction is equal, applied pressure can be defined as osmotic pressure of the dilute solution having that particular concentration of solutes.

In practical applications, pumps are used to supply the pressure to overcome osmotic pressure. The water flow rate through the membrane is dependent principally upon the net driving pressure. Solute flow rate through the membrane is dependent almost solely upon solute concentration of the feed water.

Figure 42 illustrates a typical RO installation. A single pressure vessel containing a membrane is shown, but normally a number of pressure vessels would be arranged in a series-parallel array. A pump continuously feeds the pressure vessel, and the back pressure valve on the concentrate stream controls pressure within the vessel and against the membrane. Increased pressure increases transport rate of the permeate.

Spiral-wound membranes modules (figs. 43, 44, and 45) were evaluated at the LVS test site. In this configuration, a product water collection channel is formed between two sheets of membrane by a fabric sealed in place on three of the four laminate edges. The fourth edge of the laminate is attached to a central tube, which has openings for collecting the permeate from backing material. A mesh screen over the membrane rejection surface forms a feed water flow channel through each module as the composite is wound around the central tube. An outer wrap is applied to maintain modular configuration. The result is a compact package that is efficient, economical, and applicable to a wide variety of feed waters. The spiral configuration has a high packing density, low manufacturing cost, and can be cleaned both chemically and hydraulically with relative ease. Its primary disadvantage is the need for extensive pretreatment on highly turbid feed water because feed flow passages are quite small and subject to clogging.

Figure 42 shows a basic RO system consisting of a pump or other source of pressure, membrane element, pressure container to house the membrane element and control system to regulate flows and pressures.

Normally, an RO system is employed with some form of pretreatment depending upon the particular application. Pretreatment can range from a simple cartridge filter to protect the high-pressure pump and membranes from particulate matter to coagulation, sedimentation, and filtration, followed by pH adjustment when operating on a highly turbid industrial waste stream. Pretreatment can include aeration, degasification, chlorination, pH adjustment, and even IX (ion exchange), depending on the final use of the product water.

Most RO systems in common usage today employ semipermeable membranes made of either cellulose acetate or a polyamide. Cellulose acetate membranes have a greater susceptibility to biological attack and have a limited pH range, while polyamide membranes have no tolerance for chlorine or other oxidizing agents in the feed stream. Good system design and proper feed pretreatment can overcome shortcomings, allowing both types of membranes to have extensive use in a wide area of applications.

A. Equipment Description and Operation

The ROPP (reverse osmosis pilot plant) was designed by Bureau engineers. Bureau engineers specified and purchased all piping, instruments, and equipment shown on the design drawings (figs. G-21, G-22, G-23, and G-24). The ROPP was constructed at the YDTF by PRC, the operating contractor. Final designs and revisions to the ROPP discussed later in this section were made by PRC.

1. *Membranes.* — The membrane elements shown on the top of the following page were obtained from FSD (Fluid Systems Division) of UOP, Inc., and DSI (Desalination Systems, Inc.).

2. *Flow loops.* — Figure 46 presents the RO flow schematic. Figure 47 is an exterior interconnecting piping diagram. Valve descriptions are included in table 56. The RO system consisted of feed tank TK-1 where reject brine and feed makeup were blended to produce a feed of proper TDS. Feed water was pumped via a high-pressure pump to the trailer-mounted pressure vessel arrays. The RO feed went directly into PV-1 (pressure vessel 1). Brine reject from PV-1 was fed to PV-2; similarly, PV-2 reject was fed to PV-3, and so on through PV-5. The individual vessel product streams flowed into a common header and then to waste. Pressure vessels 1 through 3, which contained FSD membranes, were steel vessels obtained from FSD. Four fiberglass vessels were obtained from DSI, two of which were used, two were spares.

3. *Cooling system.* — A stainless steel coil used for cooling the feed water was mounted inside TK-1 (capacity 3.8 m^3). Agitation provided homogenous RO feed water by mixing of reject brine and feed makeup. Cooling water circulated through the cooling coil was stored in the chilled-water tank (capacity 0.5 m^3). Water in the chilled-water tank was cooled by a chiller unit with a capacity of approximately 7.5 tons.

Supplier	Membrane model No.	No. of membranes	Maximum operating pressure (kPa)	Appendix No.[d]
Fluid Systems[e]	4600PA	1 each	[a] 4140	Table J-3
	4160S	1 each	[a] 4140	Table J-1
	4160HR	1 each	[a] 4140	Table J-2
	1001PA	2 each	[b] 6900	Table J-4
Desalination Systems, Inc.	8401	2 each	[c] 6210	Table J-6
	8402	2 each	[c] 6210	Table J-5

[a] These membranes were operated 33 to 50 percent above the manufacturer's maximum recommended operating pressure.

[b] These membranes were operated at a maximum of 90 percent of the manufacturer's maximum recommended operating pressure.

[c] These membranes were operated at less than 90 percent of the maximum recommended operating pressure during the first phase of testing and at the maximum operating pressure during the final phase of testing.

[d] See appendix J for each membrane specification.

[e] Normally 4 inches in diameter by 40 inches long and approximately 70-ft^2 area.

4. *Disinfection equipment.* — The pickling solution tank (capacity 0.5 m^3) was used only when the RO system was to be shut down for long periods. At that time, the discharge side of the solutioning pump was connected to the RO system at the feed water inlet to the RO trailer. A preservative solution was then pumped through the pressure vessels to prevent biological growth.

5. *Chemical addition.* — A turbidity monitor and a slipstream pH sensor sampled feed water to the high-pressure pump. Feed water pH was recorded on a chart recorder in the RO trailer; a controller was used to adjust sulfuric acid output that regulated feed pH. An in-line conductivity probe installed on the suction side of the high-pressure pump was connected to a meter readout and a chart recorder. To inhibit scale, SHMP (sodium hexametaphosphate) was injected into the feed makeup influent to TK-1. The final concentration of the SHMP in the makeup stream was approximately 10 g/m^3. Sodium sulfite injected into the feed water stream removed residual chlorine before the stream entered the high-pressure pump and subsequently entered RO elements.

6. *Instruments and controls.* — Stainless steel pipe and tube were used for all piping from the discharge side of the high-pressure pump to the back pressure control valve. Figures 48 and 49 show interior piping and pressure differential manifold mounted in a trailer. Pressure gages were located in the system feed and at the back pressure control valve to allow pressure to be monitored. Differential pressure gages measured pressure drop across each vessel. Rotometers measured brine reject flow and combined product flow. Individual product streams were measured volumetrically to obtain accurate flow measurements. Sample taps located on feed and reject piping of each vessel were used for sample collection. Product samples were collected from plastic tubes which conveyed product from a pressure vessel to the product header. The RO system was protected from overpressurization by two separate devices. A high-pressure sensing switch shut down the high-pressure pump when the system overpressurized. A rupture disk provided a secondary protection system in the event of failure of the pressure-sensing switch.

7. *Modifications.* — Several modifications of the original RO system were made during the course of the test. These are shown in figure 49. The 1" pneumatically controlled back pressure valve (V-27-1) did not function correctly. The valve regulator throat was too large for the system. In the automatic mode, pressure would oscillate uncontrollably. This occurred because the valve had to be almost totally closed to control the pressure. This resulted in over control. The valve closed completely and instaneously when a small correction was necessary and then opened too far and too rapidly. This situation was corrected by placing the ½" manually operated throttling valve on the downstream side of the

back pressure control valve. When this manual valve was throttled, differential pressure across the back pressure valve was effectively reduced, allowing operation in a more open position.

B. Task objective

The LVS water generally averaged 105 °F in temperature; also, during the summer days, ambient air temperatures range between 100 and 110 °F. The principal objective of task RO was to evaluate performance of polyamide RO membranes under conditions of warm temperatures, high salinity, and high water recovery.

C. Task description and performance

The following experimental tasks were to be performed:

- Task RO-I, install, hydraulically test, and shakedown operation of the ROPP
- Task Ro-II, test five membranes in series at constant temperature, flow, pressure, and brine concentration
- Task RO-III, operate the ROPP at a pressure near 6210 kPa (900 lb/m^2)

1. *Task RO-I, installation and startup.* — Construction of the ROPP was based on preliminary drawings (figs. G.21 and G.22) prepared by Bureau engineers. Final design was completed at the site by the contractor (figs. G.23 and G.24). Task RO-I was completed on July 9, 1980.

2. *Task RO-II, testing five membranes at steady-state conditions.* — Five membranes were evaluated in series at constant temperatura, flow, pressure, and brine concentration. The intent was to simulate membrane desalting conditions in a high recovery section of a three-stage RO desalting train.

a. Operating pressure. — Since operation of a typical high recovery section would be near 80 percent water recovery, desalting of L-SA-IX pretreated water would result in a brine with a concentration of approximately 33 000 mg/L of TDS in recycle tank TK-1. Brine effluent from the last vessel (ROPV-5) would be 40 000 mg/L. Using the rule of thumb of 10 lb/in^2/1000 mg/L, OP (osmotic pressure) is then:

$$OP = \left\{ 10\ \text{lb/in}^2 \times \frac{\text{TDS (40000 mg/L)}}{\text{1000 mg/L}} \right\} = 400\ \text{lb/in}^2$$

and the required operation pressure would be about 800 lb/in^2 (gage).

b. Membranes. — The following membrane were installed for task RO-II:

Pressure vessel	Membrane module
ROPV-1	4-inch UOP, model No. 4600PA, 98 percent salt rejection
ROPV-2 and -3	4-inch UOP, model No. 1001PA, 98.9 percent salt rejection
ROPV-4 and -5	4-inch DSI, model No. B402, 99 percent salt rejection

c. Control conditions. — The ROPP operated at the following controlled levels:

RO feed rate	7.6×10^{-4} m^3/s
RO feed pressure	5516 kPa
RO feed TDS	32 000 g/m^3
RO reject TDS	40 000 g/m^3
Temperature limits	30 °C
pH range	4 to 6
SHMP dosage	10.0 g/m^3
Sodium sulfite dosage	1.0 g/m^3

d. Site operations. — The ROPP was an experimental system and considerable up-and-down operations were experienced in debugging the total system. The ROPP was first brought on-line on July 9, 1980. On July 14, at 9 p.m., the system reached steady-state conditions, and the makeup flow rate was increased to 0.2×10^{-3} m^3/s to maintain recycle tank content at the desired 32 000 g/m^3 TDS concentration. The first set of conditions at steady-state were recorded on July 15. The RO system operated from July 14 (when steady-state operation was reached) through July 22, with a feed flow rate of approximately 0.7×10^{-3} m^3/s at an operating pressure of 5516 kPa. Average feed pH was 4.6 (set-point pH 4.5); average conductivity of feed was 4404 mS/n (approximately 4400 mS/m desired). Sodium hexametaphosphate was injected to attain a desired final concentration of 10 g/m^3.

On July 17, the RO chiller unit repeatedly shut down after being reset because the high ambient temperature of 112 °F overloaded thecompressor motor. The RO system was secured because of a high temperature feed to PV-1 of 32 °C. The chiller was returned to service after approximately 1 hour, but the unit again continued to shut down. The chiller unit was allowed to cool and was returned to service after 4.5 hours of off-line time. On July 18, the feed water temperature criterion for the RO system was modified. New operating conditions shown in table 57 were immediately instituted.

On July 22, the RO system was secured because of apparent biological contamination. Immediate preparations were made to disinfect and clean the system. The RO membranes were flushed with formaldehyde solution and were then left with the formaldehyde solution remaining in the membranes. The tanks and appurtenant piping of the RO system and IX system were subsequently disinfected and cleaned. Startup of the RO system was initiated on July 24, after it had been secured for approximately 43.8 hours during cleanup. The unit was operated at a low pressure for 1.6 hours to flush the formaldehyde solution from the system into the formaldehyde solution tank. The formaldehyde solution was then neutralized with concentrated hydrogen peroxide. Flow was then increased to approximately 0.7×10^{-3} m^3/s and operation was initiated.

Because of an increase in product flow rate from approximately 0.07×10^{-3} m^3/s to approximately 0.13×10^{-3} m^3/s and a high product conductivity of approximately 1700 mS/m (feed conductivity 1580 mS/m) during startup, the unit was secured and the membrane element replaced in the first pressure vessel. The failed lead element was replaced with an FSD 1001PA membrane. During membrane changeout, it was discovered that improper FSD membranes had been installed in pressure vessels 2 and 3. The following list presents the proper membrane loading requirements and the actual loading arrangement:

Pressure vessel	Membrane loaded	Membranes required by the program
PV-1	FSD 4600PA	FSD 4600PA
PV-2	FSD 4160HR	FSD 1001PA
PV-3	FSD 4160S	FSD 1001PA
PV-4	DSI B402	DSI B402
PV-5	DSI B402	DSI B402

A decision was made to operate with the FSD 4160HR and 4160S membranes in pressure vessels 2 and 3, respectively. The RO unit was flushed with a 1.0 percent formaldehyde solution and secured. On July 25, a filter and booster pump were installed on the suction side of the high-pressure pump. After a total off-line time of approximately 23.5 hours, startup procedures for bringing the RO system up to operating parameters were initiated.

On July 26, equilibrium conditions were reached. The RO unit operated from July 26 through July 29, with a feed flow rate of approximately 0.74×10^{-3} m^3/s. Average feed pH during this time was 5.3 (setpoint pH 4.5), average feed conductivity was 4382 mS/m (approximately 4400 mS/m desired). Average temperature of the feed to PV-1 was 39.0 °C (maximum allowable temperature 45 °C), average pressure of the feed to PV-1 was 5516 kPa (5516 kPa desired). Sodium hexametaphosphate was injected to attain a final concentration of 10 g/m^3.

On July 30, the RO systems were secured because of a broken variable speed belt on the high-pressure pump. A replacement belt was requisitioned. Tanks T-33 and TK-1 (figs. 46 and 47) were super-chlorinated, and the RO membranes were secured with a formaldehyde solution remaining in the membranes to protect against possible biological contamination. The RO unit remained secured until August 5, when the replacement belt for the high-pressure pump was received and installed. The membranes in pressure vessels 2 and 3 were removed. Pressure vessel 2 was left empty since only one additional FSD 1001PA membrane was available. This was loaded into pressure vessel 3, and the remainder of the RO testing was conducted with only four membranes in service.

The RO unit operated from August 6 through 25 with a feed flow rate of approximately 0.74×10^{-3} m^3/s. The average feed pH during this time was 4.3 (setpoint pH 4.5), the average feed conductivity was 4442 mS/m (approximately 4500 mS/m desired). The average temperature of the feed to PV-1 was 36.8 °C (maximum allowable temperature 45 °C), the average PV-1 feed pressure was 5518 kPa (5516 kPa desired). Sodium hexametaphosphate was injected to attain a stream concentration of 10 gal/m^3.

e. Summary of results. — Task RO-II was initially scheduled to operate at condition 1, shown in table 57. Because of operational difficulties and revision of temperature criteria, the system was also operated at conditions 2, 3, and 4. Table 58 presents a summary of operating times for the various conditions. Table 59 presents operating times for membranes used during task RO-II and -III.

Table 60 presents a summary of operating data for task RO-II and -III. Table 61 presents the concentrations of the various constituents of the process streams for test condition 1. Tables 62, 63, 64, and 65 present the concentrations of the various constituents of the process streams for test condition 4.

3. *Task RO-III, continuous operations at 6206 kPa pressure.* — Brought ROPP operation from steady state of task RO-II from a feed pressure of 5516 kPa to a feed pressure of 6206 kPa. Steady state was reestablished at the higher pressure and performance evaluated for 4 days.

a. Site operation. — The RO unit was secured on August 25 to recharge the bladder accumulator to 5516 kPa to be able to operate at the higher pressure. The RO unit operated from August 25 until August 29, with a feed flow rate of approximately 0.72×10^{-3} m^3/s. Average feed pH during this time was 4.4 (setpoint pH 4.5); average feed conductivity was 5815 mS/m (approximately 5760 mS/m desired). Average PV-1 feed temperature was 33.6 °C (maximum allowable temperature 45 °C); average PV-1 feed pressure was 6206 kPa (6206 kPa desired). Sodium hexametaphosphate was injected to attain a final stream concentration of 10 g/m^3.

b. Summary of results. — Task RO-III was scheduled as condition 5, as shown in table 57. Table 58 presents a summary of operating time for condition 5. Table 66 presents the concentrations of the various constituents of the process streams for test condition 5.

D. Membrane Performance

Table 59 presents operating time for membranes used during task RO-I and -III. Performance of the seawater membranes at the LVS site are summarized in table 67. Figure 50 is a graph of conductivity versus TDS for feed water.

Figure 51 is a graph of conductivity versus TDS for recycle product water and reject brine water. Specifications for the membranes are given in appendix J. Figures 52 through 59 present plots of the following data shown in table 60:

- — Differential pressure
- — Feed conductivity
- — Product conductivity
- — Reject conductivity
- — Product temperature
- — Product flow

Computer calculated salt transport coefficients based on the above operation data are shown as figures 60 through 64. The transport coefficient relates solute flow through the membrane to the concentration gradient based on bulk stream properties. This coefficient is corrected to 23.9 °C and is computed by the equation:

$$B = \frac{Fp\ Cp\ K^{(23.9-T)}}{C_{LM} - Cp}$$

where:
B = Salt transport coefficient, in m/s
Fp = Product water flow, in kg/s
T = Absolute temperature, in degrees Kelvin
Cp = Product stream concentration, in kg/m^3
C_{LM} = Log-mean stream concentration, in kg/m^3
K = Temperature coefficient factor (1.02 to 1.03)

The numerator consists of the total solids (or any individual ion, radical, or solids group) mass transfer (weight/time). The denominator consists of a calculation of the net driving force (difference in concentration across membrane). The rate of permeation of solute through a membrane can be determined from the net driving force and the membrane system solute transport coefficient.

1. *Fluid Systems membrane 4600PA.* — Inspection of the plots of the data for membrane 4600 PA (fig. 52) indicates a severe problem developed in this element. The differential pressure increased steadily during the period of operations. This trend indicated a high probability the membrane functioned as a filter since no prefilters were in use on the system at that time of the operation. Inspection of the membrane subsequently verified this condition. An attempt was made to clean the membrane and return it to service; however, the membrane suffered a mechanical failure and was replaced with an FSD 1001PA element (fig. 54). Membrane 4600 PA experienced a steady flux decline during the period it was in service. This may have been caused, at least partially, by the foreign material that collected on it. However, a more likely cause was severe compaction. The membrane, rated for a maximum operating pressure of 4140 kPa, operated at a constant pressure of 5516 kPa.

2. *Fluid Systems membrane 1001PA.* — After failure of membrane 4600PA, a prefilter was installed ahead of the high-pressure pump. Membrane 1001PA was installed in the lead

pressure vessel (PV-1). Data presented in figures 54 and 60 indicate a fairly uniform operation. The rise in conductivity and very slight drop in product flow for the hours of operation reflect the increase in system operating pressure from 5516 to 6210 kPa. This apparent anomaly can be seen in the appropriate plots for each of the membranes that were in service during that last phase of test condition 5 (tables 57, 58, and 66). A membrane 1001PA was loaded into pressure vessel 3, but pressure vessel 2 was left empty since the available supply of Fluid Systems membranes had been exhausted. Membrane 1001PA performed well in pressure vessel 3; no operational problems were encountered.

3. *Fluid Systems membranes 4160R and 4160S.* — At the outset of the RO testing, vessels 2 and 3 were loaded with membranes 4160HR and 4160 S, respectively. Data presented for these two membranes (figs. 55, 56, 61, and 62) indicate that a higher than expected product flow decline was experienced. This is probably attributable to excessive compaction. The membranes were operated at 5516 kPa; the manufacturer's recommended maximum operating pressure is 4140 kPa. Both membranes were removed from the pressure vessels and stored in a preservative solution (formaldehyde). This action was taken as a precaution against excessive loss of operating time because of possible membrane failure similar to that which occurred with the membrane 4600PA.

4. *Desalination Systems membranes B401 and B402.* — Pressure vessels 4 and 5 each had been loaded with membrane B401 (figs. 58 and 63) and B402 (figs 59 and 64). As can be seen from the operating data, both performed well. The data indicate that the product flow increased with time throughout the testing. What appeared to be a flow reduction during the last 90 hours of operation was caused by an increase in feed pressure and a subsequent increase in feed TDS.

Table 56. — *The RO system valve description*

Valve Number	Description
V-1	Brine tank bottom drain valve
V-2	Brine tank side drain valve
V-3	RO product to brine pond drain valve
V-4	RO reject to brine pond drain valve
V-5	RO reject to brine pond drain valve
V-6	Chilled water tank drain valve
V-7	RO feed pressure sensing line shut-off valve
V-8	RO feed shut-off valve
V-9	RO feed bypass (return to tank) valve
V-10	RO reject return to tank valve
V-11	RO reject return to tank valve
V-12	RO product cleaning solution return to tank valve
V-13	RO reject cleaning solution return to tank valve
V-14	Brine tank main feed shut-off valve
V-15	RO feed make-up flow control valve
V-16	RO feed flow meter isolation valve
V-17	RO feed flow meter bypass valve
V-18	RO feed flow meter isolation valve
V-19	RO feed make-up sample valve
V-20	Brine tank sample valve
V-21	Facility air to RO trailer shut-off valve
V-22	Potable water to RO trailer shut-off valve
V-23	RO high pressure feed pump supply valve
V-24	RO reject sample valve
V-25	RO product sample valve
V-26	RO product air break bypass valve
V-27	RO backpressure control valve assembly
V-27-1	RO backpressure control valve
V-27-2	RO backpressure manual/auto mode 3-way selector valve
V-28	ED feed shut-off valve
V-29	RO feed make-up shut-off valve
V-30	RO feedwater flush shut-off valve
V-31	ED feed (RO feed make-up) pump suction valve
V-32	Brine tank float-control shut-off valve
V-33	Cleaning solution tank drain valve
V-34	Cleaning solution pump suction valve
V-35	Cleaning solution pump discharge valve
V-36	Chilled water pump main suction shut-off valve
V-37	Cooling coil feed pump suction valve
V-38	Cooling coil feed pump discharge valve
V-39	Cooling coil feed return valve
V-40	Chiller feed pump suction valve
V-41	Chiller feed pump discharge valve
V-42	Chiller feed return valve
V-43	RO feed (to PV-1) sample valve
V-44	PV-1 rej/PV-2 feed sample valve
V-45	PV-2 rej/PV-3 feed sample valve
V-46	PV-3 rej/PV-4 feed sample valve
V-47	PV-4 rej/PV-5 feed sample valve
V-48	PV-5/System rej sample valve
V-49	PV-1 Feed ΔP 3-way selector valve
V-50	PV-1/PV-2 interstage ΔP 3-way selector valve
V-51	PV-2/PV-3 interstage ΔP 3-way selector valve
V-52	PV-3/PV-4 interstage ΔP 3-way selector valve
V-53	PV-4/PV-5 interstage ΔP 3-way selector valve
V-54	PV-5 rej ΔP 3-way selector valve
V-55	Backpressure valve assembly supply air valve
V-56	PV-1 product check valve
V-57	PV-2 product check valve
V-58	PV-3 product check valve
V-59	PV-4 product check valve
V-60	PV-5 product check valve
V-61	Sample tap isolation valve
V-62	Flow thru pH sensor shut-off valve
V-63	HP pump feed return sample tap

Table 57. — *Operating conditions, RO systems*

Condition No.	Period From	Period To	Feed temperature limit, °C	Feed pressure, kPa	Feed pH range	Feed TDS, g/m³	Vessels/ Elements utilized
1	7/11/80	7/18/80	<30	5516	4-6	31 700	V-1/4600PA V-2/4160HR V-3/4160S V-4/B402 V-5/B402
2	7/18/80	7/25/80	<42	5516	4-6	31 700	same as condition 1
3	7/25/80	8/ 5/80	<45	5516	4-6	31 700	V-1/1001PA V-2/4160HR V-3/4160S V-4/B402 V-5/B402
4	8/ 5/80	8/25/80	<45	5516	4-6	31 700	V-1/1001PA V-2/ - V-3/1001PA V-4/B402 V-5/B402
5	8/25/80	8/29/80	<45	6206	4-6	41 500	same as condition 4

Note Sodium hexametaphosphate was injected into the feed make-up stream to attain a final concentration of 10 g/m³. Sodium sulfite was injected into the feed stream to attain a final concentration of approximately 1 g/m³, for chlorine removal.

Table 58. — *Accumulated operating and steady state[*] time, RO systems*

Condition No.	Period From	Period To	A Total elapsed time, hrs	B On-line time, hrs	C Steady state time, hrs	B/Ax100, %	C/Ax100, %	C/Bx100, %
1[b]	7/11/80 @ 1440	7/18/80 @ 1500	168.3	145.4	82.2	86.4	48.8	56.5
2[c]	7/18/80 @ 1500	7/25/80 @ 1345	166.8	97.2	95.0	58.3	57.0	97.7
3[d]	7/25/80 @ 1345	8/5/80 @ 2030	270.7	105.1	85.4	38.8	31.5	81.3
4[e]	8/5/80 @ 2030	8/25/80 @ 0945	469.3	450.2	418.2	95.9	89.1	92.9
5[f]	8/25/80 @ 0945	8/29/80 @ 0720	93.5	91.2	81.8	97.5	87.5	89.7
Summary	7/11/80 @ 1440	8/29/80 @ 0720	1168.6	889.1	762.6	76.1	65.3	85.8

[*] When data were taken, feed tank conductivity and feed pressure had reached specified operating conditions.

[b] On 7/18/80, the feed temperature limit was changed from 30 °C to 42 °C.

[c] On 7/25/80, the vessel 1 membrane was replaced, and the feed temperature limit was changed from 42 °C to 45 °C.

[d] On 8/5/80, the vessel 2 membrane was removed, and the vessel 3 membrane was replaced.

[e] On 8/25/80, the operating pressure was increased from 5516 kPa to 6206 kPa.

[f] On 8/29/80, the LVS test program ended.

Table 59. — *Membrane operating time*

Membrane model No./Vessel No.	Period		Total elapsed time, hrs	Total operating time, hrs	Operating time at sample collection time, hrs.					
	From	To			7/17/80	8/7/80	8/14/80	8/21/80	8/27/80	8/29/80
UOP 4600PA/V-1	7/11/80@1440	7/24/80@1415	311.5	242.5	119.5	—	—	—	—	—
UOP 4160HR/V-2	7/11/80@1440	7/30/80@0215	443.5	347.6	119.5	—	—	—	—	—
UOP 4160S/V-3	7/11/80@1440	7/30/80@0215	443.5	347.6	119.5	—	—	—	—	—
UOP 1001PA/V-1	7/25/80@1345	8/29/80@0720	833.5	646.5	—	134.6	294.8	460.1	598.4	646.0
UOP 1001PA/V-3	8/5/80@2030	8/29/80@0720	562.8	541.4	—	29.5	189.7	355.0	493.3	540.9
DSI B402/V-4	7/11/80@1440	8/29/80@0720	1168.6	889.1	119.5	377.2	537.4	702.7	841.0	888.6
DSI B402/V-5	7/11/80@1440	8/29/80@0720	1168.6	889.1	119.5	377.2	537.4	702.7	841.0	888.6

Table 60. — *Operating data, RO system temperature variations*

Vessel/Element	Feed Pressure, 5.5 MPa			Feed Pressure, 6.2 MPa		
	Max	Min	Avg	Max	Min	Avg
Feed pressure, MPa	5.55	5.50	5.52	6.21	6.21	6.21
Product temperature, °C						
V-1/4600PA	38.5	27.0	31.5	—	—	—
V-2/4160HR	40.0	27.0	33.5	—	—	—
V-3/4160S	40.0	27.0	33.5	—	—	—
V-1/1001PA	40.0	30.5	36.4	34.0	29.9	32.3
V-3/1001PA	40.3	30.6	36.1	33.9	29.9	32.3
V-4/B402	40.4	27.0	35.0	34.2	30.0	32.3
V-5/B402	40.3	27.0	35.1	34.2	30.2	32.4
Product flow, $m^3/s \times 10^{-3}$						
V-1/4600PA	0.071	0.049	0.057	—	—	—
V-2/4160HR	0.040	0.021	0.026	—	—	—
V-3/4160S	0.032	0.018	0.023	—	—	—
V-1/1001PA	0.067	0.049	0.057	0.049	0.042	0.044
V-3/1001PA	0.062	0.048	0.054	0.045	0.040	0.042
V-4/B402	0.065	0.018	0.053	0.052	0.047	0.049
V-5/B402	0.050	0.025	0.041	0.042	0.037	0.039
Feed conductivity, S/m						
V-1/4600PA	4.79	3.80	4.40	—	—	—
V-2/4160HR	5.10	4.20	4.72	—	—	—
V-3/4160S	5.50	4.40	4.88	—	—	—
V-1/1001PA	4.65	4.25	4.42	6.00	4.25	5.81
V-3/1001PA	4.93	4.50	4.70	6.30	5.92	6.10
V-4/B402	5.70	4.60	5.03	6.51	6.25	6.41
V-5/B402	6.00	4.89	5.37	6.95	6.65	6.82
Product conductivity, S/m						
V-1/4600PA	0.10	0.07	0.08	—	—	—
V-2/4160HR	0.34	0.20	0.28	—	—	—
V-3/4160S	0.22	0.14	0.18	—	—	—
V-1/1001PA	0.16	0.09	0.11	0.15	0.13	0.14
V-3/1001PA	0.12	0.07	0.09	0.12	0.11	0.11
V-4/B404	0.40	0.05	0.24	0.63	0.55	0.60
V-5/B402	0.38	0.04	0.16	0.43	0.36	0.40

Table 61. — *Reverse osmosis process streams composition, samples collected during test condition 1 (sample date 7/17/80)*

Source		Combined product	Reject	Feed make-up	Feed to PV-1	Feed to PV-2	Feed to PV-3	Feed to PV-4	Feed to PV-5	Product of PV-1	Product of PV-2	Product of PV-3	Product of PV-4	Product of PV-5
Date		7/17/80	7/17/80	7/17/80	7/17/80	7/17/80	7/17/80	7/17/80	7/17/80	7/17/80	7/17/80	7/17/80	7/17/80	7/17/80
Time collected		0830	0830	0830	0830	0830	0830	0830	0830	0830	0830	0830	0830	0830
Data analyzed at Yuma	Units	8/13/80	8/13/80	8/13/80	8/13/80	8/13/80	8/13/80	8/13/80	8/13/80	8/13/80	8/13/80	8/13/80	8/13/80	8/13/80
pH*		4.0	4.7	4.0	4.5	4.6	4.7	4.7	4.7	3.5	4.3	4.5	5.1	5.2
TDS (calculated)[b]	g/m³	413	40 309	8 627	30 127	32 775	34 069	35 457	37 989	220	983	596	292	185
TDS (evaporated @ 105 °C)*	g/m³	495	40 200	8 556	30 480	32 745	33 968	35 813	37 476	241	964	644	319	262
Conductivity @ 25 °C*	mS/m	120	5 300	1 450	4 200	4 500	4 610	4 740	5 000	90	200	120	80	30
Temperature*	°C	29.0	30.0	33.0	29.5	30.0	30.0	30.0	30.0	29.0	29.0	29.0	29.0	29.0
Silica*	g/m³	<1.0	9.7	2.2	7.6	9.5	8.5	8.9	9.1	<1.0	<1.0	<1.0	<1.0	<1.0
Calcium	g/m³	<0.4	7.9	5.7	6.0	6.5	6.8	7.0	7.4	<0.4	<0.4	<0.4	<0.4	<0.4
Magnesium	g/m³	0.18	175	46.3	132	143	149	155	166	0.18	0.15	0.20	0.19	0.10
Sodium	g/m³	153	13 870	2 970	10 360	11 250	11 740	12 250	13 080	78.9	369	223	106	65.6
Potassium	g/m³	9.1	770	153	548	592	650	670	740	5.8	25.4	15.0	5.0	3.8
Iron, total	g/m³	<0.10	0.22	<0.10	0.20	0.20	0.40	0.18	0.20	<0.10	<0.10	<0.10	<0.10	<0.10
Manganese, total	g/m³	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30	<0.30
Strontium	g/m³	<0.3	0.4	<0.3	0.3	0.4	0.4	0.4	0.5	<0.3	<0.3	<0.3	<0.3	<0.3
Bicarbonate[b]	g/m³	2.4	31.7	9.8	26.8	29.3	29.3	31.7	31.7	ND	9.8	9.8	12.2	12.2
Carbonate*	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Sulfate	g/m³	10	9 740	2 050	7 120	7 940	8 100	8 550	9 050	5	11	13	10	11
Chloride	g/m³	239	15 720	3 395	11 940	12 802	13 400	13 800	14 920	130	573	340	165	98
T alkalinity as $CaCO_3$*	g/m³	2.0	26.0	8.0	22.0	24.0	24.0	26.0	26.0	ND	8.0	8.0	10.0	10.0
P alkalinity as $CaCO_3$*	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Hydroxide*	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		3.4	7.6	5.9	7.2	7.3	7.4	7.5	7.6	2.4	4.9	5.0	3.6	6.2
E.F. [TDS(evaporated)/cond.]*		4.1	7.6	5.9	7.3	7.3	7.4	7.6	7.5	2.7	4.8	5.4	4.0	8.7
Σ Anions	eq/m³	6.99	646.86	138.64	485.58	527.52	547.22	567.91	609.92	3.77	16.55	10.02	5.06	3.19
Σ Cations	eq/m³	6.90	637.83	137.20	475.[illegible]4	516.61	539.92	563.12	601.94	3.60	16.71	10.10	4.76	2.96
Control value	eq/m³	+0.42	+0.89	+0.64	+1.28	+1.32	+0.85	+0.54	+0.83	+1.03	-0.44	-0.30	+1.64	+1.47

*Analysis conducted at LVSTS

[b] YDTF used only one-half the bicarbonate in calculating the TDS

Table 62. — *Reverse osmosis process streams composition, samples collected during test condition 4 (sample date 8/7/80)*

				Feed to				Product of			
Source		Feed make-up	Reject	PV-1	PV-3	PV-4	PV-5	PV-1	PV-3	PV-4	PV-5
Date		8/7/80	8/7/80	8/7/80	8/7/80	8/7/80	8/7/80	8/7/80	8/7/80	8/7/80	8/7/80
Time collected		0300	0300	0300	0300	0300	0300	0300	0300	0300	0300
Data analyzed at Yuma	Units	9/5/80	9/5/80	9/5/80	9/5/80	9/5/80	9/5/80	9/5/80	9/5/60	9/5/80	9/5/80
pH[a]		3.8	4.8	4.5	4.6	4.8	4.7	3.6	3.7	5.3	5.4
TDS (calculated)[b]	g/m^3	8 780	46 410	32 953	36 078	38 863	43 012	496	359	1 229	734
TDS (evaporated @ 105 °C)[a]	g/m^3	8 864	47 224	33 303	36 347	39 446	43 743	452	367	1 219	738
Conductivity @ 25 °C[a]	mS/m	1 600	6 080	4 600	4 930	5 300	5 700	88	70	212	135
Temperature[a]	°C	35.5	37.5	38.0	38.2	38.2	38.5	36.9	37.0	37.0	37.0
Silica[a]	g/m^3	9.4 (1.9)[c]	2.1 (8.6)[c]	6.8	7.3	7.7	8.7	<1.0	<1.0	<1.0	<1.0
Calcium	g/m^3	7.3	12.7	9.1	9.7	10.2	11.0	<0.40	<0.40	<0.40	<0.40
Magnesium	g/m^3	40.9	204	144	156	170	188	0.52	0.22	0.39	0.23
Sodium	g/m^3	3 020	15 990	11 390	12 370	13 430	14 800	187	135	464	279
Potassium	g/m^3	175	890	599	700	750	830	11.8	8.4	21.3	13.2
Iron, total	g/m^3	<0.04	0.31	0.23	0.24	0.26	0.28	0.06	<0.04	<0.04	<0.04
Manganese, total	g/m^3	<0.02	0.02	0.02	0.02	<0.02	0.02	<0.02	<0.02	<0.02	<0.02
Strontium	g/m^3	<0.03	0.3	<0.3	<0.3	<0.4	<0.5	<0.30	<0.30	<0.30	<0.30
Bicarbonate[a]	g/m^3	ND	4.9	3.9	6.3	7.8	9.3	ND	ND	7.3	9.8
Carbonate[a]	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Sulfate	g/m^3	2 035	10 900	7 660	8 700	9 120	10 520	6.0	4.0	4.4	4.4
Chloride	g/m^3	3 500	18 400	13 140	14 140	15 380	16 660	291	211	735	432
T-alkalinity as $CaCO_3$[a]	g/m^3	ND	4.0	3.2	5.2	6.4	7.6	ND	ND	6.0	8.0
P-alkalinity as $CaCO_3$[a]	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Hydroxide[a]	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		5.5	7.6	7.2	7.3	7.3	7.5	5.6	5.1	5.8	5.4
E.F. [TDS(evaporated)/cond.][a]		5.5	7.8	7.2	7.4	7.4	7.7	5.1	5.2	5.8	5.5
Σ Anions	eq/m^3	141.12	746.19	530.30	580.21	623.97	689.26	8.33	6.04	20.95	12.44
Σ Cations	eq/m^3	139.57	735.74	523.08	569.32	617.88	681.05	8.48	6.11	20.76	12.49
Control value	eq/m^3	+0.68	+0.90	+0.87	+1.20	+0.62	+0.76	-0.61	-0.35	+0.43	-0.18

[a] Analysis conducted at LVSTS
[b] YDTF used only one-half the bicarbonate in calculating the TDS
[c] Analysis rerun at YDTF

Table 63. — *Reverse osmosis process streams composition, samples collected during test condition 4 (sample date 8/14/80)*

					Feed to				Product of			
Source		Combined product	Reject	Feed makeup	PV-1	PV-3	PV-4	PV-5	PV-1	PV-3	PV-4	PV-5
Date		8/14/80	8/14/80	8/14/80	8/14/80	8/14/80	8/14/80	8/14/80	8/14/80	8/14/80	8/14/80	8/14/80
Time collected		0645	0415	0415	0415	0415	0415	0415	0415	0415	0415	0415
Data analyzed at Yuma	Units	10/17/80	10/17/80	10/17/80	10/17/80	10/17/80	10/17/80	10/17/80	10/17/80	10/17/80	10/17/80	10/17/80
pH[a]		3.8	4.7	5.3	4.4	4.6	4.7	4.7	3.5	3.6	5.2	5.2
TDS (calculated)[b]	g/m³	855	43 715	9 090	29 822	32 959	36 091	39 109	438	316	1 417	933
TDS (evaporated @ 105 °C)[a]	g/m³	880	43 618	8 986	30 258	33 222	36 083	39 016	430	363	1 431	969
Conductivity @ 25 °C[a]	mS/m	200	5 200	1 400	3 900	4 200	4 500	4 700	100	100	300	300
Temperature[a]	°C	35.0	37.5	35.4	37.2	37.6	36.8	36.5	37.5	36.2	36.6	36.5
Silica[a]	g/m³	<1.0	9.7	2.7	7.2	7.6	7.9	8.8	<1.0	<1.0	<1.0	<1.0
Calcium	g/m³	<0.4	12.2	4.9	8.6	9.4	10.3	10.8	<0.4	<0.4	<0.4	<0.4
Magnesium	g/m³	0.21	129	113.4	89.6	98.4	108	116	0.22	0.20	0.36	0.20
Sodium	g/m³	321	15 000	3 240	10 380	11 420	12 470	13 510	160	115	532	358
Potassium	g/m³	16.5	860	170	574	640	710	760	10.6	7.8	25.0	16.4
Iron, total	g/m³	<0.04	0.63	<0.04	0.56	0.44	0.39	0.71	<0.04	<0.04	<0.04	<0.04
Manganese, total	g/m³	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02
Strontium	g/m³	<0.3	0.5	0.4	0.5	0.4	0.4	0.5	0.3	0.3	0.3	0.3
Bicarbonate[a]	g/m³	ND	6.8	7.8	4.9	6.8	8.3	6.3	ND	ND	6.8	6.3
Carbonate[a]	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Sulfate	g/m³	3.2	10 700	2 020	7 100	7 920	8 780	9 500	4.4	3.2	6.0	8.4
Chloride	g/m³	514	17 000	3 635	11 660	12 860	14 000	15 200	262	190	850	547
T-alkalinity as $CaCO_3$[a]	g/m³	ND	5.6	6.4	4.0	5.6	6.8	5.2	ND	ND	5.6	5.2
P-alkalinity as $CaCO_3$[a]	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Hydroxide[a]	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		4.3	8.4	6.5	7.6	7.8	8.0	8.3	4.4	3.2	4.7	3.2
E.F. [TDS(evaporated)/cond.][a]		4.4	9.4	6.4	7.8	7.9	8.0	8.3	4.3	3.6	4.8	3.1
Σ Anions	eq/m³	14.57	702.6	144.7	476.9	527.9	578.0	626.8	7.48	5.43	24.21	15.71
Σ Cations	eq/m³	14.41	685.7	146.6	474.0	521.7	570.0	617.2	7.26	5.25	23.82	16.02
Control value	eq/m³	+0.48	+1.54	-0.81	+0.39	+0.75	+0.88	+0.98	+1.02	+0.95	+0.81	-0.89

[a] Analysis conducted at LVSTS

[b] YDTF used only one-half the bicarbonate in calculating the TDS

Table 64. — *Reverse osmosis process streams composition, samples collected during test condition 4 (sample date 8/21/80)*

					Feed to				Product of			
Source		Feed make-up	Combined product	Reject	PV-1	PV-3	PV-4	PV-5	PV-1	PV-3	PV-4	PV-5
Date		8/21/80	8/21/80	8/21/80	8/21/80	8/21/80	8/21/80	8/21/80	8/21/80	8/21/80	8/21/80	8/21/80
Time collected		0315	0415	0315	0315	0315	0315	0315	0315	0315	0315	0315
Data analyzed at Yuma	Units	10/21/80	10/21/80	10/21/80	10/21/80	10/21/80	10/21/80	10/21/80	10/21/80	10/21/00	10/21/80	10/21/80
pH[a]		5.9	3.8	4.6	4.5	4.6	4.7	4.6	3.5	3.5	5.1	5.1
TDS (calculated)[b]	g/m^3	8 565	1 148	45 789	32 464	35 404	37 602	41 175	516	385	2 144	1 348
TDS (evaporated @ 105 °C)[a]	g/m^3	8 775	1 188	46 336	33 251	35 598	38 597	42 553	598	365	2 168	1 316
Conductivity @ 25 °C[a]	mS/m	1 500	180	5 600	4 400	4 700	4 970	5 300	96	72	340	220
Temperature[a]	°C	31.5	33.0	35.0	34.5	34.0	34.5	34.5	33.5	33.8	34.0	34.0
Silica	g/m^3	4.76	1.42	20.0	15.0	15.8	16.6	18.3	1.12	1.12	2.16	1.24
Calcium	g/m^3	6.50	<0.50	11.4	8.10	8.70	9.50	10.5	0.50	0.50	0.50	0.50
Magnesium	g/m^3	60.4	0.48	303	216	234	250	272	0.64	0.21	0.90	0.36
Sodium	g/m^3	2 940	426	15 300	11 000	11 900	12 800	13 900	185	140	797	498
Potassium	g/m^3	179	23.8	950	700	740	780	870	14.5	10.1	42.2	26.1
Iron, total	g/m^3	<0.04	<0.04	0.86	0.66	0.65	0.75	0.81	0.14	<0.04	<0.04	<0.04
Manganese, total	g/m^3	<0.02	<0.02	0.03	0.02	0.02	0.02	0.02	<0.02	<0.02	<0.02	<0.02
Strontium	g/m^3	0.30	0.30	0.60	0.50	0.60	0.60	0.60	0.40	0.50	0.50	<0.20
Bicarbonate	g/m^3	8.3	ND	7.8	9.3	8.8	10.7	6.8	ND	ND	7.3	6.8
Carbonate	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Sulfate	g/m^3	1 920	8.0	12 100	8 320	9 300	9 440	10 600	4.0	<2.0	7.2	2.4
Chloride	g/m^3	3 450	688	17 100	12 200	13 200	14 300	15 500	310	233	1 290	816
T-alkalinity as $CaCO_3$	g/m^3	6.8	ND	6.4	7.6	7.2	8.8	5.6	ND	ND	6.0	5.6
P-alkalinity as $CaCO_3$	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Hydroxide	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		5.7	6.4	8.2	7.4	7.5	7.6	7.8	5.4	5.4	6.3	6.1
E.F. [TDS(evaporated)/cond.][a]		5.8	6.6	8.3	7.6	7.6	7.8	8.0	6.2	5.1	6.4	6.0
Σ Anions	eq/m^3	137.5	19.58	734.6	517.6	566.2	600.2	658.2	8.83	6.57	36.66	23.18
Σ Cations	eq/m^3	137.8	19.19	715.3	514.6	556.3	597.8	649.8	8.50	6.40	35.86	22.38
Control value	eq/m^3	-0.14	+0.95	+1.67	+0.37	+1.12	+0.26	+0.81	+1.33	+0.82	+1.19	+1.71

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[a] Analysis conducted at LVSTS.

[b] YDTF used only one-half the bicarbonate in calculating the TDS.

Table 65. — *Reverse osmosis process streams composition, samples collected during test condition (sample date 8/27/80)*

Source		Feed make-up	Combined product	Reject	Feed to PV-1	Feed to PV-3	Feed to PV-4	Feed to PV-5	Product of PV-1	Product of PV-3	Product of PV-4	Product of PV-5
Date		8/27/80	8/27/80	8/27/80	8/27/80	8/27/80	8/27/80	8/27/80	8/27/80	8/27/80	8/27/80	8/27/80
Time collected		0710	0710	0710	0710	0710	0710	0710	0710	0710	0710	0710
Data analyzed at Yuma	Units	10/21/80	10/21/80	10/21/80	10/21/80	10/21/80	10/21/80	10/21/80	10/21/30	10/21/80	10/21/80	10/21/80
pH[a]		5.4	3.8	4.6	4.5	4.6	4.6	4.6	3.4	3.5	5.1	5.1
TDS (calculated)[b]	g/m³	8 508	2 153	61 910	47 236	49 920	54 114	58 377	923	682	4 149	2 694
TDS (evaporated @ 105 °C)[a]	g/m³	8 517	2 128	65 187	47 742	51 703	54 110	58 599	960	635	4 141	2 630
Conductivity @ 25 °C[a]	mS/m	1 450	315	7 300	5 900	6 300	6 500	6 950	140	108	570	380
Temperature[a]	°C	29.5	31.0	32.0	32.0	31.7	31.8	32.0	31.0	31.3	31.1	31.0
Silica	g/m³	3.5	1.5	14.6	12.1	15.8	15.8	16.8	1.1	<1.0	2.4	1.4
Calcium	g/m³	9.60	<0.50	67.0	51.0	54.0	67.0	62.0	<0.50	<0.50	0.60	0.50
Magnesium	g/m³	77.6	1.14	563	418	445	476	523	0.82	0.27	2.28	0.83
Sodium	g/m³	2 910	802	20 500	15 600	16 600	17 900	19 300	345	250	1 550	1 010
Potassium	g/m³	172	42.4	1 260	950	1 000	1 060	1 170	22.9	17.8	83.0	49.6
Iron, total	g/m³	<0.04	<0.04	1.32	1.15	1.18	1.21	1.24	<0.04	<0.04	<0.04	<0.04
Manganese, total	g/m³	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	<0.02
Strontium	g/m³	0.20	<0.20	0.50	0.30	0.30	0.40	0.40	<0.20	<0.20	<0.20	<0.20
Bicarbonate	g/m³	10.2	ND	10.2	9.3	9.8	9.8	9.3	ND	ND	8.8	9.8
Carbonate	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Sulfate	g/m³	1 910	6.2	18 100	13 600	14 300	15 900	17 600	6.4	6.2	6.6	7.2
Chloride	g/m³	3 420	1 300	21 400	16 600	17 500	18 700	19 700	547	408	2 500	1 620
T-alkalinity as $CaCO_3$	g/m³	8.4	ND	8.4	7.6	8.0	8.0	7.6	ND	ND	7.2	8.0
P-alkalinity as $CaCO_3$	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Hydroxide	g/m³	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		5.9	6.8	8.5	8.0	7.9	8.3	8.4	6.6	6.3	7.3	7.1
E.F. [TDS(evaporated)/cond.][a]		5.9	6.8	8.9	8.1	8.2	8.3	8.4	6.9	5.9	7.3	6.9
Σ Anions	eq/m³	136.4	36.80	980.9	751.7	791.7	858.9	922.5	15.56	11.64	70.81	46.01
Σ Cations	eq/m³	137.9	36.06	973.6	739.8	787.0	847.8	915.6	15.66	11.35	69.76	45.30
Control value	eq/m³	-0.64	+1.03	+0.47	+1.01	+0.38	+0.83	+0.48	-0.27	+1.00	+0.87	+0.87

[a] Analysis conducted at LVSTS.

[b] YDTF used only one-half the bicarbonate in calculating the TDS

Table 66. — *Reverse osmosis process streams composition, samples collected during test condition 5 (sample date 8/29/80)*

					Feed to				Product of			
Source		Feed make-up	Combined product	Reject	PV-1	PV-3	PV-4	PV-5	PV-1	PV-3	PV-4	PV-5
Date		8/29/80	8/29/80	8/29/80	8/29/80	8/29/80	8/29/80	8/29/80	8/29/80	8/29/80	8/29/80	8/29/80
Time collected		0645	0645	0645	0645	0645	0645	0645	0645	0645	0645	0645
Data analyzed at Yuma	Units	10/21/80	10/21/80	10/21/80	10/21/80	10/21/80	10/21/90	10/21/80	10/21/80	10/21/80	10/21/80	10/21/80
pH^a		5.4	3.8	4.6	4.5	4.5	4.6	4.4	3.5	3.5	5.1	5.1
TDS (calculated)^b	g/m^3	8 594	2 225	63 051	47 492	50 733	54 011	58 510	919	693	4 288	2 765
TDS (evaporated @ 106 °C)^a	g/m^3	8 610	2 237	62 866	48 360	50 637	53 922	62 969	913	790	4 322	2 841
Conductivity @ 25 °C^a	mS/m	1 450	325	7 020	5 820	6 100	6 400	6 720	145	116	608	400
Temperature^a	°C	30.2	30.5	32.5	32.5	33.0	33.0	33.0	31.6	31.8	31.9	32.0
Silica	g/m^3	3.0	1.3	13.5	11.4	11.9	12.1	13.0	1.0	2.4	3.2	1.8
Calcium	g/m^3	11.9	0.60	111	86.0	92.0	97.0	106	0.60	<0.50	0.80	0.60
Magnesium	g/m^3	43.1	0.99	521	390	415	447	486	0.85	0.30	2.36	0.89
Sodium	g/m^3	2 970	832	21 200	15 700	16 800	17 800	19 200	343	252	1 600	1 040
Potassium	g/m^3	151	42.7	1 200	900	1 010	1 150	1 200	23.1	18.0	91.0	50.4
Iron, total	g/m^3	<0.04	<0.04	0.92	1.12	0.98	0.76	0.42	<0.04	<0.04	<0.04	<0.04
Manganese, total	g/m^3	<0.02	<0.02	<0.02 90	0.80	0.70	0.80	0.90	<0.20	<0.20	<0.20	<0.20
Bicarbonate	g/m^3	9.8	ND	8.3	7.3	7.3	7.8	7.8	ND	ND	7.3	7.3
Carbonate	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Sulfate	g/m^3	2 010	7.0	18 900	14 000	15 000	16 100	17 500	7.6	7.0	6.8	7.4
Chloride	g/m^3	3 400	1 240	21 100	16 400	17 400	18 400	20 000	543	413	2 580	1 660
T-alkalinity as $CaCO_3$	g/m^3	8.0	ND	6.8	6.0	6.0	6.4	6.4	ND	ND	6.0	6.0
P-alkalinity as $CaCO_3$	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
Hydroxide	g/m^3	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND
E.F. [TDS(calculated)/cond.]		5.9	6.8	9.0	8.2	8.3	9.4	8.7	6.3	6.0	7.1	6.9
E.F. [TDS(evaporated)/cond.]^a		5.9	6.9	9.0	8.3	8.3	8.4	9.4	6.3	6.8	7.4	7.1
Σ Anions	eq/m^3	137.9	37.95	989.0	754.4	803.4	854.6	928.9	15.48	11.80	73.04	47.10
Σ Cations	eq/m^3	137.2	37.40	1 001	742.4	795.4	845.3	911.2	15.61	11.45	72.16	46.63
Control value	eq/m^3	+0.33	+0.79	-0.79	+1.02	+0.64	+0.69	+1.22	-0.35	+1.21	+0.71	+0.56

[a] Analysis conducted at LVSTS

[b] YDTF used only one-half the bicarbonate in calculating the TDS

Table 67. — *Performance of high temperature sea water membranes*

Element model		4600PA		4160PA		1001PA		4160HR		4160S		1001PA		B401		B402	
Manufacturer		Fluid Systems		Fluid Systems		Fluid Systems		Fluid Systems		Fluid Systems		Fluid Systems		Desalination Systems		Desalination Systems	
Type		Spiral wound		Spiral wound		Spiral wound		Spiral wound		Spiral wound		Spiral wound		Spiral wound		Spiral wound	
Percent water recovery	Units	80	90	80	90	80	90	80	90	80	90	80	90	80	90	80	90
Pressure vessel number		ROPV-1		ROPV-1		ROPV-1		ROPV-2		ROPV-3		ROPV-3		ROPV-4		ROPV-5	
Feed pressure, average	kPa	5520		5220		5520	6210	5520		5520		5520	6210	5520	6210	5520	6210
Differential pressure																	
Start	kPa	13				21	20	5		2		6	13	120	110	50	64
End	kPa	80				20	18	10		5		13	14	112	120	63	75
Feed temperature																	
Maximum	°C	37				37	35	37		37		37	35	37	34	375	34
Average	°C	33				33	32	34		34		35	33	33	33	35	33
Operating time (accumulative)	hours	243		243		580	770	330		330		470	660	825	900	825	900
Product flow:																	
Start	$m^3/s \times 10^{-6}$	60				63	55	30		25		54	52	20	48	30	36
End	$m^3/s \times 10^{-6}$	50				50	44	22		18		52	42	55	50	45	36
Average	$m^3/s \times 10^{-6}$	57				57	50	26		22		53	47	51	49	43	36
Brine recycle flows																	
Average	$m^3/s \times 10^{-6}$	760															
Operating conditions	figure	52		53		54	54	55		56		57	57	58	58	59	59
Salt transport coefficient	figure					60		61		62				63			64
Average TDS																	
Reject	$g/m^3 \times 10^3$	40.2		37		39	53	38		40		40	63	43	55	47	61
Product	g/m^3	250		350		600	700	1300		850		400	550	1500	1500	600	3800
Feed	$g/m^3 \times 10^3$	30.5		34		34	36	36		38		37	52	40	53	43	48

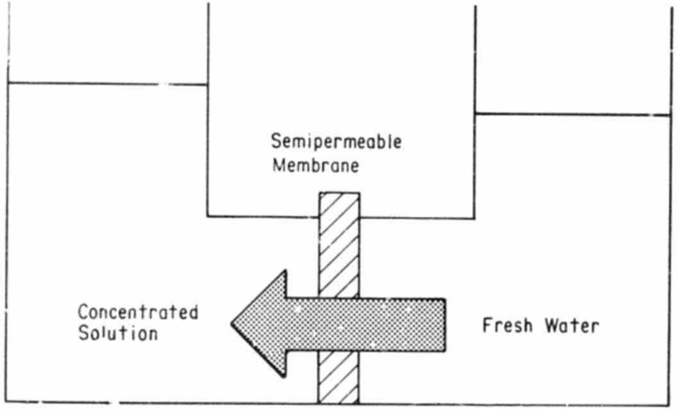


Figure 40. – Osmosis with normal flow from low-concentration solution to high-concentration solution.

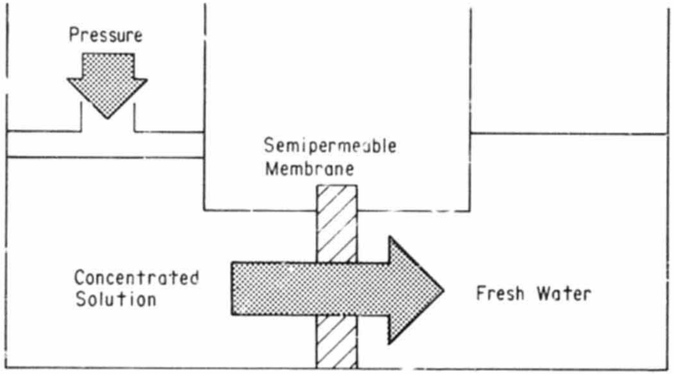


Figure 41. — Reverse osmosis with flow reversed by application of pressure to high-concentration solution.

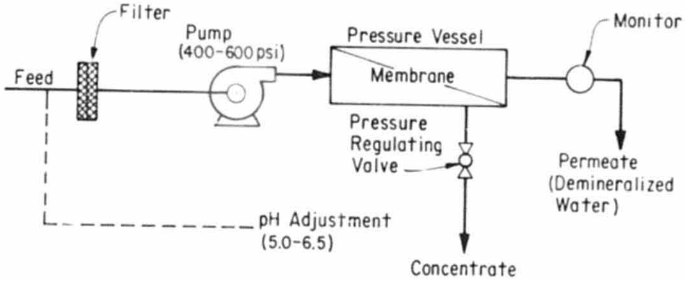


Figure 42 — Schematic of reverse osmosis system.

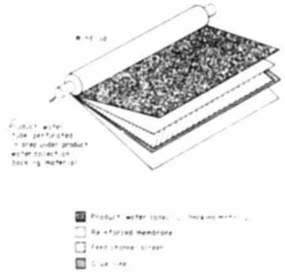


Figure 43 — Schematic assembly diagram for
ROGA spiral-wound module

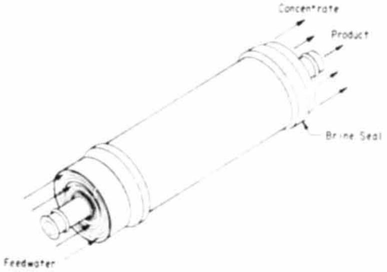


Figure 44 — Flow paths in spiral wound element

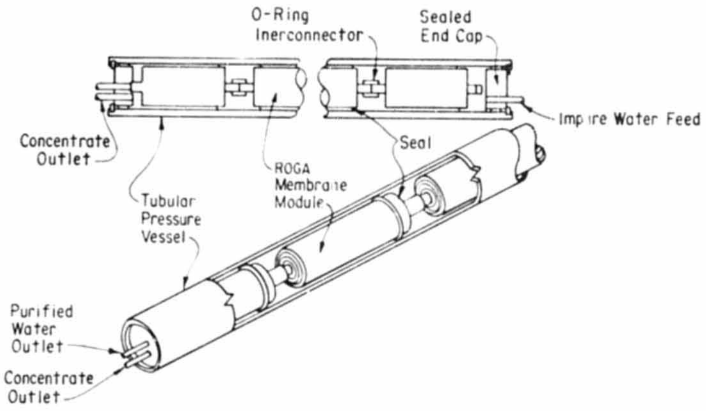


Figure 45 — Pressure vessel containing ROGA modules

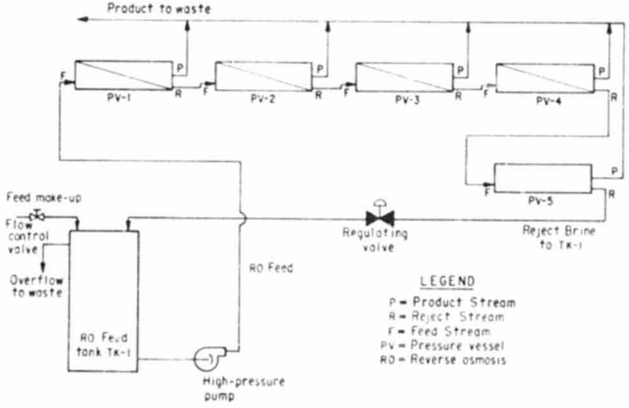
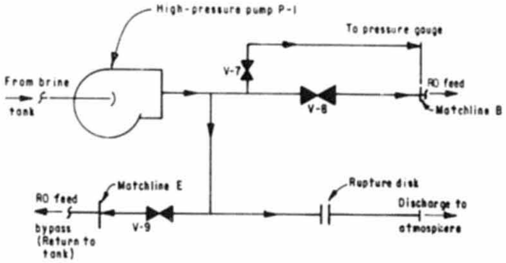


Figure 46 — The RO flow schematic



DETAIL A

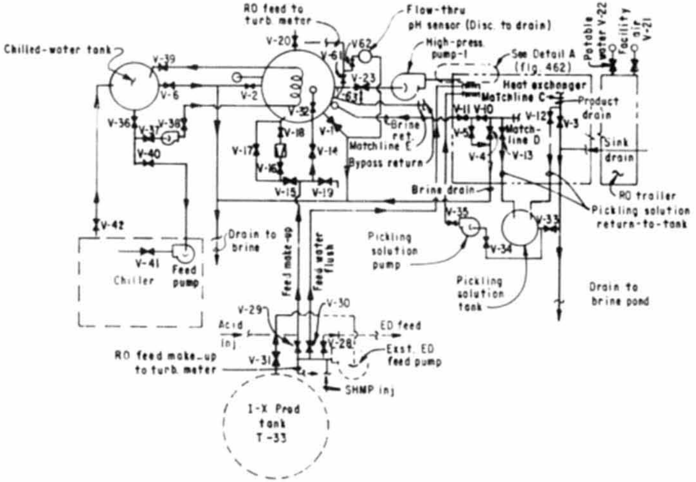
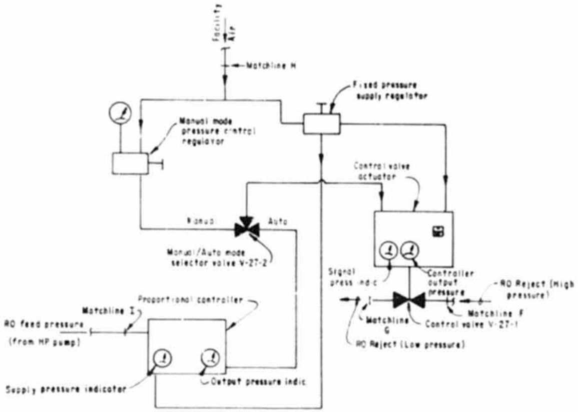


Figure 47. — Exterior piping diagram.



DETAIL A
Backpressure Control Valve Assembly V-27

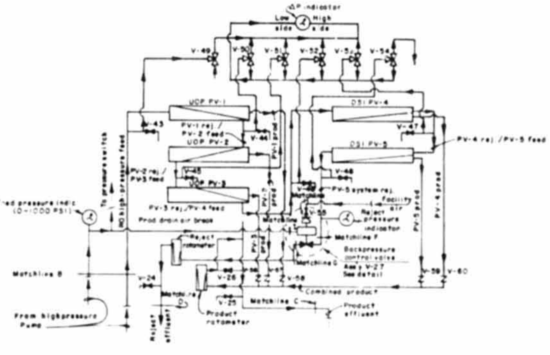


Figure 48 — Interior piping diagram

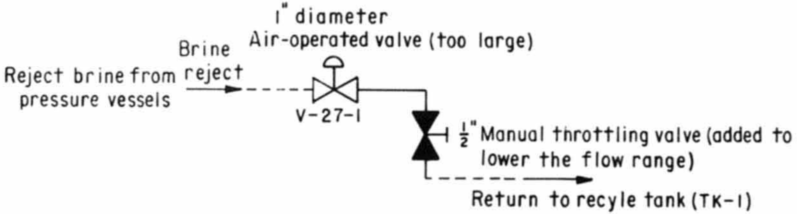


Figure 49 — Modification of reject brine flow control

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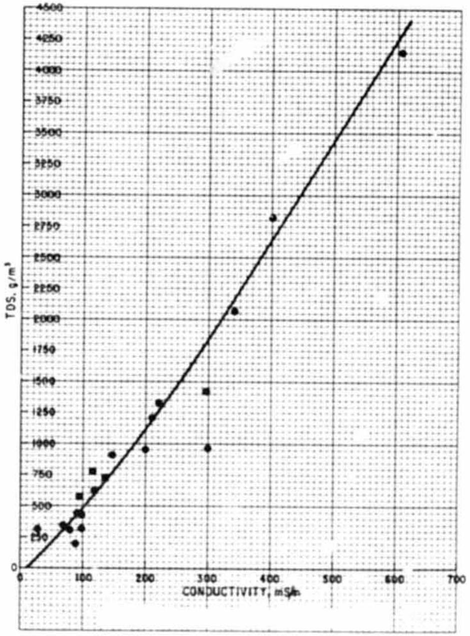


Figure 50 — The RO product water

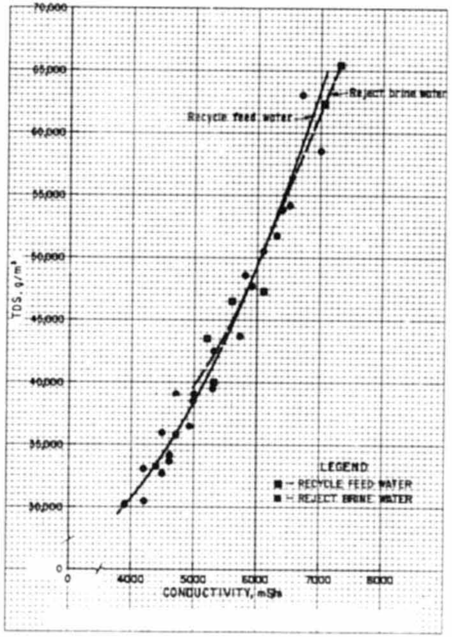


Figure 51 — The RO recycle fe[illegible]rine

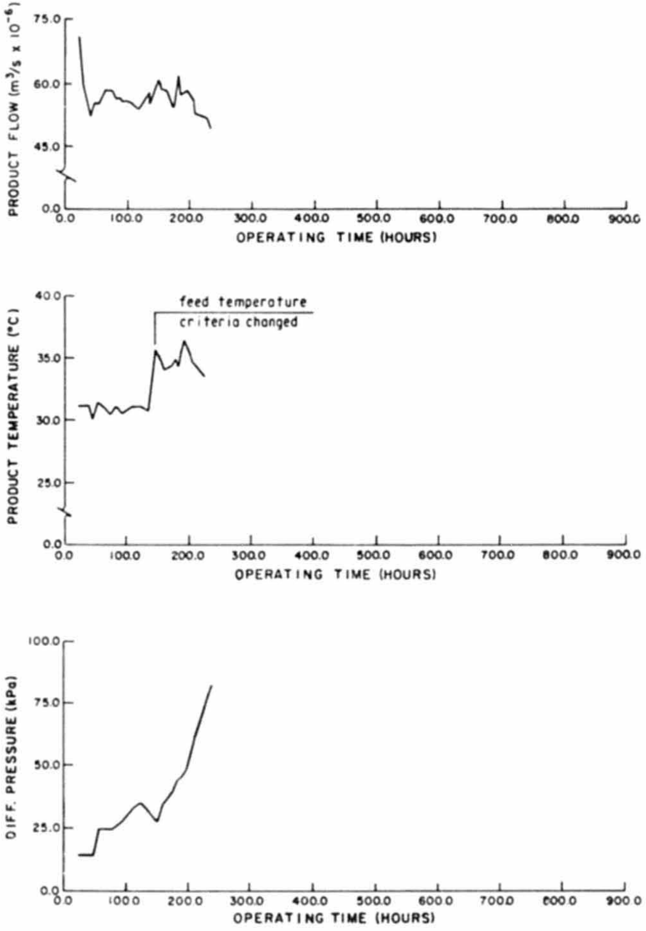


Figure 52. — The UOP membrane 4600 PA (PV-1) performance

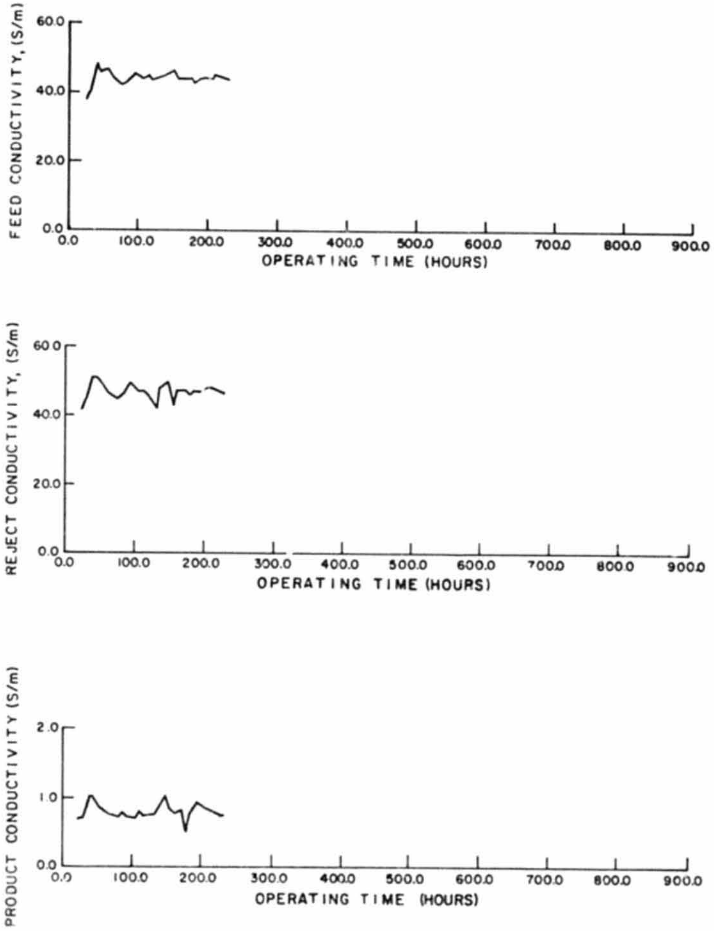


Figure 53. — The UOP membrane 4160 PA (PV-1) performance

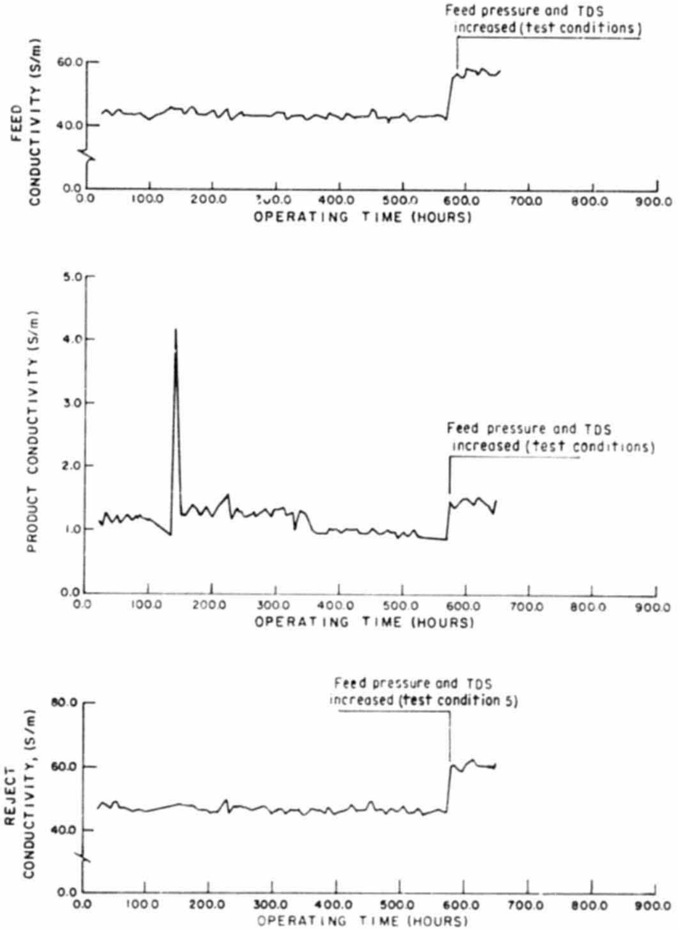


Figure 54 — The UOP membrane 1001 PA (PV-1) performance (sheet 1 of 2)

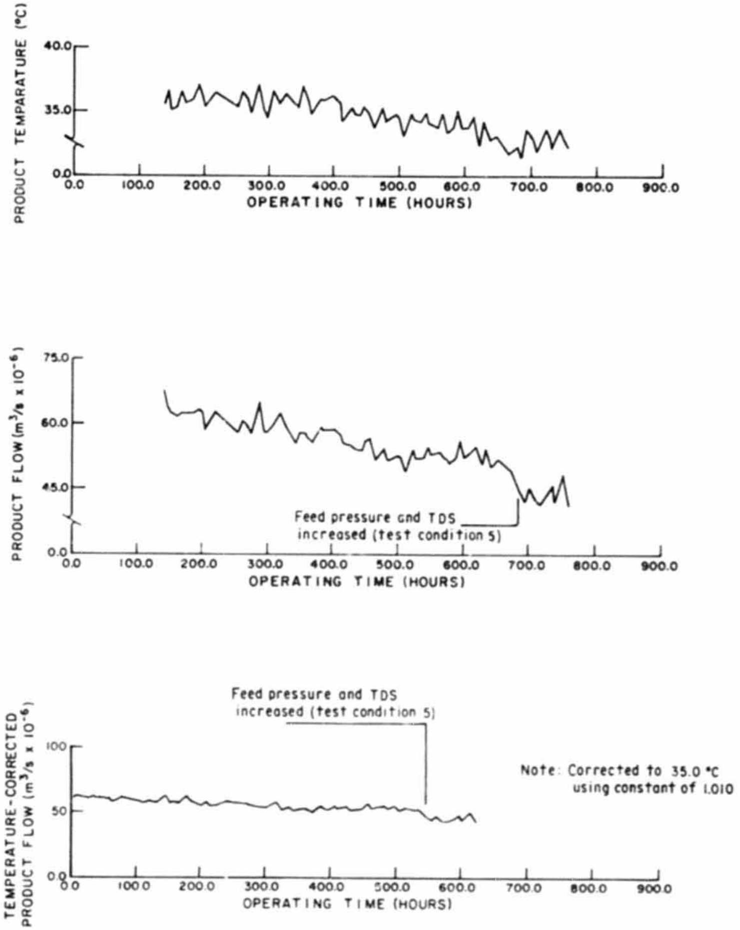


Figure 54 — The UOP membrane 1001 PA (PV-1) performance (sheet 2 of 2)

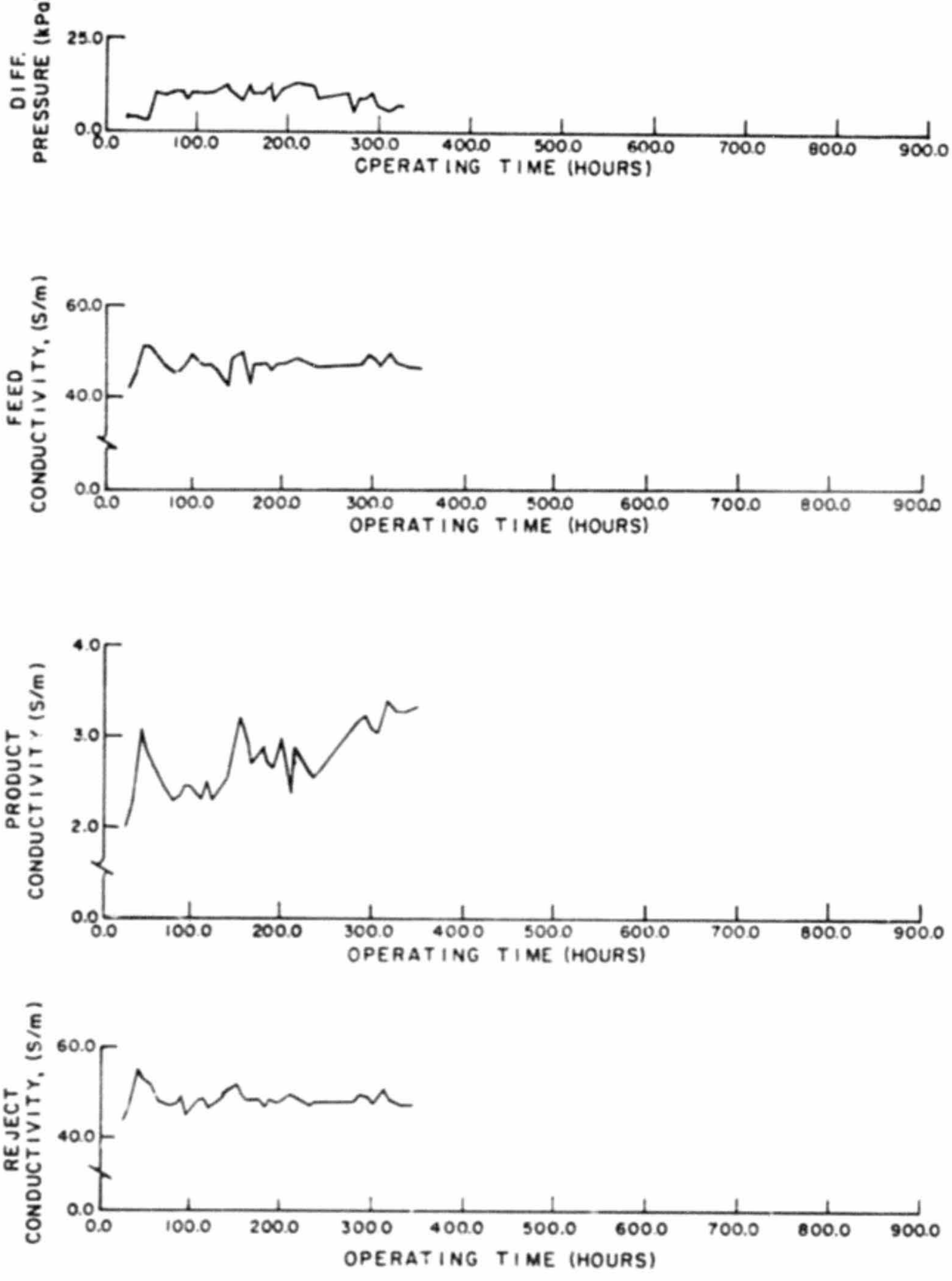
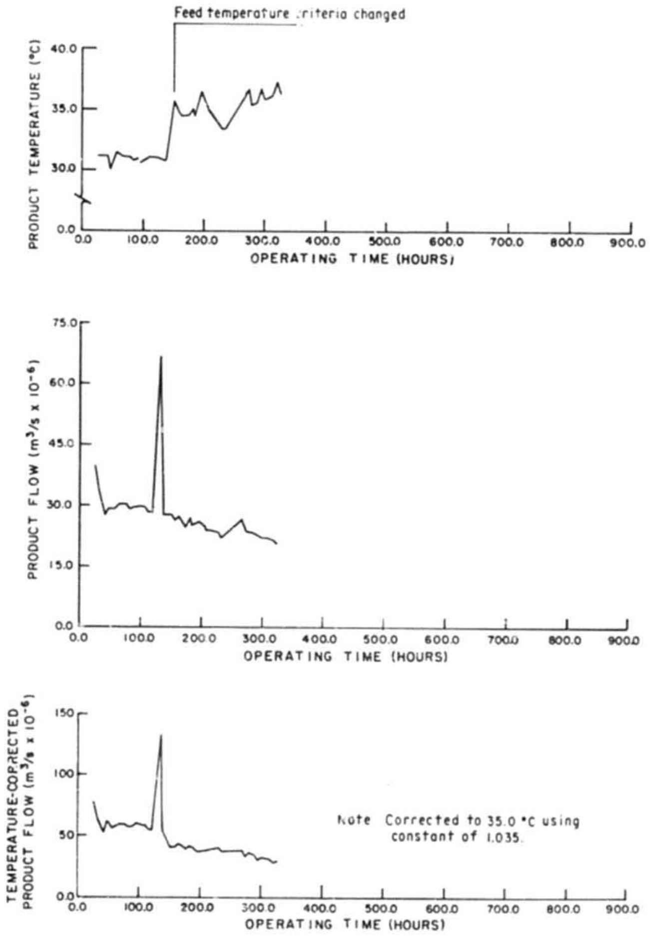


Figure 55 — The UOP membrane 4160 HR (PV-2) performance (sheet 1 of 2).



Note Corrected to 35.0 °C using constant of 1.035.

Figure 55. — The UOP membrane 4160 HR (PV-2) performance (sheet 2 of 2).

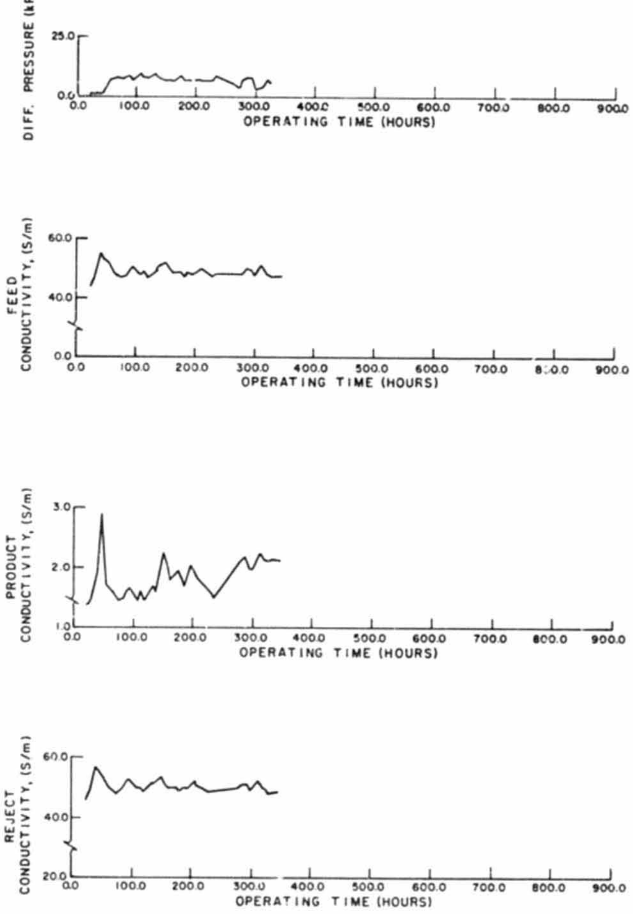


Figure 56. — The UOP membrane 4160 S (PV-3) performance (sheet 1 of 2).

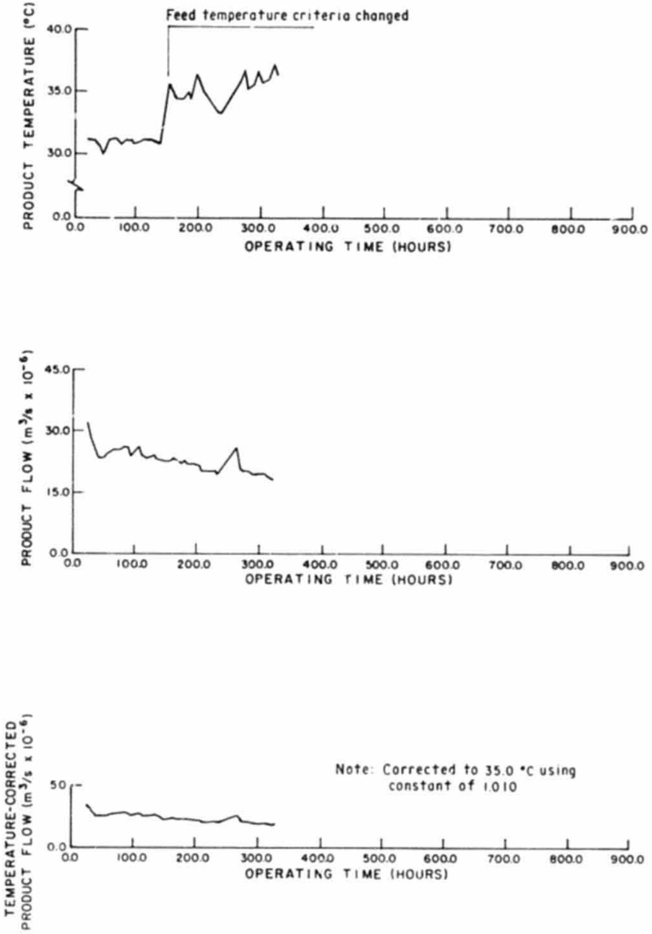


Figure 56. — The UOP membrane 4160 S (PV-3) performance (sheet 2 of 2).

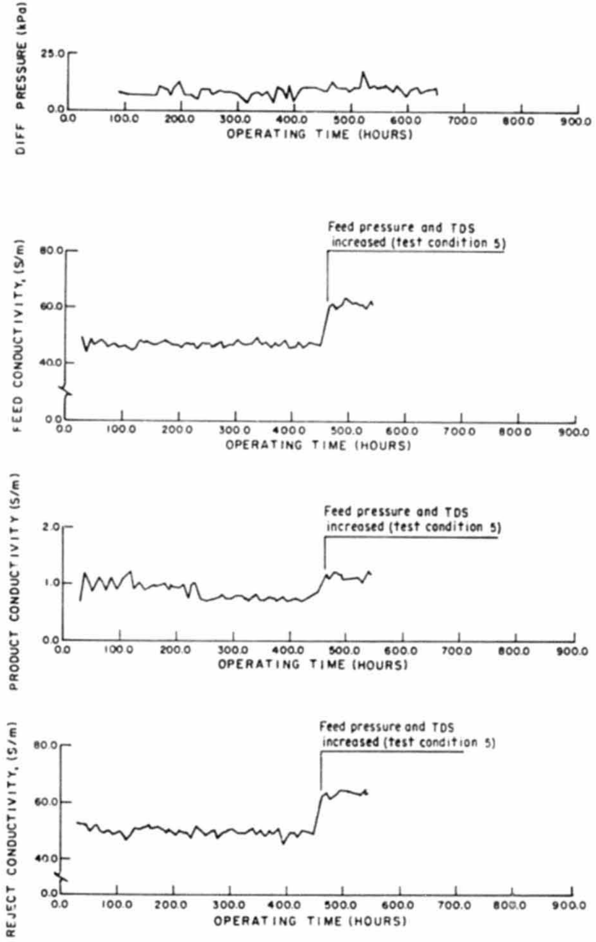


Figure 57. — The UOP membrane 1001 PA (PV-3) performance (sheet 1 of 2).

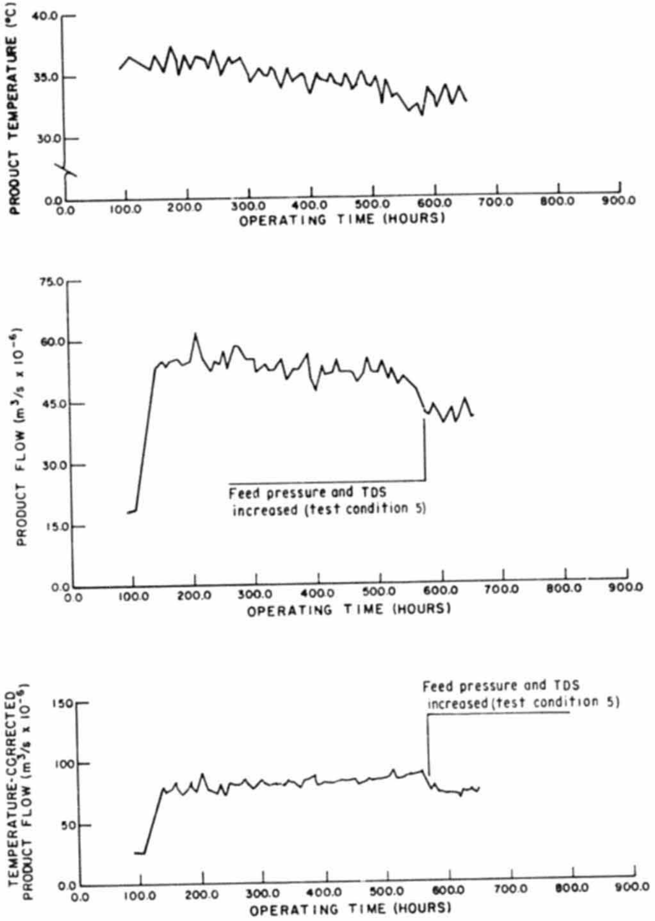


Figure 57. — The UOP membrane 1001 PA (PV-3) performance (sheet 2 of 2).

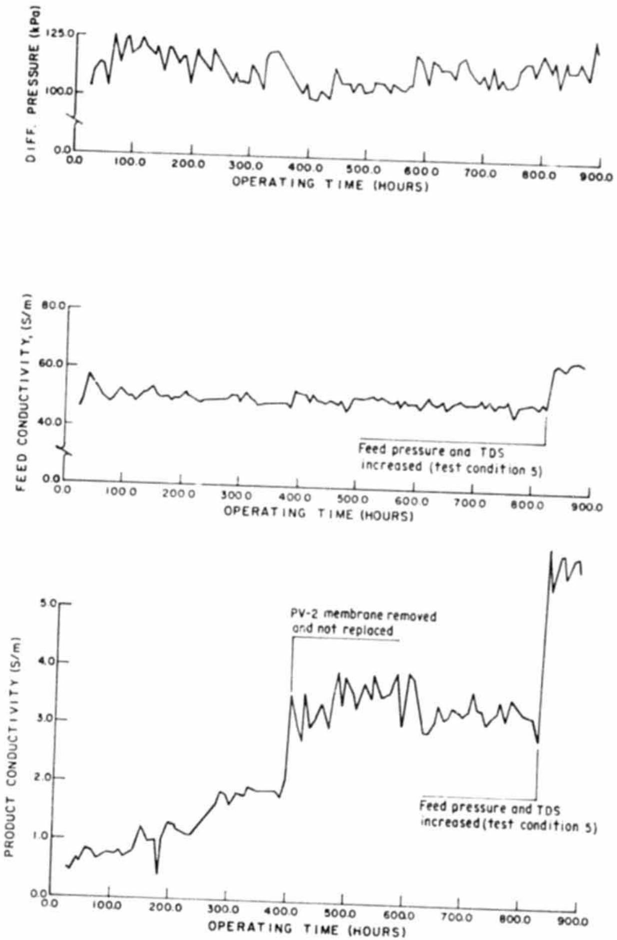


Figure 58. — The DSI membrane B401 (PV-4) performance (sheet 1 of 2).

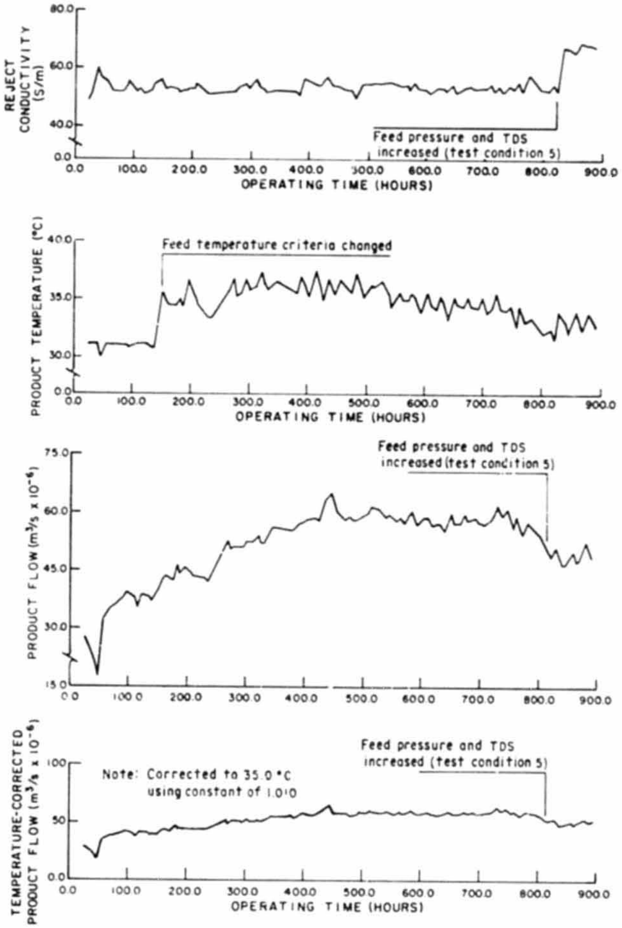


Figure 58 — The DSI membrane B401 (PV-4) performance (sheet 2 of 2)

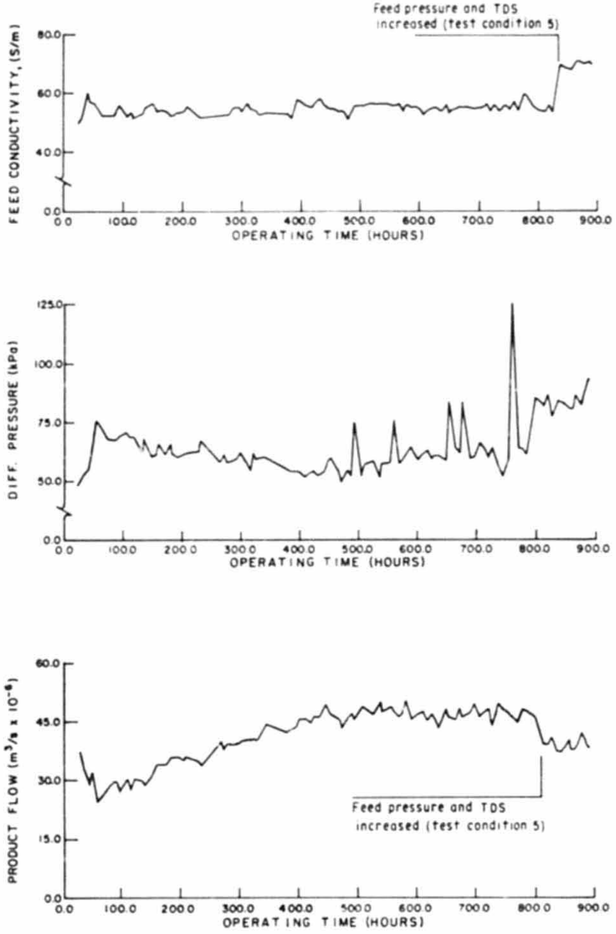


Figure 59 — The DSI membrane B402 (PV-5) performance (sheet 1 of 2)

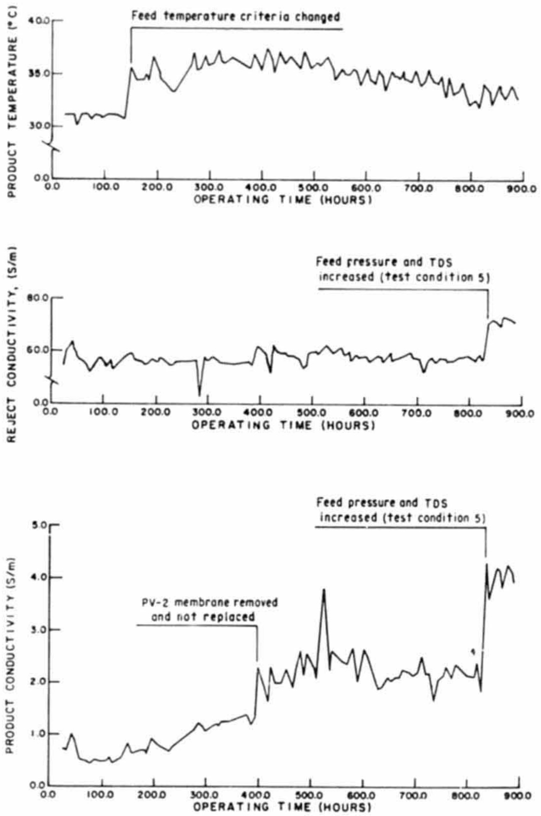


Figure 59 — The DSI membrane B402 (PV-5) performance (sheet 2 of 2)

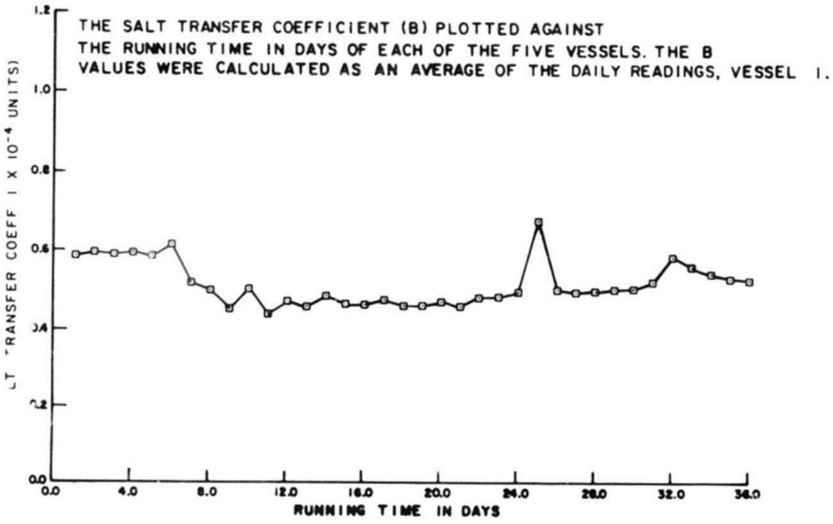


Figure 60 — Fluid Systems membrane 1001 PA salt transfer

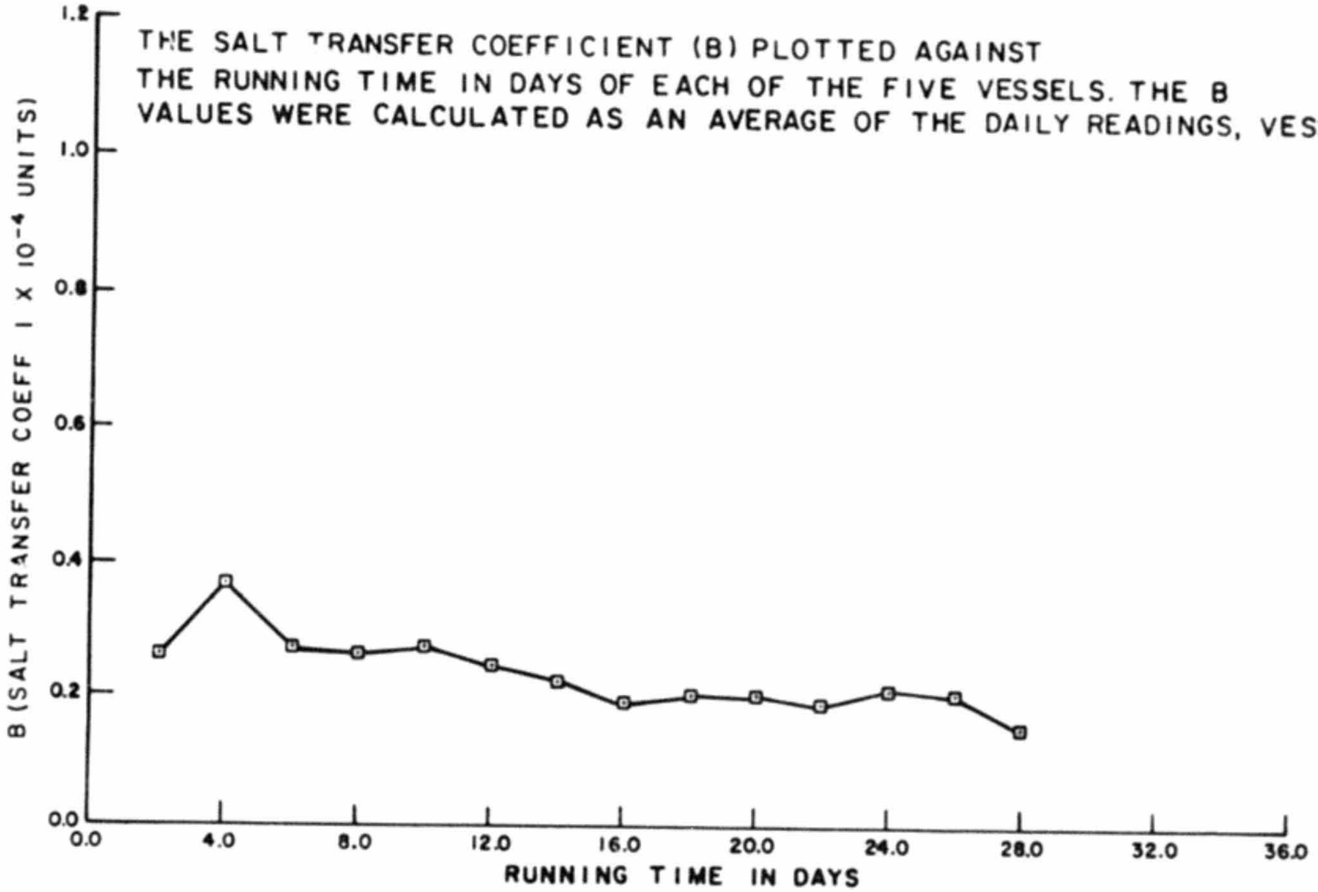


Figure 61. — Fluid Systems membrane 41 60HR salt transfer.

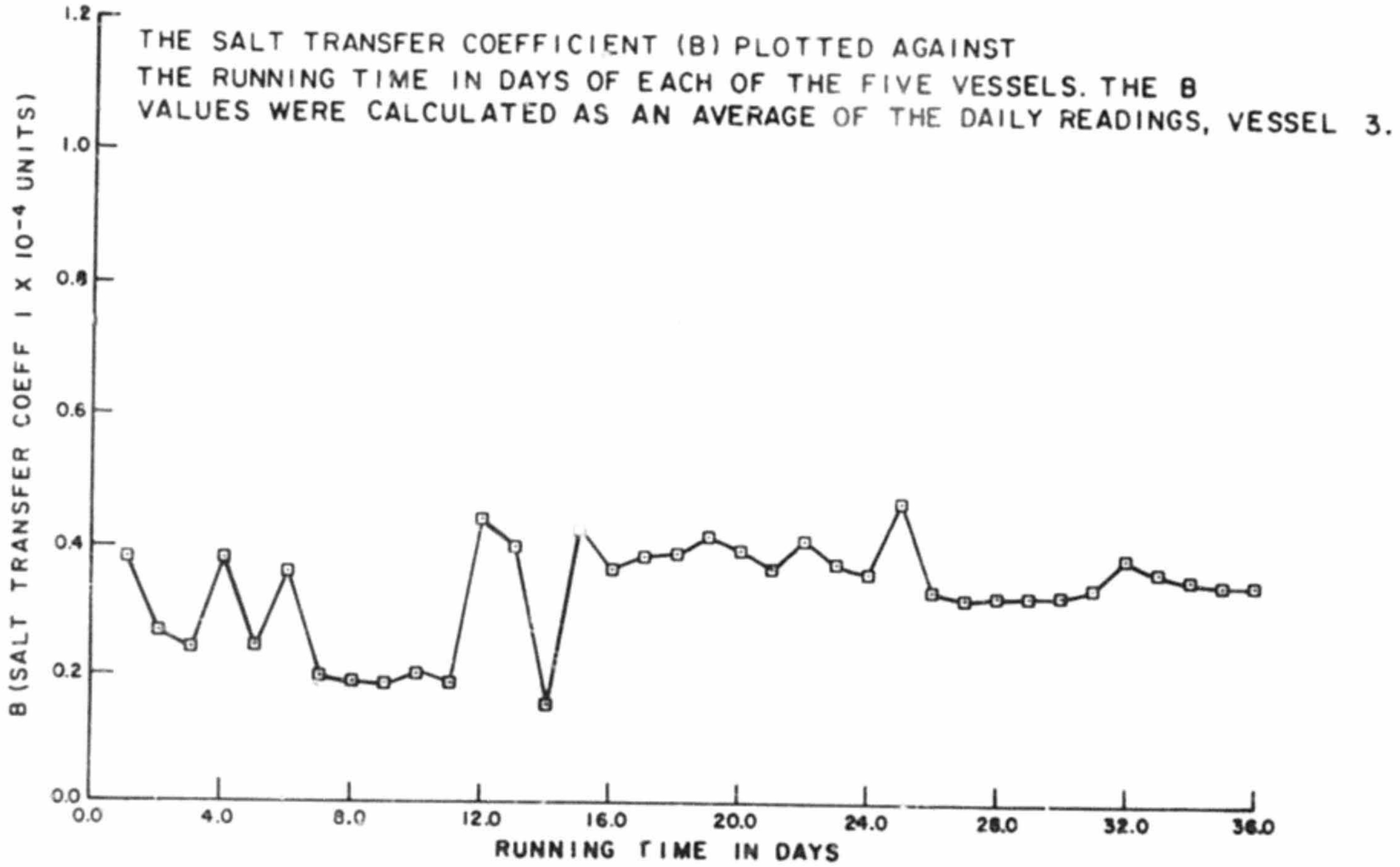


Figure 62. — Fluid Systems membrane 41 60 S salt transfer.

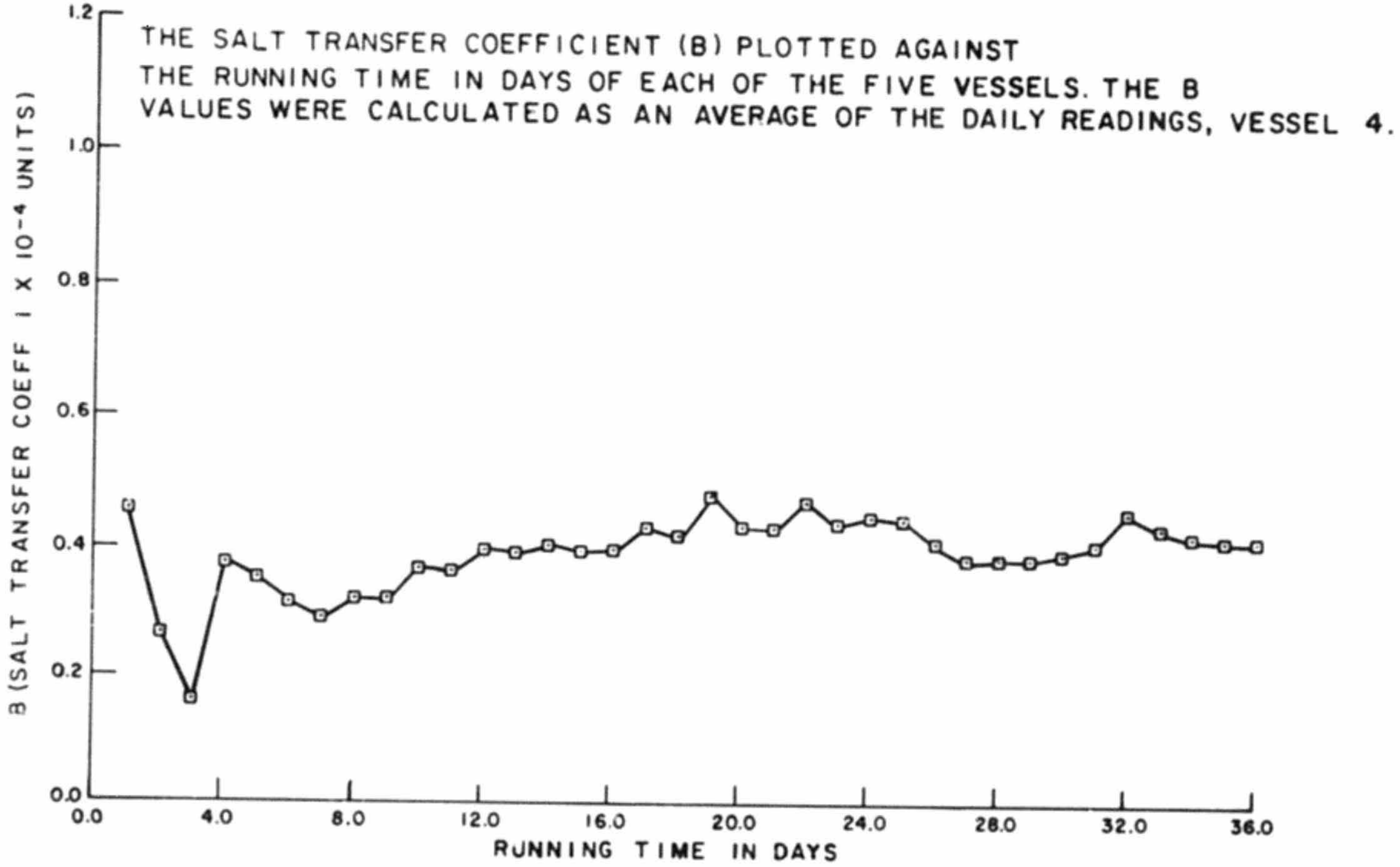


Figure 63. — Desalination Systems membrane 401 salt transfer.

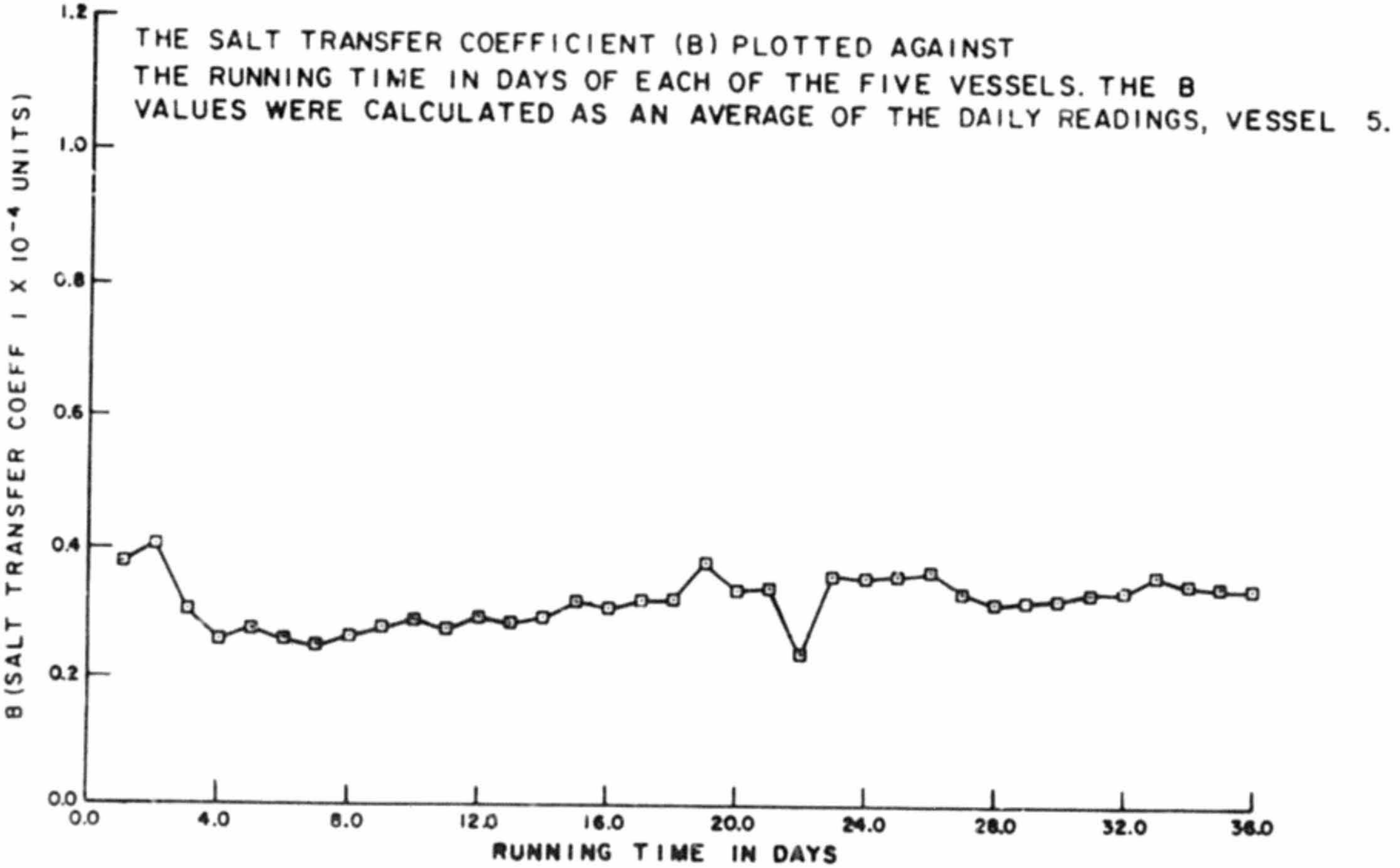


Figure 64 — Desalination Systems membrane 402 salt transfer

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[13] The Permutit Company, "Permutit Spiractor," bulletin No. 5852, Paramus, New Jersey 07652, 1975.

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[15] Buckley, John D., "Reverse Osmosis: Moving from Theory to Practice," *Consulting Engineer*, vol. 45, No. 5, pp. 55-61, November 1975.

[16] Clark, Robert M., *Estimating Water Treatment Cost, Volume 2*, Drinking Water Research Division, Environmental Protection Agency, EPA-600/2-79-162b, August 1979.

APPENDIXES

APPENDIX A

Descriptive inventory of all equipment and trailers at the LaVerkin Springs Site on October 18, 1979

Item	Name	Description
1	Control laboratory trailer	Complete mobile analytical control laboratory and chemist office; trailer 8 feet wide by 31 feet long by 7½ feet internal height and 11 feet external height
2	EDPP (electrodialysis pilot plant) trailer	Ionics, Inc., Aquamite V-1 electrodialysis package assembly; trailer 8 feet by 35 feet long and 8 feet internal height and 12½ feet external height
3	OCC (operators control center) trailer	Contains control panels for the chemical treatment pilot plant, engineering drawing area, bathroom, and coffee area; trailer 8 feet wide by 31 feet long by 7½ feet internal height and 11 feet external height
4	IXPP (ion exchange pilot plant) trailer	Dual 14-inch-diameter ion exchange columns with microprocessor hydraulic controls; trailer 8 feet wide by 40 feet long by 8 feet internal height and 12½ feet external height
5	Shop and electrical control center trailer	Contains maintenance work benches, supply storage cabinet, 7.5-hp air compressor, and the electrical control center for the chemical treatment and the EDPP; trailer 8 feet wide by 35 feet long by 8 feet internal height and 12½ feet external height
6	Supply trailer	When brought to LVS site, contained all the pumps and pipes required for the reassembly of the IX and ED systems; trailer (nothing mounted on floor or walls) 8 feet wide by 28 feet long by 8-foot internal height and 12-foot external height
7	Office trailer	Contains three offices, reproduction room, supply room, office supply room, complete bathroom, and hallways; trailer 12 feet wide by 50 feet long by 8-foot internal height and 11½ feet external height; also has a 8- by 12-foot telescoping room.
8	CTPP-A (chemical treatment pilot plant A), flatbed open trailer	Contains aeration tank, four chemical additive mix tanks, space for future solids contact reactor, and four dual-media filters; flatbed trailer 8 feet wide by 40 feet long by 4 feet high
9	CTPP-B (chemical treatment pilot plant B), flatbed open trailer	Contains rapid mix tank, flocculation tank, and clarifier tank; flatbed trailer 8 feet wide by 24 feet long by 4 feet high
10	Ground set large auxiliary tanks:	
a.	Lime treated water clearwell*	FRP (fiberglass reinforced), flatbottom closed tank with top-mounted agitator support; ⅛ inch thick by 8 feet diameter by 8-foot OAL (overall length)

Item	Name	Description
b.	Ion exchange water clearwell*	FRP flatbottom, closed tank with top-mounted agitator support; ⅜ inch thick by 8 feet diameter by 8-foot OAL
c.	Electrodialysis water clearwell*	FRP flatbottom, open top tank; ⅜ inch thick by 7 feet diameter by 6-foot OAL
d.	Ion exchange reject brine mix tank*	FRP flatbottom, open-top tank; ⅛ inch thick by 4 feet diameter by 11-foot OAL
e.	Calcium sulfate precipitation tank*	FRP skirted, dish-bottom, open-top tank; ⅜ inch thick by 5 feet diameter by 10-foot OAL
f.	Filter backwash tank A*	FRP flatbottom, dished top, closed tank; ⅜ inch thick by 7 feet diameter by 14½-foot OAL
g.	Filter backwash tank B*	FRP flatbottom, open-top tank; ⅛ inch thick by 8½ feet diameter by 8-foot OAL

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* Tanks are named here for reference purposes; tanks might be used in other service for LVS test

APPENDIX B

Table B.1. — *The YDTF laboratory analytical methods summary*

						Detection limit	
Analysis	Method	Source[1]	Use	Instrumentation	Remarks	**Actual**	Reported
Acidity, total	Potentiometric titration	A	Lab	Automatic microburet, 0.05 mL graduations	Titration to pH 8.3 using phenolphthalein indicator. Routine aliquot 25 mL, 1 mL titrant = 1 mg CaCO$_3$	2 g/m^3 as CaCO^3	<2 g/m^3 as CaCO$_3$
Alkalinity, Phenolphthalein (P-alk.)	Potentiometric titration	A	Lab	Automatic microburet, 0.05 mL graduation	Titrated to pH 6.3 employing phenolphthalein indicator. Routine aliquot 25 mL; 1 mg titrant = 1 mg CaCO$_3$	2 g/m^3 as CaCO$_3$	<2 g/m^3 as CaCO$_3$
			OCC		(Same as above)	4 g/m^3 as CaCO$_3$	0 g/m^3 as CaCO$_3$
Alkalinity, total (T-alk.)	Potentiometric titration	A	Lab	Automatic microburet, 0.05 mL graduations	Titrated to pH 6.3 employing bromocresol green-methyl red indicator. Routine aliquot 25 mL; 1 mL titrant = 1 mg CaCO$_3$	2 g/m^3 as CaCO$_3$	<2 g/m^3 as CaCO$_3$
			OCC	Automatic buret, 0.1 mL graduations	(Same as above)	4 g/m^3 as CaCO$_3$	0 g/m^3 as CaCO$_3$
Ammonia-N	Titrimetric (acidimetric)	I, A	Lab	Distillation apparatus; automatic microburet, 0.05 mL gradations	Preliminary distillation. Max. aliquot 300 mL; 1 mL titrant = 0.28 mg NH^3-N	0.05 g/m^3	<0.05 g/m^3
Barium	Automic emission	B, A	Lab	IL 551 AA/AE	By special request only Analyzed by flame emission	Has not been determined	Approx. 0.1 g/m^3
Bicarbonate	Calculation	A	Lab	—	Based on T-alk., P-alk. relationship	0 g/m^3	N.D.
			OCC		(Same as above)	0 g/m^3	0 g/m^3
Calcium	Atomic absorption	B, A	Lab	IL 551 AA/AE	Routine analysis at 1:2 dilution or greater, 0.1 percent La and Cs added	0.05 g/m^3	<0.10 g/m^3
	Atomic absorption	C, A	Lab	Pye-Unicam SP1950	(Same as above)	0.05 g/m^3	<0.1 g/m^3
	EDTA titrimetric	A	OCC	Automatic buret 0.1 mL graduations	Routine aliquot 10 mL, 1 mL titrant = 0.4 mg Ca	4 g/m^3	<4 g/m^3
Calcium hydroxide percent available	AWWA standard for Quickline and Hydrated Lime ANSI/AWWA B202-77	F	Lab	Automatic microburet, 0.05 mL graduations	0.5 g sample size; 1 mL titrant = 1.32 percent avail. Ca(OH)$_2$	0.07%	<0.1%
Carbonate	Calculation	A	Lab	—	Based on T-alk., P-alk. relationship	0 g/m^3	N.D.
			OCC		(Same as above)	0 g/m^3	0 g/m^3
Carbon, total organic	Catalytic combustion, IR detection	A, G	Lab	Beckman 915B TOC analyzer, Beckman 865 IR detector	Calibration 0-20 g/m^3, repeatability ±1 percent full scale	0.2 g/m^3	<0.2 g/m^3

Table B.1. — *The YDTF laboratory analytical methods summary — Continued*

						Detection limit	
Analysis	Method	Source[1]	Use	Instrumentation	Remarks	**Actual**	Reported
		A, O	Lab	Beckman 915 TOC analyzer	Calibration 0-50 g/m^3	0.5	<1.0
Chloride	Argentometric	A	Lab	Automatic microburet, 0.05 mL graduations	Routine aliquot 25 mL, 1 mL titrant = 1 mg Cl	2 g/m^3	<2 g/m^3
	Mercuric nitrate	A	Lab	(Same as above)	(Same as above)	2 g/m^3	<2 g/m^3
Chlorine, total and free	DPD ferrous titrimetric	A	OCC	Automatic buret, 0.1 mL graduations	Routine aliquot 100 mL, 1 mL titrant = 0.2 mg chlorine	0.02 g/m^3	N.D.
Density, weight of 100 mL	Weight determination	-	Lab	Analytical balance	Approximate resolution of 0.1 g/100 mL	100.0 g (no solids present)	100.0 g (no solids present)
Dissolved solids, total	Calculation	-	Lab		Summation of major anions and cations + silica		
	Evaporation	A	Lab	Steam bath, oven, analytical balance	Evaporation on steam bath, dried at 103-105 °C, 100 mL aliquot	Approx. 10 mg/L	<10 mg/L
Electrical conductivity	Wheatstone bridge	A, H	Lab OCC	Beckman RC 18 A, with glass, platinum-electrode cell	Cell constant = 10.0, values corrected to 25 °C	Approx. 0.1 mS/m	<0.1 mS/m
Formaldehyde	MBTH	L	Lab	Pye-Unicam SP1950 UV/visible spectrophotometer	Max. aliquot 50 mL, approx. 2.5- g min detectable quantity with 80- L auto-cell	0.05	<0.05
	Modified Shiff's Test	M	Lab	(Same as above)		Has not been determined	Approx. 0.1 g/m^3
	Sodium sulfite		Lab	Automatic microburet, 0.05 mL graduations	For CH$_2$O > 40 g/m^3. Based on 200 mL max. aliquot. Resolution approx. ±7.5	N/A	N/A
Hardness	EDTA titrimetric	A	OCC	Automatic buret, 0.1 mL graduations	Routine aliquot 5 mL; 1 mL titrant = 1 mg CaCO$_3$	20 g/m^3 as CaCO$_3$	<20 g/m^3 as CaCO$_3$
Hydroxide	Calculation	A	Lab	—	Based on T-alk., P-alk. relationship	0 g/m^3	N.D.
			OCC		(Same as above)	0 g/m^3	0 g/m^3
Insoluble in HCl	Calculation	-	Lab	Gooch crucible, analytical balance	Weight loss calculated after treatment with conc. HCl. One gram sample. Computed to nearest 0.1 percent	0.1%	<0.1%
Iron	Atomic absorption	B	Lab	IL 551 AA/AE	Background correction employed	0.04 g/m^3	<0.04 g/m^3
	Tripyridine	K	Lab	SP1700 UV/visible spectrophotometer	5 g min detectable quantity with 80 L auto-cell and 50 mL max. aliquot	0.10 g/m^3	<0.10 g/m^3
	Phenanthroline	D	OCC	Hach color comparator	0.1 g/m^3 resolution	0.1 g/m^3	<0.1 g/m^3

Table B.1 — *The YDTF laboratory analytical methods summary — Continued*

Analysis	Method	Source[1]	Use	Instrumentation	Remarks	Detection limit	
						Actual	Reported
Magnesium	Atomic absorption	B	Lab	IL 551 AA/AE	Background correction employed. Routine analysis at 1:2 dilution or greater, 0.1 percent La and Cs added.	0.005 g/m^3	<0.01 g/m^3
	Atomic absorption	B	Lab	Pye-Unicam SP1950	(Same as above, but without background correction).	0.010 g/m^3	<0.02 g/m^3
	Calculation	A	OCC	—	Calculated from the EDTA calcium and hardness titrations.	3.9 g/m^3	<4 g/m^3
Manganese	Atomic absorption	B, A	Lab	IL 551 AA/AE	Background correction employed.	0.02 g/m^3	<0.02 g/m^3
	Persulfate	A	Lab	Pye-Unicam SP1750 UV/visible spectrophotometer	10 g min. detectable quantity with 80 L auto-cell and 100 mL max. aliquot.	0.10 g/m^3	<0.10 g/m^3
	Sodium periodate	D	OCC	Hach color comparator	0.1 g/m^3 resolution.	0.1 g/m^3	<0.1 g/m^3
Nitrate-N	Ultraviolet spectrophotometric	A	Lab	Pye-Unicam SP 1750 UV/visible spectrophotometer	2 g NO_3-N min. detectable quantity with 80 L auto-cell and 50 mL max. aliquot.	0.04 g/m^3 NO_3-N 0.20 g/m^3 NO_3	<0.04 g/m^3 NO_3-N <0.20 g/m^3 NO_3
Oxygen dissolved	Winkler (Iodometric)	A	Lab	Automatic microburet, 0.05 mL graduations	Azide and alum flocculation modifications employed as required; 1 mL titrant = mg/L DO	0.1 g/m^3	<0.1 g/m^3
	Meter		Lab	YSI Model 54A	Calibrated by comparison to Winkler method.	0.1 ppm	<0.1 ppm
pH	Electronic meter	A	Lab OCC	Corning 125 pH meter	2-point calibration; read to nearest 0.1 unit.	0.01 unit	<0.1 unit
Phosphate	Stannous Chloride	A	Lab	Pye-Unicam SP1750 UV/visible spectrophotometer	1 g min. detectable quantity with 80 L auto-cell and 100 mL max. aliquot.	0.01 g/m^3	<0.01 g/m^3
Potassium	Atomic absorption	B, A	Lab	IL 551 AA/AE	Routine analysis at 1:2 dilution or greater; 0.1 percent La and Cs added.	0.020 g/m^3	<0.04 g/m^3
	Atomic absorption	C, A	Lab	Pye-Unicam SP1950	(Same as above)	0.06 g/m^3	<0.12 g/m^3
Residue, fixed (FR)	Residue	A	Lab	Analytical balance, laboratory furnace	Calculated from NFR and VR	0.01 g/L	<0.10 g/L
Residue, non-filtrable (NFR)	Residue	A	Lab	Analytical balance, laboratory oven	Glass fiber filter, dried at 103-105 °C.	0.01 g/L	<0.10 g/L
Residue, volatile (VR)	Residue	A	Lab	Analytical balance, laboratory furnace	Residue (NFR) ignited at 550 °C.	0.01 g/L	<0.10 g/L

Table B.1 — *The YDTF laboratory analytical methods summary — Continued*

Analysis	Method	Source	Use	Instrumentation	Remarks	Detection limit	
						Actual	Reported
Silica	Heteropoly Blue (Modified Molybdolosilicate)	A	Lab	Pye-Unicam SP1750 UV/visible spectrophotometer	5 g min. detectable quantity with 80 L auto-cell and 50 mL max. aliquot, at 650 nm. Measures molybdate-reactive silica	0.10 g/m^3	<0.10 g/m^3
	Molybdosilicate	A	Lab	(Same as above)	Approx 50 g min. detectable quantity with 80 L auto-cell and 50 mL max. aliquot, at 410 nm. Measures molybdate-reactive silica.	1.0 g/m^3	<1.0 g/m^3
Sodium	Atomic absorption	B, A	Lab	IL 551 AA/AE	Routine analysis at 1:2 dilution or greater, 0.1 percent La and Cs added.	0.10 g/m^3	<0.20 g/m^3
	Atomic absorption	C, A	Lab	Pye-Unicam SP1950	(Same as above)	0.20 g/m^3	<0.4 g/m^3
Strontium	Atomic absorption	B, A	Lab	IL 551 AA/AE	Routine analysis at 1:2 dilution or greater, 0.1 percent La and Cs added.	0.020 g/m^3	<0.04 g/m^3
	Atomic absorption	C, A	Lab	Pye-Unicam SP1950	(Same as above)	0.03 g/m^3	<0.06 g/m^3
Sulfate	THQ (Tetrahydroxyquinone)	E	Lab	Automatic microburet, 0.05 mL graduations	Max. aliquot 25 mL titrant = 1 mg SO_4 Subject to increased error for samples <20 g/m^3 SO_4	2 g/m^3	<2 g/m^3
	Turbidimetric	A	Lab	Pye-Unicam SP1750 UV/visible spectrophotometer	Approx. min. detectable quantity 100 g with 80 L auto-cell and 100 mL max. aliquot (accord. to Standard Methods)	Has not been determined	Approx <1 g/m^3
Sulfuric acid, percent	Specific gravity	J	Lab	Hydrometers	Determined from sp.g. tables supplied by CITCO	0.0%	N.D.
Turbidity	Nephelometric	N, A	OCC	Monitek 150 turbiometer	Two point calibration with 3 additional check standards	0.1 JTU	<0.1 JTU

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APPENDIX C

Appraisal Level Design With Cost Estimates of a Lime-Soda Ash Pretreatment Plant at LaVerkin Springs

A. Introduction

1. *Purpose*

The purpose of this appraisal level design was to provide support data to USBR (Bureau of Reclamation) planners for preparation of an updated report dealing with the LaVerkin Spring salinity problem. This particular design was specifically requested by USBR and was based on the data obtained from the pilot program established by USBR.

2. *Scope*

This analysis was limited to the pretreatment processes and does not include any provisions for the collection and pumping of the raw spring water or product water storage and pumping. However, dewatering and ultimate disposal of the waste sludge were considered in this analysis. The components of the pretreatment process included aeration for removal of carbon dioxide, lime-soda softening, and filtration. These processes provided for the removal of temporary and permanent hardness.

A brief description of the major process components of a proposed lime-soda softening plant are presented along with cost estimates.

The LaVerkin Springs Unit, Utah, Feasibility Report, July 1973 [8], was used to determine the plant capacities, number of solids contact reactors, and attendant equipment. Other sources of data are referenced where applicable.

B. Conceptual Design

1. *Basis of Project Design*

The results of the chemical analyses of the LVS raw water before and after carbon dioxide stripping and the pilot data derived from the operation of the LaVerkin Springs Test Site were utilized as the basis for sizing the various component systems of the proposed softening plant. These data are presented in the LaVerkin Springs Test Site final report [5].

The design capacity of the proposed softening pretreatment plant was 0.34 m^3/s and was based upon an average annual spring flow of 0.33 m^3/s. The total waste sludge flow after vacuum filtration was estimated at 4.4×10^{-4} m^3/s and was composed of 80 to 85 percent calcium carbonate. The sludge contained radium-226, a naturally occurring radioisotope in the raw water. The average concentration in the raw water was approximately 33 pCi/L (33×10^{-12} Curies per liter). The analysis indicated that approximately 95 percent of the radium-226 would end up in the sludge by the pretreatment process. The concentrations in the sludge had been estimated by Eisenhauer [2] to be 10 pCi/g of 60 weight percent sludge. The disposal of this sludge may present environmental problems that will need to be considered in any future development of the LVS site.

2. *Design of Pretreatment Plant*

This section describes the components of the proposed treatment plant. A process flow diagram for the proposed LaVerkin Springs lime-soda softening pretreatment plant is shown in figure C-1. The plant consisted of air stripping for carbon dioxide removal, lime-soda softening, filtration, and sludge thickening-dewatering. Figure C-2 illustrates a typical layout of an A-L-SA (aeration - lime-soda ash) softening pretreatment facility with a capacity of 0.34 m^3/s. This layout is provided for general planning purposes since final layout should be based on site characteristics.

a. *Aeration system.* — The aeration system consisted of three forced-draft aeration towers. Each tower had a capacity of 50 percent of the total design flow, or 0.17 m^3/s. This allowed one tower to be offline for cleaning purposes without a reduction of the plant inflow rate. The height of a typical tower ranged from 6 to 7 m.

Because of the severe scaling nature of the raw water after removal of CO_2, the towers were constructed of acid-resistant materials so that each tower can be acid cleaned to remove the calcium carbonate scale deposits. A special acid circulation and storage system was required for this system. All acid wastes and rinse waters were disposed of in the sludge disposal system.

b. *Upflow solids contact reactor.* — Three solids contact reactors, each rated at 50 percent of the total design flow, were selected for lime-soda softening. Each reactor had a design capacity of 0.17 m^3/s and an upflow rate of approximately

1.18×10^{-3} m/s. Total reactor hydraulic detention time was 2.25 hours. The reactors each had a diameter of 18 m and a surface area of approximately 250 m^2. Normal operation was with two reactors on-line and the third in a standby mode.

Soda ash, lime, ferric sulfate, and chlorine were added to the reactor which provided mixing as well as clarification. The majority of the sludge in the lime-soda softening process was calcium carbonate which is produced by the reaction between the lime-soda and the source water calcium.

c. *Chemical handling and storage facilities.* — Facilities were provided for receiving and storing a maximum 30-day supply of bulk chemicals used in the softening process. Table C-1 presents the assumed chemical dosages based on the pilot test program and the total monthly and yearly chemical usage quantities required. Capital costs include provisions for the 30-day storage of all chemicals.

d. *Filtration.* — Sulfuric acid was injected into the solids contact reactor effluent to reduce the scaling tendency of the water prior to filtration. Lowering the pH protected the filter media from calcium carbonate scaling which decreased filter performance. Dual media filters were used for filtration of the softened water. The total filter area required was 165 m^2 for a hydraulic loading rate of 2.04×10^{-3} m/s. The maximum surface area per filter was 50 m^2.

Therefore, a minimum of four filters, each with a surface area of approximately 42 m^2, were required. A fifth filter with the same dimensions was provided as a standby unit. The filter media depth was assumed to be 0.8 m.

A backwash water storage tank, and a surge/decant tank were included with the filter. A 10-minute backwash time was assumed. The volume of each of these storage tanks was approximately 310 m^3. The backwash loading rate was approximately 12.2×10^{-3} m/s.

e. *Sludge disposal.* — Backwash waste water was decanted as the initial solids concentration process. Sludge from the solids contact reactor and the gravity filter decant tank was thickened and dewatered. The solids content was increased by the use of a vacuum filter, and the sludge was ultimately to be disposed of in a landfill. Approximately 13 900 m^3 of concentrated sludge (approximately 60 percent by weight) would be produced per year.

f. *Design criteria summary.* — Table C-2 presents a summary of the design criteria of the proposed A-L-SA pretreatment softening plant. Included are the aeration system, solids contact reactor, gravity filters, and sludge disposal.

C. Cost Estimates

This section presents both capital and O&M cost data for the proposed A-L-SA softening pretreatment plant at LaVerkin Springs.

1. *Construction Costs*

Table C-3 presents construction cost data for the components of the proposed lime-soda softening plant. These costs have been adopted from the EPA document "Estimating Water Treatment Costs" [16]. The costs presented for LVS are based on an Engineering New Record Construction Cost Index value of 297.73 which is representative of June 1980. Since the sludge produced in the process will be radioactive to a small extent, an additional 25 percent was added to the cost of vacuum filtration and landfill disposal. This allowed for protective handling and special procedures which would be required during sludge handling.

2. *O&M Costs*

Table C-4 presents O&M cost data for the components of the proposed pretreatment plant. Labor and energy are expressed in manhours/yr and kWh/yr, respectively.

Maintenance materials are based upon an updated Producer Price Index of 242.1 for June 1980. Chemical costs are based upon estimates by McKesson Chemicals in Phoenix, Arizona. The chemical price estimates reflect delivered prices to Hurricane, Utah. A summary of the unit chemical prices is shown below:

	$/metric ton
a. Chlorine (1-ton cylinders)	317.51
b. Sulfuric acid	45.36
c. Ferric sulfate	272.00
d. Soda ash	110.67
e. Quicklime (as CaO)	56.62

3. *Total Costs*

Table C-5 presents a cost estimate summary of the proposed A-L-SA pretreatment plant. Included are total annual capital, total annual O&M, and unit costs. The analysis shows the unit cost of water for the proposed lime-soda softening pretreatment plant is approximately $301/acre-foot, (or $0.244/$m^3$). An average labor rate of $12 per hour and an electrical power rate of $0.05 per kWh were assumed in preparation of the total annual O&M costs.

An estimated annual O&M cost of $61,000 for vacuum filtration and sludge disposal (table C.5) would be approximately $4.40/$m^3$ of sludge.

Table C-1. — *Chemical storage quantities*

Chemical	Dosage	Monthly usage (metric tons/month)	Yearly usage (metric tons/year)
Lime (as CaO)	473 g/m³	416.8	5 002
Soda ash	1200 g/m³	1057	12 684
Ferric sultate	7.5 g/m³	6.6	79.2
Sulfuric acid	1325 L/day	39.7	476.4
Chlorine	1 g/m³	0.88	10.56

Table C.2. — *Design criteria summary of the proposed LaVerkin Springs aeration-lime-soda ash softening pretreatment plant*

Design Flow Rate	$0.34\ m^3/s$
Aeration System	
Number of towers	3
Capacity of each tower	$0.17\ m^3/s$
Typical tower height	6.5 m
Solids Contact Reactor	
Clarification flow rate	$1.18 \times 10^3\ m/s$
Total reactor detention time	2.25 h
Number of reactors (includes one standby)	3
Area of each reactor	$250\ m^2$
Diameter of each reactor	18m
Gravity Filters	
Hydraulic loading rate	$2.04 \times 10^{-3}\ m/s$
Number of filters (includes one standby)	5
Area of each filter	$42\ m^2$
Filter media depth	0.8 m
Nominal backwash loading rate	$12.2 \times 10^{-3}\ m/s$
Backwash time	10 min
Volume of backwash storage tank	$310\ m^3$
Volume of surge/decant tank	$310\ m^3$
Sludge Disposal	
Method	Vacuum filtration and landfill disposal
Total sludge volume (concentrated to approximately 60 percent solids by weight)	$13\ 900\ m^3/yr$

Table C.3. — *Pretreatment plant construction costs*

	Component	Cost
1.	Aeration towers	$ 675,000
2.	Lime feed system	123,000
3.	Soda ash feed system	185,000
4.	Ferric sulfate feed system	28,000
5.	Chlorine feed system	32,500
6.	Sulfuric acid feed system	25,000
7.	Upflow solids contact reactors	595,000
8.	Gravity filters (includes media, housing, surge/decant tank, and backwash storage water tank)	1,380,000
9.	Administration, laboratory, and maintenance	88,000
10.	Vacuum filtration and landfill disposal of sludge	356,000
11.	Total construction cost	$3,487,500

Based on a design flow rate of $0.34\ m^3/s$.

Table C.4. — *Pretreatment O&M costs*

	Component	Labor (manhours/hr)	Energy (kWh/hr)	Maintenance materials ($/yr)	Chemicals ($/yr)
1.	Aeration towers	400	40 000	4 900	—
2.	Lime feed system	2 350	63 000	1 760	287 200
3.	Soda ash feed system	2 850	78 000	2 400	1 424 000
4.	Ferric sulfate feed system	285	15 000	230	21 900
5.	Chlorine feed system	470	3 560	2 800	3 400
6.	Sulfuric acid feed system	190	3 680	140	21 900
7.	Upflow solids contact reactor	2 190	190 000	3 800	—
8.	Gravity filters	3 500	463 000	13 500	—
9.	Administration, laboratory, and maintenance	4 800	190 000	4 500	—
10.	Vacuum filtration and landfill disposal of sludge	2 940	212 500	15 600	—
	Total	19 975	1 258 740	49 630	1 758 400

Based on a design flow rate of $0.34\ m^3/s$.

Table C.5. — *Pretreatment total cost summary*

I.	**Capital Costs**	
	1. Construction costs	$3,437,500
	2. Contingency factor 32.5 percent of construction (includes engineering, legal, fiscal administrative, etc.)	1,133,400
	3. Total capital cost (1) + (2)	4,620,900
	4. Expected life of facility	25 years
	5. Interest rate	10 percent
	6. Capital recovery factor	0.1101
	7. Total annual capital cost (6) x (3)	$508,760/yr
II.	**O&M Costs**	
	8. Maintenance materials	$ 49,630/yr
	9. Chemicals	$1,758,400/yr
	10. Labor (19,975 manhours/yr at $12/manhour)	$239,700/yr
	11. Energy (1,258,740 kWh/yr at $0.05 kWh)	$62,900/yr
	12. Total annual O&M costs	$2,110,630/yr
III.	**Total Annual Costs**	
	13. Total annual cost (7) + (12)	$2,619,390/yr
	14. Unit costs:	
	a. $/1,000 gallons	0.92
	b. $/acre-foot	301.3
	c. $/m^3	0.244

Based on a design flow rate of 0.34 m^3/s.

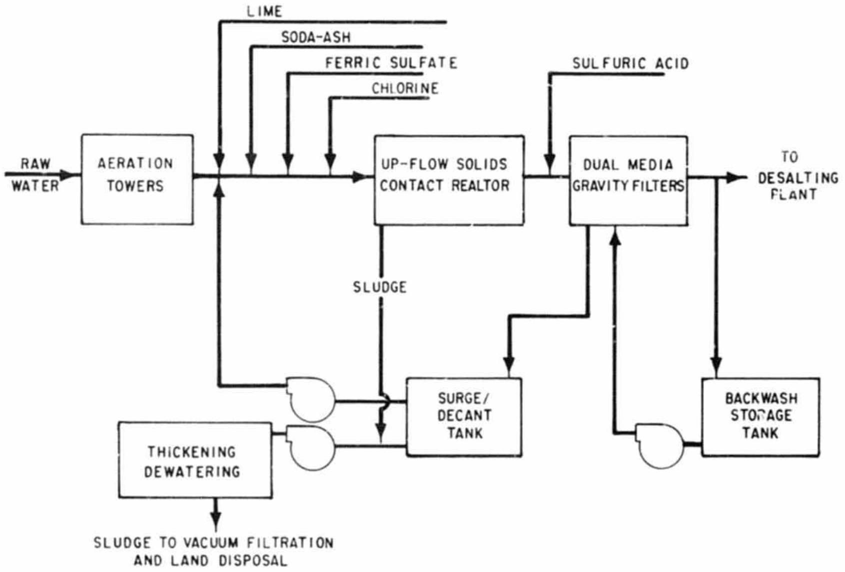


Figure C.1. — Process flow diagram for the proposed LaVerkin Springs lime-soda softening pretreatment plant.

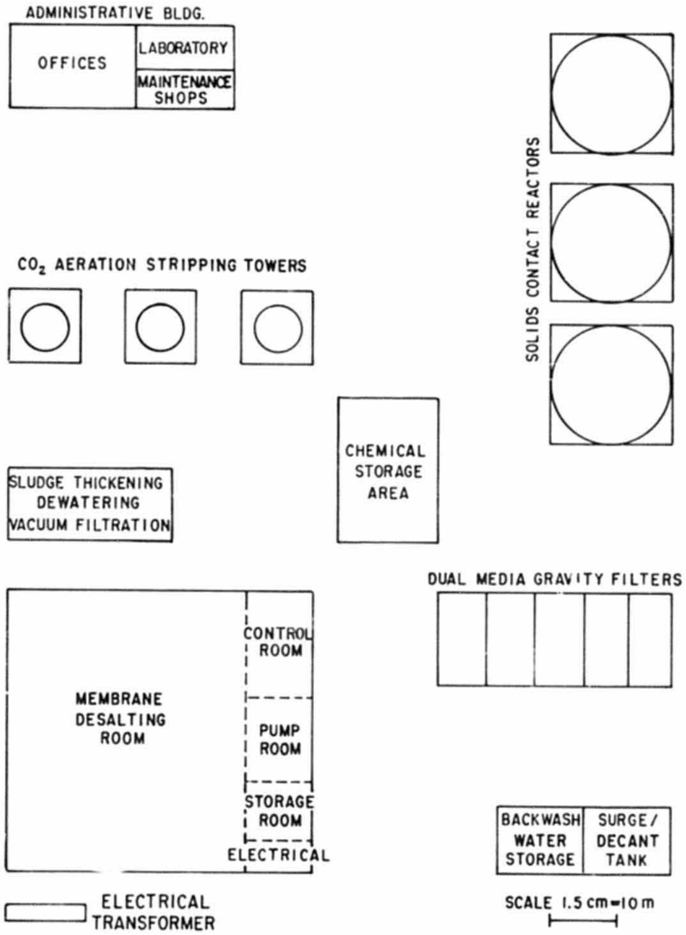


Figure C.2. — Schematic layout of the proposed LaVerkin Springs lime-soda softening pretreatment plant.

APPENDIX D

Appraisal Level Design With Cost Estimates of a Lime-ion Exchange Pretreatment Plant at LaVerkin Springs

A. Introduction

1. *Purpose*

See appendix C.

2. *Scope*

The same as appendix C, except a lime-ion exchange process is proposed instead of a lime-soda ash process.

B. Conceptual Design

1. *Basis of Project Design*

The same as appendix C, except disposal of waste sludge will be reduced due to the deletion of soda ash treatment and an additional disposal requirement will occur due to the generation of ion-exchange spent regenerant and rinses. The total waste sludge flow after vacuum filtration resulting from aeration and lime treatment was estimated a 2.1×10^{4} m^3/s, composed of 80 to 85 percent calcium carbonate. The total flows of IX spent regenerants and rinses to waste were estimated as 0.08 m^3/s. The chemical composition of the IX spent water waste can be determined from table 32. Ion exchange softening with brine regeneration is a viable pretreatment option only if a desalting plant is built at the LVS site. The reject brine normally requiring disposal at the LVS site desalting plant would be used to regenerate the IX resin and ultimately be disposed of as waste IX spent regenerants and rinses.

2. *Design of Pretreatment Plant*

A process flow diagram for the proposed LaVerkin Springs A-L-IX softening pretreatment plant is shown in figure D.1. The plant consists of air stripping of carbon dioxide for removal of calcium solubilized by carbon dioxide, lime treatment for removal of calcium associated with bicarbonate, dual media filtration, sludge thickening-dewatering, and IX treatment to remove calcium in solution as chloride salt.

a. The aeration system was considered identical to the system described in appendix C.

b. The upflow solids contact reactors were the same as described in appendix C.

c. The chemical handling and storage facilities were the same as appendix C, except soda ash was deleted. The initial charge of IX resin was considered a capital expenditure. However, the chemical storage quantities in table D.1 included resin makeup for attrition and carryout losses. About 1 metric ton per year of resin would adequately take care of system losses. Since the IX process proposed will utilize reject brine from the desalting process, the chemical storage did not include any commercial sodium chloride for regeneration of the resin.

d. Filtration was essentially the same. Similar size filters would be required for the A-L-IX process as was specified in appendix A for the A-L-SA process.

e. The quantity of sludge for disposal was 50 percent of the amount in appendix A, or 7000 m^3 per year.

f. Design criteria summary. — Table D-2 presents a summary of the design parameters of the proposed A-L-IX pretreatment softening plant. Included are the aeration system, solids contact reactors, gravity filters, sludge disposal, and IX columns.

C. Cost Estimates

This section presents both capital and O&M cost data for the proposed A-L-IX softening pretreatment plant at LaVerkin Springs.

1. *Construction costs*

Table D-3 presents construction cost data for the components of the proposed lime-IX softening plant. These costs have been adopted from the EPA document "Estimating Water Treatment Costs" [13]. The costs presented for LVS are based on an Engineering News Record Construction Cost Index value of 297.73, representative of June 1980. Since the sludge produced in the process will be radioactive to a small extent, an additional 25 percent has been added to the cost of vacuum filtration and landfill disposal. This allows for protective handling and special procedures which would be required during sludge handling.

2. *O&M Costs*

Table D-4 presents O&M cost data for the components of the proposed pretreatment plant. Labor and energy are expressed in manhours/yr and kWh/yr, respectively. Maintenance materials are based upon an updated Producer Price Index of 242.1 for June 1980. Chemical costs are based upon estimates by McKesson Chemicals in Phoenix, Arizona. The chemical price estimates reflect delivered prices to Hurricane, Utah. A summary of the unit chemical prices is shown below:

	\$/metric ton
Chlorine (1-ton cylinders)	317.51
Sulfuric acid	45.36
Ferric sulfate	272.00
Quicklime (as CaO)	56.62

3. *Total Costs*

Table D-5 presents a cost estimate summary of the proposed LVS pretreatment plant. Included are total annual capital, total annual O&M, and unit costs. The analysis shows that the unit cost of water for the proposed A-L-IX softening pretreatment plant is approximately \$146/acre-ft (or \$0.118/m^3). An average labor rate of \$12/h and an electrical power rate of \$0.05/kWh have been assumed in preparation of the total annual O&M costs.

Annual O&M cost of approximately \$61,000 for vacuum filtration and sludge disposal is included in item 12 of table C-5. This represents approximately \$4.40/m^3 of sludge.

Table D.1 — Chemical storage quantities

Chemical	Dosage	Monthly usage (metric tons/mo)	Yearly usage (metric tons/yr)
Lime (as CaO)	473 g/m^3	416.8	5002
Ion exchange resin (makeup)			1
Ferric sulfate	7.5 g/m^3	6.6	79.2
Sulfuric acid	1325L/day	39.7	476.4
Chlorine	1g/m^3	0.88	10.56

Table D2. — *Design criteria summary of the proposed LaVerkin Springs aeration-lime-IX softening pretreatment plant*

Design Flow Rate	0.34 m^3/s
Aeration System	
Number of towers	3
Capacity of each tower	0.17 m^3/s
Typical tower height	6.5 m
Solids Contact Reactor	
Clarification flow rate	1.18×10^{-3} m/s
Total reactor detention time	2.25 h
Number of reactors (includes one standby)	3
Area of each reactor	250 m^2
Diameter of each reactor	18 m
Gravity Filters	
Hydraulic loading rate	2.04×10^{-3} m/s
Number of filters (includes one standby)	5
Area of each filter	42 m^2
Filter media depth	0.8 m
Nominal backwash loading rate	12.2×10^{-3} m/s
Backwash time	10 min
Volume of backwash storage tank	310 m^3
Volume of surge/decant tank	310 m^3
Sludge Disposal	
Method	Vacuum filtration and landfill disposal
Total sludge volume (concentrated to approximately 60 percent solids by weight)	13 900 m^3/yr
IX System	
Plant flow rate	0.34 m^3/s (7.76 Mgal/d)
Initial charge of resin	78 m^3 (2756 ft^3)
Calcium removal	20.0 eq/m^3 of water
Number of contactors	6
Diameter of contactors	3.6 m (12 ft)
Housing	410 m^2 (4690 ft^2)

Table D3. — *Pretreatment plant construction costs*

Component	Cost
1. Aeration towers	$ 675,000
2. Lime feed system	185,000
3. Ferric sulfate feed system	28,000
4. Chlorine feed system	33,000
5. Sulfuric acid feed system	25,000
6. Upflow solids contact reactors	595,000
7. Gravity filters (includes media, housing, surge/decant tank, and backwash storage water tank	1,380,000
8. IX system	856,000
9. Administration, laboratory, and maintenance	88,000
10. Vacuum filtration and landfill disposal of sludge	180,000
11. Total construction cost	$4,045,000

Table D4. — *Pretreatment O&M costs*[1]

Component	Labor (manhours/yr)	Energy (kWh/yr)	Maintenance materials ($/yr)	Chemicals ($/yr)
1. Aeration tower	400	40 000	4,900	287,200
2. Lime feed system	2 350	63 000	1,760	21,900
3. Ferric sulfate feed system	290	15 000	230	3,400
4. Chlorine feed system	470	3 560	2,800	21,900
5. Sulfuric acid feed system	190	3 680	140	—
6. Upflow solids contact reactor	2 190	190 000	3,800	—
7. Gravity filters	3 500	463 000	13,500	—
8. Administration, laboratory, and maintenance	4 800	190 000	4,500	—
9. Vacuum filtration and landfill disposal of sludge	2 100	110 000	15,600	—
10. IX system	2 200	280 600	3,500	--
Total	18 490	1 358 840	50,730	334,400

[1] Based on a design flow rate of 0.34 m^3/s.

Table D5. — *Pretreatment total cost summary*[1]

I. **Capital Costs**	
A. Construction costs	$4,045,000
B. Contingency factor of 32.5 percent of construction (includes engineering, legal, fiscal administrative)	$1,314,620
C. Total capital cost (A) + (B)	$5,359,620
D. Expected life of facility	25 years
E. Interest rate	10 percent
F. Capital recovery factor	0.1101
G. Total annual capital cost (F) x (C)	$ 590,090/yr
II. **O&M Costs**	
H. Maintenance materials	$ 50,730
I. Chemicals	$ 334,400/yr
J. Labor (18,490 manhours/yr at $12./manhour)	$ 221,880/yr
K. Energy (1,358,840 kWh/yr at $0.05 kWh)	$ 67,940/yr
L. Total annual O&M costs	$ 674,950/yr
III. **Total Annual Costs**	
M. Total annual cost (G) + (L)	$1,265,040/yr
N. Unit costs:	
a. $/1,000 gallons	0.44
b. $/acre-foot	145.5
c. $/m^3	0.118

[1] Based on a design flow rate of 0.34 m^3/s.

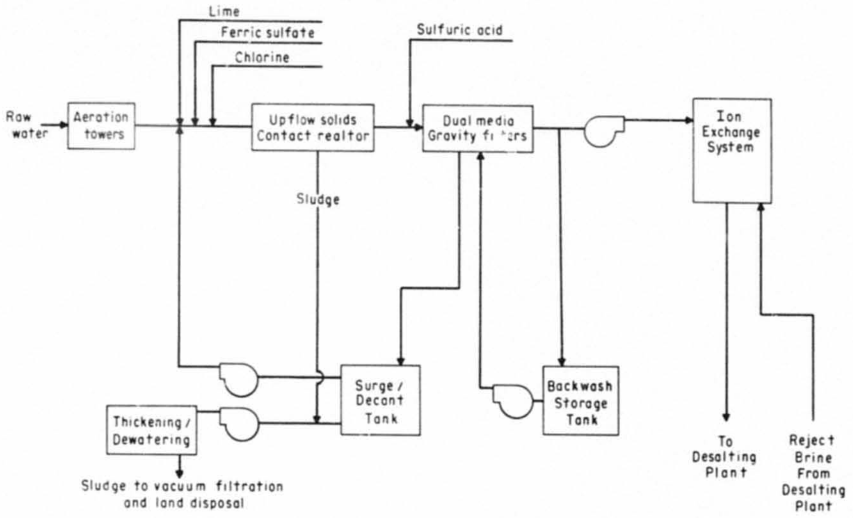


Figure D.1. — Flow diagram for the proposed La Verkin Springs Aeration-Lime-Ion exchange softening pretreatment plant.

APPENDIX E

Appraisal Level Design With Cost Estimates of a Lime Pretreatment Plant at LaVerkin Springs[1]

Table E.1. — *Pretreatment plant construction costs*

Component	Cost
1. Aeration towers	$ 675,000
2. Lime feed system	123,000
3. Ferric sulfate feed system	28,000
4. Chlorine feed system	33,000
5. Sulfuric acid feed system	25,000
6. Upflow solids contact reactors	595,000
7. Gravity filters (includes media, housing, surge/decant tank, and backwash storage water tank)	1,380,000
8. Admnistration, laboratory, and maintenance	88,000
9. Vacuum filtration and landfill disposal of sludge	180,000
10. Total construction cost	$3,127,000

[1] Based on a design flow rate of 0.34 m^3/s

Table E.2. — *Pretreatment O&M costs*[1]

Component	Labor (manhours/hr)	Energy (kWh/hr)	Maintenance materials ($/yr)	Chemicals ($/yr)
1. Aeration towers	400	40 000	4,900	—
2. Lime feed system	2 350	63 000	1,760	287,200
3. Ferric sulfate feed system	290	15 000	230	21,900
4. Chlorine feed system	470	3 560	2,800	3,400
5. Sulfuric acid system	190	3 680	140	21,900
6. Upflow solids contact reactor	2 190	190 000	3,800	—
7. Gravity filters	3 500	463 000	13,500	—
8. Administration, laboratory, and maintenance	4 800	190 000	4,500	—
9. Vacuum filtration and landfill disposal of sludge	2 100	110 000	15,600	—
10. Total	16 290	1 078 240	47,230	334,400

[1] Based on a design flow rate of 0.34 m^3/s

Table E.3. — *Pretreatment total cost summary*[1]

I.	**Capital Costs**	
	A. Construction costs	$3,127,000
	B. Contingency factor of 32.5 percent of construction (includes engineering, legal, fiscal administrative)	$1,016,280
	C. Total capital cost (A) + (B)	$4,143,280
	D. Expected life of facility	25 years
	E. Interest rate	10 percent
	F. Capital recovery factor	0.1101
	G. Total annual capital cost (F) x (C)	$ 456,180/yr
II.	**O&M Costs**	
	H. Maintenance materials	$ 47,230/yr
	I. Chemicals	$ 334,400/yr
	J. Labor (16,290 manhours/yr at $12/manhour)	$ 195,480/hr
	K. Energy (1,078,240 kWh/yr at $0.05 kWh)	$ 53,910/hr
	L. Total annual O&M costs	$ 631,020/hr
III.	**Total Annual Costs**	
	M. Total annual cost (G) + (L)	$1,087,200/hr
	N. Unit costs:	
	1. $/100,000 gallons	0.38
	2. $/acre-foot	125.0
	3. $/$m^3$	0.101

[1] Based on a design flow rate of 0.34 m^3/s

APPENDIX F

Appraisal Level Design With Cost Estimates of an Aeration Pretreatment Plant

Table F.1. — *Pretreatment plant construction costs*[1]

	Component	Cost
1.	Aeration towers	$ 675,000
2.	Ferric sulfate feed system	28,000
3.	Chlorine feed system	33,000
4.	Circular clarifier (140 ft)	524,770
5.	Administratrion, laboratory, and maintenance	88,000
6.	Vacuum filtration and landfill disposal of sludge	119,000
7.	Total construction cost	$1,467,770

[1] Based on a design flow rate of 0.34 m^3/s

Table F.2. — *Pretreatment O&M costs*

Component	Labor (manhours/hr)	Energy (kWh/hr)	Maintenance materials ($/hr)	Chemicals ($/hr)
1. Aeration towers	400	40 000	4,900	—
2. Ferric sulfate feed system	290	15 000	230	21,900
3. Chlorine feed system	470	3 560	2,800	3,400
4. Circular clarifier	360	10 450	2,300	—
5. Administration, laboratory, and maintenance	4800	190 000	4,500	—
6. Vacuum filtration and landfill disposal of sludge	2100	110 000	15,600	—
7. Total	8420	369 010	30,330	35,300

[1] Based on a design flow rate of 0.34 m^3/s

Table F.3. — *Pretreatment total cost summary*[1]

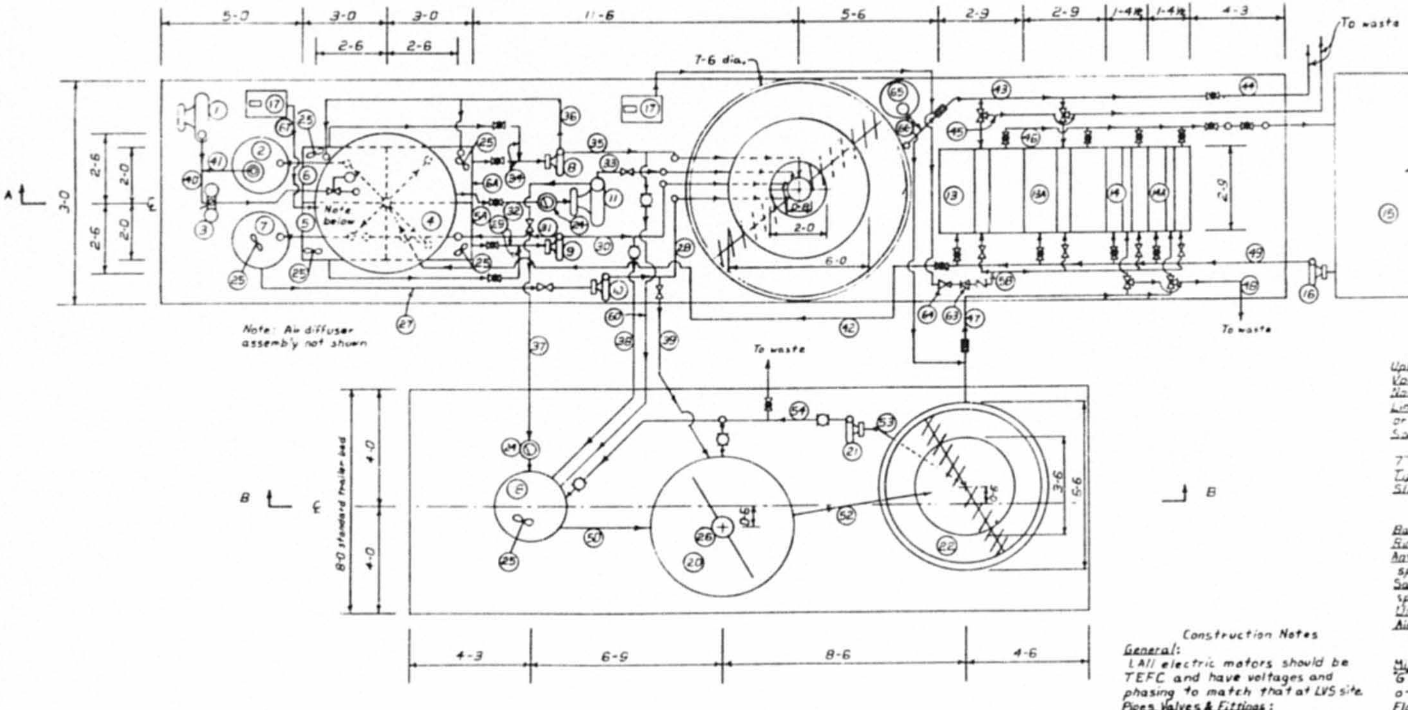
I.	**Capital Costs**	
	A. Construction costs	$1,467,700
	B. Contingency factor 32.5 percent of construction (includes engineering, legal, fiscal administrative)	$ 477,000
	C. Total capital cost (A) + (B)	$1,944,770
	D. Expected life of facility	25 years
	E. Interest rate	10 percent
	F. Capital recovery factor	0.1101
	G. Total annual capital cost (F) x (C)	$ 214,119/yr
II.	**O&M Costs**	
	H. Maintenance materials	$ 30,330
	I. Chemicals	$ 35,330/hr
	J. Labor (8,420 manhours/yr at $12/manhour)	$ 101,040/yr
	K. Energy (369,010 kWh/hr at $0.05 kWh)	$ 18,450/yr
	L. Total annual O&M costs	$ 185,150/yr
III.	**Total Annual Costs**	
	M. Total annual cost (G) + (L)	$ 399,270/yr
	N. Unit costs:	
	1. $/1,000 gallons	0.14
	2. $/acre-foot	42
	3. $/$m^3$	0.037

[1] Based on a design flow rate of 0.34 m^3/s

APPENDIX G

FIGURES

LVS Site Test Design and Process Drawings



SYMBOLS

- Gate valve
- Rotameter
- Flow tube, magnetic
- Ball valve, true union
- Three-way valve
- Pinch valve
- Pressure reducing valve
- Check valve
- In-line static mixer

PLAN

LIME-SODA ASH WATER PRETREATMENT PILOT PLANT
SOLIDS CONTACT REACTOR AND IN-LINE SYSTEMS

NO SCALE

Construction Notes

General:

1. All electric motors should be TEFC and have voltages and phasing to match that at LVS site.

Pipes, Valves & Fittings:

1. Where possible use cemented connections.
2. On NPT connections use thread-in nipples with cemented unions.

Materials of Construction For Process Brine wetted surfaces:

1. Don't use carbon steel, 304 SS, or 400 series SS.
2. Use epoxy coated carbon steel, FRP, PP, PVC, CPVC, 316 SS, ni-aluminum bronze, or bronze

Process Notes

Solids Contact Reactor

Upflow rate @ ½ liquid depth; 0.53 gpm/ft² @ 20gpm
Volume in cone: 644 gal., 32 min retention @ 20gpm
Nominal water feed; 20gpm
Lime Feed; 1 to 12 lbs/hr. of 95% $Ca(OH)_2$ @ 1lb/gal. or 215 mg/L of 85% CaO
Soda ash; 1 to 12 lb/hr of 95% Na_2CO_3 @ 1lb/gal

Marine type Impeller

7"ø × 6" pitch @ 250 rpm
Tip speed; 7.6 ft/sec.
Sludge rake; 7-2, 0.2-12 rpm; tip speed, 5-30 ft/min

Filter

Backwash rate; 20-27 gpm/ft² for 8 min
Rinse rate 3 gal/min for 10 min
Anthracite; effective size, 1.0 to 1.2 mm, specific gravity, 1.57; uniformity coefficient, 1.7 max.
Sand; effective size, 0.55mm; uniformity coefficient, 1.7 max specific gravity 2.62
Under drain strainer; 0.5mm openings
Air scour; 6.6 scfm/ft² for 5 min.

In-Line Lime-soda ash Pretreatment Unit

Mixing tank: 1 min detention @ 10gpm
"G" values 500/sec to 1000/sec.; "Gt" product between 150,000 and 660,000.
Flocculation tank; 45 min detention @ 10gpm
"G" values, 10/sec to 100/sec.; "Gt" product, 18,000 to 270,000
Clarification tank; 145 min detention @ 10gpm; upflow rate, 0.54 gpm/ft² @ 10gpm

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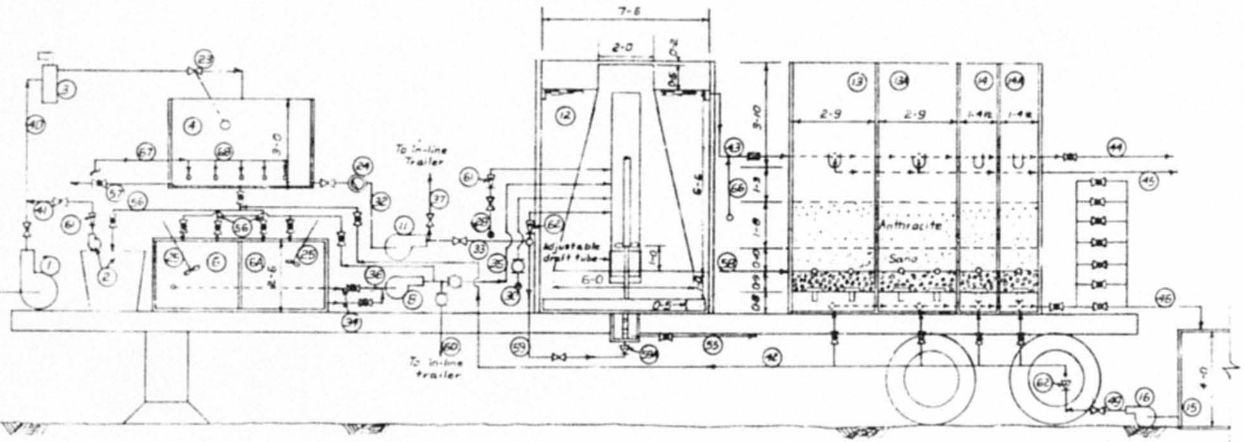
TITLE II

LIME-SODA ASH
WATER PRETREATMENT PILOT PLANT
GENERAL PLAN ARRANGEMENT

CHIEF, DIVISION OF DESIGN
DENVER, COLORADO SEPT. 6, 1978
1293-D-1

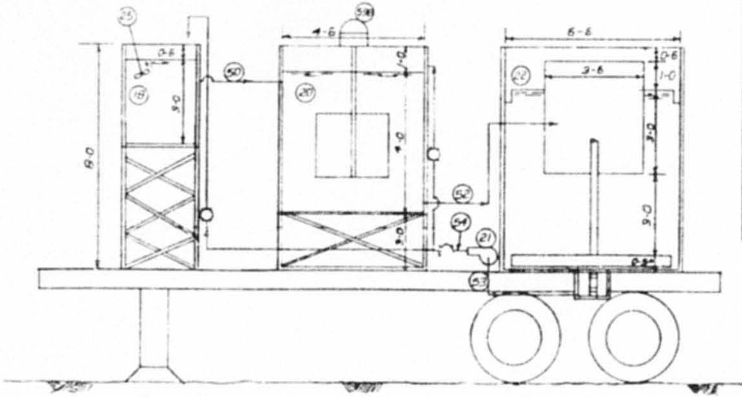
Figure G.1. — Aeration - lime-soda ash pretreatment trailer layouts.



SECTION A-A

SOLIDS CONTACT REACTOR TRAILER

ITEM NO.	ITEM NAME	DESCRIPTION
58	Air scour supply line	PVC, sch 40, 1"
59	Backflush line	PVC, sch 40, ½"
59A	Rotameter	1gal/min capacity, ±2% Full scale accuracy
59B	Flocculator turbine	G factor = 10/sec to 100/sec, variable speed turbine
60	In-line soda ash line	Flexible plastic tubing, ½"
61	Rotameter	25 ml/min. capacity, ±2% full scale accuracy
62	Rotameter	175 gal/min capacity, ±2% full scale accuracy
63	Gas rotameter	50 cfm capacity, ±2% full scale accuracy
64	Pressure regulating valve	4 psi constant outlet pres. 50 cfm capacity
65	H_2SO_4 tank & metering pump	pH adjustment of SCR effluent.
66	H_2SO_4 feed line	Polyethylene tubing
67	Aeration feed line	PVC, sch 40, 1"
68	Air diffuser assembly	PVC, sch 40, 1"



SECTION B-B

IN-LINE PRETREATMENT TRAILER

NO SCALE

SCHEDULE OF EQUIPMENT

ITEM NO.	ITEM NAME	DESCRIPTION
1	Raw water feed pump	1½"×1½", 30gpm @ 28TDH, material of construction – CPVC, or material resistant to acidic waters.
2	Hypochlorinator system	8gpd pump, 30gal. tank
3	Duplex strainer	Tate-Temco, 10-10F-150, size 1½", screwed ends, cast steel, 40mesh stainless steel strainer
4	Raw water feed tank	5'⌀×3', 440gal capacity
5,5A	Lime slurry feed tanks	2'×3'×2½', 110gal capacity each
6,6A	Soda ash slurry feed tanks	2'×3'×2½', 110gal capacity each
7	Chemical feed tank	2'⌀×2', 50gal capacity
8	Soda ash feed pump	A.R. Wilfley, model "K" sand pump, 1750rpm, 30gpm @ 44 TDH, white iron
9	Lime feed pump	See item 8.
10	Chemical feed pump	Wallace & Tiernan model 44-111, 9-44gpd capacity, ¼"×⅜"
11	Reactor feed pump	See item 1.
12	Solids contact reactor	7'6"⌀×8', upflow rate 0.55 gpm/ft² @ 22gpm
13,13A	Dual media filters	2-9×2-9, 3gpm/ft² @ 20gpm
14,14A	Dual media filters	2-9×1-4½, 3gpm/ft² @ 10gpm
15	Clearwell	10'×10'×4', 2800 gal capacity
16	Backwash pump	2"×2", 180 gpm @ 20TDH, centrifugal pump
17	Air blower	50cfm, 10psi, 1hp., 3600 rpm
18	Mixing tank	2½'⌀×3½', 110gal. cap., 11min. detention
20	Flocculator tank	4'⌀×5', 45 min. detention @ 10gpm
21	Sludge pump	Progressive cavity pump, variable speed
22	Clarification tank	6½'⌀×9', 145 min detention @ 10gpm
23	Water level control valve	Float control, 1½"
24	Water meter	Disc type, 1½", Badger Meter model SC-ER
25	Clamp mount mixer	½ h.p., 1750 rpm
27	Chemical tank-pump line	PVC, sch 40, ⅜"
28	Chemical pump-reactor line	PVC, sch 40, ¼"
29	Lime slurry pump line	PVC, sch 40, ¾"
30	Lime slurry reactor line	PVC, sch 40, ¾"
31	Lime slurry recycle line	PVC, sch 40, ½"
32	Raw water pump line	PVC, sch 40, 1½"
33	Reactor raw water line	PVC, sch 40, 1½"
34	Soda ash slurry pump line	PVC, sch 40, ¾"
35	Soda ash slurry reactor line	PVC, sch 40, ¾"
36	Soda ash slurry recycle line	PVC, sch 40, ½"
37	In-line raw water feed line	Flexible plastic tubing, 1"
38	In-line lime slurry feed line	Flexible plastic tubing, ½"
39	In-line soda ash feed line	Flexible plastic tubing, ½"
40	Raw water tank feed line	PVC, sch 40, 1½"
41	Chlorine feed line	Black polyethylene, ⅜"
42	Clearwell-chemical line	PVC, sch 40, 1½"
43	Reactor-filter feed line	PVC, sch 40, 2"
44	Reactor waste line	PVC, sch 40, 2"
45	Backwash drain line	PVC, sch 40, 2"
46	Clearwell feed line	PVC, sch 40, 2"
47	Clarifier-filter line	PVC, sch 40, 1½"
48	Backwash drain line	PVC, sch 40, 2"
49	Backwash feed line	PVC, sch 40, 2"
50	Mixing tank-flocculator line	PVC, sch 40, 2"
52	Flocculator-clarifier line	PVC, sch 40, 2"
53	Clarifier-pump sludge line	PVC, sch 40, 1"
54	Sludge return line	PVC, sch 40, 1"
55	Sludge waste line	Flexible plastic tubing, 2"
56	Chemical mixing water lines	PVC, sch 40, 1½"
57	Raw water drain line	PVC, sch 40, 1½"

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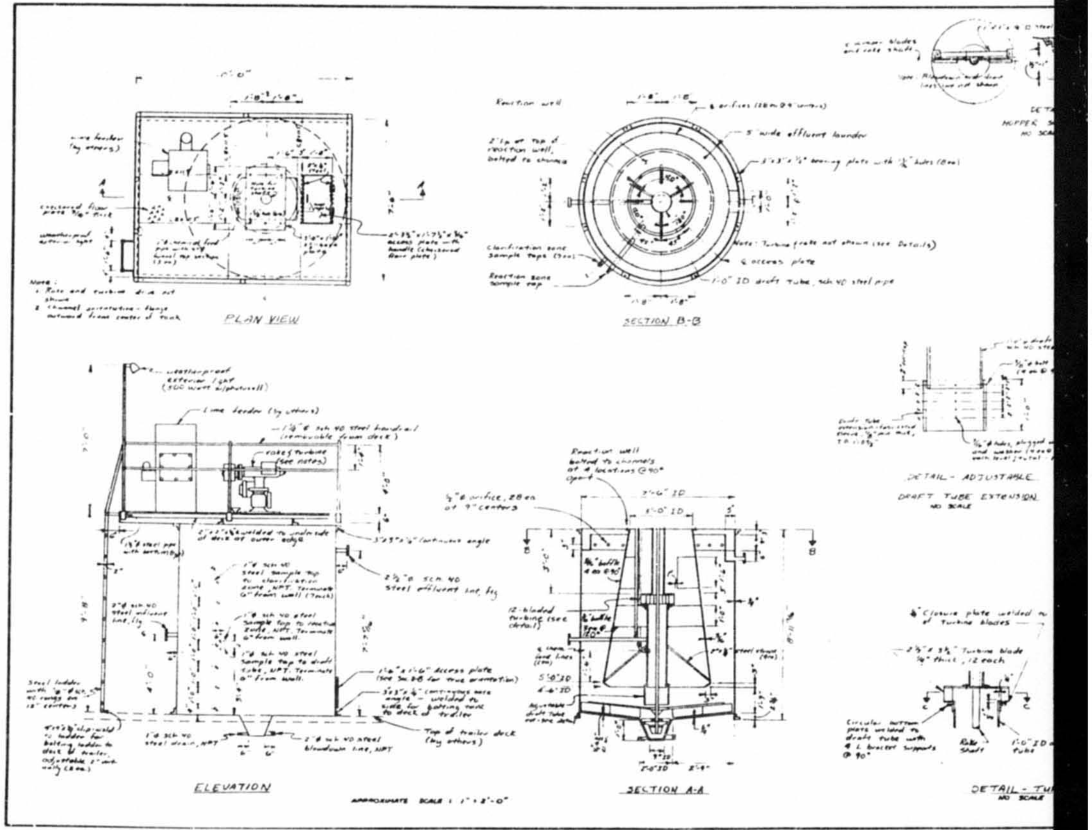
LIME-SODA ASH
WATER PRETREATMENT PILOT PLANT
GENERAL ARRANGEMENT SECTIONS

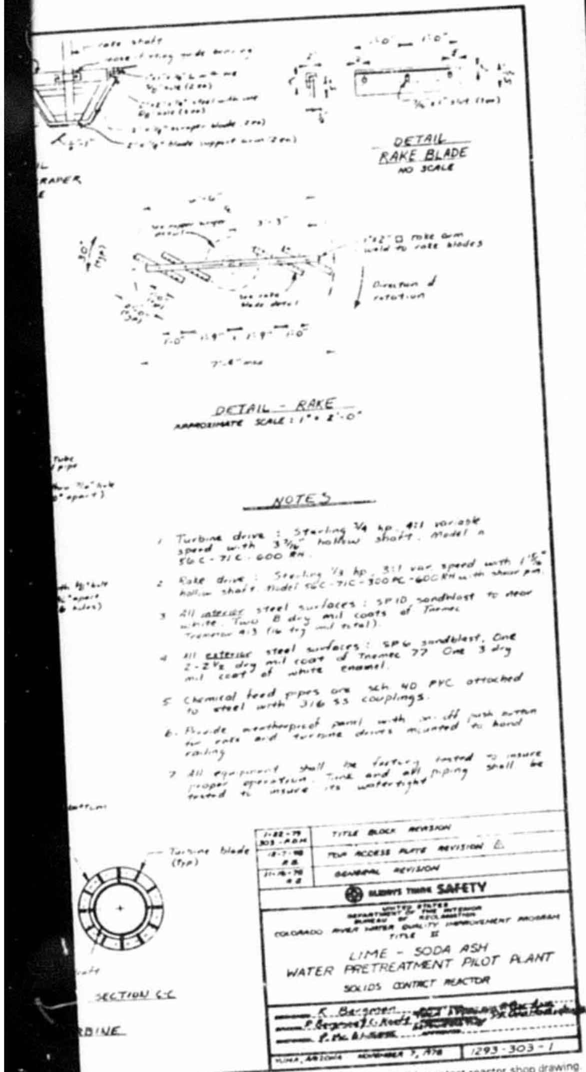
CHIEF, DIVISION OF DESIGN

DENVER, COLORADO SEPT. 6, 1978
1293-D-2

Figure G.2. — Aeration - lime-soda ash treatment pilot plant elevation.

223


PLAN VIEW
Lime feeder (by others)
Checkered floor plate 3/16" thick
weatherproof exterior light
1'-0" 1'-6"
7'-0"
2'-3½" x 1'-7½" x 3/16" access plate with handle (checkered floor plate)
Note:
1. Rake and turbine drive not shown
2. Channel orientation - flange outward from center of tank
SECTION B-B
Reaction well
2" lip at top of reaction well, bolted to channels
6 orifices (28 ea @ 9" centers)
5" wide effluent launder
3"x3"x½" bearing plate with ½" holes (8 ea)
Note: Turbine grate not shown (see Details)
Clarification zone sample taps (3 ea)
Reaction zone sample tap
access plate
1'-0" ID draft tube, sch 40 steel pipe
1'-8" 1'-8"
ELEVATION
weatherproof exterior light (300 watt w/photocell)
Lime feeder (by others)
1½" ø sch 40 steel handrail (removable from deck)
rake & turbine (see notes)
2"x2"x¼" welded to underside of deck at outer edge
3"x3"x¼" continuous angle
1½" ø steel pipe with fittings
1" ø sch 40 steel sample tap to clarification zone, NPT. Terminate 6" from wall (3 each)
2½" ø sch 40 steel effluent line, flg
2" ø sch 40 steel influent line, flg
1" ø sch 40 steel sample tap to reaction zone, NPT. Terminate 6" from wall.
1" ø sch 40 steel sample tap to draft tube, NPT. Terminate 6" from wall.
1'-6" x 1'-6" access plate (see Sec B-B for true orientation)
3"x3"x¼" continuous wire angle - welded to side for bolting tank to deck of trailer
Steel ladder with ¾" ø sch 40 rungs on 12" centers
4"x4"x⅜" clip-weld to ladder for bolting ladder to deck of trailer, adjustable 2" vertically (2 ea.)
1" ø sch 40 steel drain, NPT
2" ø sch 40 steel blowdown line, NPT
Top of trailer deck (by others)
7'-0"
4'-0"
APPROXIMATE SCALE: 1" = 2'-0"
SECTION A-A
Reaction well bolted to channels at 4 locations @ 90° apart
½" ø orifice, 28 ea at 9" centers
3'-6" ID
1'-0" ID
5"
3'-0"
12-bladed turbine (see detail)
4 chem. feed lines (2 ea)
5'-0" ID
4'-6" ID
Adjustable draft tube ext - see detail
2'-0" ID
2'-4"
8'-11⅝"
B
B
DETAIL - ADJUSTABLE DRAFT TUBE EXTENSION
NO SCALE
½" ø holes, plugged
HOPPER
NO SCALE
¼" Closure plate welded to of turbine blades
2½" x 3½" Turbine blade ¼" thick, 12 each
Circular bottom plate welded to draft tube with 4 L bracket supports @ 90°
Rake Shaft
1'-0" ID tube
C
C
DETAIL - TU
NO SCALE



NOTES

1. Turbine drive: Sterling 3/4 hp 4:1 variable speed with 3 7/16" hollow shaft. Model n 56C-71C-600 RH.
2. Rake drive: Sterling 1/3 hp, 3:1 var. speed with 1 15/16" hollow shaft. Model 56C-71C-300 RC-600 RH with shear pin.
3. All interior steel surfaces: SP10 sandblast to near white. Two 8 dry mil coats of Tnemec Tnemetar 413 (16 dry mil total).
4. All exterior steel surfaces: SP6 sandblast. One 2-2½ dry mil coat of Tnemec 77. One 3 dry mil coat of white enamel.
5. Chemical feed pipes are sch 40 PVC attached to steel with 316 SS couplings.
6. Provide weatherproof panel with on-off push button for rake and turbine drives mounted to hand railing.
7. All equipment shall be factory tested to insure proper operation. Tank and all piping shall be tested to insure its watertight.

Figure G.3. — Solids contact reactor shop drawing.

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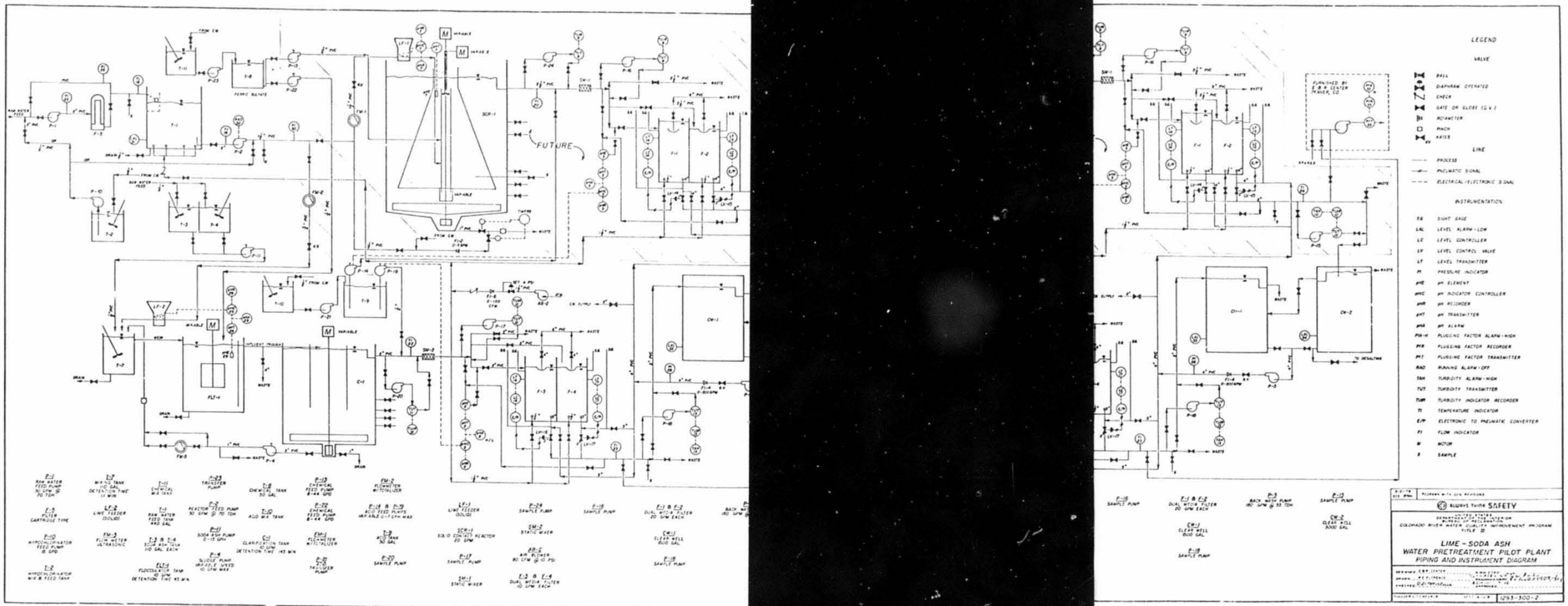

227

Figure G.4. — Equipment, piping and instrument diagram.

227A

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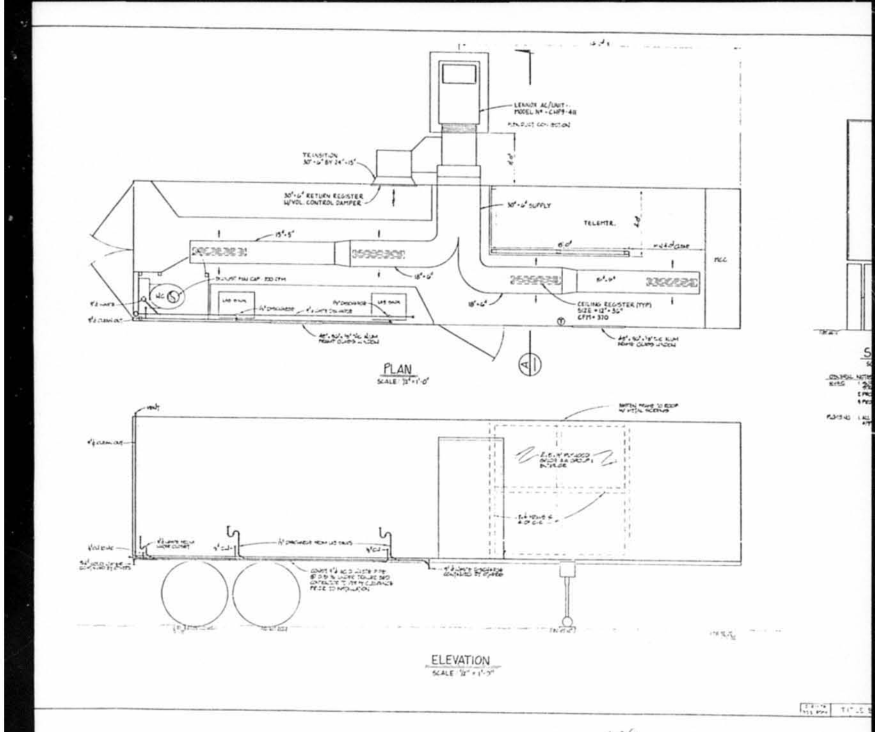


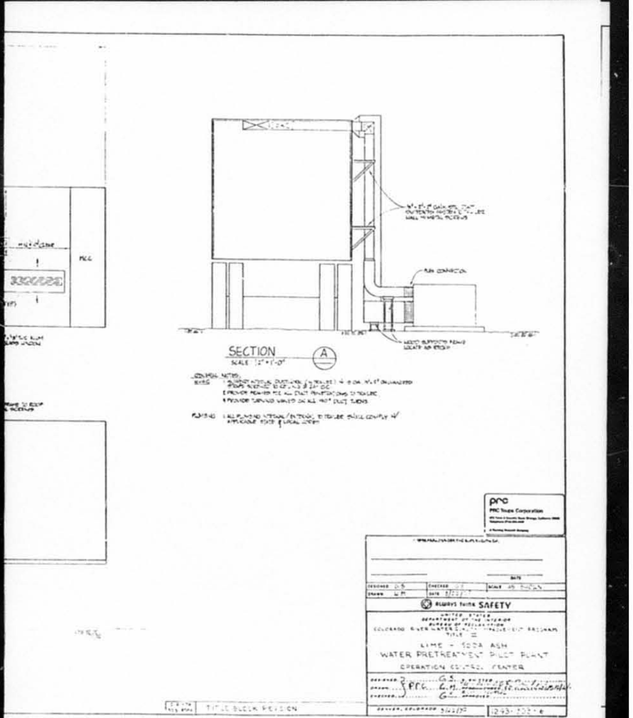


Figure G.5. — Operations control center.

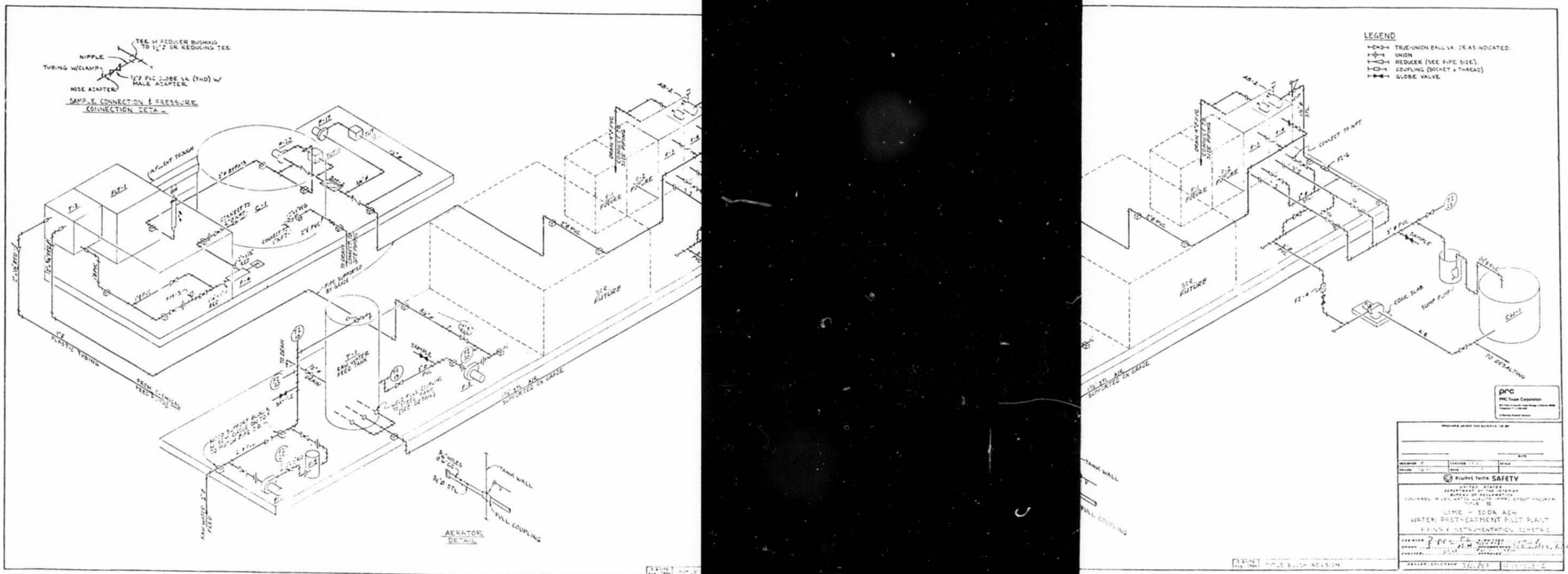
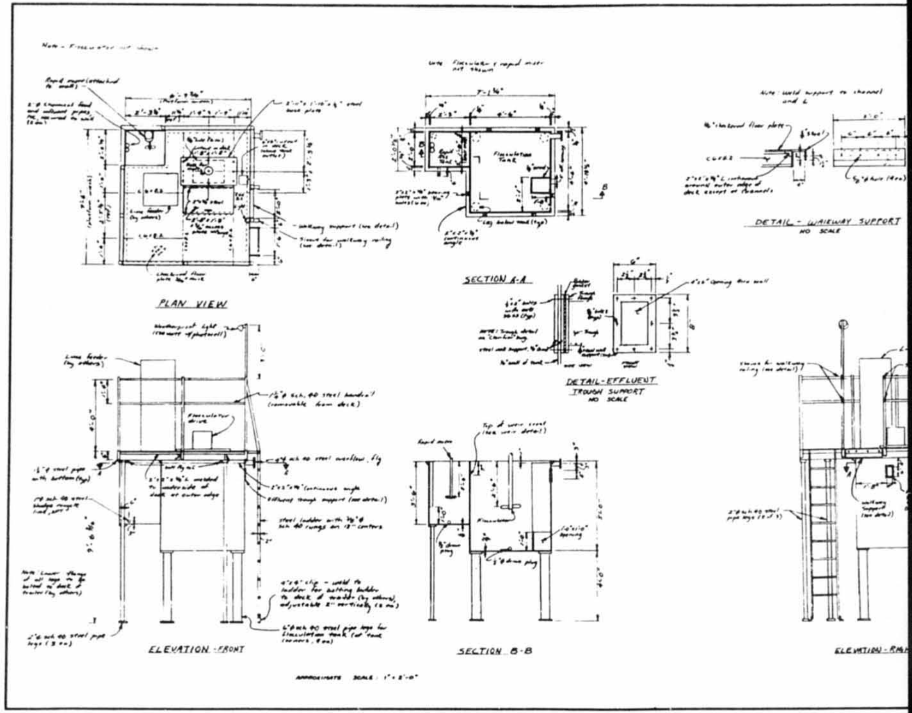
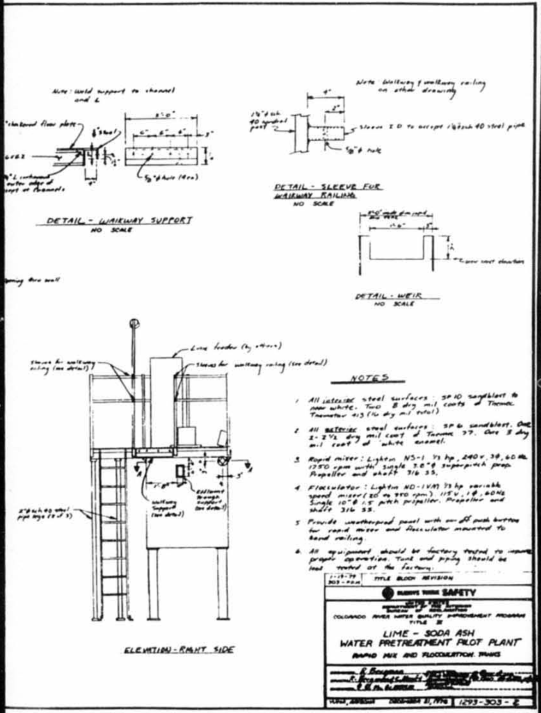


Figure G.6 — Equipment, piping and instrument isometric diagram





NOTES

1. All interior steel surfaces: SP10 sandblast to near white. Two 8 dry mil coats of Tnemec Tnemetar 413 (16 dry mil total)
2. All exterior steel surfaces: SP6 sandblast. One 2-2½ dry mil coat of Tnemec 77. One 3 dry mil coat of white enamel.
3. Rapid mixer: Lightnin NS-1 ⅓ hp, 240v, 3φ, 60 Hz 1750 rpm with single 3.0"φ superpitch prop. Propeller and shaft 316 SS.
4. Flocculator: Lightnin ND-1VM ⅓ hp variable speed mixer (20 to 350 rpm) 115V, 1φ, 60Hz Single 10"φ 1.5 pitch propeller. Propeller and shaft 316 SS.
5. Provide weatherproof panel with on-off push button for rapid mixer and flocculator mounted to hand railing.
6. All equipment should be factory tested to insure proper operation. Tank and piping should be leak tested at the factory.

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COLORADO RIVER WATER QUALITY IMPROVEMENT PROGRAM
TITLE II
LIME - SODA ASH
WATER PRETREATMENT PILOT PLANT
RAPID MIX AND FLOCCULATION TANKS

1293-303-2

Figure G.7. — Rapid mix tank and flocculator tank shop drawing.

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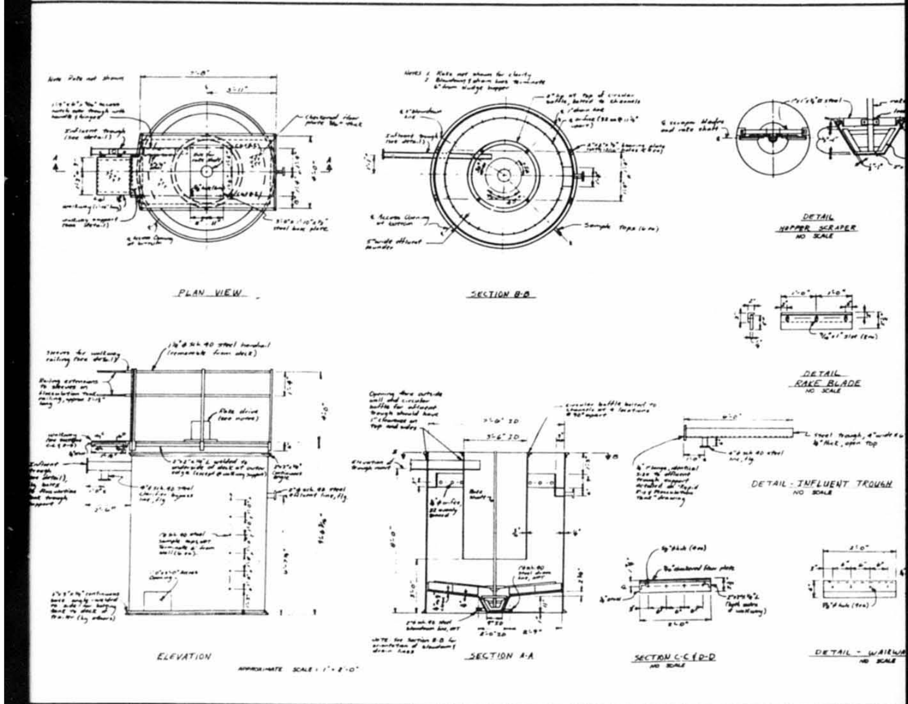

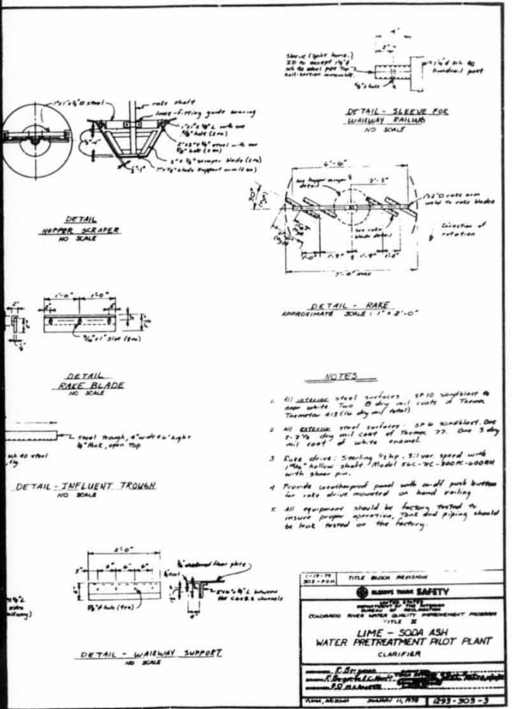

NOTES

1. All interior steel surfaces SP 10 sandblast to near white. Two 8 dry mil coats of Therma Thermotar 413 (16 dry mil total).
2. All exterior steel surfaces SP 6 sandblast. One 2-2½ dry mil coat of Therma 77. One 3 dry mil coat of white enamel.
3. Rake drive: Sterling ⅓ hp, 3:1 var. speed with 1 7/16" hollow shaft. Model SLC-3C-300K-600RH with shear pin.
4. Provide weatherproof panel with on-off push button for rake drive mounted on hand railing.
5. All equipment should be factory tested to insure proper operation. Tank and piping should be leak tested at the factory.

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TITLE II

LIME - SODA ASH
WATER PRETREATMENT PILOT PLANT
CLARIFIER

R. Bergman
R. Bergman & L.C. Hoff
P.D. [illegible]

YUMA, ARIZONA
JANUARY 11, 1978
1293-303-3

Figure G.8. — Clarifier shop drawing

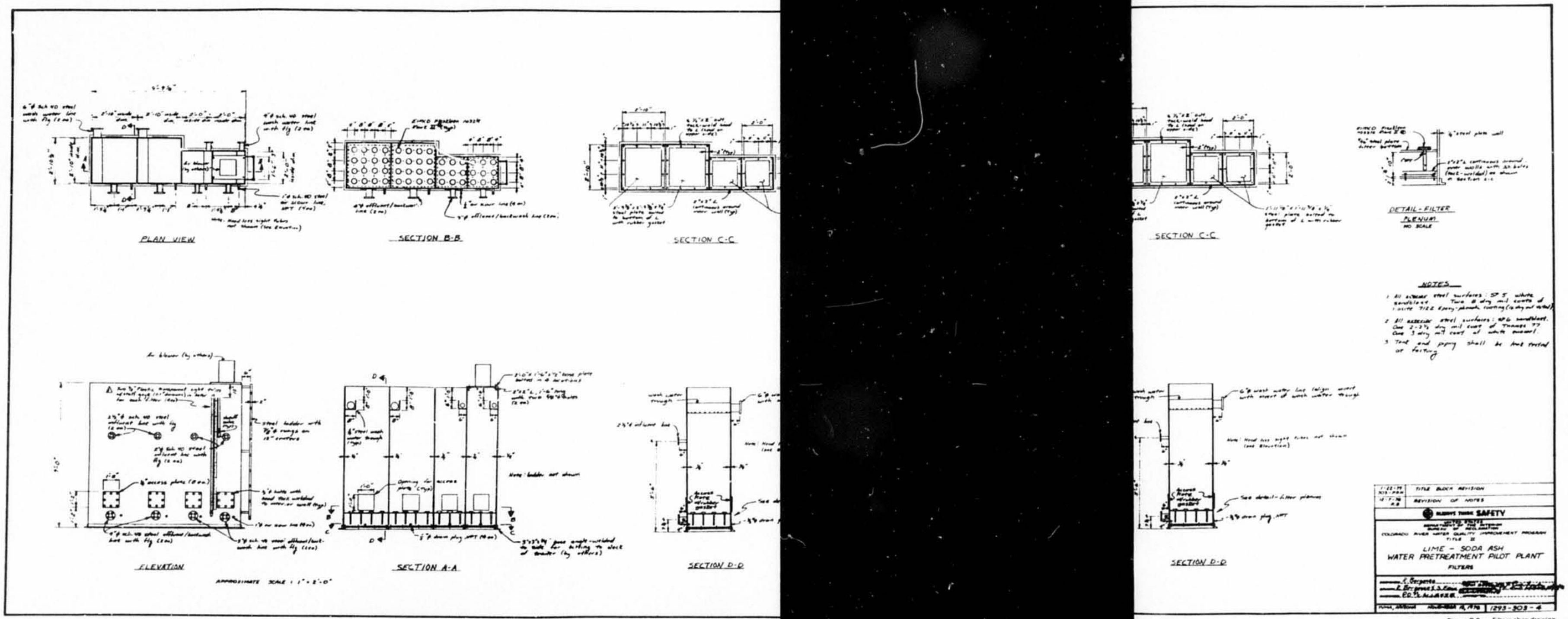


Figure G.9. — Filters shop drawing.

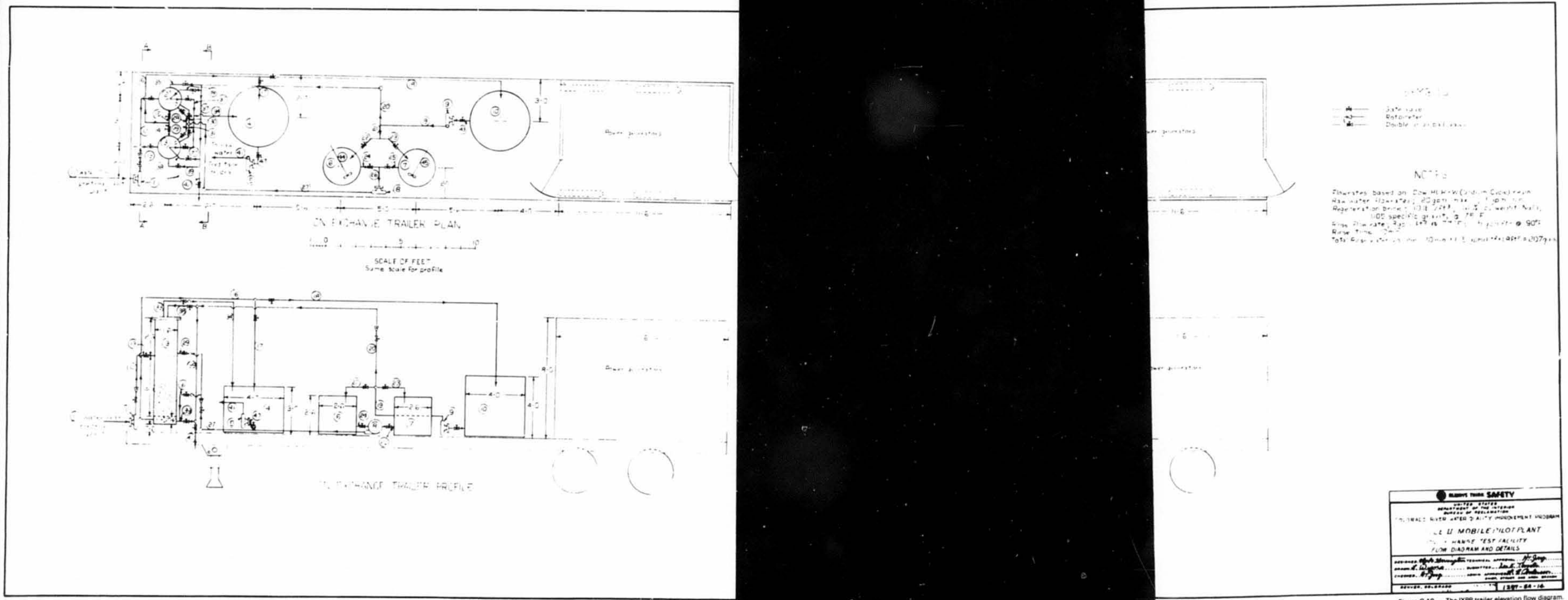
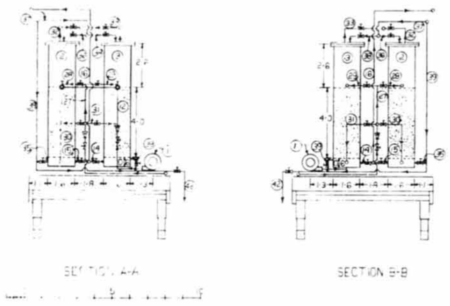


Figure G.10. — The IXPP trailer elevation flow diagram.



SECTION A-A

SECTION B-B

SCHE

ITEM NO.	TYPE OR FUNCTION
	Ion exchange feed pump
2,3	Ion exchange column
4	Backwash receiving tank
5	Backwash receiving pump
6,7	Brine supply tank
8	Brine supply pump
9	Clearwell pump
10	Clearwell tank
11	Pretreated water feed line
12,13	Ion exchange column feed line
14,15,16	Ion exchange product water line
17	Product water-backwash line
18	Product water-clearwell line
19	Clearwell supply line
20	Backwash/rinse water line
21,22,23	Brine tank feed line
24,25,26	Brine tank-pump feed line
27,28,29	Brine regeneration feed line
30,31	" " "
32,33,34	Backwash drain line
35,36,37	I.X. column backwash/rinse line
38,39	" " " "
40	Backwash receiving pump line
41	" " "
42	Drain line
43	Clearwell pump supply line
44,45	Mixer, clamp mount

SCHEDULE OF EQUIPMENT

ITEM NO.	TYPE OR FUNCTION	DESCRIPTION
	Ion exchange feed pump	Chemtrol model CP-9034, 1725 rpm, 25 gpm @ 18-ft TDH [illegible]
2,3	Ion exchange column	1 ft 6 in diameter, 7 ft height
4	Backwash receiving tank	4 ft diameter, 3 ft height, 280 gal capacity
5	Backwash receiving pump	Chemtrol model CP-9034, see item 1 for specifications
6,7	Brine supply tank	2 ft 6 in diameter, 2 ft 6 in height, 92 gal capacity
8	Brine supply pump	Chemtrol model CP-9034, see item 1 for specifications
9	Clearwell pump	Chemtrol model CP-9034, see item 1 for specifications
10	Clearwell tank	4 ft diameter, 4 ft height, 375 gal capacity
11	Pretreated water feed line	PVC, size 1¼" schedule 40
12,13	Ion exchange column feed line	" " 1" " "
14,15,16	Ion exchange product water line	" " 1" " "
17	Product water-backwash line	" " 1" " "
18	Product water-clearwell line	" " 1" " "
19	Clearwell supply line	" " 1" " "
20	Backwash/rinse water line	" " 1" " "
21,22,23	Brine tank feed line	" " 1" " "
24,25,26	Brine tank-pump feed line	" " 1¼" " "
27,28,29	Brine regeneration feed line	" " 1" " "
30,31	" " "	" " 1" " "
32,33,34	Backwash drain line	" " 1" " "
35,36,37	I.X. column backwash/rinse line	" " 1" " "
38,39	" " " "	" " 1" " "
40	Backwash receiving pump line	" " 1¼" " "
41	" " "	" " 1" " "
42	Drain line	" " 1" " "
43	Clearwell pump supply line	" 1¼
44,45	Mixer, clamp mount	Lightnin, NS-2, 1750 rpm

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UNITED STATES
DEPARTMENT OF THE INTERIOR
BUREAU OF RECLAMATION

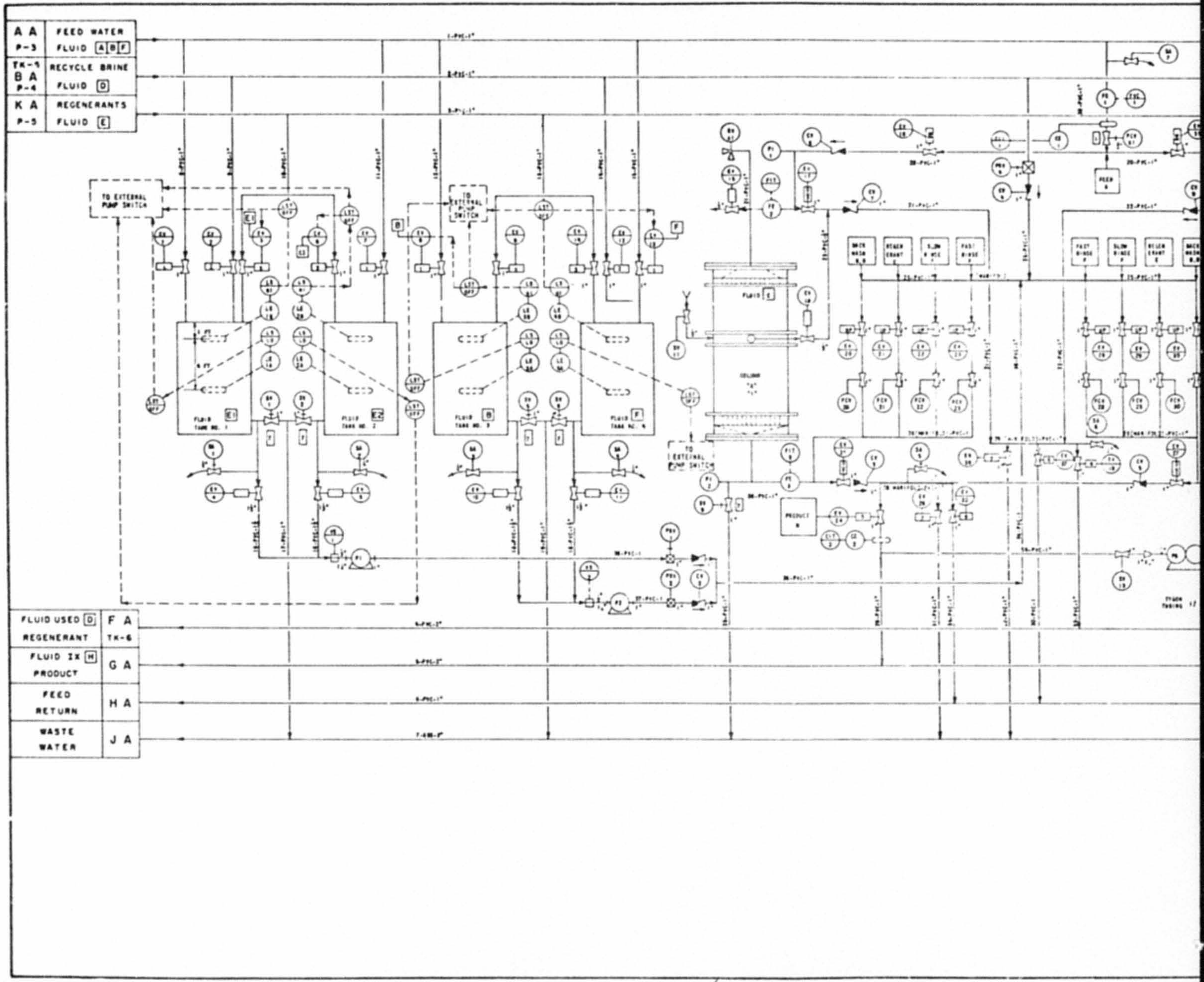
COLORADO RIVER WATER QUALITY IMPROVEMENT PROGRAM
TITLE II MOBILE PILOT PLANT
ION EXCHANGE TEST FACILITY
FLOW DIAGRAM AND DETAILS

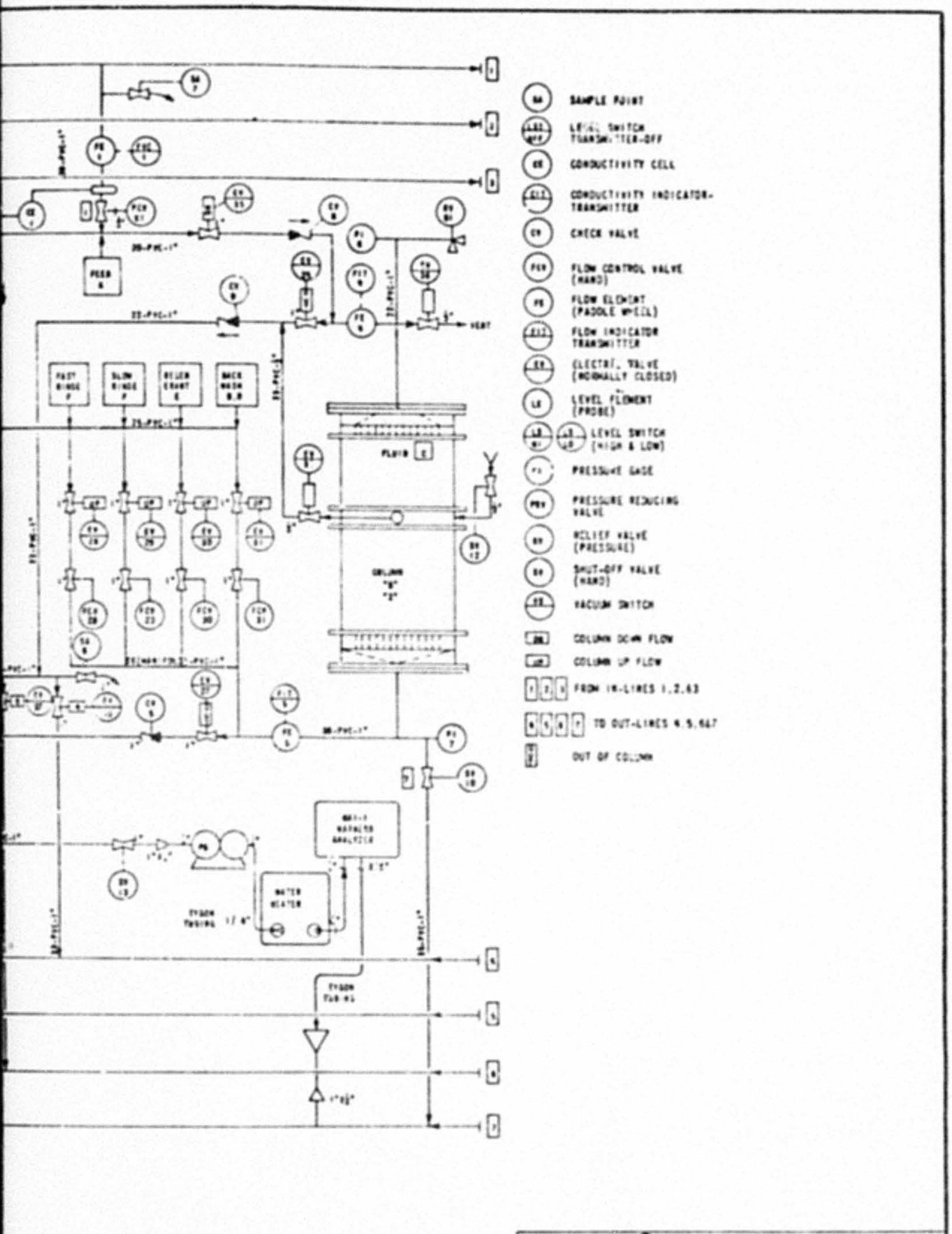
DESIGNED [illegible] TECHNICAL APPROVAL [illegible]
DRAWN L. Wayne SUBMITTED [illegible]
CHECKED [illegible] APPROVED [illegible]

DENVER, COLORADO 1297-5A-15

Figure G.11. — The IXPP Columns elevation flow diagram.

BEST COPY AVAILABLE


AA
P-3
FEED WATER
FLUID A B F
TK-5
BA
P-4
RECYCLE BRINE
FLUID D
KA
P-5
REGENERANTS
FLUID E
1-PVC-1"
2-PVC-1"
3-PVC-1"
TO EXTERNAL PUMP SWITCH
TO EXTERNAL PUMP SWITCH
E1
C1
B
F
LST OFF
LS
LE
4 FT
FLUID TANK NO. 1
E1
FLUID TANK NO. 2
E2
FLUID TANK NO. 3
B
FLUID TANK NO. 4
F
FLUID E
COLUMN "A"
TO EXTERNAL PUMP SWITCH
BACK WASH
REGEN ERANT
SLOW RINSE
FAST RINSE
FAST RINSE
SLOW RINSE
REGEN ERANT
BACK WASH
FEED
PRODUCT
PI
FIT
FE
PCV
FCV
PSV
CV
EV
SA
DV
P1
P2
TYGON TUBING
FLUID USED D
REGENERANT
FA
TK-6
FLUID IX H
PRODUCT
GA
FEED
RETURN
HA
WASTE
WATER
JA
4-PVC-2"
5-PVC-2"
6-PVC-1"
7-ABS-2"



SAFETY

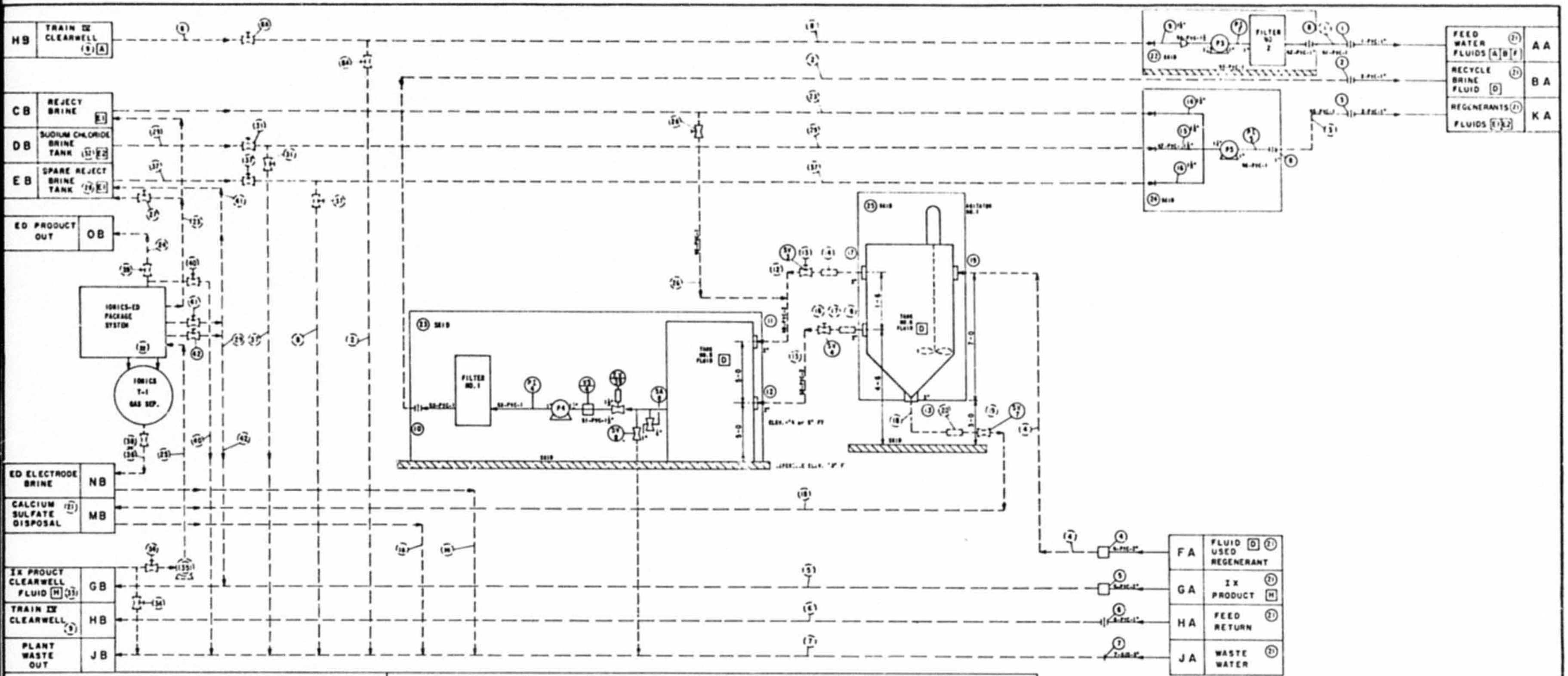
UNITED STATES
DEPARTMENT OF THE INTERIOR
BUREAU OF RECLAMATION
YUMA HIGH RECOVERY AND LAVERKIN SPRINGS

PILOT PLANT TEST PROGRAMS
P AND ID FOR MOBILE IXPP-I
ION EXCHANGE PRETREATMENT SYSTEM

DESIGNED .... TECHNICAL APPROVAL ....
DRAWN .... SUBMITTED ....
CHECKED .... APPROVED ....
DENVER, COLORADO

1297-D-68

Figure G.12. — The IXPP equipment, piping and instrument diagram (ERC).



CONNECTIONS PROVIDED BY E AND R CENTER (SOLID LINES)

①	UNION, FNPT, 1",PVC	⑭	NIPPLE, MNPT, 1½",PVC
②	UNION, FNPT, 1",PVC	⑮	NIPPLE, MNPT, 1½",PVC
③	UNION, FNPT, 1",PVC	⑯	NIPPLE, MNPT, 1½",PVC
④	COUPLINGS, FNPT, 2",PVC	⑰	BULKHEAD FTG., FNPT, 2",PP
⑤	COUPLINGS, FNPT, 2",PVC	⑱	BULKHEAD FTG., FNPT, 2",PP
⑥	UNION, FNPT, 1",PVC	⑲	BULKHEAD FTGS., FNPT, 2",PP
⑦	NIPPLE, MNPT, 3",ABS	⑳	BULKHEAD FTG., FNPT, 2",PP
⑧	UNION, FNPT, 1",PVC	㉑	IX UNIT (7 PIPE CONNECTIONS)
⑨	NIPPLE, MNPT, 1½",PVC	㉒	PUMP (2 PIPE CONNECTIONS)
⑩	UNION, FNPT, 1",PVC	㉓	PUMP & TANK (3 PIPE CONNECTIONS)
⑪	BULKHEAD FTG, FNPT, 2",PP	㉔	PUMP (4 PIPE CONNECTIONS)
⑫	BULKHEAD FTG, FNPT, 2",PP	㉕	AGITATED TANK (4 PIPE CONNECTIONS)
⑬	BULKHEAD, FTG, FNPT, 2",PP	㉖	UNION, FNPT, 1",PVC
		㉗	VALVE, FNPT, 1",PVC

CONNECTIONS PROVIDED BY LCR AND ON IONICS UNIT

(1)	PIPE, SCH80, 1",PVC	(13)(16)(19)	SHUT-OFF VALVES, 2",PLASTIC	(32)	AGITATED TANK,~ 200 GAL., PLASTIC
(2)	PIPE, SCH80, 1",PVC	(14)(17)(20)	SIGHT GLASS, 2",GLASS-PVC	(33)	HOLDING TANK,~ 3000 GAL., RUBBER
(3)	PIPE, SCH80, 1",PVC	(21)	TEMPORARY HOLDING TANK, DELETED	(34)	2 SHUT-OFF VALVES, 1½",PLASTIC
(4)	PIPE, SCH80, 2",PVC	(22)	IONICS MANUFACTURE TO ED UNIT, SHIP TO YUMA	(35)	GOOD SUCTION PUMP, 10 GPM
(5)	PIPE, SCH80, 2",PVC	㉓	PIPE & CONNECTIONS, 1",PVC	(36)	PIPE TO WASTE, ½", SEE IONICS FLOW DIAG. FM-1
(6)	PIPE, SCH80, 1",PVC	(24)	PIPE & CONNECTIONS, 1½",PVC	(37)	VALVE & CONNECTING PIPE, 1½",PVC
(7)	PLANT WASTE OUT	(25)	PIPE & CONNECTIONS, 1½",PVC	(38)	½" DRAIN VALVE ON IONICS T-1
(8)	PIPE, SCH80, 1½",PVC	(26)	SHUT-OFF VALVE & CONNECTING PIPE, 1",PVC	(39)	1½" VALVE FOR PRODUCT TO STORAGE ON IONICS UNIT
(8A)	2 SHUT-OFF VALVES, 1½",PVC	(27)	SHUT-OFF VALVE & CONNECTING PIPE, 1",PVC	(40)	1½" VALVE & PIPE FOR OFF-PRODUCT TO WASTE FROM IONICS UNIT
(9)	TANK,~ 3000 GAL., NON-CORRODING	(28)	TANK, 1500 GAL., NON-CORRODING	(41)	1" VALVE & PIPE FOR BRINE BLOW DOWN FROM IONICS UNIT
(10)	LIME SOFTENED WATER TRANSFER FROM TRAIN IV	(29)	PIPE, SCH80, 1½",PVC	(42)	PROVIDE STEPS FOR ACCESS TO IX-PILOT PLANT
		(30)	DELETE CENT. PUMP, 20 GAL.		
		㉛	2 SHUT-OFF VALVES, 1½",PVC		
⑫⑮⑱	PIPE, SCH80, 2",PLASTIC				

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UNITED STATES
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BUREAU OF RECLAMATION
YUMA HIGH RECOVERY AND LAVERKIN SPRINGS
PILOT PLANT TEST PROGRAMS
DETAIL COMPOSITE P AND ID
INTERCONNECTION OF PILOT PLANTS

DESIGNED .... TECHNICAL APPROVAL
DRAWN .... SUBMITTED
CHECKED .... ADMIN APPROVED ....
DENVER, COLORADO Revisions ①-4/1/78 | 1297-D-69

Figure G.13. — The E.P. & I.D. for interconnection of IXPP.

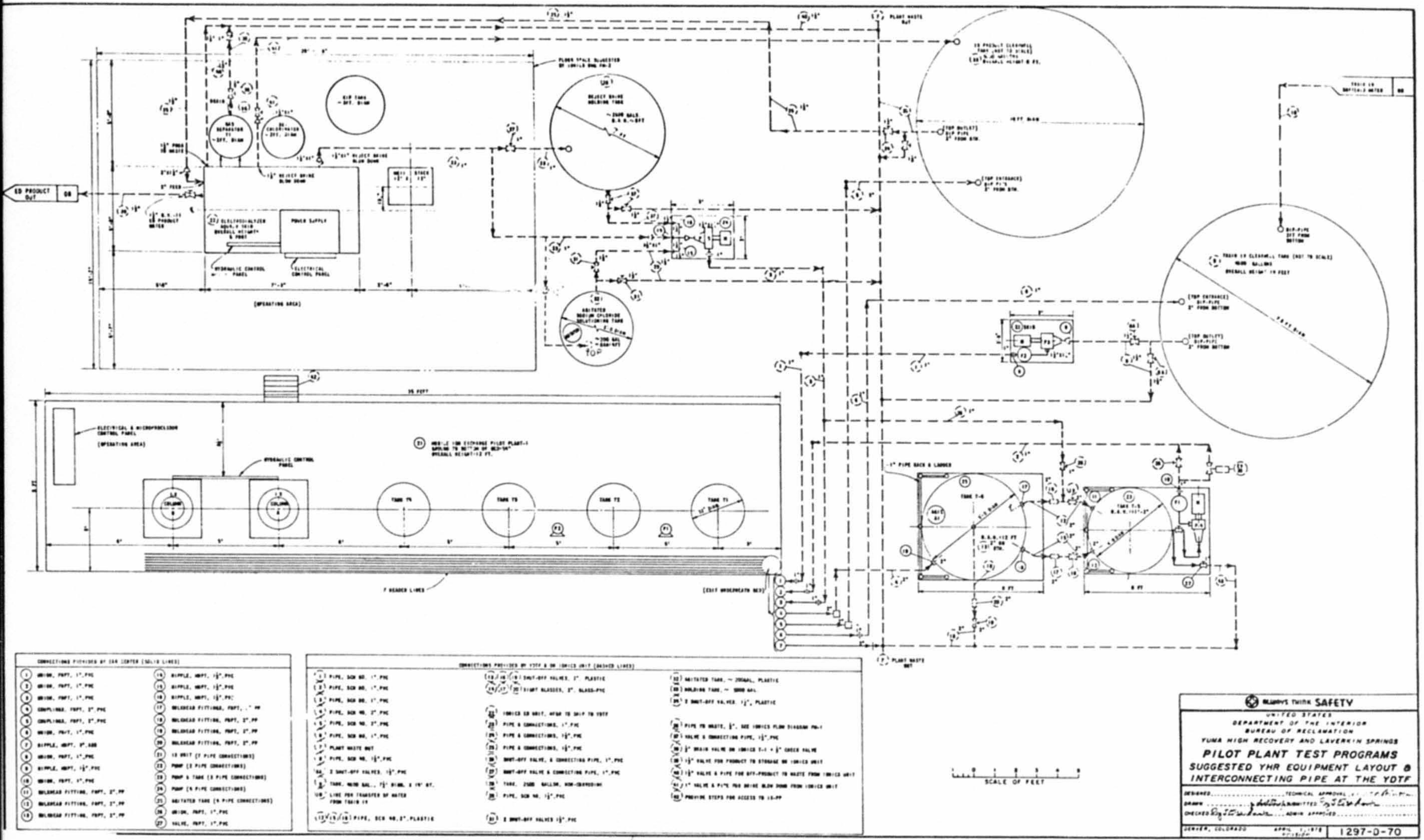

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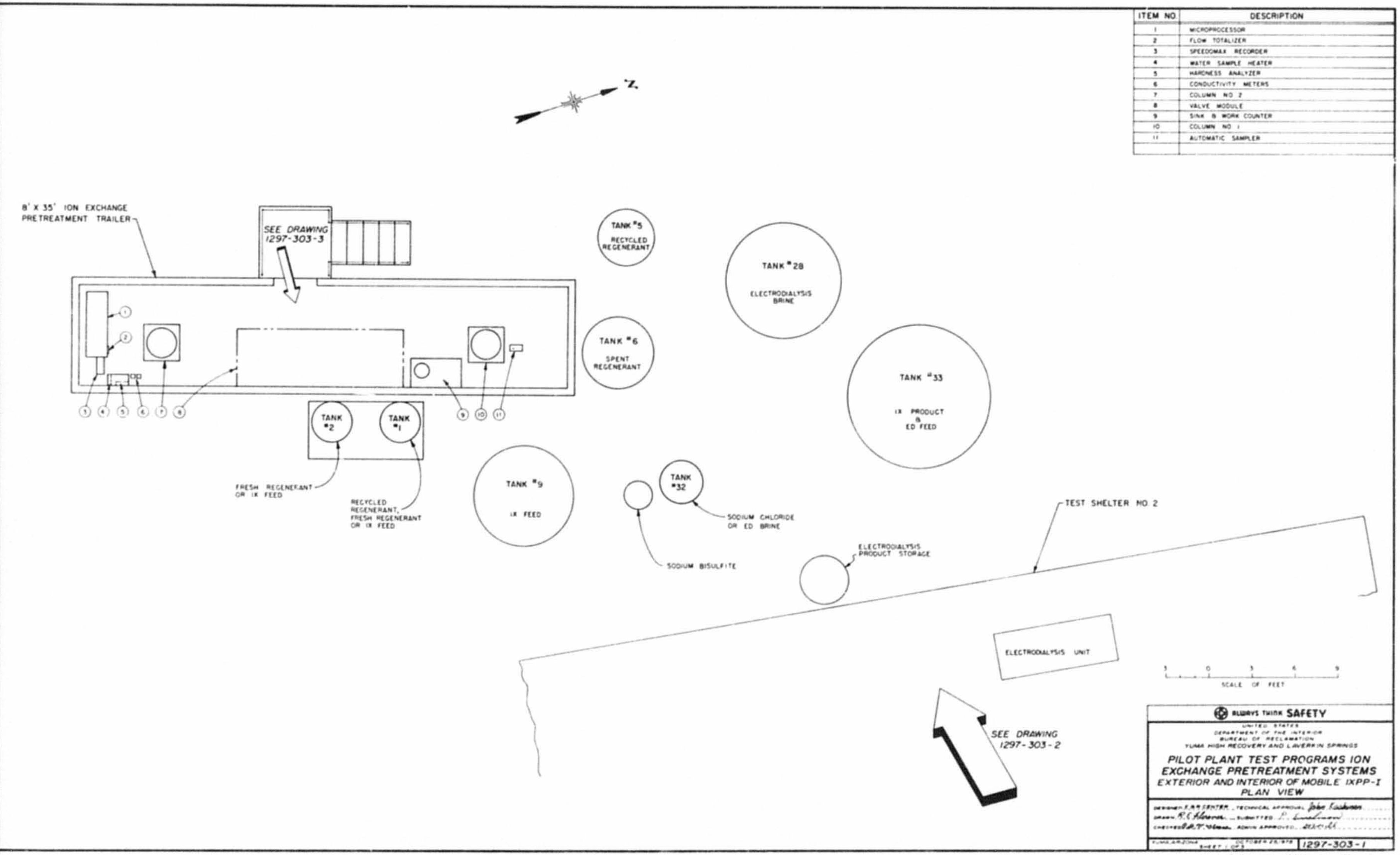
Figure G.14. — The ERC recommended layout of IXPP system.



ITEM NO.	DESCRIPTION
1	MICROPROCESSOR
2	FLOW TOTALIZER
3	SPEEDOMAX RECORDER
4	WATER SAMPLE HEATER
5	HARDNESS ANALYZER
6	CONDUCTIVITY METERS
7	COLUMN NO. 2
8	VALVE MODULE
9	SINK & WORK COUNTER
10	COLUMN NO. 1
11	AUTOMATIC SAMPLER

Figure G.15. — The IXPP system YDTF layout

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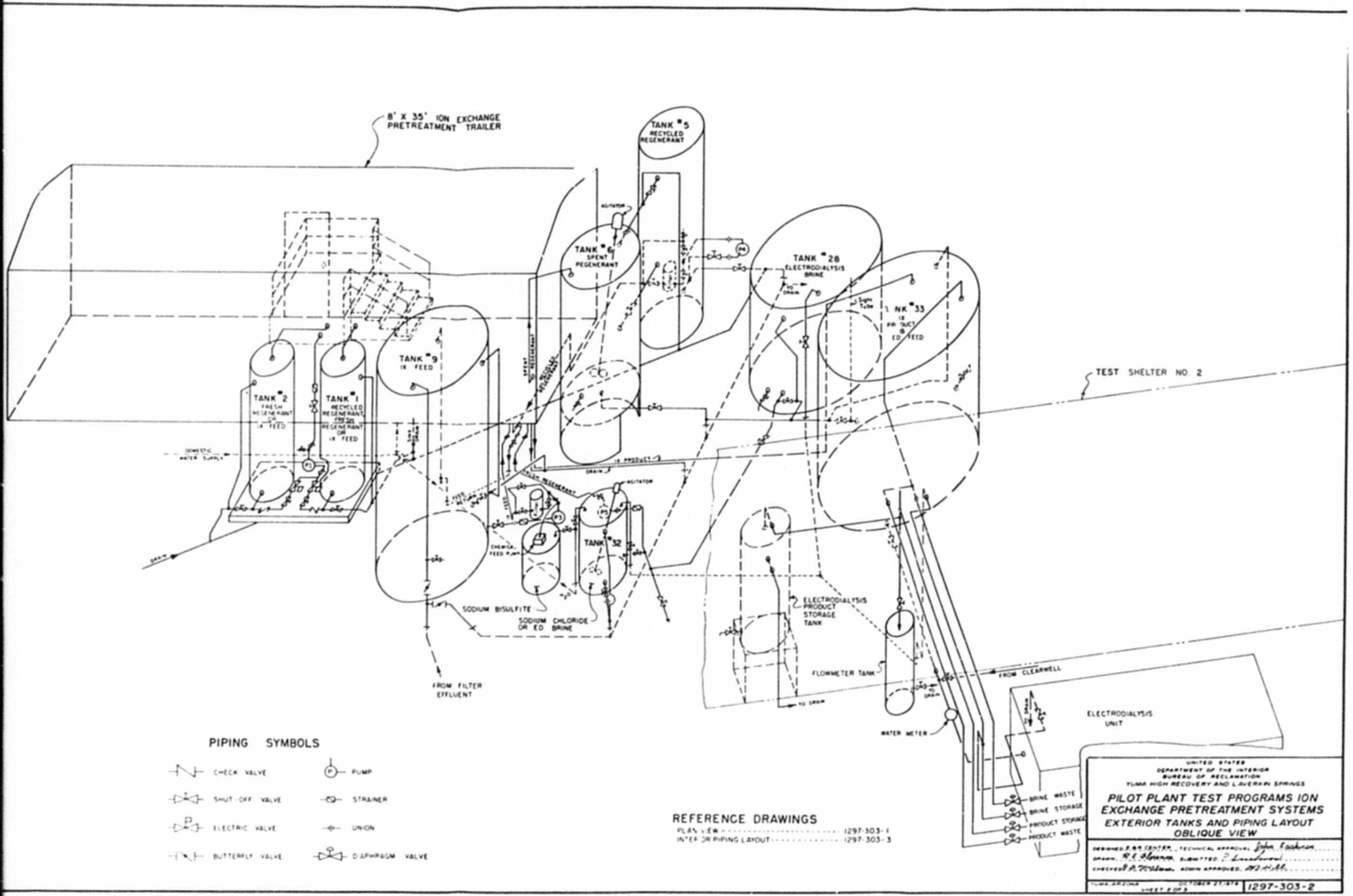


Figure G.16. — The IXPP system YDTF layout (oblique view).

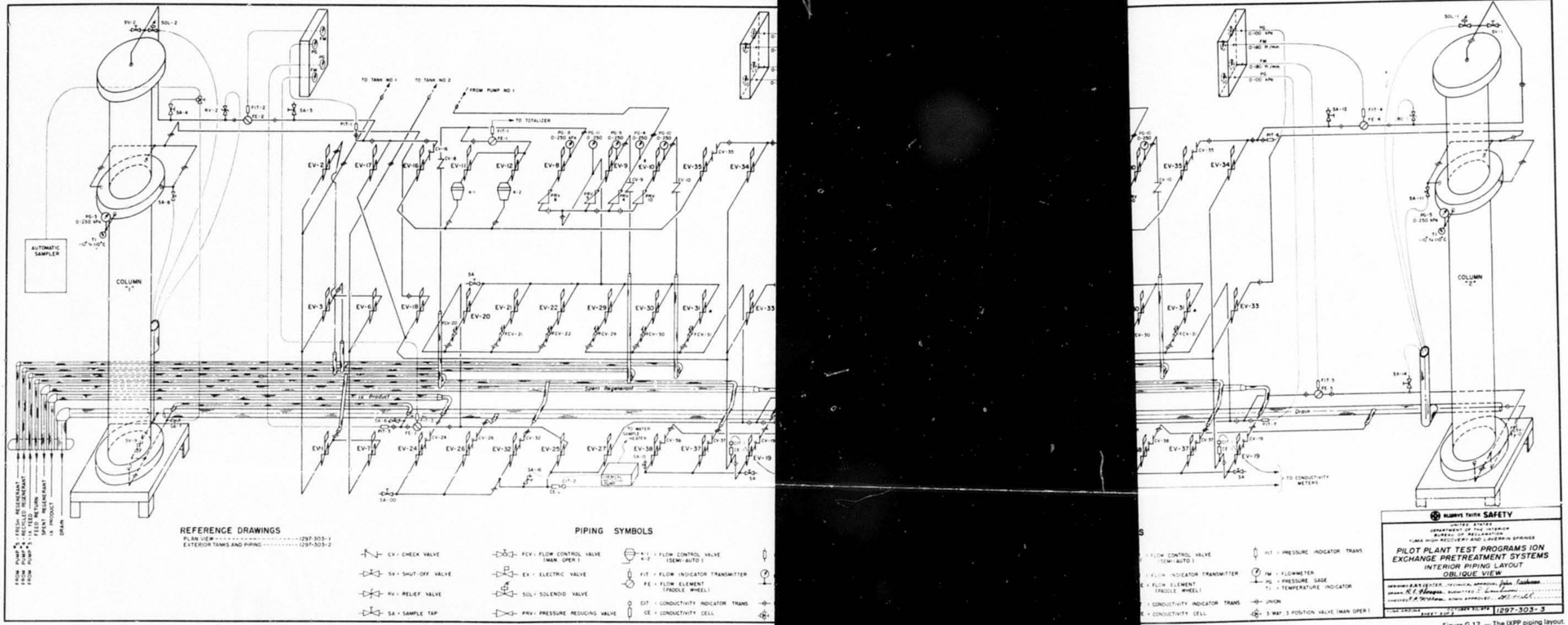


Figure G.17. — The IXPP piping layout.

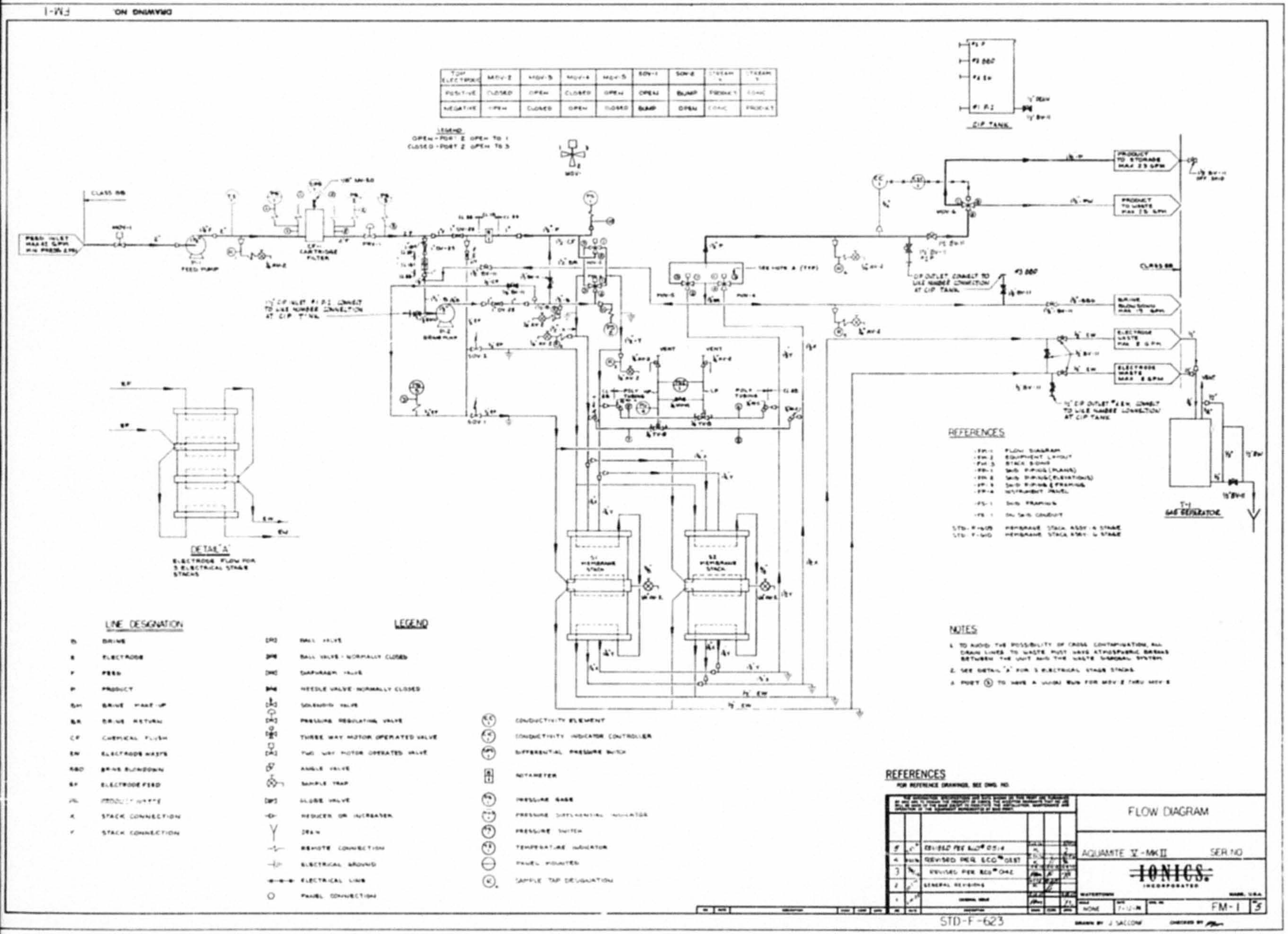
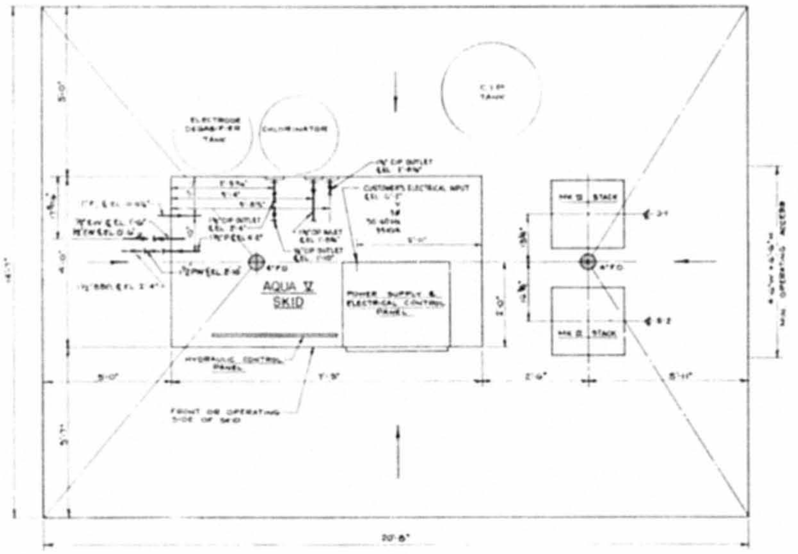


Figure G.18. — The EP & ID for electrodialysis unit.



NOTES
1. LIGHTING REQUIREMENTS - GENERAL LABORATORY LEVEL - 50 FOOT CANDLES
2. MINIMUM CEILING HEIGHT = 8'-0".
3. MAXIMUM FLOOR LOAD = 400 LBS/FT² AVERAGE.
4. FEED WATER TO BE SUPPLIED BY CUSTOMER AT [illegible] PSIG MIN
5. FLOOR DRAINS TO BE ACID RESISTANT, DURIRON OR EQUAL.

LEGEND
F - FEED
P - PRODUCT
EW - ELECTRODE WASTE
PW - PRODUCT WASTE
BBD - BRINE BLOWDOWN

FLOOR DOWNSLOPE [illegible]/FOOT
MINIMUM, INDICATED BY ⟵

REFERENCES
FOR REFERENCE DRAWINGS, SEE DWG. NO.

2		REVISED PER [illegible]	
1		[illegible]	

EQUIPMENT LAYOUT
AQUAMITE V2-MK II SER NO
IONICS
INCORPORATED
WATERTOWN MASS. U.S.A.
-FM- 2
STD-F-628
DRAWN BY T.H.L. CHECKED BY

Figure G.19. — Equipment layout for electrodialysis unit.

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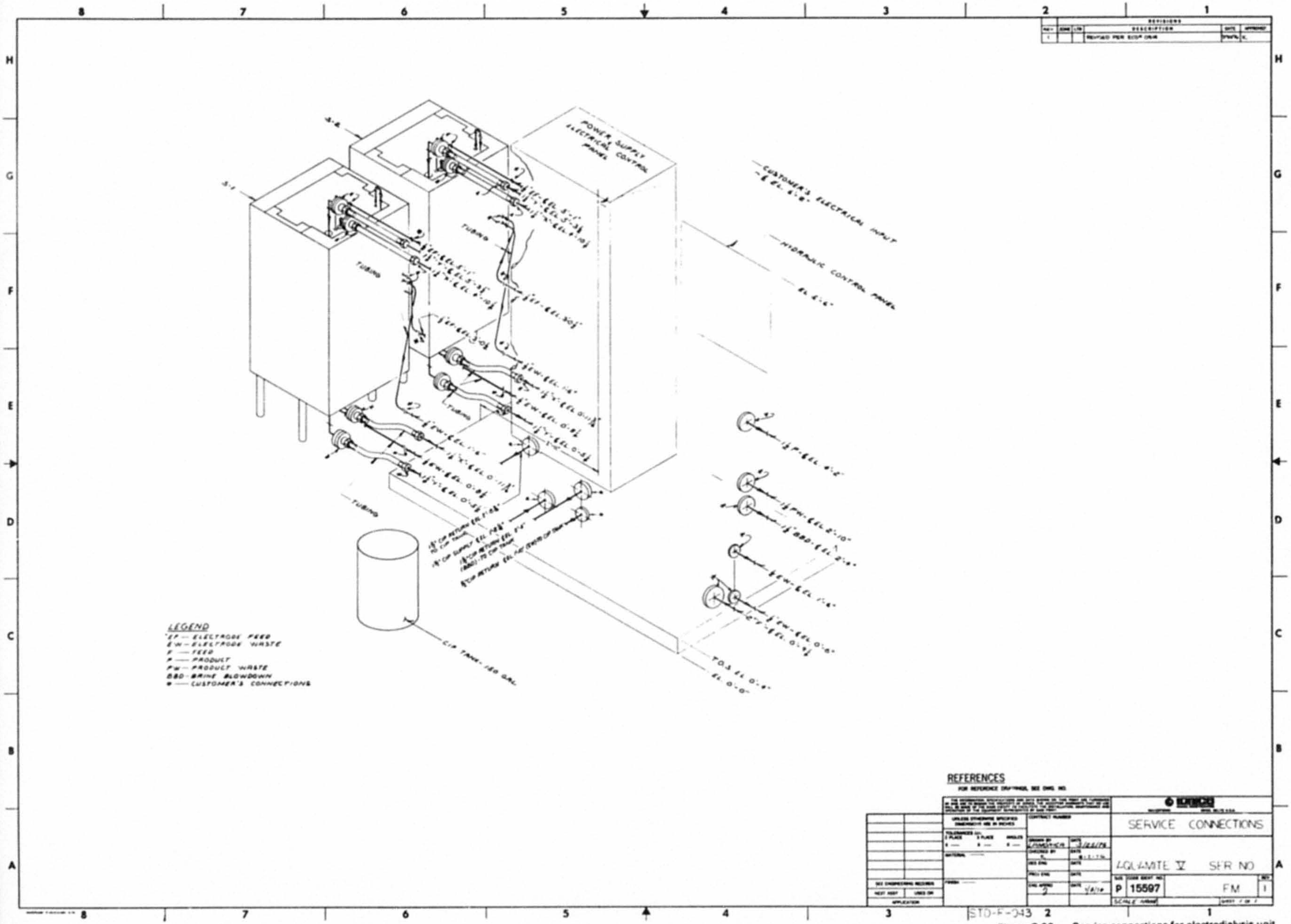

Figure G.20. — Service connections for electrodialysis unit.

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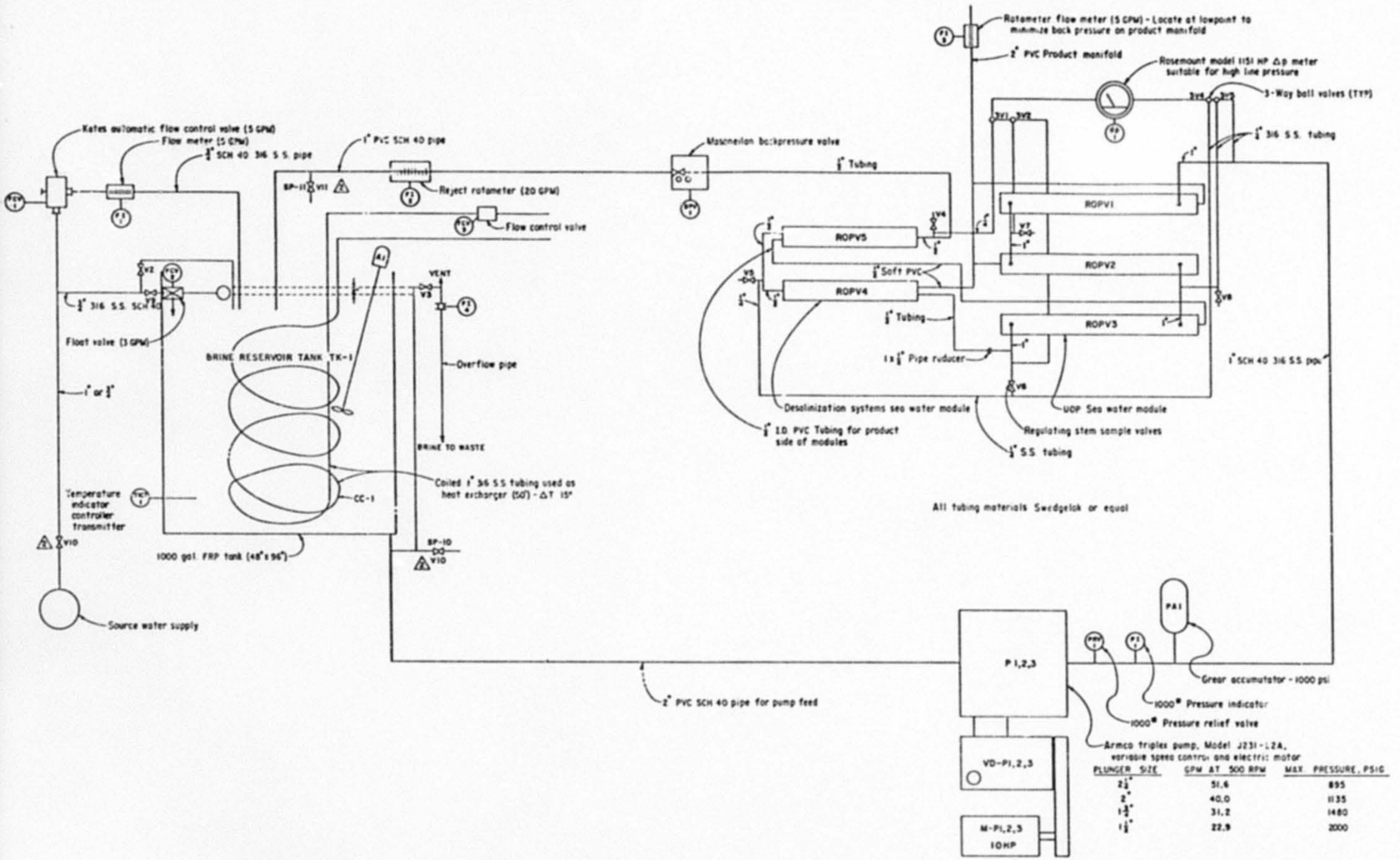

PLUNGER SIZE	GPM AT 500 RPM	MAX. PRESSURE, PSIG
2 1/4"	51.6	895
2"	40.0	1135
1 3/4"	31.2	1480
1 1/2"	22.9	2000

REVISION		
NO.	DESCRIPTION	DATE
1	TAG NO'S ADDED	3-21-80
2	VALVES ADDED: V1, V9, V10, V11	3-25-80
3	SP (SAMPLE POINT) LOCATIONS	3-25-80
4	ADDED F I 4, FLOW SIGHT GLASS	4-4-80

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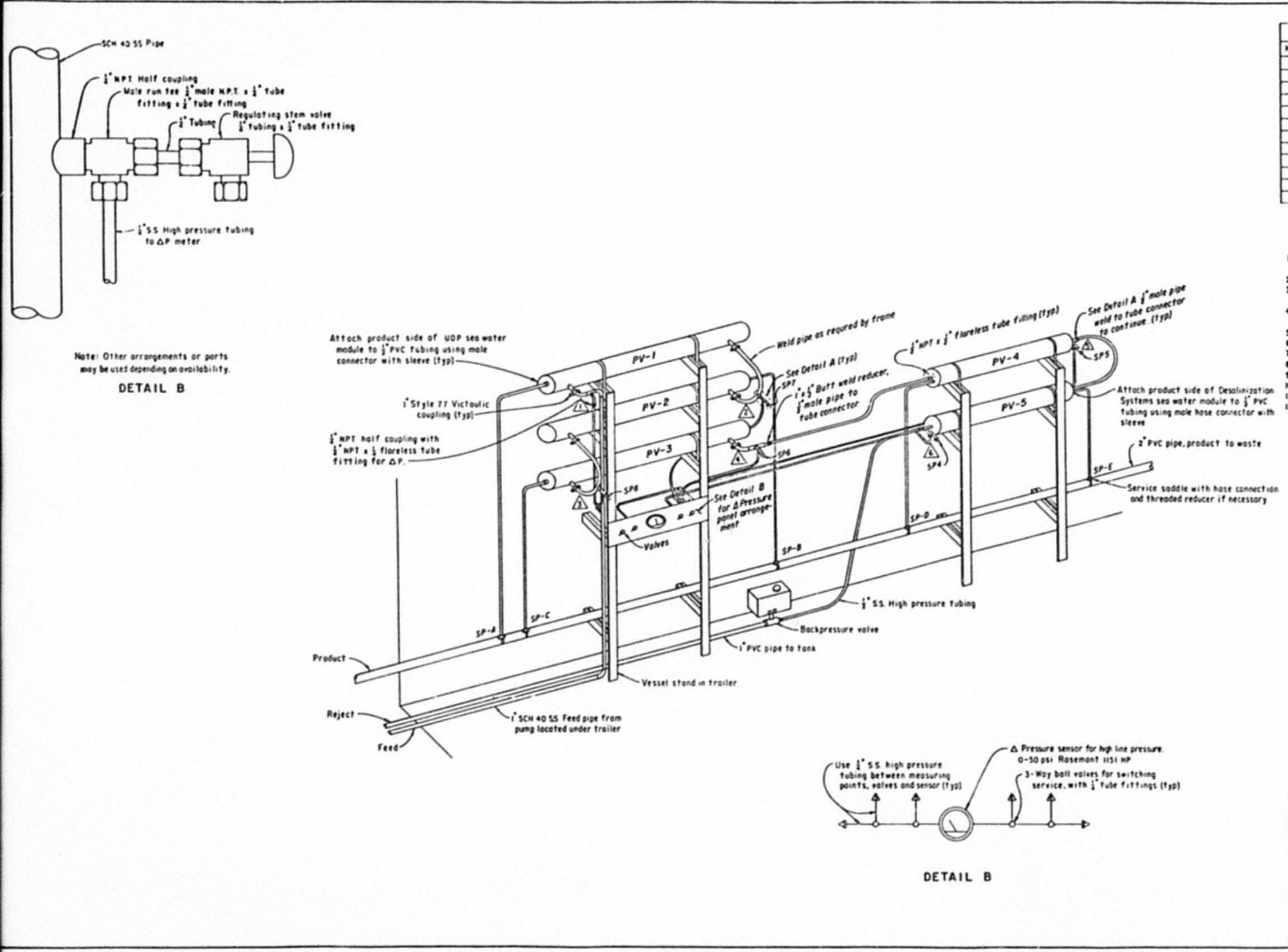
DIVISION OF RESEARCH
LA VERKIN SPRINGS, HURRICANE, UTAH

REVERSE OSMOSIS BRINE RECYCLE SYSTEM
PIPING AND INSTRUMENT FLOW DIAGRAM
SEAWATER MEMBRANE TEST UNIT

DESIGNED.. P. SABEL .... TECHNICAL APPROVAL ....
DRAWN .... SUBMITTED ....
CHECKED .... APPROVED ....
DENVER, COLORADO FEB 15, 1980 1292-D-1744-1

Figure G.21. — Proposed EP & ID RO system.



REVISION		
NO	DESCRIPTION	DATE
1	SP (SAMPLE POINT) LOCATIONS	3-25-80

1. Arrangement or materials may be changed to conform with availability of parts.
2. Other equipment may be added as required.
3. Pipes and modules shall be clamped or secured as required for 1000 psi working pressure.
4. All Victaulic couplings shall be coated with corrosion resistent coating.
5. All pipe shall be SCH 40 unless noted.
6. All pipe is to be welded.
7. All welds shall be made as required.
8. All tube fittings and valves shall be stailess steel type 316
9. All $\frac{1}{4}$" S.S. tubing shall be minimum wall thickness of .035"
10. All $\frac{1}{2}$" S.S. tubing shall be minimum wall thickness of .049"
11. All tube fillings and valves shall be of floreless ferrule type
12. All tube fillings and valves shall be rated for 1000 psi minimum working pressure.

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DIVISION OF RESEARCH
LA VERKIN SPRINGS HURRICANE, UTAH

REVERSE OSMOSIS BRINE RECYCLE SYSTEM
ISOMETRIC ELEVATION DRAWING
SEAWATER MEMBRANE TEST UNIT

DESIGNED R. HARRIS TECHNICAL APPROVAL
DRAWN J. Wilstorf SUBMITTED
CHECKED APPROVED
DENVER, COLORADO FEB 13, 1982 1292-D-1744-2

Figure G.22. — Isometric elevation drawing, RO system.

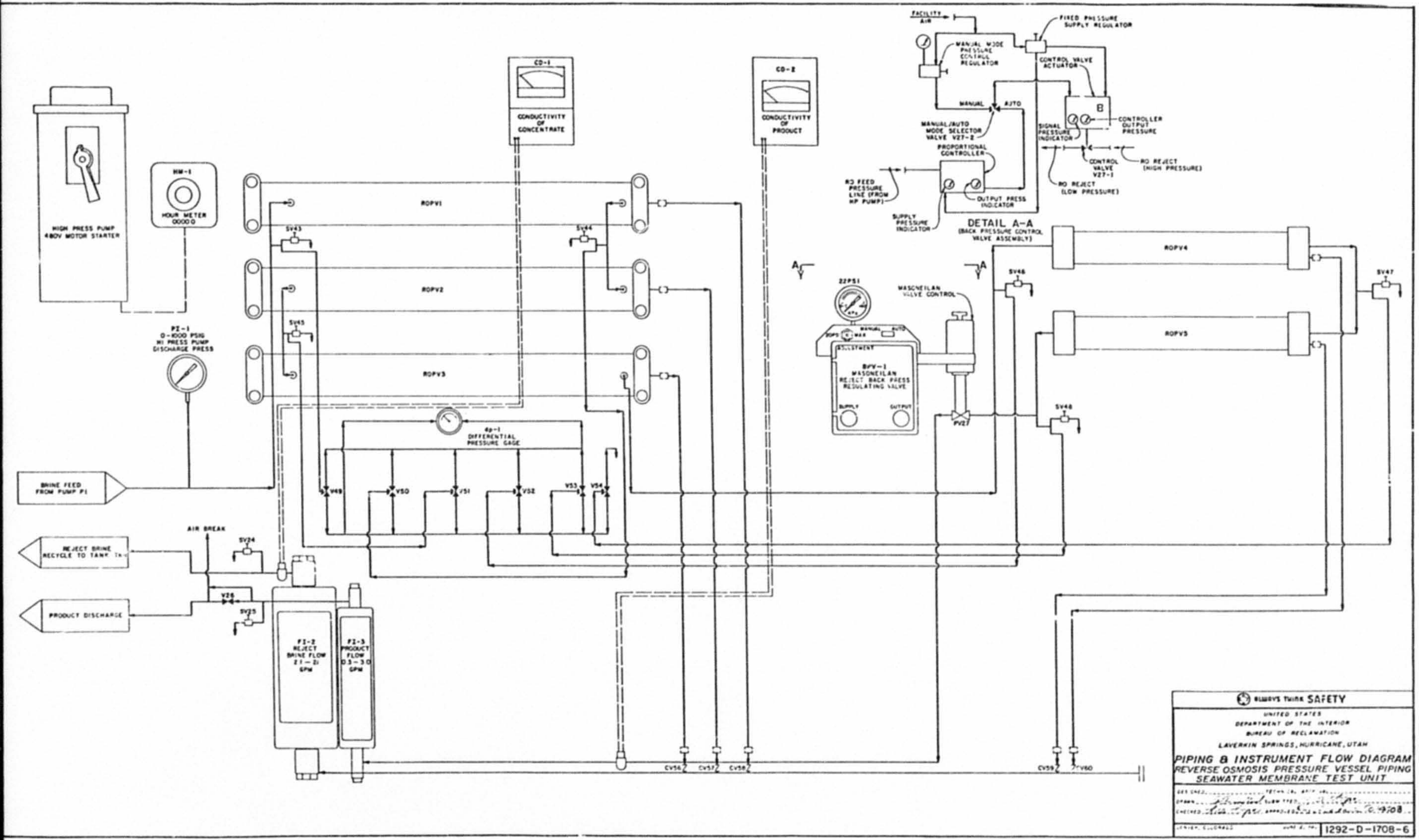
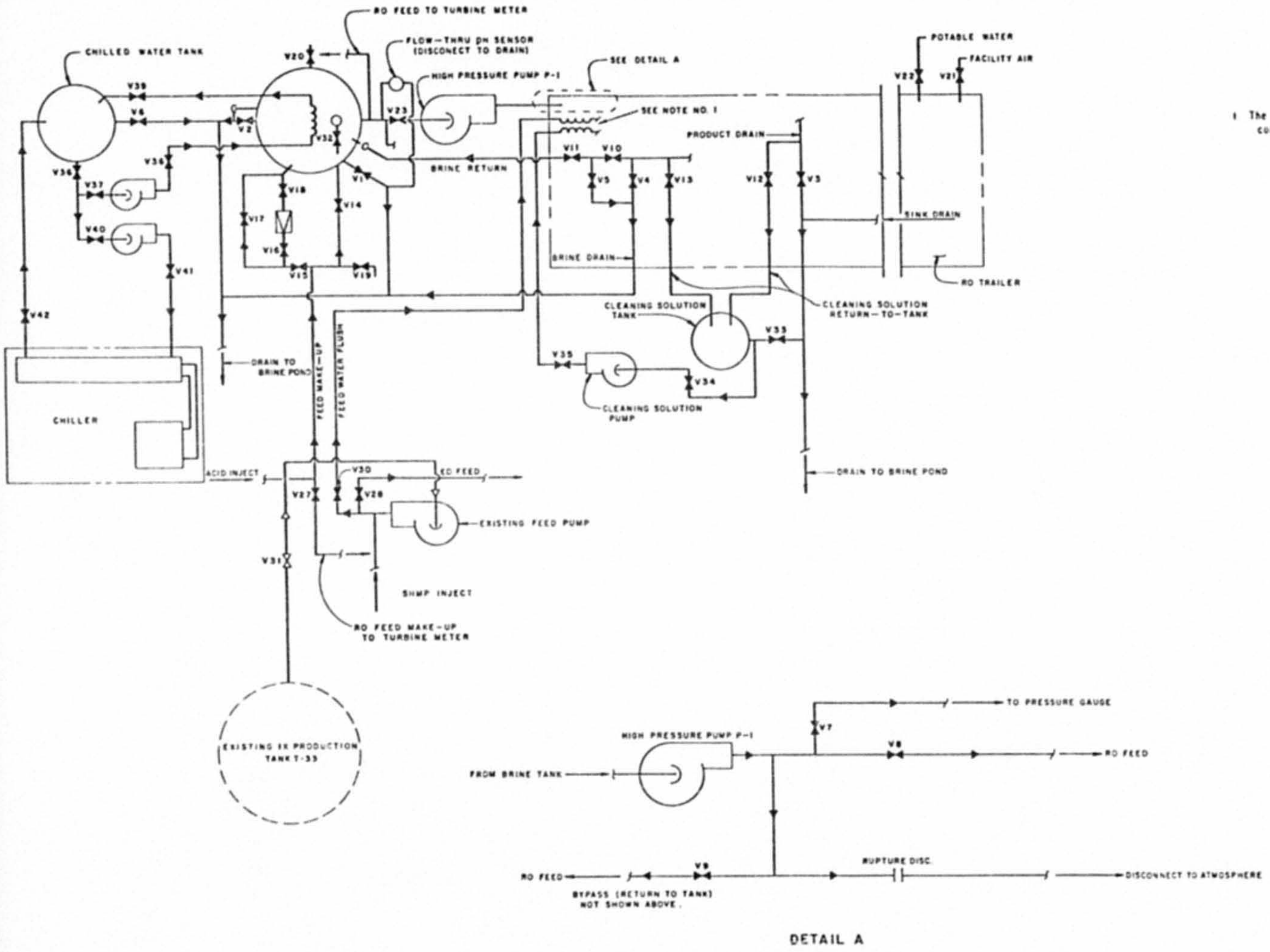


Figure G.23. — The EP & ID RO recycle system.



NOTES

1. The feed water flush and cleaning solution feed lines are connected into the RO feed system only when needed.

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LAVERKIN SPRINGS, HURRICANE, UTAH

PIPING & INSTRUMENT FLOW DIAGRAM
REVERSE OSMOSIS BRINE RECYCLE SYSTEM
SEAWATER MEMBRANE TEST UNIT

DESIGNED........ TECHNICAL APPROVAL........
DRAWN L. Hill........ SUBMITTED........
CHECKED........ APPROVED........
DENVER, COLORADO JULY 16, 1990 1292-D-1708-7

Figure G.24. — The EP & ID RO recycle system.

APPENDIX H

Operating Instructions for Water Conditioning Equipment

Date — June 27, 1980
Job No. A141J 20 360-9B
Prepared for U.S. Department of Reclamation

SPIRACTOR

LIST OF CONTENTS

SPECIFICATIONS AND DATA

Model	Special Drawing 157-32310
Purpose	Softening
Number of Units	1

Principal Dimensions

Top Diameter	12"
Bottom Diameter	3"
Vertical Height	13' - 8"
Vertical Height (overall)	14' - 3"
Flow Rate	10 GPM
Flow Operating Range	50% to 100% capacity. Do not operate below 50% capacity.
Operating Weight	lbs.

Chemical Dosages

Hydrated Lime	1.06 lbs. per HOUR
Soda Ash Chemical Feed System	5.3 lbs. per HOUR
Catalyst Material Type	Catalyst sand Hydro No. 5 as furnished by Florida Silica & Sand Company.
Volume of Catalyst initial charge	½ FULL
Catalyst Removal	Excess removal of catalyst manually controlled by opening flush-out valve to drain pit. Removal determined by carryover at No. 4 sample port.

Sample Connections:

No.	Locations
1	above the bottom straight of Spiractor
2	above the bottom straight of Spiractor
3	above the bottom straight of Spiractor
4	above the bottom straight of Spiractor
5	Effluent Nozzle

NOTE: All five (5) sample connections are valved for chain wheel operation from ground level.

Chemical Test: (Instructions included with test kits)

Test Set No. 12 — Alkalinity
Test Set No. 73 — Total hardness, calcium hardness, and magnesium hardness.

Catalyst Charging Hopper

Number of Units:	NONE
Dimensions:	
Top Diameter:	
Bottom Diameter:	
Height	
Eductor:	NONE

Reference Drawings: Layout of Spiractor
Assembly of Catalyst Hopper ....

CAPACITY

The Spiractor will give a constant flow of 5 gpm.

PRINCIPAL OF OPERATION

The plant is operated for the purpose of reducing the hardness and iron content of the raw water. The softening is accomplished by the addition of lime.

Raw water is admitted at a constant flow rate at the bottom of the Spiractor and passes through a catalyst bed in a tangential upward flow.

Lime and, if required, trisodium phosphate are added just at point of entry.

Reaction of the Chemical with the raw water causes precipitates to form which deposit on the particles of catalyst, causing the particles to "grow."

This reaction may be written as:

$$CaO + H_2O \underline{\qquad\qquad\qquad} Ca(OH)_2$$
$$Ca(OH)_2 + Ca(HCO_3) \underline{\qquad\qquad} 2CaCO_3 + 2H_2O$$

The addition of trisodium phosphate is useful in treating certain waters. The action of this chemical is to slow down the reaction between the lime and alkalinity resulting in the formation of more dense grains of "grown catalyst," and also prevents the formation of small particles which would tend to carry over.

The treated water then leaves the Spiractor and flows to service.

In operation, it is necessary to periodically remove a portion of the "grown" catalyst from the catalyst bed, to prevent the catalyst bed from building up to a point where catalyst will carry over from the Spiractor. A certain amount of fresh catalyst must also be added to prevent the average size of the catalyst particles from increasing excessively.

FLOW CONTROL

The flow to the Spiractor is controlled by an inlet valve. The spiractor must operate at a constant rate of low. It will operate effectively at any rate between about 50% and 100% of its maximum rate, but must be held constant at the rate selected.

SAMPLING CONNECTIONS

Five (5) sample connections at various levels are used in operation of the Spiractor. Refer to page 2 for locations and designations.

For convenience in designation, the top sampling point on the vessel is referred to as sampling Point No. 4. The next lower point is called No. 3 and so forth down to sample point No. 1. The presence of catalyst grains in sample No. 4 indicates that the top of the catalyst bed is riding too high, either as a result of too high a flow rate or excess catalyst in the unit. Sampling point No. 5 is in the effluent nozzle.

PLACING IN OPERATION

Be sure that inside of tank is clean and free of debris. Install an initial charge of catalyst. Add catalyst until the bed level is between sampling points No. 6 and 7. This should require about 5 Cu. Ft. of catalyst. With the rated flow of 5 gpm flowing through the unit, the bed should be washed to waste for a few minutes to remove excessively fine grains.

Stop flow and check bed level. If level of catalyst is above sampling point No. 7 close inlet valve and remove excess catalyst. Place the lime feed in operation (if required).

CONTROL OF CATALYST BED

When operating at the design flow rate of 5 GPM the top of the expanded bed should be held between sampling point No. 9 and sampling point No. 10. The velocity of flow through top of the expanded bed should be approximately 11 to 12 gallons per minute per square foot of bed area. If a lower flow rate is used the bed should be lowered so that the velocity of flow through the surface of the bed (expressed as gpm per sq. ft.) is about the same as above. The chemical feed should, of course, be correspondingly reduced.

Operation of the unit should be such that a sample taken from point No. 3 shows the presence of catalyst material, while a sample from point No. 4 contains no catalyst.

If the sample from point No. 3 contains no catalyst, the flow rate may not be high enough or the quantity of catalyst may not have grown enough to show at this point.

If the sample from point No. 4 contains carried-over material, the flow rate may be too high, or there is an excess of catalyst in the unit and some catalyst must be blown out of the unit. Make adjustments as necessary.

IMPORTANT: NEVER ALLOW CATALYST TO CARRY OVER INTO **EFFLUENT WATER**

It is estimated that the individual catalyst grains will grow from 3 to 4 times their original diameters. The amount of replacement catalyst needed will vary from about 1/25th to about 1/50th of the grown catalyst removed.

When operating at lower rates, as described above, the amount of grown catalyst to be removed will, of course, be reduced. At lower flow rates it will also be necessary to increase the ratio of replacement catalyst to grown catalyst removed so that the average catalyst grain size will be smaller than when operating at full rate of flow. If catalyst outlet becomes plugged, a manual pressure supply is provided to flushback and free catalyst outlet.

Alternately the level of the catalyst bed may be determined by sounding from the top of the spiractor with a weighted disc suspended on a cord, by actually feeling the catalyst level when the spiractor is in operation.

ANTICIPATED OPERATING RESULTS

It is possible that changes will occur in the composition of the raw water. Changes in the chemical dosages will have to be made to meet changes in the composition of the raw water.

CHEMICAL CONTROL

As a matter of routine control, the following tests should be made several times a day on both the raw and treated water and then results recorded.

Total Hardness (T.H.)
Methyl Orange Alkalinity (Alk. A)
Phenolphthalein Alkalinity (Alk. B)

To convert the total hardness the treated water in ppm to grains per gallon use the following formula.

T.H. in ppm (As $CaCo_3$) = (17.1)(T.H. in grains per gallon)

Substitute the ppm values of Alk. A and Alk. B obtained for the treated water in the following equation.

(2 x Alk. B) - A = Alk. C

If the result of the above equation is between +5 and -5 the operation is satisfactory. If the result is less than -5 the lime feed should be increased. If the result is greater than +5 the lime feed should be decreased. These values are intended to be *general guidelines*.

After making any changes in the chemical dosage, allow 20 to 30 minutes to elapse so that the new conditions will become stabilized through the Spiractor before taking a sample for new tests.

EDUCTOR

PURPOSE

The eductor carries the fresh catalyst from the hopper up into the Spiractor.

OPERATING PROCEDURE

The inlet valve to the eductor should be opened. The dilution water line should then be opened, and the catalyst fed into the hopper. It will be carried from there up into the Spiractor.

When sufficient catalyst has been withdrawn from the hopper by the eductor to fill the Spiractor up to the level between sample connections No. 1 and 2 the dilution line should be closed, followed by the eductor inlet valve, in that order. If too much time is allowed to elapse between the closing of the discharge and the inlet valves, the hopper will overflow.

TRI-SODIUM PHOSPHATE FEED (IF REQUIRED)

Tri-Sodium phosphate solution should be made and pumped by a positive displacement pump. The phosphate solution is introduced into the raw water entering the Spiractor.

A weighed quantity of chemical is placed in the feed tank and water is slowly admitted, thus dissolving the charge. Agitate if necessary to dissolve all the phosphate.

The pump should deliver a measured volume at a constant rate. The amount of chemical feed may be varied by varying the strength of the mixture. If the pump runs, however, at a constant number of strokes per minute, the volume of solution delivered can be varied by adjusting the stroke. Keep in mind that any change in the flow rate must be accompanied by a corresponding change in chemical feeding.

EFFLUENT CONTROL

Liquid flowing into the effluent pocket and thence to service is controlled by means of a weir.

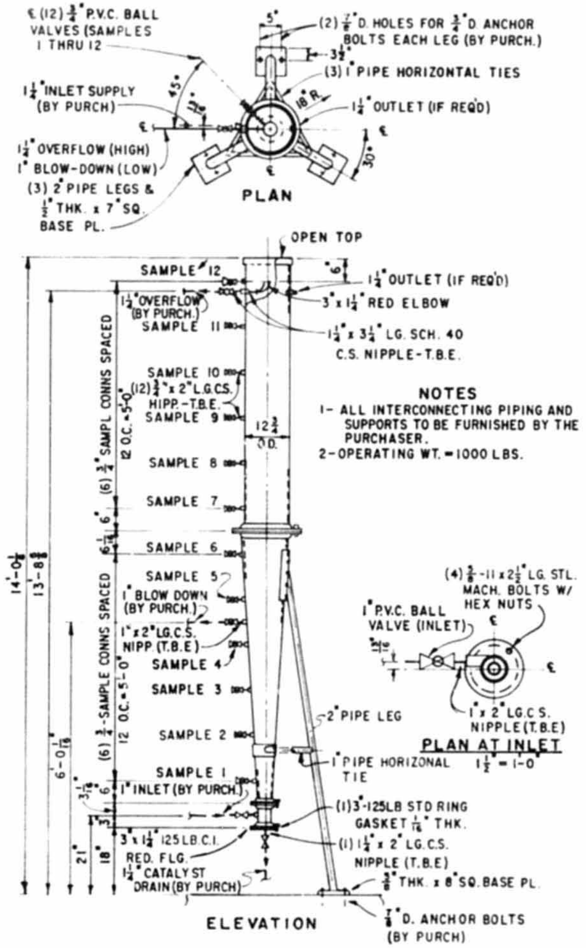

Figure H.1. — Spiractor assembly

BILL OF MATERIAL
U.S. BUREAU OF RECLAMATION
YUMA, ARIZONA

(1) 12" DIA. TOP X 3" DIA. BOTTOM
x 14'-0" HG. SPIRACTOR PILOT PLANT

JOB A141J-20360-98
SHEET 1 OF 2

	Drawing	Quantity	Description	Acct. 070-61-33
1	157-32310		12" Top Dia. x 3" Bottom Dia. x 14'-0"	
2			Gravity Spirator Assembly (W. Legs)	
3				
4	129-25785	1	12" Dia. Top x 3" Dia. Bottom x 12'6" Hg. Stl.	
5			Gravity Spiractor Tank Det. Complete Per	
6			Dwg., Notes & Specs on Dwg	
7				
8			**Inlet Piping**	
9		1	1" PVC SCR. End Single Union Manual Ball Valve	
10				
11		1	1" x 2" Lg. Sch. 40 C.S. Nipple — TBE	
12				
13			**Over Flow Piping**	
14		1	1¼" PVC SCR. End Single Union Manual Ball Valve	
15				
16		1	3" x 1¼" C.I. SCR. Red. Elbow — 90°	
17		2	1¼" x 3¼" Lg. Sch. 40 C.S. Nipple — TBE	
18				
19			**Catalyst Blow-Down & Drain**	
20		1	1¼" PVC SCR. End Single Union Manual Ball Valve	
21		1	1¼" x 2" Lg. Sch. 40 C.S. Nipple — TBE	
22				
23			**Blow-Down Piping**	
24		1	1" PVC SCR. End Single Union Manual Ball Valve	
25		1	1" x 2" Lg. Sch. 40 C.S. Nipple — TBE	
26				
27				

Figure H 2 — Spiractor bill of materials

JOB A141J-20350-93
SHEET 2 OF 2

	Drawing	Quantity	Description	ACCT. 070-31-33
1			**Sample Piping**	
2		12	¾" PVC SCR. End Single Union Manual Ball Valves	
			Tag: Sample 1 Thru 12	
4				
5		12	¾" x 2" Lg. Sch. 40 C.S. Nipples — TBE	
6				
7			**Chemical Feed**	
8		3	Milton Roy Chempac 50 Consisting of the following:	
9			Steel Tank, Pump Mounting Platform Beneath Tank,	
10			Hinged Lid Suction Piping, Shut-off Valve,	
11			Strainer & Sight Level Indicator.	
12				
13			Milton Roy Diaphragm Pump — MROY Model R130A-37,	
14			Cast Iron Construction, Std. ¼ H.P., 1725 R.P.M.,	
15			1 Phase, 60 Hz. 115 Volt Tenv Motor Integrally Mounted.	
16			Pump is Factory Mounted Under Above Tank.	
17				
18			Agitator Having Shaft & Propeller PVC Coated, Tenv.	
19			Motor for 1 PH. 60 Hz. 115 Volt, Chemical Mixer is	
20			Mounted on Above Tank by Factory.	
21				
22	187-16635	1	¾" Badger Disc Meter	
23		1	Pulse Transmitter Model MS-E1	
24				
24	187-12120	3	Auto. Time Control Timers	
26				
27				

Figure H 2 — Spiractor bill of materials (continued)

APPENDIX I

Table I.1 — *Ionics Incorporated product specification for electrodialysis membrane Type 204-SXZL-386*

PROPERTIES AND CHARACTERISTICS

Many standard Ionics electrodialysis reversal ("EDR") plants are furnished with 204-SXZL-386 Modacrylic fiber-backed anion transfer membranes for the transport of anions while retarding cations. The property data are typical values only and no warranty as to such properties is given.

Reinforcing Fabric: Modacrylic (copolymer of vinyl chloride and acrylonitrile)
Weight: 4 oz/yd^2 Specific Weight: 13.7 mg/cm^2
Membrane Thickenss: 20 mils (0.5 mm)
Burst Strength (Mullen): 100 psi (7.0 kg/cm^2)
Water Content: 46% of wet resin only
Capacity: 2.20 meq/dry gram resin (minimum)

VARIOUS ELECTROCHEMICAL PROPERTIES

Concentration	0.01 N NaCl	0.1 N NaCl	1.0 N NaCl	3.0 N NaCl
Area Specific Resist. (ohm-cm^2)	14.0	11	5	2
Specific Conductance mho/cm	3.6 x 10^{-3}	4.5 x 10^{-3}	10 x 10^{-3}	25 x 10^{-3}
Current Efficiency (Fraction of current carried by anions only)	0.99	0.96	0.88	—

OTHER PROPERTIES
Water Transport: 0.120 liters per Faraday in 0.6 NaCl @ 16 ma/cm^2
Sucrose Transport: 11.5 grams per Faraday from 30% sucrose in 0.2 N KCl into 0.02 N KCl @ 16 ma/cm^2

Table I.2. — *Ionics Incorporated product specification for electrodialysis membrane Type 61CZL-183*

PROPERTIES AND CHARACTERISTICS

Ionics' Dynel-backed cation-transfer membrane 61CZL183 has a tighter matrix than membrane 61AZL183. Therefore, it is suggested for use in electrodialysis as a means for transporting cations while excluding anions in applications where loss of solvent and of non-ionized product must be reduced below that obtainable with membrane 61AZL183.

Backing — Type: Dynel Specific Weight: 14 mg/cm^2
Weight: 4 oz/yd^2 Content: 34 wt% (dry)
Membrane Thickness: 24 mils (0.6 mm)
Burst Strength (Mullen): 115 psi (8 kg/cm^2)
Water Content: 40% of wet resin only
Capacity: 2.7 meq/dry gram resin

	0.01 N NaCl	0.1 N NaCl	1.0 N NaCl	3.0 N NaCl
Area Specific Resist. (ohm cm^2)	13	11	8	5
Spec. Conductance (mho/cm)	5 x 10^{-3}	6 x 10^{-3}	8 x 10^{-3}	12 x 10^{-3}

Sucrose Transport @ 16 ma/cm^2, 30% sucrose in 0.2 N KCl/0.02 N KCl (g/Faraday)

Table I.3. — *Ionics Incorporated product specifications for electrodialysis membrane Type 61AZL386*

PROPERTIES AND CHARACTERISTICS

Many standard Ionics electrodialysis reversal ("EDR") plants are furnished with 61-AZL-386 Modacrylic fiber-backed cation-transfer membranes for the transport of cations while retarding anions. The property data are typical values only and no warranty as to such properties is given.

Reinforcing Fabric: Modacrylic (copolymer of vinyl chloride and acrylonitrile)
Weight: 4 oz/yd^2 Specific Weight: 13.7 mg/cm^2
Membrane Thickness: 20 mils (0.5 mm)
Burst Strength (Mullen): 75 psi (5.25 kg/cm^2)
Water Content: 46% of wet resin only
Capacity: 2.44 meq/dry gram resin (minimum)

VARIOUS ELECTROCHEMICAL PROPERTIES

Concentration	0.01 N NaCl	0.1 N NaCl	1.0 N NaCl	3.0 N NaCl
Area Specific Resist. (ohm-cm^2)	16.7	11	4	2
Spec. Conductance mho/cm	3×10^{-3}	4.5×10^{-3}	12.5×10^{-3}	25×10^{-3}
Current Efficiency + (Fraction of current carried by cation only)	0.98	0.92	0.86	—

OTHER PROPERTIES

Water Transport: 0.200 liters per Faraday in 0.6 N NaCl @ 16 ma/cm^2
Sucrose Transport: 30 grams per Faraday from 30% sucrose in 0.2N KCl into 0.2N KCl @ 16 ma/cm^2

APPENDIX J

Table J-1. — *Fluid systems product specification for ROGA spiral-wound reverse osmosis element model 4160S*

Individual element performance under the following test conditions:

1. 2000 ppm NaCl solution
2. 420 lb/in^2 applied pressure
3. Solution temperature 77 °F
4. 10 percent water recovery
5. 30 minutes of operation prior to data collection
6. pH 5.0 to 6.0

Operating under the above conditions, the following initial performance can be expected:

1. Salt rejection based on chloride ion	Minimim 92.5 percent Average 95 percent
2. Product flow rate for a single element	1200 gal/day

Additional design information:

Maximum operating pressure	600 lb/in^2
Recommended initial operating pressure	400 to 450 lb/in^2
Maximum recommended feed flow rate per element	12 gal/min
Maximum pressure drop per element	12 lb/in^2
Maximum pressure drop per six-element pressure tube	60 lb/in^2
Maximum feedwater turbidity	1 JTU
Maximum pressure drop per element during cleaning	10 lb/in^2
Permeate flow to concentrate flow design ratio for the last element in a pressure tube	1.6
Maximum feedwater chlorine concentration	10 ppm
Maximum feedwater temperature	86 °F
Recommended feedwater pH range	4.0 to 6.0
Interconnector	Bonded to element
O-Rings, part number 10-0243	One supplied
Anti-telescoping/centering device	Bonded to element

Table J-2. — *Fluid systems product specification for ROGA spiral-wound reverse osmosis element model 4160HR*

Individual element performance under the following test conditions

1. 2000 ppm NaCl solution
2. 420 lb/in^2 applied pressure
3. Solution temperature 77 °F
4. 10 percent water recovery
5. 30 minutes of operation prior to data collection
6. pH 5.0 to 6.0

Operating under the above conditions, the following initial performance can be expected

1. Salt rejection based on chloride ion	Minimum 96 percent Average 97 percent
2. Product flow rate for a single element	1000 gal/day

Additional design information

Maximum operating pressure	600 lb/in^2
Recommended initial operating pressure	400 to 450 lb/in^2
Maximum recommended feed flow rate per element	12 gal/min
Maximum pressure drop per element	12 lb/in^2
Maximum pressure drop per six-element pressure tube	60 lb/in^2
Maximum feedwater turbidity	1 JTU
Maximum pressure drop per element during cleaning	10 lb/in^2
Permeate flow to concentrate flow design ratio for the last element in a pressure tube	1.6
Maximum feedwater chlorine concentration	1.0 ppm
Maximum feedwater temperature	86 °F
Recommended feedwater pH range	4.0 to 6.0
Interconnector	Bonded to element
O-Rings, part number 10-0243	One Supplied
Anti-telescoping/centering device	Bonded to element

Table J-3. — *Fluid systems product specification for ROGA spiral-wound reverse osmosis element model 4600PA*

Individual element performance under the following test conditions:

1. 2000 ppm NaCl solution
2. 420 lb/in^2 applied pressure
3. Solution temperature 25 °C
4. 10 percent water recovery
5. 30 minutes of operation prior to data collection
6. pH 5.0 to 6.0

Operating under the above conditions, the following initial performance can be expected:

1. Salt rejection based on chloride ion	Minimum 97 percent Average 98 percent
2. Product flow rate for a single element	1250 gal/day

Additional design information:

Maximum operating pressure	60 lb/in^2
Recommended initial operating pressure	400-450 lb/in^2
Maximum pressure drop per element	12 lb/in^2
Maximum feedwater turbidity	1 JTU
Permeate flow to concentrate flow design ratio for the last element in a pressure tube	1.6
Maximum feedwater chlorine concentration	0.0 ppm
Maximum feedwater temperature	45 °C
Recommended feedwater pH range	4.0 to 6.0
Interconnector, part number 05-0815	One Supplied
O-Rings, part number 10-0248	Two Supplied
Anti-telescoping/centering device	Bonded to element

Table J-4. — *Fluid systems product specification for ROGA spiral-wound reverse osmosis element model 1001PA*

Individual element performance under the following test conditions:

1. 35000 ppm total dissolved solids typical seawater
2. 800 lb/in^2 applied pressure
3. Solution temperature 25 °C
4. 7 percent water recovery
5. 30 minutes of operation prior to data collection
6. pH 5.0 to 6.0

Operating under the above conditions, the following initial performance can be expected:

1. Salt rejection based on chloride ion	Minimum 98.5 percent Average 98.9 percent
2. Product flow rate for a single element	800 gal/day

Additional design information:

Maximum operating pressure	1000 lb/in^2
Recommended initial operating pressure	800-850 lb/in^2
Maximum pressure drop per element	12 lb/in^2
Maximum feedwater turbidity	1 JTU
Permeate flow to concentrate flow design ratio for the last element in a pressure tube	1.10
Maximum feedwater chlorine concentration	0.0 ppm
Maximum feedwater temperature	45 °C
Recommended feedwater pH range	4.0 to 6.0
Interconnector, part number 05-0815	One supplied
O-Rings, part number 10-0244	Two supplied
Anti-telescoping/centering device	Bonded to element

Table J-5. — *Desalination Systems Inc., specification for B-400 Series spiral-wound industrial reverse osmosis cartridges*

Rated performance:

Model number	Output (gal/day)	Salt rejection (chloride ion)
B-402	700	99+

Production specification:

Membrane type	DESAL 1
Recommended operating pressure	400 to 800 lb/in^2
Maximum operating pressure	900 lb/in^2
Maximum feed flow rate	16 gal/min
Minimum concentrate flow rate	4.0 gal/min
Maximum pressure drop per cartridge	8 lb/in^2
Maximum pressure drop per vessel	40 lb/in^2
Maximum allowed short time temperature	158 °F (70 °C)
Maximum recommended continuous operating temperature	122 °F. (50°C)
pH Range, continuous operation cleaning or sterilization	1 to 8
(short time)	1 to 12
Chlorine tolerance	0

Table J-6. — *Desalination Systems Inc., specification for B-400 Series spiral-wound industrial reverse osmosis cartridges*

Rated performance:

Model number	Output (gal/day)	Salt rejection (chloride ion)
B-401	1000	99+

Production specification:

Membrane type	DESAL 1
Recommended operating pressure	400 to 800 lb/in^2
Maximum operating pressure	900 lb/in^2
Maximum feed flow rate	16 gal/min
Minimum concentrate flow rate	4.0 gal/min
Maximum pressure drop per cartridge	8 lb/in^2
Maximum pressure drop per vessel	40 lb/in^2
Maximum allowed short time temperature	158 °F (70 °C)
Maximum recommended continuous operating temperature	122 °F. (50°C)
pH Range, continuous operation cleaning or sterilization	1 to 8
(short time)	1 to 12
Chlorine tolerance	0

Mission of the Bureau of Reclamation

The Bureau of Reclamation of the U.S. Department of the Interior is responsible for the development and conservation of the Nation's water resources in the Western United States.

The Bureau's original purpose "to provide for the reclamation of arid and semiarid lands in the West" today covers a wide range of interrelated functions. These include providing municipal and industrial water supplies; hydroelectric power generation; irrigation water for agriculture; water quality improvement; flood control; river navigation; river regulation and control; fish and wildlife enhancement; outdoor recreation; and research on water-related design, construction, materials, atmospheric management, and wind and solar power.

Bureau programs most frequently are the result of close cooperation with the U.S. Congress, other Federal agencies, States, local governments, academic institutions, water-user organizations, and other concerned groups.

A free pamphlet is available from the Bureau entitled "Publications for Sale." It describes some of the technical publications currently available, their cost, and how to order them. The pamphlet can be obtained upon request from the Bureau of Reclamation, Attn D-922, P O Box 25007, Denver Federal Center, Denver CO 80225-0007.

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