

# **FLUORIDE REDUCTION in COMMUNITY WATER SUPPLIES**

**prepared for the :**

**STATE OF SOUTH CAROLINA  
DEPARTMENT OF HEALTH & ENVIRONMENTAL CONTROL  
WATER SUPPLY DIVISION**

**A JOINT VENTURE**

**J. E. SIRRINE COMPANY & AWARE, INC.**

**Summerville, South Carolina**

**Brentwood, Tennessee**

**volume ONE**

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IN  
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## ABSTRACT

Forty-three community water supplies located along the east coast of South Carolina contain fluoride concentrations which exceed the limit established by law. During the course of this study, each high fluoride supply was evaluated in an effort to identify one or more viable alternatives that might be implemented to achieve compliance.

The primary option proposed for the majority of systems was blending. By constructing shallow wells that tap a low-fluoride aquifer and mixing their yield with that of the existing deep wells, fluoride concentrations can be attenuated to acceptable levels in the combined supplies. For blending to be successful, twenty-five communities would be required to drill 204 new wells. The total estimated capital cost of blending is \$15,305,000.

The remaining communities can achieve compliance by implementing one of four other alternatives. Those alternatives and their attendant capital costs are as follows: seven would replace existing wells with new ones at a cost of \$385,000, eight would abandon existing supplies and purchase water from an adjacent utility at a cost of \$200,000, two would employ activated alumina treatment to reduce fluoride at a cost of \$10,472,000, and the remaining one would construct a separate distribution system to serve exempt users exclusively at a cost of \$230,000.

Implementation of the primary alternative by all forty-three communities would require a state-wide capital expenditure of \$26,592,000.

## ACKNOWLEDGEMENTS

This report represents the combined efforts of a great many individuals. It would be impractical to express appreciation to every contributor; however, the authors wish to acknowledge the assistance of Bob Heater of Heater Well Company, Larry West of the S. C. Water Resources Commission, and Al Zack of U.S. Geological Survey in providing groundwater data. The authors also wish to express their gratitude to the various equipment vendors, especially Hydronautics Water Systems and UOP Fluid Systems Division, for the invaluable information that they provided on water treatment processes . A special thanks is due the personnel that manage and/or operate the water systems included in this report for the information and advice that they offered. Two of those managers, Ron Bycroft of Mt. Pleasant and Robert Winfield of Conway, made significant contributions to this study by providing accommodations and facilities for the bench scale testing that was conducted. Another noteworthy contributor to the testing effort was Culligan Water Conditioning, Inc. who provided the equipment. Finally the authors thank Fred Soland of the South Carolina Department of Health and Environmental Control for his valuable guidance and suggestions.

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CHAPTER 1  
FLUORIDE REDUCTION  
IN SOUTH CAROLINA

INTRODUCTION

As of the 24th of June, 1977, water supplies throughout the United States were required to comply with the Environmental Protection Agency (EPA) National Interim Primary Drinking Water Standards. Those standards established a maximum contaminant level (MCL) for fluoride and nine other inorganic chemicals. Enforcement responsibility for the standards was requested by, and subsequently granted to, the South Carolina Department of Health and Environmental Control (SC DHEC). As an initial step toward bringing South Carolina water supplies into compliance with the law, the State authorized this study of fluoride reduction alternatives. During the course of the study, which was funded by EPA and conducted under the auspices of SC DHEC, forty-three community water supplies were evaluated and conceptual solutions for reducing fluoride concentrations were formulated for each of the affected systems. For systems where treatment of an existing or alternate water source was indicated, desk-top system designs were developed; where fluoride attenuation by blending the existing supply with the yield from proposed shallow wells was indicated, designs were based upon estimates of water quality and quantity obtained from well drillers, the South Carolina Department of Water Resources, and United States Geological Survey personnel. Planning-level cost estimates were prepared in 1980 dollars for all alternatives and served as a basis for ranking the various solutions in their order of viability.

Individual reports documenting the results of the study, as applicable to a specific system, were prepared and transmitted to the respective purveyor responsible for each supply determined to be in non-compliance. Copies of those reports are contained in Volume Two of this document.

#### PROPOSED FLUORIDE REDUCTION METHODS

All high fluoride water utilized as a community source of supply in South Carolina is drawn from wells. Reduction of the fluoride concentration in the yield from those wells can be achieved in one of three ways. The water can be treated to remove a portion of the fluoride, it can be blended with a low fluoride source, or the source can be abandoned in favor of a more acceptable supply. Depending upon a host of site-specific conditions, variations of all three basic fluoride reduction methods can be implemented by the various purveyors to bring their respective systems into compliance.

##### Blending

Blending low fluoride water with existing supplies to achieve an acceptable mix was determined to be a viable option for twenty-seven systems. Of those, blending ranks as the least expensive solution for twenty-five. Implementation of the primary alternative in all systems would require the construction of 204 new shallow blending wells having a combined capacity of 11,573 gallon per minute (GPM).

The cost of designing and constructing the blending facilities was estimated at \$15,305,000. The blended supply would be distributed to 29,465 consumers (approximately 100,000 people).

Water Quantity. The scope of this study did not include the construction of test wells; consequently, availability and capacity were estimated from existing information obtained from private contractors and governmental agencies. Some capacity variation will occur in all wells constructed in response to the information provided in this report. However, the projected yield should be attainable in most instances.

Water Quality. Blending shallow well water with existing supplies will create iron-related aesthetic problems in most communities. The severity of those problems will vary with the actual iron concentrations that are encountered and with the blending ratios required to reduce the fluoride content of the combined supply to acceptable levels.

Lacking accurate quality data, the assumption was made that sequestering would provide a relatively inexpensive means of controlling red water in most systems. Iron removal was included in the blending scenario for three communities where the success of sequestering is doubtful. Those communities are the City of Conway, Bulls Bay Rural Community Water District, and North Myrtle Beach.

Shallow wells are generally more susceptible to bacterial contamination than are deep wells. Consequently, gas chlorination equipment and contact tanks were included in the estimated cost of several specific blending alternatives.

#### Abandonment

Abandonment of existing wells in favor of another source of supply was determined to be a viable option for twenty-five communities. Purchas-

ing water from another utility and constructing new wells were the two variations of abandonment found to be feasible.

Purchase. Buying water from an adjacent utility was presented as a primary option for eight systems. Of those, five would be able to obtain a low-fluoride supply, the remaining three would tie to a larger system that is in noncompliance. The rationale behind connecting to a larger non-complying utility lies in the economy-of-scale that would be realized by the large purveyors in reducing fluoride as opposed to the small system operator. For example, Crystal Lakes Mobile Home Park would incur an annual per consumer cost increase of \$540.00 by installing an activated alumina treatment system on its existing supply. However, by purchasing water from the City of Myrtle Beach, which would also employ activated alumina treatment, Crystal Lakes could reduce its annual per consumer cost increase to \$133.11. Comparison of the two treatment alternatives in terms of cost per unit of installed capacity verifies the economy-of-scale to which the significant decrease in consumer cost is attributed. That comparison is as follows:

Crystal Lakes Mobile Home Park

● Installed A.A. system capacity	125 GPM
● Annual cost increase	\$108,000
● Annual cost increase per GPM of installed capacity	<u>\$864.00</u>

City of Myrtle Beach

● Installed A.A. treatment capacity	9460 GPM
● Annual cost increase	\$2,222,800
● Annual cost increase per GPM of installed capacity	<u>\$234.00</u>

Implementation of the primary alternative in all eight systems would result in the abandonment of 1127 GPM of water supply capacity. The service to 1378 consumers, approximately 4,800 people, would shift to a larger municipal supply.

Construction of New Wells was indicated as the primary option for seven communities. As was the case with blending wells, quality and quantity were estimated from existing information. Therefore, actual conditions encountered will vary somewhat from results predicted in this study utilizing assumed data.

The new wells, if constructed, will replace 775 GPM of existing capacity which will be abandoned. The replacement facilities will serve 305 consumers, approximately 1,100 people.

#### Treatment

Initially, two methods of water treatment were identified as possible fluoride reduction alternatives. One was reverse-osmosis, the other was an ion-exchange process utilizing activated alumina as the exchange medium. Further efforts, which included a complete review of available fluoride treatment literature and two limited bench-scale tests that were conducted as part of this study, resulted in the selection of activated alumina as the most viable process for fluoride reduction treatment.

It was noted that both treatment processes have demonstrated technical feasibility in many instances. However, full scale fluoride reduction is a relatively new area of endeavor for municipal water purveyors.

Therefore, the reader is cautioned that any design of full scale treatment facilities should be preceded with adequate bench scale and pilot plant testing of each source of supply.

Activated alumina treatment was determined to be a possible solution for twenty communities. Of those, only two, the City of Myrtle Beach and the Town of Edisto Beach, exhibit existing situations that favor the selection of treatment as the primary fluoride reduction alternative.

#### Unique Solutions

Of the forty-three communities evaluated, only the Town of Kingstree presented a situation requiring a unique solution. The primary alternative for the community proposes the construction of a separate distribution system which would serve exempt users exclusively. Five major consumers two schools, two industries, and a hospital are located in close proximity to the only municipal well that contains excessive concentrations of fluoride. By constructing a mini-system to serve the exempt users from the high fluoride well, Kingstree can achieve compliance without constructing treatment facilities or shallow blending wells.

#### Regional System

In addition to the individual alternatives that were presented for each community, a regional water system concept was developed for Horry and Georgetown Counties. The source of supply for the system would be the Great Pee Dee River. A 35 MGD conventional surface water treatment facility would be located on the northeast side of the river along U.S. Highway 701.

Transmission mains varying in diameter from 14 to 60 inches would convey the water in a westerly direction as far as Conway, in a southerly direction as far as Pawley's Island, and in a northerly direction as far as North Myrtle Beach.

Management and operation of the proposed system would be effected under a joint agreement of all political subdivisions involved.

Each community system served would purchase water on a bulk basis for resale to its consumers.

The capital cost of the regional system has been estimated at \$85,000,000. Assuming that the system operated during calendar 1980, annualized costs would total \$11,695,000. The bulk water purchase rate for that period was calculated to be \$2.95 per 1000 gallons. That computation was made as follows:

● Debt Service at 12% for 30 Years		\$10,551,900
● Estimated Operating Cost		<u>1,143,100</u>
Annualized Cost		\$11,695,000
● Current Annual Demand (in thousands)		
(10,860)(365) =		3,963,900
● Cost Per 1000 Gallons		
$\frac{\$11,695,000}{3,963,900}$ =		<u>\$2.95</u>

Design, securing permits and approvals, solicitation of proposals, contract negotiation and award, and construction of the proposed regional system can be accomplished within 60 months of completion of required referendums, rate structure studies, funding procurement, etc.

TABLE 1-1

## REGIONAL WATER SYSTEM DATA

COMMUNITY	CURRENT DEMAND (MGD)	1990 DEMAND (MGD)
North Myrtle Beach	1.571	5.100
Briarcliffe	0.120	0.309
Myrtle Beach	5.700	13.330
Myrtle Beach Air Force Base	0.173	0.311
Crystal Lakes MHP	0.032	0.058
GSW&SA Socastee Area	0.338	3.300
GSW&SA Garden City Area	0.135	0.385
Forest Acres Trailer Park	0.003	0.007
Platt Water Company	0.085	0.323
Surfside Beach	0.336	1.400
Garden City	0.402	2.160
Pawley's Island/ Murrell's Inlet	0.500	3.820
Wagon Wheel Farms	0.008	0.018
Inlet Oaks Trailer Park	0.020	0.036
Conway & Conway Rural	1.330	4.000
Mike Williamson MHP	0.008	0.014
Oakey Swamp Trailer Park	0.008	0.012
Bucksport Water Com- pany	<u>0.061</u>	<u>0.220</u>
	10.860	34.803



A summary of the community data utilized in developing the regional water system concept is tabulated in Table 1-1.

#### FLUORIDE REDUCTION COST

The financial information presented in this report is based upon 1979 consumer levels and 1980 planning-level cost data. The annual costs presented do not represent the total cost of any alternative. They do represent the increase that would be incurred if the proposed facilities were constructed and became operational during calendar 1980. For example, blending alternatives include the estimated annualized capital cost and the estimated operating cost for all proposed facilities. They do not include similar costs for existing facilities which will remain in operation. Consequently, the annual cost data presented in this document represents the increase that will result from fluoride reduction, not the total operating cost that will be borne by the community.

Another noteworthy item is that all capital expenses were assumed to be financed completely with long term loans repayable over a 30-year period at a 12% rate of interest. The assumption was made primarily to place the alternatives on an equal basis for comparison. Secondly, evaluation of the fiscal position of each purveyor included in the study was not possible within the scope of this project. In actuality, the funding program selected by each community will have a significant impact on the per consumer cost increases listed throughout this writing. In most cases - grants, utilization of existing cash reserves, and lower interest rates will act singularly or in concert to reduce the stated incremental cost increases. For example, at an interest rate of 12% for 30 years, annual debt service on \$1,000,000

would be \$124,140. By reducing the interest rate to 8% for 30 years, the annualized expense would be reduced to \$80,590.

Implementation of the primary alternative by all communities included in this study would result in a state-wide capital expenditure of \$26,592,000. That translates to approximately \$177 for each of the 150,000 people who would have the fluoride concentration of their water supply reduced as a result of the proposed changes.

#### SUMMARY

The results of this study are summarized in the following table.

TABLE 1-2

SUMMARY OF FLUORIDE REDUCTION ALTERNATIVES  
FOR COMMUNITY WATER SYSTEMS IN SOUTH CAROLINA

<u>COMMUNITY</u> and <u>ALTERNATIVES</u>	RANKING	NUMBER OF CONSUMERS (1979 data)	SYSTEM CAPACITY (gpm)	CAPITAL COST (thousands dollars)	ANNUAL COST INCREASE		
					OPERATING COST (thousands dollars)	DEBT SERVICE (thousands dollars)	CONSUMER INCREASE (dollars)
<u>American Heritage MHP</u> Construct shallow well and blend	1	65	44	50	0.1	6.3	97.05
<u>Aynor</u> Construct shallow wells and blend	1	325	150	90	0.8	11.2	36.92
Treat existing deep well supply with activated alumina	2		150	340	27.0	42.2	212.95
<u>Belmont S/D</u> Purchase water from Summerville	1	32	N/A	35	1.6	4.4	187.50
<u>Briarcliffe S/D</u> Construct shallow wells and blend	1	417	300	110	1.2	13.7	35.63
Purchase water from Myrtle Beach	2		N/A	35	70.3	4.3	178.91
Treat existing deep well supply with activated alumina	3		300	397	38.5	49.3	210.51
<u>Bucksport Water Company</u> Construct shallow wells and blend	1	303	355	180	0.5	22.3	75.39
Purchase water from Conway	2		N/A	100	40.7	12.4	175.43
Tie to regional system	3		150	453	6.0	56.2	205.00
Treat existing deep well supply with activated alumina	4		175	470	46.7	58.3	346.53
Construct surface water treatment facility on Pee Dee River	5		175	500	77.9	62.1	462.05

TABLE 1-2

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<u>COMMUNITY</u> and ALTERNATIVES	RANKING	NUMBER OF CONSUMERS (1979 data)	SYSTEM CAPACITY (gpm)	CAPITAL COST (thousands dollars)	ANNUAL COST INCREASE		
					OPERATING COST (thousands dollars)	DEBT SERVICE (thousands dollars)	CONSUMER INCREASE (dollars)
<u>Bulls Bay Water District</u> Construct shallow wells with iron treatment facilities and blend	1	856	500	580	18.0	72.0	105.14
<u>Central MHP</u> Construct new well and abandon existing supply	1	36	14	45	0.1	5.6	156.58
<u>Conway</u> Construct shallow wells with iron treatment facilities and blend	1	3,930	3,000	2,340	22.8	290.5	79.72
Treat existing deep well supply with activated alumina	2		3,000	2,270	207.0	281.8	124.38
Tie to regional system	3		2,777	10,427	138.7	1,294.5	364.67
<u>Crystal Lakes MHP</u> Purchase water from Myrtle Beach	1	200	N/A	30	22.9	3.7	133.11
Treat existing deep well supply with activated alumina	2		125	355	63.9	44.1	540.00
<u>Edisto Beach</u> Treat existing deep well supply with activated alumina	1	750	360	1,077	71.3	133.7	273.33
Treat abandoned brackish well yield with reverse-osmosis	2		360	1,100	185.4	136.5	429.33
<u>Forest Acres MHP</u> Purchase water from GSW&SA	1	14	8	20	1.6	2.5	290.07
Construct new wells and abandon existing supply	2		N/A	45	0.1	5.6	402.64

TABLE 1-2

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FOR COMMUNITY WATER SYSTEMS IN SOUTH CAROLINA

<u>COMMUNITY</u> and <u>ALTERNATIVES</u>	RANKING	NUMBER OF CONSUMERS (1979 data)	SYSTEM CAPACITY (gpm)	CAPITAL COST (thousands dollars)	ANNUAL COST INCREASE		
					OPERATING COST (thousands dollars)	DEBT SERVICE (thousands dollars)	CONSUMER INCREASE (dollars)
<u>Garden City Beach</u>							
Construct shallow wells and blend	1	1,319	1,580	1,000	4.0	124.2	97.16
Tie to regional system	2		1,500	3,134	41.7	389.0	326.54
Treat existing deep well supply with activated alumina	3		1,580	2,890	216.2	357.5	434.97
<u>GCW&amp;SA Pawley's/Murrell's</u>							
Drill shallow wells and blend	1	2,555	1,810	900	2.7	111.7	44.80
Tie to regional system	2		2,653	2,139	28.5	265.5	115.05
Treat existing deep well supply with activated alumina	3		1,810	2,255	208.0	279.9	190.97
<u>GSW&amp;SA Garden City Area</u>							
Construct shallow wells and blend	1	800	400	310	1.5	38.5	49.90
Tie to regional system	2		267	1,058	14.1	131.3	181.70
Treat existing deep well supply with activated alumina	3		400	840	62.0	104.3	207.85
<u>GSW&amp;SA Socastee</u>							
Construct shallow wells and blend	1	1,571	1,500	900	3.4	111.8	73.28
Treat existing deep well supply with activated alumina	2		1,500	1,590	156.0	197.4	224.94
Tie to regional system	3		2,292	2,650	35.2	328.9	231.80
<u>Hemmingway</u>							
Construct shallow wells and blend	1	532	785	205	0.3	25.5	48.30
Treat existing deep well supply with activated alumina	2		500	435	52.0	54.0	199.25

TABLE 1-2

## SUMMARY OF FLUORIDE REDUCTION ALTERNATIVES

## FOR COMMUNITY WATER SYSTEMS IN SOUTH CAROLINA

<u>COMMUNITY</u> and ALTERNATIVES	RANKING	NUMBER OF CONSUMERS (1979 data)	SYSTEM CAPACITY (gpm)	CAPITAL COST (thousands dollars)	ANNUAL COST INCREASE		
					OPERATING COST (thousands dollars)	DEBT SERVICE (thousands dollars)	CONSUMER INCREASE (dollars)
<u>INLET OAKS MHP</u> Construct shallow wells and blend	1	95	38	40	0.2	5.0	54.38
Purchase water from GCW&SA	2		N/A	25	21.8	3.1	262.15
<u>Isle of Palms Water Company</u> Construct shallow wells and blend	1	1,439	1,000	340	0	42.2	29.47
Treat existing deep well supply with activated alumina	2		1,000	734	72.5	82.6	107.71
<u>Jamestown</u> Install pump in existing shallow well and blend	1	90	180	28	0.1	3.5	39.22
<u>Kingstree</u> Isolate well and construct mini- system to serve exempt consumers	1	1,800	500	230	0	28.6	15.86
Construct shallow wells and blend	2		785	250	0.5	31.0	17.52
<u>Lane</u> Construct shallow well and blend	1	135	145	60	0.1	7.5	55.93
<u>Little River</u> Construct shallow wells and blend	1	290	259	110	1.0	13.7	50.54
Purchase water from GSW&SA	2		N/A	0	21.4	0	73.71
Treat existing deep well supply with activated alumina	3		325	331	26.0	41.1	231.34
<u>Loris</u> Connect selected existing wells and blend	1	870	440	100	3.3	12.4	18.06
Treat existing low fluoride wells to remove iron	2		450	400	19.5	49.7	79.49
Treat existing high fluoride wells with activated alumina	3		400	405	40.5	50.3	104.34

TABLE 1-2

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## FOR COMMUNITY WATER SYSTEMS IN SOUTH CAROLINA

<u>COMMUNITY</u> and <u>ALTERNATIVES</u>	RANKING	NUMBER OF CONSUMERS (1979 data)	SYSTEM CAPACITY (gpm)	CAPITAL COST (thousands dollars)	ANNUAL COST INCREASE		
					OPERATING COST (thousands dollars)	DEBT SERVICE (thousands dollars)	CONSUMER INCREASE (dollars)
<u>Mike Williamson MHP</u>							
• Construct new well	1	40	15	45	0.1	5.6	142.18
• Purchase water from Conway	2		N/A	5	5.6	0.6	156.38
<u>Mt. Pleasant</u>							
Construct shallow wells and blend	1	5,200	2,460	1,470	0	182.5	35.09
• Treat existing deep well supply with activated alumina	2		2,250	1,650	167.2	204.8	71.54
<u>Myrtle Beach</u>							
• Treat existing deep well supply with activated alumina	1	10,000	9,460	9,395	1,056.5	1,166.3	222.28
• Tie to regional system	2		9,257	44,665	593.9	5,543.5	613.75
<u>Myrtle Beach Air Force Base</u>							
• Construct shallow wells and blend	1	865	325	380	0.2	47.2	54.74
• Purchase water from Myrtle Beach	2		N/A	45	102.0	5.6	124.35
<u>North Myrtle Beach</u>							
• Construct shallow wells and blend	1	4,700	4,400	5,100	32.0	633.2	141.52
• Treat existing deep well supply with activated alumina	2		4,400	4,915	461.0	610.2	227.90
• Tie to regional system	3		3,542	12,535	166.7	1,556.1	366.55
<u>North Tranquil Acres S/D</u>							
• Purchase water from Summerville	1	175	N/A	13	7.1	1.6	49.71
<u>Oakey Swamp MHP</u>							
• Purchase water from Conway	1	20	N/A	4	2.8	0.4	162.60

TABLE 1-2

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FOR COMMUNITY WATER SYSTEMS IN SOUTH CAROLINA

<u>COMMUNITY</u> and <u>ALTERNATIVES</u>	RANKING	NUMBER OF CONSUMERS (1979 data)	SYSTEM CAPACITY (gpm)	CAPITAL COST (thousands dollars)	ANNUAL COST INCREASE		
					OPERATING COST (thousands dollars)	DEBT SERVICE (thousands dollars)	CONSUMER INCREASE (dollars)
<u>Pine Ridge MHP</u> urchase water from regional water ystem. Note: regional system is n planning stage. Cost and imple- entation schedule not available at his time.	1	22	N/A				
shallow well and blend	2		33	29	0.1	3.6	164.77
<u>Plantersville Water System</u> nstruct shallow wells and blend	1	240	150	55	0.3	6.8	29.70
urchase water from Brown's Ferry stem	2		N/A	0	8.0	0	33.22
<u>Platt Water Company</u> struct shallow wells and blend	1	450	270	160	0.7	20.0	45.56
eat existing deep well supply with ivated alumina	2		200	350	29.6	43.5	162.22
to regional system	3		224	666	8.9	82.7	203.39
<u>Red Hill Water System</u> struct new well	1	40	28	35	0.1	4.3	111.13
chase water from Brown's Ferry tem	2		N/A	45	1.3	5.6	172.93
<u>Rock Bluff S/D</u> ruct new well	1	21	8	45	0.1	5.6	268.24
<u>Rose Hill Water System</u> ruct new well	1	42	32	35	0.1	4.3	106.07
hase water from Brown's Ferry	2		N/A	42	1.5	5.2	158.95



TABLE 1-2

SUMMARY OF FLUORIDE REDUCTION ALTERNATIVES  
FOR COMMUNITY WATER SYSTEMS IN SOUTH CAROLINA

<u>COMMUNITY</u> and ALTERNATIVES	RANKING	NUMBER OF CONSUMERS (1979 data)	SYSTEM CAPACITY (gpm)	CAPITAL COST (thousands dollars)	ANNUAL COST INCREASE		
					OPERATING COST (thousands dollars)	DEBT SERVICE (thousands dollars)	CONSUMER INCREASE (dollars)
<u>Sangaree S/D</u> Construct regional system	1	815	N/A	(See Note	Pine	Ridge MHP)	
Purchase water from CCPW	2		N/A	96	27.5	11.5	47.85
<u>South Tranquil Acres S/D</u> Purchase water from Summerville	1	100	N/A	2.0	2.9	0.3	31.00
<u>Town of Stuckey</u> Construct new well	1	86	200	135	0	16.7	194.31
<u>Sullivans Island</u> Construct additional shallow wells and blend	1	790	500	18	0	2.2	2.83
Construct new deep well and treat with activated alumina	2		500	865	50.7	107.4	200.00
<u>Surfside Beach</u> Construct shallow wells and blend	1	1,485	1,167	700	3.4	86.9	60.78
Treat existing deep well supply with activated alumina	2		1,170	1,115	102.8	138.4	162.40
Tie to regional system	3		972	2,632	35.0	326.8	243.63
<u>Sycamore Acres S/D</u> Transfer water from adjacent system and blend	1	91	175	50	0	6.2	68.22
<u>Wagon Wheel Farms</u> Construct new well	1	40	19	45	0	5.6	140.23
Purchase water from GCW&SA	2		N/A	4	5.4	0.5	147.93

## CHAPTER 2

### TREATMENT

#### FOR FLUORIDE REDUCTION

Previous experiences with the removal of fluoride were reviewed in order to evaluate methods which have been used to remove fluoride from drinking water. The purpose of this review was to define the process capabilities and limitations of demonstrated systems, and to establish the most appropriate technologies for the systems being evaluated on this project.

#### INCIDENTAL FLUORIDE REMOVAL

There are a number of processes employed in water treatment systems which may incidentally remove fluoride. The following technologies fall into this category and are briefly discussed with respect to their defluoridation capabilities:

- Removal of fluoride utilizing the lime softening process.
- Removal of fluoride using carbon adsorption.
- Removal of fluoride utilizing ion exchange resins.
- Precipitation of fluoride using ferric salts.
- Fluoride removal utilizing zeolites.

#### Lime Softening

Lime is best considered a viable defluoridation alternative for harder waters having moderately low fluoride levels, between 3.0 and 4.0 mg/l.<sup>1</sup>

Laboratory Experiments. In 1934 Boruff<sup>2</sup> discovered that the addition of lime to fluoride spiked waters resulted in partial fluoride removal, reducing 5.0 mg F<sup>-</sup>/l to 3.0 mg F<sup>-</sup>/l. Co-precipitation of the fluoride was found to occur when the carbonate and magnesium hardness precipitated and the phenolphthalein alkalinity was greater than or equal to the total alkalinity. A detailed analysis of the mechanism of fluoride reduction during lime softening was made by Scott.<sup>3</sup> Scott observed that fluoride was adsorbed on the gelatinous magnesium hydroxide precipitate formed in the lime/magnesium reaction. Scott also found a mathematical relation between magnesium and fluoride removal which holds that the fluoride reduction is approximately equal to 7 percent of the initial fluoride multiplied by the square root of the magnesium removed. Consequently, only a high degree of magnesium removal will result in significant reductions of fluoride. Strict control of pH and caustic alkalinity must be maintained to secure proper magnesium precipitation.

Installations Employing Lime Softening for Defluoridation. Early reports of fluoride reduction by three small lime softening plants<sup>3</sup> showed modest removal capabilities, as shown in Table 2-1. Culp and Stoltenberg<sup>4</sup> ran experiments on a local water to determine the increase in fluoride reduction when adding sufficient magnesium in a lime softening process. Their magnesium dosages, according to Scott's formula, resulted in good fluoride removals (from 3.2 mg F<sup>-</sup>/l influent to less than 1 mg F<sup>-</sup>/l effluent). However, scale-up calculations indicated that the additional magnesium requirement was excessive. Several plants in states with hard water wells, such as Ohio, Indiana and Illinois were able to achieve modest fluoride

TABLE 2-1

FLOURIDE REDUCTION AT THREE LIME-SOFTENING PLANTS (Scott, et al)

	Plant 1		Plant 2		Plant 3	
	Raw	Eff	Raw	Eff	Raw	Eff
Fluoride	1.5	0.7	2.1	1.6	1.7	0.95
Magnesium	46	14	23	9	42.7	16
Total Hardness	523	121	234	65	449	99
Alkalinity	337	42	270	101	373	88
pH	7.8	10.3	7.7	10.0	7.1	10.0

levels economically<sup>5</sup> because these waters contained adequate amounts of magnesium. The operating costs of these softening plants were much lower than any of the more specialized processes, but the initial capital required to build such a plant was high.

#### Activated Carbon Adsorption

Few studies have been done on the potential of carbon adsorption because of the high costs associated with lowering (and subsequently raising) the pH of the water.<sup>6</sup> In 1934, McKee and Johnston experimented with four different types of carbon. One carbon type was successful in removing fluoride, an acid-treated carbon discarded from a soda pulp industry. The carbon removal efficiency was excellent on those waters having a pH of 3.0 or less. During this study fluoride levels were reduced from 8.0 mg F<sup>-</sup>/l to less than 1 mg F<sup>-</sup>/l. Tests on water at pH values above 3 showed no fluoride reduction.

#### Ion Exchange Using Synthetic Resins

The Rohm & Haas Amberlite IR-4B was first thought to be promising for the removal of fluorides. However, later studies showed that its capacity was entirely due to the precipitated aluminum oxide which had formed in the bed during alum regeneration. The floc caused severe flow restriction through the column and the process had to be abandoned.<sup>8</sup> Thompson and McGarvey<sup>9</sup> discovered that the strong basic Amberlite XE-75 (also made by Rohm & Haas) was quite effective for removing fluorides. The efficiency of the column was observed to be a function of the ratio of fluorides to the amount of total anions in the raw water as seen in Figure 2-1. Maier<sup>8</sup> calculated that

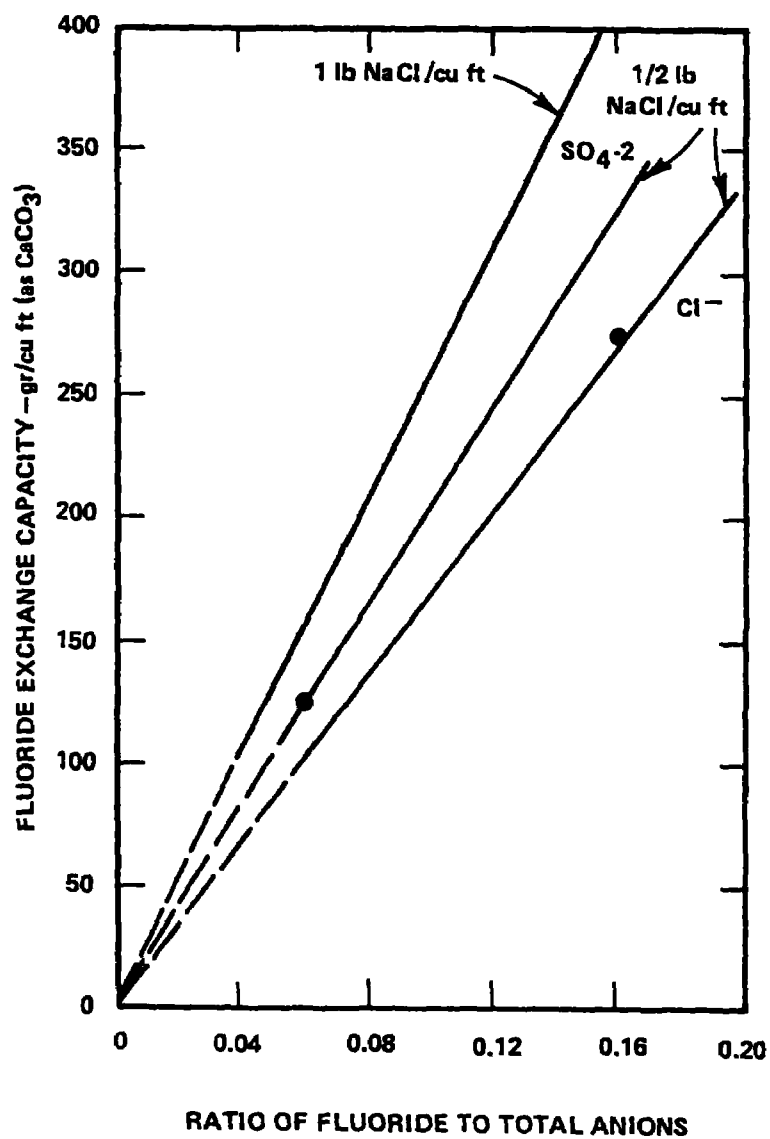


FIG. 2-1 . EFFECT OF INFLUENT-WATER COMPOSITION ON FLUORIDE EXCHANGE CAPACITY UTILIZING AMBERLITE XE-75

if the water being treated at Bartlett, Texas (with a fluoride-total anion ratio of 0.0147) were treated with this resin, the salt regeneration alone would cost about \$200/mil gal (1953 dollars). He also pointed out that most high-fluoride waters have very low fluoride-total anion ratios, exemplified by the 0.0083 ratio for Britton, South Dakota water which contained an average of 6.7 mg F<sup>-</sup>/l. Due to practically obtainable capacities ranging from 100-400 grains F<sup>-</sup>/cu ft,<sup>9</sup> resinous ion exchange is not economically competitive with the more fluoride-specific removal processes.<sup>1</sup>

#### Ferric Salts

Experiments by Boruff<sup>2</sup> were performed to determine removal of fluoride by complexing using ferric salts. The floc produced is a ferric fluoride complex which is slightly ionized. Applications of 34 to 85 mg/l of ferric salts on waters containing from 1.8 to 5.0 mg F<sup>-</sup>/l at pH 7.2 achieved minimal fluoride removal. When this treatment was accompanied by 170 to 340 mg/l lime, waters containing 1.8 to 5.0 mg F<sup>-</sup>/l were reduced to 1.6 and 4.7 mg F<sup>-</sup>/l, respectively. Ferric salts were found to be largely ineffective for practical fluoride removal.

#### Zeolite

Boruff<sup>2</sup> determined that natural zeolite (the particular type was not specified) has a small removal capacity for fluoride, based on laboratory experiments with a 5.0 mg F<sup>-</sup>/l water. Using a miniature contact bed, Boruff found that fluoride residuals of 0.6 mg F<sup>-</sup>/l were obtainable after passage of 6 liters of the stock water. After the first regeneration with

a saline solution, the minimum effluent fluoride concentration was 0.9 mg F<sup>-</sup>/l in the 4- to 6-liter fraction and 4.0 mg F<sup>-</sup>/l in the 44- to 47-liter fraction. Subsequent saline regenerations showed rapidly deteriorating capacity. Such a limited removal capability precludes zeolite from consideration as a viable treatment alternative.

#### FLUORIDE REMOVAL USING BONE-RELATED MEDIA

Bone and its related materials have demonstrated fluoride removal capability. Fluoride removal efficiencies are found to be much greater than those found with the more general treatment methods already discussed.

##### Bone

The use of bone as a substrate for defluoridating water was the earliest reported methodology shown to be effective, but it imparted an unwelcome flavor to the water. Bone is essentially a calcium phosphate,  $\text{Ca}_3(\text{PO}_4)_2$ , in which the carbonates are replaced by fluoride when fluoride-bearing waters are contacted with the media. The resulting fluorapatite compound is later flushed with a caustic solution to remove the attached fluoride, leaving behind a hydroxyapatite compound available for further fluoride sorption. Matters of economy led to the obsolescence of bone as a defluoridation media, as the use of charred bone and synthetic tricalcium phosphate/hydroxyapatite mixtures became widespread.<sup>1</sup>

##### Bone Char

Bone char is more economical and asthetically pleasing than bone. Consequently, it replaced bone as a media in contact filters. The concept of



using bone char was borrowed from the sugar industry, where it had been employed in the decolorization of syrups. Charred bone, which is an apatite and carbon mixture,<sup>1</sup> has proved its viability in many years of use as the media in the Britton, South Dakota plant and five California plants.

Laboratory Experiments. Burwell, et al<sup>10</sup> ran field tests on 30-40 mesh virgin bone black and obtained excellent fluoride removals on waters containing 5.0 mg F<sup>-</sup>/l. At an average flow rate of 20 gal/hr, 700 gal of the water could be passed through a 10 lb bed before effluent fluoride levels reached 1 mg F<sup>-</sup>/l. The authors found in a separate test that fluoride removal efficiencies quadrupled when 900 mg/l bicarbonate was reduced (by the addition of hydrochloric acid) to 100 mg/l bicarbonate. Trisodium phosphate regenerant was used instead of the common, but difficult to handle, caustic soda solution for easier handling in a small plant application. This method was compared to the caustic/hydrochloric acid method. It was found that although trisodium phosphate and sodium acid phosphate were more expensive, their superior performance and ease of handling justified these costs. However, other reports<sup>8</sup> reported that this pair of regenerant compounds reduced bed capacities by 12 percent after the initial regeneration. Regardless of the choice of regenerant, other studies have shown that bone char is not very durable and periodic media replacement is required.<sup>11</sup>

Installations Employing Bone Char. Bone char has been found to be a successful defluoridation media in several small installations in California.<sup>12</sup> The Camp Irwin plant serves a 200-family housing unit and a 50-space trailer court, treating a natural water with fluoride levels

between 9 and 12 mg F<sup>-</sup>/l. The plant has two separate contact units. Each unit is capable of treating 15,000 gal/cycle at an average flow of 10 gpm. Automatic controls enable the second unit to cut on when the first unit becomes exhausted. Tests taken during the first 30 days' operation showed that the fluoride content of the treated water was reduced to 0.6 to 0.8 mg F<sup>-</sup>/l. Sodium hydroxide and phosphoric acid are employed as regenerant chemicals.<sup>13</sup> Regeneration costs were high (\$1.20/1,000 gal in 1955). Media loss of 2 to 4 in. annually has been a problem, along with decreasing cycle times with each regeneration. The Camp Irwin plant changed to an activated alumina media in July of 1959, but returned to bone char in December of 1960.<sup>12</sup>

The oldest continuously operated defluoridation plant in the United States ran from 1948 to 1971 at Britton, South Dakota; this plant began operation with bone char in 1953.<sup>14</sup> The supply consisted of three wells containing 6.7 mg/l F<sup>-</sup> which were pumped to an enclosed steel tank at rates varying from 38 to 206 gpm. The contact tank contained 300 cu ft of 28-48 mesh bone char, resting on 12 in. of gravel. Regeneration was found to be necessary about every 450,000 gal. The media would first be backwashed with treated water (370 gpm for approximately 18 min) to remove the accumulated sand followed by 5,000 gal of a 1 percent caustic solution. The caustic was then removed by rinsing with raw water (7,000 gal at 150 gpm) and applying a weak carbon dioxide solution to the bed. When the pH of the effluent approached that of the raw water, the end of the cycle was reached.<sup>5</sup> After it was demonstrated through laboratory investigations that lowering the pH of the raw water could optimize the use of bone char, the Britton plant included such a modification in its process. An existing

sulfuric acid dilution system was altered to feed acid continuously to the raw water. However, due to the solubility of bone char in acids, the pH was maintained at levels of 8.4 to 7.1 which are above the range in which optimal removal is accomplished.<sup>14</sup>

### Synthetic Bone

Tricalcium phosphate and hydroxyapatite can be formed by reacting phosphoric acid and lime. The use of this material as a fluoride removal media stemmed from the need for a cheaper alternative to natural bone or bone char. In powder form, this synthetic bone can be added to raw fluoride-bearing water to precipitate the insoluble apatites. The material can also be made in the form of coarse granules for use as a contact medium in a filter.<sup>1</sup>

Laboratory Experiments. Investigations conducted by Adler et al<sup>15</sup> on tricalcium phosphate showed promise for its defluoridation capabilities. In column tests on a 30 mg F<sup>-</sup>/l water, the synthetic bone lowered the fluoride concentration to 0.3 mg F<sup>-</sup>/l in the first 3-liter fraction. Adler et al found that neither influent fluoride levels nor raw water pH have any significant effect on the fluoride removal capacity of tricalcium phosphate. Influent water hardness was found to have a slight effect (lower amounts being optimal). Comparative tests indicated that, pound for pound, tricalcium phosphate had roughly twice the capacity of activated alumina. Testing of several regenerants indicated that sodium hydroxide followed by hydrochloric acid was most cost effective. However, this method resulted in a loss of 2.5 to 3 percent of the tricalcium phosphate per regeneration.

Citing this loss of media, Behrman and Gustafson<sup>16</sup> conducted tests to study regenerant effects on the tricalcium phosphate. A fluoride-bearing water (5 or 10 mg F<sup>-</sup>/l) was passed downward through a 10-in. depth of 20 to 40 mesh tricalcium phosphate granules, at the rate of 1 gal/sq ft/min. Regeneration was accomplished by passing 1 liter of sodium hydroxide solution successively through the unit at 0.61 gal/sq ft/min. Behrman and Gustafson found that the fluoride removal efficiency of the bed became unacceptable in 25-35 cycles. Loss of efficiency occurred even though the 2 percent media loss per regeneration was being replaced at regular intervals. The actual overall loss per cycle was far greater than the 2 percent visible shrinkage observed by Adler et al. Careful examination of the tricalcium phosphate showed that the bulk density had increased from the original value of 35 lb/cu ft to about 68 lb/cu ft, and that the original porosity of 72 percent had decreased to about 45 percent. Furthermore, the average particle size of the spent material was appreciably greater than that of the original. Upon examination of the spent regenerating solutions, Behrman and Gustafson hypothesized that the hydrochloric acid had dissolved certain constituents from the upper part of the bed. These materials were being partly redeposited in the lower portion of the bed, with a consequent increase in density and particle size and a corresponding decrease in porosity. A modified method of tricalcium phosphate regeneration was proposed, which neutralized with carbon dioxide instead of with hydrochloric acid. More tests were done with the same bed, increasing the water wash to 2 liters, and substituting the hydrochloric acid with 3 liters of an aqueous solution of carbon dioxide containing 1.5 g CO<sub>2</sub>/l. The rate of flow during the application of the carbon dioxide solution was

increased to 1.8 gal/sq ft/min. At the end of one hundred thirty cycles, the following conditions were observed:

- The original volume of tricalcium phosphate had not decreased.
- The fluoride-removal capacity had remained the same.
- There was no apparent change in the density, porosity, or particle size of the granules.
- The fluoride concentration was reduced from 10 mg F<sup>-</sup>/l to 0.1 mg<sup>-</sup>/l after 2,377 B.V.

The city water in which these tests were conducted had a pH of 7.8 to 8.1 and an alkalinity of 110 to 115 mg/l.

Studies were also made by Goodwin and Litton,<sup>17</sup> who ran pilot plant experiments on natural waters containing 5.2 mg F<sup>-</sup>/l fluoride. The pilot plant consisted of a 1.3 cu ft bed of a tricalcium phosphate/hydroxyapatite mixture, through which the raw water was passed at the rate of 1.5 gpm. Goodwin and Litton achieved an effluent fluoride level of 0.42 mg F<sup>-</sup>/l, averaged over 50 runs. It was found that approximately 1 lb of caustic was required for each cubic foot of phosphate, and the authors estimated that a phosphate replacement of 5 percent might be necessary for every three hundred cycles of operation.

Installations Employing Synthetic Bone. Synthetic bone had its first full-scale application in 1937 at the Climax, Colorado water plant. Scooba, Mississippi followed in 1940. Both plants were abandoned in 1949.<sup>1</sup> The Climax plant, which treated a 14 mg F<sup>-</sup>/l water, consisted of four tanks, each charged with 45 in. of tricalcium phosphate. The combined capacity of the four units was a net of 300,000 gal of treated water in 24

hours. One or two of the units were regenerated each day, using sodium hydroxide and carbonic acid (carbon dioxide in solution). Loss of media from the filter was reported to be negligible.<sup>18</sup>

The previously described plant at Britton, South Dakota, began its operations in 1948 using Fluorex, a synthetic hydroxyapatite, as its media. The existing water supply system at Britton consisted of three 1,000-ft deep wells, drawing water with an exceptionally high mineral content, and a fluoride concentration of 6.7 mg F<sup>-</sup>/l.<sup>14</sup> Regeneration was performed every third day with a caustic solution followed by neutralization with carbonic acid.<sup>19</sup> The Fluorex media, although efficient for fluoride removal, was subject to high losses through regeneration. These losses through attrition reached 52 percent per year. Subsequently, the plant was forced to change to a bone char media in 1953.<sup>14</sup>

#### SORPTION USING ALUMINUM COMPOUNDS

A variety of compounds containing aluminum have been proposed for removing fluorides from water, including bauxite, sodium aluminate, aluminum sulfate, and activated alumina. Early research indicated bauxite and sodium aluminate had minimal removal capacity.<sup>2</sup> Aluminum Sulfate and activated alumina will be discussed with respect to their potential as viable full-scale defluoridation media.

##### Aluminum Sulfate

Aluminum sulfate has a high fluoride absorption capacity, but would be considered an effective treatment methodology only for certain waters. Coagulant aids such as clays and activated silicas may increase aluminum

sulfate capacity.<sup>1</sup> Certain cations also reduce capacity. In many cases, the reduction in capacity renders this method economically impractical.

Laboratory Studies. Regardless of potential cation interference, several laboratory experiments have demonstrated the effectiveness of aluminum sulfate in defluoridation. Kempf et al<sup>20</sup> added anhydrous alum,  $Al_2(SO_4)_3$ , to a city water containing 7.5 mg  $F^-/l$ . The pH of the raw water had been adjusted from 8.4 to 7.15. The vessel was stirred for 30 min and allowed to stand, samples being removed at intervals. Fluoride levels were found to be 0.85 mg  $F^-/l$  after two hours, and 0.40 mg  $F^-/l$  one day later. Tests were made on a separate city water with a fluoride concentration of 8.5 mg  $F^-/l$  to determine the effect of pH on defluoridation. The results of varying alum doses at two pH levels are shown in Table 2-2. Fluoride levels were measured 2.5 hr after mixing.

Boruff<sup>2</sup> found that a dose of 170 mg/l aluminum sulfate was necessary to reduce fluoride concentrations from 5.0 mg  $F^-/l$  to 1.0 mg  $F^-/l$  in a stock water. The alum was added to a vessel, mixed for 30 min and allowed to stand. The optimum pH range for the removal of fluorides with aluminum sulfate was found to lie between 6.25 and 7.0. Boruff observed that the chemical composition of the water was not of great importance in fluoride removal efficiencies. Boruff found that even chloride and sulfate levels as high as 1,000 mg/l had no effect on defluoridation. Treatment of waters containing 256 mg/l sodium silicate was found to decrease the removal capabilities of aluminum sulfate.

Kempf et al<sup>21</sup> published a report two years after their original study describing successful fluoride removal in a pilot plant using aluminum

TABLE 2-2

BENCH-SCALE STUDY OF ALUMINUM SULFATE (Kempf, et al)

Aluminum Sulfate (mg/l)	pH = 7.88 (mg F <sup>-</sup> /l)	pH = 6.95 (mg F <sup>-</sup> /l)
110	6.5	5.85
220	3.75	3.25
440	1.13	0.45



sulfate. The plant, attached to the mains in a school, consisted of three mixing tanks, a settling basin, and sand filters. Fluoride levels, which originally ranged from 7 to 10 mg F<sup>-</sup>/l, were reduced to between 1.5 and 2 mg F<sup>-</sup>/l with alum doses of approximately 340 mg/l.

Scott et al<sup>3</sup> treated stock waters of various fluoride concentrations with aluminum sulfate, thoroughly mixing each flask and allowing it to settle. The results of this work indicated that required alum doses exceeded those found adequate by Boruff.<sup>2</sup> Scott et al lowered initial fluoride levels of 1.7, 3.0 and 6.0 mg F<sup>-</sup>/l to 1.0 mg F<sup>-</sup>/l by the addition of 200, 340, and 890 mg/l alum, respectively.

Later work by Boruff et al<sup>22</sup> investigated a discrepancy noted between fluoride removal in natural waters and the removal earlier demonstrated with his laboratory-made sodium fluoride solutions, the natural waters showing a decrease in removal efficiencies. Tap water samples were prepared with 4 mg/l calcium fluoride, magnesium, fluoride, aluminum fluoride, sodium fluoride, and sodium fluorosilicate, respectively. All samples were treated with 9.72 grains of Al(SO<sub>4</sub>)<sub>3</sub> 18H<sub>2</sub>O per gal (167 mg/l), mechanically stirred, and allowed to settle. This study indicated that the cation associated with the fluoride ion in water greatly affected the degree of its removal by alum floc. The magnesium and aluminum fluoride salts exhibited a modest removal rate; sodium fluoride salt demonstrated a higher degree of defluoridation. The study results are summarized in Table 2-3.

TABLE 2-3

REMOVAL OF FLUORIDES FROM SALTS IN WATER AT  
VARIOUS pH VALUES BY ALUM FLOC

Salt	pH 7.4 <sup>a</sup>		pH 6.4 <sup>a</sup>		pH 6.6 <sup>a</sup>		pH 6.6 <sup>b</sup>	
	I <sup>c</sup>	F	I	F	I	F	I	F
NaF	4.0	1.7	4.0	1.8	4.0	1.6	4.0	1.0
CaF <sub>2</sub>	3.7	1.7	--	--	4.0	3.0	4.0	2.3
MgF <sub>2</sub>	3.9	1.5	3.9	2.2	4.0	3.5	4.0	3.3
AlF <sub>3</sub>	3.6	1.6	3.6	2.8	4.0	3.4	4.0	2.6
Na <sub>2</sub> SiF <sub>6</sub>	4.0	1.9	4.0	2.2	4.0	3.1	4.0	1.2

<sup>a</sup>Coagulant added over a period of 2-3 min.

<sup>b</sup>Coagulant added over a period of 30 min.

<sup>c</sup>I = initial fluoride concentration in mg/l; F = final fluoride concentration in mg/l.

### Activated Alumina

Known chemically as gamma aluminum oxide ( $\gamma\text{-Al}_2\text{O}_3$ ), activated alumina is manufactured both in gram quantities as an analytical reagent and in large quantities as a desiccant. It becomes "activated" when neutral alumina is treated with acid, commonly sulfuric or hydrochloric. The crystal structure of activated alumina is known to contain gaps in its cation lattice which result in local sites of negative charge. However, the availability of positively-charged sites is also indicated by the electroneutrality of the substance. Said sites accommodate anion "adsorption". Ionic adsorption of cations and anions may take place simultaneously but one path of exchange usually dominates. Experimental data presented in 1947 indicated that activated alumina was preferential to particular anions over others. In this hierarchy fluoride ranks third, after hydroxide and phosphite. Fluoride was followed by sulfite, ferrocyanate, chromate, sulfate, ferrocyanite, dichromate, nitrite, bromide, chloride, nitrate, et al, respectively. In treatment processes, the preferred ions can be used to displace (or regenerate) lesser-preferred ions. Activated aluminas' high preference for fluoride is a direct reversal of fluorides' low selectivity with most synthetic anion exchange resins. Since many ions are usually present to compete for adsorption sites, activated alumina is more efficient for fluoride uptake than any other anion-exchange media.<sup>23</sup>

In all applications of activated alumina this preferred-ion substitution principal holds true. If, for example, the media was manufactured with hydrochloric acid activation, the initial exchange upon water application would be fluoride taking the place of chloride, as long as the most-

preferred hydroxides are absent. Addition of hydroxide, which is more preferred than fluoride, is necessary to regenerate the fluoride-containing adsorbent. The hydroxide added at this point must be in dilute solution because strong caustic solution has been found to dissolve the alumina. Finally, applying dilute acid restores the fluoride removal capacity and readys the acidic alumina for another adsorption cycle.<sup>23</sup>

This procedure is summarized as follows:

- $\text{Alumina-HCl} + \text{NaF} \quad \text{Alumina} - \text{HF} + \text{NaCl}$
- $\text{Alumina-HF} + \text{NaOH} \quad \text{Alumina} - \text{NaOH} + \text{NaF} + \text{H}_2\text{O}$
- $\text{Alumina-NaOH} + 2 \text{HCl} \quad \text{Alumina} - \text{HCl} + \text{NaCl} + \text{H}_2\text{O} \quad (23)$

Maier<sup>1</sup> states that activated alumina can be reused indefinitely, but minimal replacement, 3 percent annually, of the media has been found to be required.<sup>11</sup>

Bench Scale Testing. Various laboratory experiments substantiated chemists' early claims of activated alumina's affinity for fluoride. In 1934, Boruff<sup>2</sup> reduced the fluoride concentration of 30 liters of water from 5.0 to 1.6 mg F<sup>-</sup>/l by passing the liquid through 300 g of activated alumina in a contact filter. Results of the 2 1/2-hour run indicated a bed capacity of 119 gr F<sup>-</sup>/cu ft. A total of nine such runs were made, during which regenerants and application rates were varied. Boruff also conducted tests on the effectiveness of adding powdered activated alumina to a stirred reactor containing water with 5 mg F<sup>-</sup>/l. The addition of as much as 50 mg/l powdered alumina failed to reduce the fluoride concentration below 2 mg F<sup>-</sup>/l.

Two years later, Fink and Lindsay<sup>24</sup> achieved a fluoride removal capacity of 334 gr/cu ft in a slightly larger contact apparatus. In their experiment, 250 gal of 5 mg F<sup>-</sup>/l water was passed through the unit before the effluent reached 1 mg F<sup>-</sup>/l. Fink and Lindsay tried different regenerant solutions, but found that only 8 percent NaOH accomplished complete renewal of the activated alumina.

Swope and Hess<sup>25</sup> reduced a 6.4 mg F<sup>-</sup>/l water to 1.1 mg F<sup>-</sup>/l using the same apparatus as that described by Fink and Lindsay. Interestingly, they achieved their highest bed capacities at their most rapid application rate (12 gal/hr), after the bed had been regenerated with 5 percent caustic. In addition, Swope and Hess attributed a significant reduction in water hardness to contact with the activated alumina.

In their laboratory studies on activated alumina, Savinelli and Black<sup>26</sup> used aluminum sulfate as a regenerant. Figures 2-2, 2-3, and 2-4 show relationships found for exchange capacity and influent fluoride concentration, influent alkalinity, and influent pH. Increasing regenerant dosages up to 12 lb alum/cu ft resulted in rapidly increasing fluoride removal efficiencies, approaching 2,000 gr/cu ft. Likewise, increasing regeneration time from 1 hr to 5 hr increased bed capacity from 600 to 1,750 gr/cu ft with the same regenerant dose. Concentration of the alum regenerant was found to have no effect on fluoride exchange capacity. No relationship was found for influent sulfate or chloride ion concentration and fluoride removal efficiency. Savinelli and Black determined that the most critical parameter affecting bed capacity was the influent alkalinity. Their data indicates that a fluoride removal efficiency of 3,400 gr/cu ft is achievable when the pH of the influent water is lowered to 5.6.

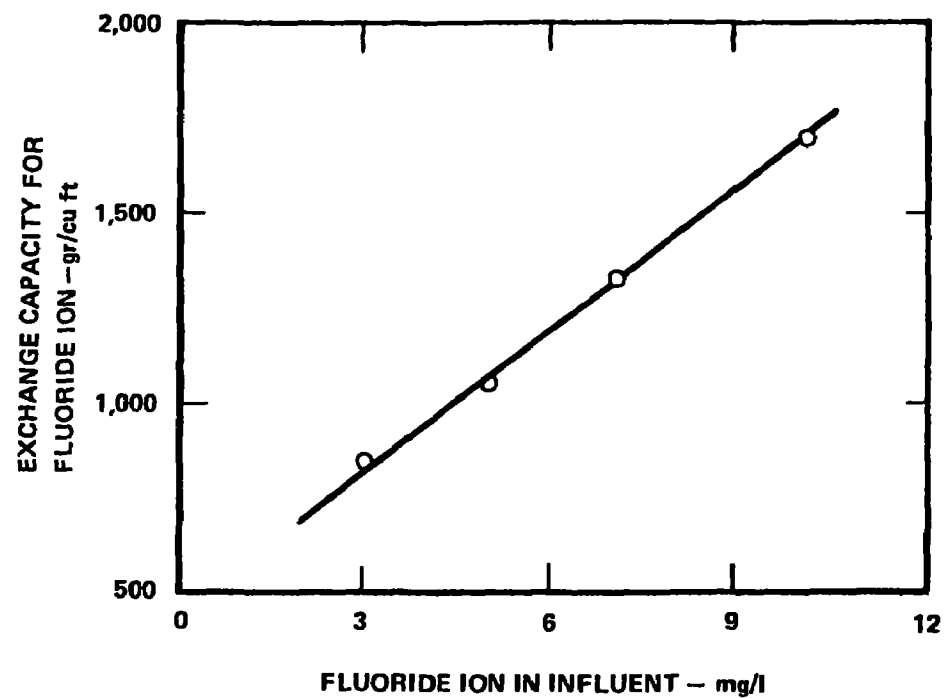


FIG. 2-2 . RELATION OF EXCHANGE CAPACITY AND FLUORIDE ION CONCENTRATION IN INFLUENT UTILIZING ACTIVATED ALUMINA

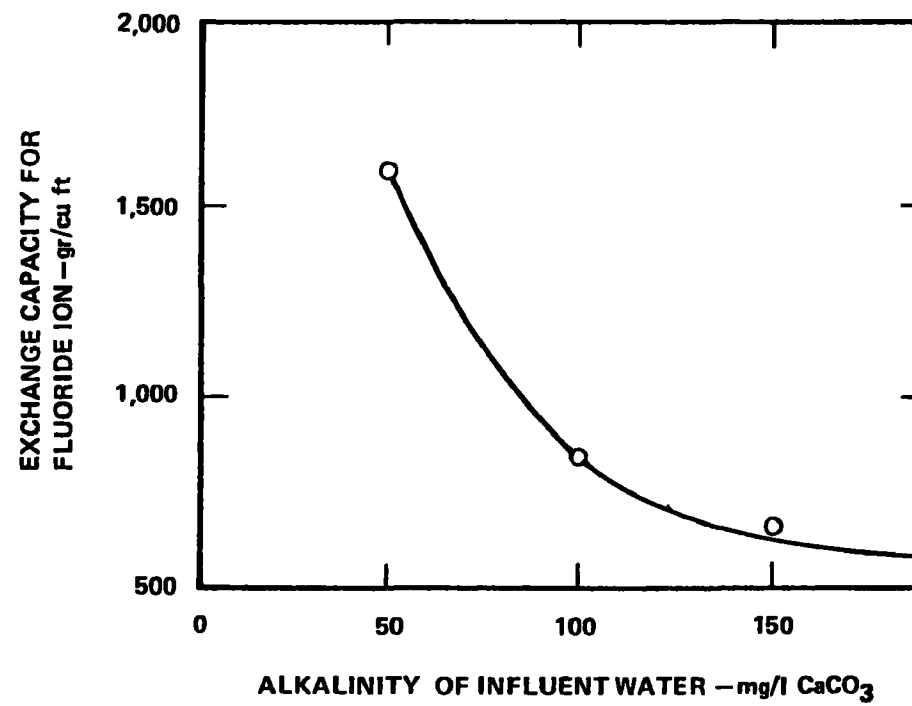


FIG. 2.3 . RELATION OF EXCHANGE CAPACITY AND ALKALINITY OF INFLUENT UTILIZING ACTIVATED ALUMINA

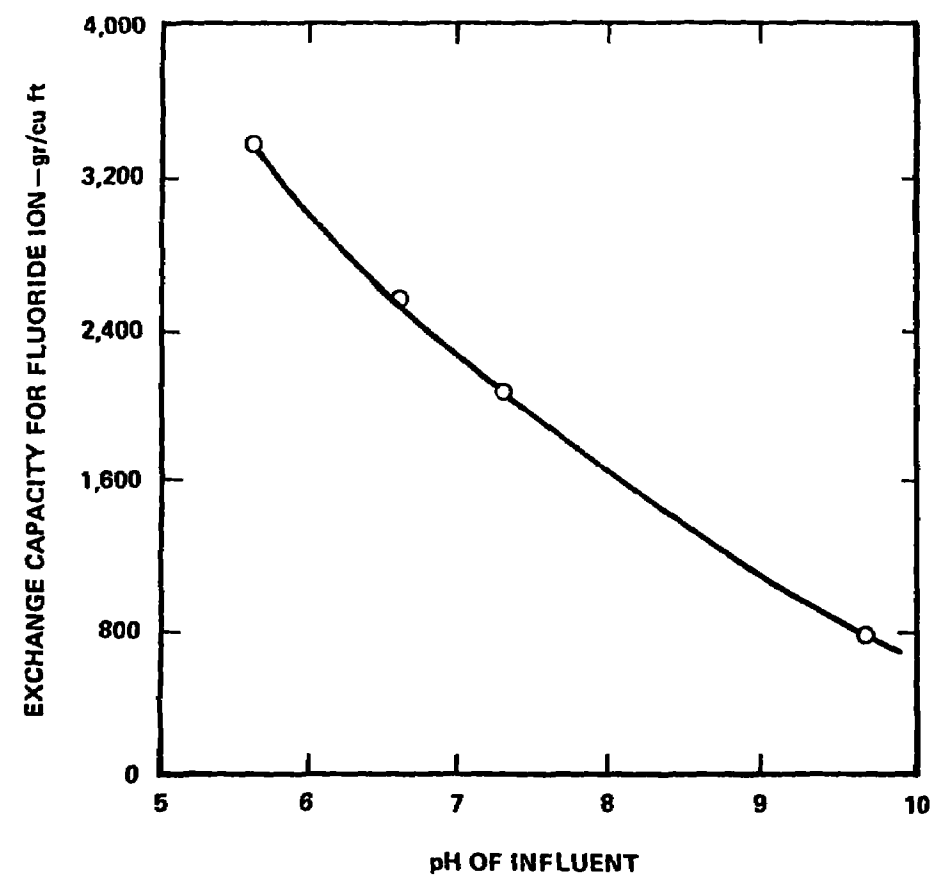


FIG. 24 .RELATION OF EXCHANGE CAPACITY AND pH OF INFLUENT UTILIZING ACTIVATED ALUMINA



The small bench column used by Bishop and Sansoucy<sup>27</sup> achieved a capacity of 1,424 gr/cu ft when reducing an influent 10 mg F<sup>-</sup>/l to 0.5 mg F<sup>-</sup>/l. The capacity was 600 gr/cu ft when treating a 5 mg F<sup>-</sup>/l influent to the same level. This finding illustrated the increasing difficulty in removing smaller quantities of fluoride. Their studies also confirm the feasibility of a fluidized activated alumina reactor. The increased surface area of the alumina granules provides greater exchange capacity than column applications while allowing high flow rates. Wu<sup>28</sup>, also experimenting with a fluidized system, added activated alumina to stirred 2-liter reactors containing fluoride solutions. Wu found that the ratio of fluoride concentration to activated alumina dose was an important factor in removal efficiency. He also determined that removals increase with lower fluoride to activated alumina ratios. Fluoride uptake was also found to be most efficient at a pH of 5 in this type of system.

Installations Employing Activated Alumina. Few full-scale fluoride removal plants exist. Those which have incorporated activated alumina technology have shown great improvement over the years in optimizing bed capacities. Three decades ago, bed capacities of 400<sup>1</sup> and 500<sup>29</sup> gr/cu ft were considered acceptable. Recently, capacities of over 3,000 gr/cu ft<sup>28, 30</sup> have been reported.

Activated alumina filtration was employed at the Bartlett, Texas plant from the time the plant opened in 1952<sup>5</sup> until its shutdown in 1977.<sup>11</sup> The plant was designed to reduce the fluoride concentration from 8.0 mg F<sup>-</sup>/l to an average of 1.0 mg F<sup>-</sup>/l and was capable of treating 400 gpm well water.<sup>8</sup> The 500 cu ft of activated alumina would require regeneration after approximately 450,000 gal of the fluoride-bearing water had been treated. A

1 percent sodium hydroxide solution was applied countercurrently to the bed at the rate of 235 gpm, and discarded to the sewer. After a rinse of 28,000 gal raw water, dilute sulfuric acid was passed through the media until the effluent alkalinity was the same as the influent. The total amount of water used per regeneration amounted to approximately 68,500 gal. In 1960, a new well was drilled at Bartlett that contained 3.0 mg F<sup>-</sup>/l fluoride; this extended the average cycle length to 1.5 mil gal. However, the overall bed capacity was reduced from 700 gr F<sup>-</sup>/cu ft to 400 gr F<sup>-</sup>/cu ft.<sup>5</sup>

The Army plant at Camp Irwin, California (normally employing bone char) temporarily used activated alumina as its filter media, but with little success.<sup>12</sup> It reportedly achieved an initial bed capacity of 500 gr F<sup>-</sup>/cu ft, but this dropped sharply after several regenerations. Large amounts of rinse water were mandatory to maintain acceptable effluent fluoride levels. The effective capacity was reduced to approximately 100 gr/cu ft after 30 to 35 cycles. Lee and Haras state that the plant study was similar to laboratory tests projecting a capacity loss rate of about 30 gr F<sup>-</sup>/cu ft per cycle for activated alumina.<sup>11</sup>

Although reports of capacity loss are uncommon, consideration should be given to frequent citations of bed cementing and blinding. Several authors<sup>11, 12, 27</sup> have noted a tendency of activated alumina beds to become clogged, thus creating shorter cycle times and highly inefficient removal. Harmon and Kalichman<sup>12</sup> cite the Elsinore, California plant as suffering from this "cementing" from an early point in its operation. Plant engi-

neers found it necessary to remove, dry, and crush all the media in the unit, as filter fines had aggregated to form a wet, grey plastic material. The problem at Elisnore was remedied by increasing caustic and acid doses.

Rubel and Woosley<sup>30</sup> dismiss these failures of activated alumina to ignorance and/or shortcutting by plant operators. They recently studied three fluoride removal plants with widely different raw water characteristics and outlined a program by which activated alumina's routine removal capability could be maintained above 2,000 gr/cu ft. Rubel and Woosley's recommendations for maximizing removal efficiency are reported to involve installation and operating costs compatible with limited public budgets and funding programs. Their optimization plan focuses mainly on six aspects of operation:

- Treatment Mode. Rubel and Woosley have determined that a pH of 5.5 is the optimum pH for activated alumina to attract fluoride ions, and advise that raw water pH be carefully held to this value. This recommendation is paramount to process optimization. If raw water pH goes above 6.0 or below 5.0, experiments have shown activated alumina bed capacity to decrease to 500 gr/cu ft. Earlier breakthrough also occurs when pH is not held at the optimum, serving as further incentive for careful pH control.
- Backwash Mode. The authors recommend backwashing with raw water prior to regeneration for two reasons:
  1. to remove those suspended solids which tend to blind the filter
  2. expand the bed to break up any channeling or wall effects. They cite an adequate backwash rate of 5 to 6 mm/s, which will expand the bed approximately 50 percent.
- Regeneration Mode. The most complete regeneration should proceed as follows:
  1. Upflow the regenerant while the bed is still expanded from backwashing.
  2. Follow regeneration with an upflow rinse, draining to the top of the media bed
  3. Downflow the regenerant and immediately follow with neutralization

Regeneration, say Rubel and Woosley, employs 1 percent sodium hydroxide solution flowing at 1.7 mm/s, and assuming a standard bed depth of 1.5 m, each step takes 35 min. Chemical requirements in each regeneration step are 27 liters of 50 percent NaOH per cubic meter of media (0.2 gal/cu ft). The intermediate upflow rinse flows at a rate of 3.4 mm/s for 30 min. The regenerate solution is usually an in-line dilution of 50 percent NaOH with raw water.

- Neutralization Mode. As soon as the downflow regeneration is completed, raw water with pH adjusted to 2.5 is fed downflow at the normal treatment flow rate. As the effluent pH neutralizes down to 9.0 to 9.5, the raw water pH is adjusted up to 4.0. When the treated water pH reaches 7.5, the final adjustment of the influent is made to 5.5 and carefully maintained.
- Initial Start-up Mode. Proper start-up procedures are critical when the activated alumina is first put on line. Rubel and Woosley strongly recommend that the contact vessel be half-filled with water before any bed material is placed into it. The presence of the water serves several important purposes. First, it can prevent cementing of the bed because it helps to dissipate the heat generated when the alumina becomes wet. Fines present in the bed material are separated out by the water from the more granular component, initiating stratification. Also, the water will protect the underdrain assembly from impact by the filling of the bed with media. An initial backwashing is also recommended to flush out all of the alumina fines. The backwashing should be carried out for an extended period of time.
- Blending. A large portion of the run time shows an effluent fluoride concentration far below the required level; likewise, there is a gradual increase to beyond maximum limits of concentration as breakthrough occurs. As a very cost effective method of increasing cycle time and cutting operating expenses, blending fluoride-bearing waters to obtain an average acceptable concentration is not a new concept.<sup>8,11,14</sup> For the three plants they studied, Rubel and Woosley found that the treated water can still be allowed to flow to storage or distribution until its fluoride concentration reaches one and one-half to two times the maximum allowable level. In systems without a large reservoir in which the major volume of a treatment run can be stored, blending can still be accomplished by staggered regenerations in plants that have two or more treatment units.

Other recommendations for the optimization of activated alumina technology include the hiring of fully qualified operators, proper selection of construction materials, and the disposal of wastewaters in accordance with local discharge standards.<sup>30</sup>

## DEMINERALIZATION BY REVERSE OSMOSIS

The reverse osmosis process was reported to be employed in more than five hundred plants having capacities greater than 25,000 gpd in 1977.<sup>31</sup> The principle underlying desalination by reverse osmosis (RO) is that fresh water will diffuse out of a brine solution across a membrane when the pressure applied to the brine side is greater than the osmotic pressure.<sup>32, 33, 34, 35</sup> The applied pressure, the ambient water temperature, and the permeability of the membrane are the major factors determining the quality and quantity of the product water. As an indication of fresh water flux across the membrane, approximately 1,000 mg/l of total dissolved solids (TDS) is equivalent to 10 psig osmotic pressure; the volume of desalinated effluent will be directly proportional to the difference between the externally applied pressure and this osmotic head.<sup>31</sup>

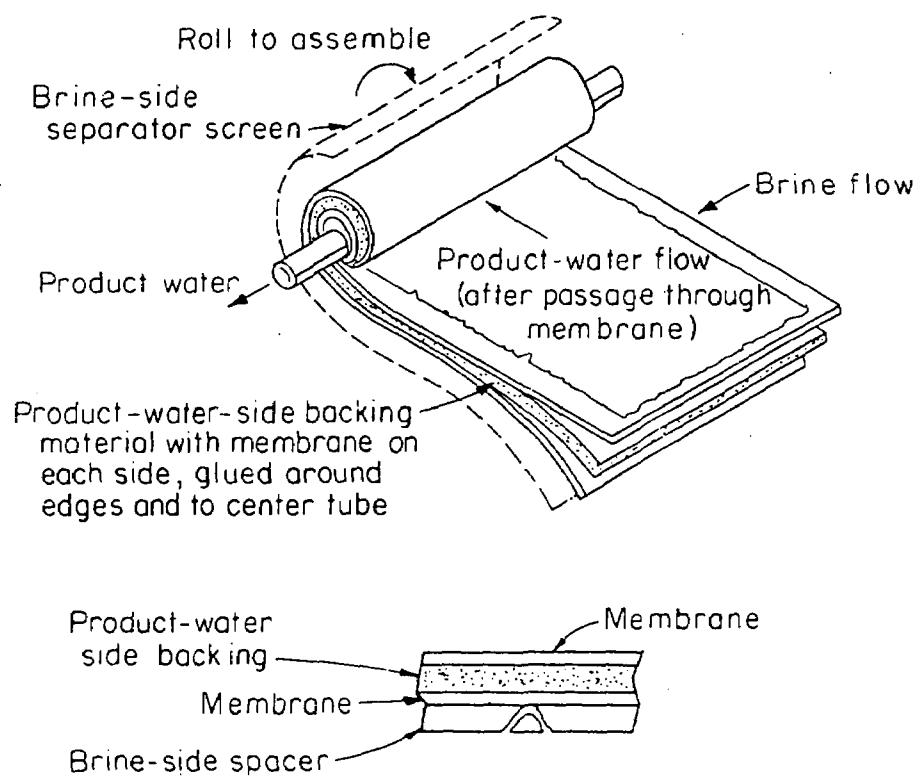
### Configurations of Membranes

In terms of its effect on overall process performance, the membrane is the most critical cost-determining item in an RO installation.<sup>33, 34</sup> Several different membrane configurations are in common use at reverse osmosis facilities: spiral wound, tubular, hollow fiber, and plate and frame. Whatever the physical arrangement, the membrane must be extremely well supported in order to withstand the high pressure drop across it.<sup>32</sup> There are also a large number of available membrane materials, although not all are readily convertible to any desired configuration of the membrane.<sup>33</sup> Commercially available membranes are commonly categorized as either cellulosic or non-cellulosic, with the most popular cellulose acetate being responsible for the recent strides in making RO an economically competitive technology.<sup>36, 37</sup> Reinforcing material or backing is compulsory for

adequate mechanical support of the film as the semipermeable cellulose acetate is often made in sheets only 1,000 Å thick. This backing material must consequently have pores less than 1,000 Å in diameter. Certain noncellulosic materials like DuPont's polyamidic nylon can be formed into hollow fibers that do not require any external support to withstand 1,000 psi or more.<sup>33</sup> Applicability of the different membrane materials also depends on certain influent water quality parameters, with the cellulosic membrane systems being much more constrained by their limited pH tolerance (approximately 3.5 to 7.5) than the non-cellulose forms (tolerating 3 to 11). However, these latter forms which include mostly polyamides, are unable to withstand free chlorine.<sup>36</sup>

Often employing cellulosic materials, the spiral wound RO element is composed of two reinforced membrane sheets sandwiching a product water collection tube, rolled up as illustrated in Figure 2-5. These spiral RO elements are assembled in series to produce the required flow. This particular configuration is known to have fairly good resistance to fouling, fair immunity to plugging, good production per unit volume, and high flux and rejection ranges. A broad span of operating pressures is attainable with spiral wound elements, and their costs/gal are relatively low.<sup>36</sup>

The tubular configuration utilizes its tube wall as a pressure vessel and the tube surface as its membrane support structure, thus combining two functions. The membrane is usually placed on the inside of the tube, and product water passes through to the outside.<sup>34</sup> Tubular systems have the widest range of operating pressures for reverse osmosis, from 400 to 1,000 psig. They are also the easiest of all configurations to clean and



**FIG. 2-5 . RO SPIRAL WOUND MODULE CONFIGURATION**

have a high resistance to plugging and fouling. These qualities make the tubular system the best choice for waters which contain very high levels of suspended solids. Compared to the spiral wound, however, production per unit volume is poor and they have a high capital cost/gal of water produced.<sup>36</sup>

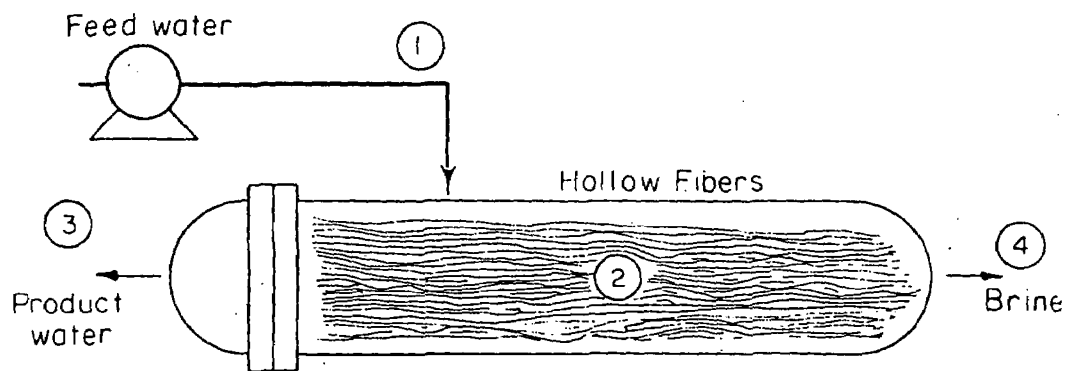
When membranes are cast in the form of hollow fibers, product water is collected in the inside, where it flows out axially to a collection manifold. This structural form may range from the size of a human hair (the polyamide hollow fiber) to four times larger (cellulose acetate hollow fibers).<sup>34, 36</sup> Sometimes referred to as "permeators", the hollow fiber configuration manifests high salt rejection, fair resistance to fouling, and very good tolerance toward chemical cleaning agents. Their low flux is compensated for by the extremely large surface area, which makes the gross production per unit very high. Costs/gal of water produced are relatively low for hollow fiber membranes.<sup>36</sup> The hollow fiber configuration is presented in Figure 2-6.

In plate and frame construction, membranes are mounted on both sides of solid plates into which product water channels have been cut. These plates alternate with brine feeding frames with the entire array housed in a pressurized vessel.<sup>34</sup> This configuration, which resembles a filter press, is not as widely employed as the other forms.

#### Prevention of Fouling and Pretreatment Requirements

There are two major causes of membrane fouling:





- ① Brackish feed water
- ② Product water collected in hollow fibers
- ③ Product water removed
- ④ Brine removed

Typical - OD:ID - 44.6 $\mu$ :24.9 $\mu$

**FIG. 2-6 . RO HOLLOW FINE FIBER MODULE CONFIGURATION**

- The presence of colloidal material and/or certain dissolved salts in the untreated water.<sup>38,39</sup> (However, these materials can be dealt with through pretreatment, which will be discussed later.)
- The lack of maintenance of a minimum brine flow rate, a parameter which is commonly reduced with the intention of increasing product water flow.<sup>31,40</sup> (This practice, however, robs the brine membrane surface of the turbulence necessary to prevent precipitation (scaling) and solids buildup, thereby causing what is known as concentration polarization.)

Polarization encourages dissolved ion breakthrough and deteriorating product water quality. However, as pointed out previously, certain membrane configurations are more inherently prone to this problem than other forms.<sup>40</sup>

Precipitates that are most apt to foul RO membranes include calcium carbonate, calcium sulfate, and various iron and manganese oxides.<sup>40</sup> Silica, humic acids, bacteria, and fine clay particles are also potential foulants.<sup>38, 39</sup> The expediency of pretreatment for RO is dependent upon the ease and effectiveness of periodic membrane flushing or cleaning, the frequency of such cleaning, and the membrane's resistance to cleaning agents. The need for pretreatment is indicated when the frequency of cleaning is unacceptable to the customer, or if the raw water contaminants cause irreversible fouling. Typical pretreatment consists of micron filtration for suspended solids removal, pH adjustment or softening, and some times activated carbon for organic removal.<sup>38</sup> Controlling pH to 5.0 to 6.0

will minimize the calcium carbonate scale. Calcium sulfate solubility can be increased by sequestering the calcium with the addition of sodium hexametaphosphate at temperatures below 96°F. Manganese levels to 1.5 mg/l and iron levels of 4.0 mg/l can be economically controlled with very small quantities of commercially available chelants, dispersants and sequestering agents.<sup>40</sup>

### Bench-Scale Testing

Pilot plant data has yielded much encouraging information concerning the efficacy and economy of reverse osmosis plants. Cruver and Sleigh<sup>41</sup> conducted studies at a 1,000 gpd sea water pilot plant at Sea World in San Diego, California to compare single-stage and two-stage reverse osmosis schemes. The authors admit that a single-stage system initially appeared more cost effective, since 20 percent less equipment was required. Their data, however, showed that the capital cost savings would be completely offset by the higher membrane costs. They found that their single-stage system required a membrane with 10 times the salt rejection capability of a two-stage system to achieve the same product water quality, one that would reject 99 percent of the salt. Such a membrane is certainly available, but the price is prohibitively high. The decision by Cruver and Sleigh to choose a double-pass system was also made on the basis of the increased reliability margin with such a system.

Schmitt and Hurley<sup>42</sup> report that the U.S. Army is testing portable RO units on surfacewaters to develop a transportable packaged system suitable for emergency use. Being employed in the studies are cellulose acetate tubular modules, rated at 10,000 gpd with 90 percent salt rejection. In tests on

Potomac River water, excellent results were found with these units, except for three instances of stoppage due to damaged parts. Separate experiments on spiral wound units showed a greater tendency toward fouling by suspended material. The authors admit that further study is needed before a recommendation can be made on a unit suitable for field army use.

### Installations Employing RO

Reverse osmosis systems have been installed to provide suitable water supply for both domestic and industrial use. It is currently employed at the Manassas, Virginia IBM semiconductor plant. The modules are used for demineralization of raw water prior to being fed to a mixed bed demineralizer. The influent water is drawn from a municipal surface water reservoir, having between 95 and 180 mg/l TDS. Pretreatment for the polyamide RO units includes acidification, cationic polymer coagulation, sand filtration, and carbon adsorption. The reverse osmosis system has been operating at the semiconductor facility since 1972 without any major problems.<sup>43</sup> Crabbe<sup>44</sup> described the success of a double-stage RO system for the production of water for pharmaceutical use. He cites the major reason for selecting the double-pass system as added protection from biological contamination. The 25 gpm installation includes softening, activated carbon, and a 10 micron filter as pretreatment. The influent water characteristics include 207 mg/l bicarbonates and 330 mg/l TDS, which are reduced to 4.9 and 7.3 mg/l, respectively.

A reverse osmosis system at a utility company is used as an effective pretreatment for the cation-anion demineralizer train for condensate water makeup. Wadlington<sup>45</sup> reports that the installation of the three-stage RO

unit (and its own pretreatment units) has resulted in substantial savings for the utility in regenerant chemicals. The number of pressure vessels in the three stages is staggered 7-4-3 (70 membrane modules) in the summer and altered 8-6-3 (185 membrane modules) in the winter to reduce increased system pressure. Membranes were replaced at this facility after two years of operation when water quality began to deteriorate. A similar experience is noted by Hollier<sup>46</sup> in his observations on an RO demineralizer for boiler feedwater at a Louisiana power station. Product water flux and quality had been seen to deteriorate, with the original output of 250 gpm dropping in 2½ years to 200 gpm, and conductivities from both stages increasing approximately 75 percent over that time period. The problem was found to be caused by iron oxides which were causing the membranes to foul. This was remedied later by acid pretreatment of the wellwater and installation of a different media in the prefilter.

Wastewater is being reclaimed with the aid of reverse osmosis in the Fountain Valley, California "Water Factory 21" plant. The final effluent from this 15 mgd facility is injected into wells, to be withdrawn later for irrigation, domestic, and industrial use. One-third of the flow passes through the RO unit, which utilizes large spiral wound cellulose acetate modules in a three-stage process. Six parallel pressure vessel assemblies each contain 35 vessels arranged in a 20-10-5 array, enabling the plant to attain 85 percent water recovery. Sodium, sulfate, chloride, electrical conductivity and COD are all reduced approximately 95 percent.<sup>47</sup>

Responsible for desalinating Red Sea water, the 3.2 mgd RO plant at Jeddah, Saudia Arabia is reported by Al-Gholaikah et al<sup>48</sup> to be performing well.

Spiral wound polyamide membrane modules are employed in a two-stage process, the first-stage units running at a 30 percent recovery and the second stage at 85 percent recovery. Pretreatment equipment for the reverse osmosis system includes acid storage and injection pumps, hexametaphosphate injection pumps, and cartridge filters. Waste brine from the first stage is discharged back into the Red Sea; reject from the second stage, being cleaner than the raw influent, is recycled to the front of the plant.

The Yuma, Arizona desalting plant is designed to operate on a 3,200 mg/l TDS feedwater. It will use reverse osmosis to produce a low TDS product water that can be blended with raw water to achieve the allowed effluent solids limits. 1 to 5 mgd will be treated by membrane desalination, with product water salinity approximately 386 mg/l TDS. The reject brine flow will be passed through an energy-recovery turbine and then routed to the Santa Clara Slough in Mexico.<sup>49</sup>

## SUMMARY

The review of demonstrated treatment technologies for the removal of fluoride indicate that there are three processes which have the capability to provide required water quality:

- Activated alumina.
- Bone char.
- Reverse osmosis.

The relative costs for the three processes have been prepared by the EPA<sup>50</sup> and The State of Arizona<sup>51</sup>. The EPA comparative costs for each of the three processes utilizing a 500 gpm system are presented in Table 2-4.

The Arizona Report indicates a total cost of 35¢/1,000 gal utilizing activated alumina, 95¢/1,000 gal utilizing bone char, and 90¢/1,000 gal utilizing reverse osmosis for a 500 gpm system.

Based on the outlined costs, and operational considerations, activated alumina has been chosen as the most viable technology for fluoride removal.

TABLE 2-4  
EPA COMPARATIVE COSTS

Process	Construction Costs (\$)	O&M Costs (\$/yr)
Activated Alumina	100,000	19,000
Bone Char	130,000	20,000
Reverse Osmosis	450,000	100,000



## CHAPTER 3

### ACTIVATED ALUMINA PILOT STUDIES

The findings of the literature survey, conversations with operators and engineers of existing fluoride removal facilities, and general cost surveys indicated that the activated alumina process is the most viable process for fluoride removal. Significant quantities of information are available on the use of activated alumina for fluoride removal. In order to confirm the application of activated alumina for fluoride removal, pilot studies were conducted at two sites in South Carolina.

The purpose of these pilot studies was to establish the fluoride removal capacities for a high and low fluoride water. The two pilot studies were conducted at the communities of Mt. Pleasant and Conway. Both sites were selected based on the fluoride content of the raw water. Mt. Pleasant's water supply contains a relatively high fluoride level, 4.3 mg F<sup>-</sup>/l. Conway's wells yield a relatively low fluoride level, 2.2 mg F<sup>-</sup>/l.

The methods utilized, experimental results, and conditions are presented below.

#### EXPERIMENTAL PROCEDURES

Identical columns were employed for both tests. Each column had a diameter of 4.5 in. and was charged with sufficient activated alumina to provide a bed depth of 2 ft. The total bed volume for each column was 0.8836 cu ft.

Each column was prepared by partially filling the vessel with water and adding a predetermined quantity of activated alumina. The virgin column contents were neutralized by downflowing water with a pH of 4-5 units in the Culligan laboratory. The pH was controlled by the addition of muriatic acid. When the pH of the column effluent reached the level of 8.0, the columns were transported to the test site. The time required to reach an effluent pH of 8.0 was 45-90 min equivalent to 87-175 Bed Volumes (BV) at a flow of 1.72 gpm.

The column was operated on-site at a pH of 5.5 for a period of 30 min at the test flow. At this time, the initial column operating conditions were established. Each column, which had a detention time of 5 min, was operated until fluoride breakthrough occurred. Throughout the run flow, influent fluoride, effluent fluoride and pH data was collected. Analytical procedures utilized through the course of study conformed to Standard Methods for the Examination of Water and Wastewater, 14th Edition, 1975.

After column exhaustion, the unit was transported to the Culligan laboratory for regeneration and neutralization. The column was first backwashed in the upflow mode for a period of 30 min at a flow of 10 gpm, equivalent to approximately 23 gpm/sq ft. The backwash rate forced the media out of the column into a device which allowed collection and continued backwash of the media.

The column was recharged with the backwashed media and regenerated with approximately 1 percent caustic. Caustic regeneration was conducted for a period of 30 min at a flow of 1.72 gpm. This flow is equivalent to a loading rate of 3.9 gpm/sq ft.

The column was again backwashed (rinsed) for a period of 30 min after regeneration. The backwash flow was 5 gpm, equivalent to a loading rate of 11.3 gpm/sq ft. Following backwash the column was neutralized in the downflow mode utilizing raw water with a pH of 4-5 units using the same procedure as outlined previously.

#### EXPERIMENTAL RESULTS

Experimental operating data is summarized in Tables 3-1 through 3-4 for each of the four experimental pilot runs. These data have been plotted to determine capacities for the two waters (see Figures 3-1 and 3-2). The test results indicate that approximately 10,000 gal of Mt. Pleasant raw water was treated without exceeding the MCL for fluoride utilizing 1 cu ft of media. The fluoride exchange capacity at Mt. Pleasant was approximately 0.424 lb F<sup>-</sup>/cu ft of activated alumina.

The Conway data indicates that 8,500 gal of raw water was treated successfully with 1 cu ft of media. The fluoride exchange capacity at Conway was 0.18 lb F<sup>-</sup>/cu ft of activated alumina.

TABLE 3-1  
MT. PLEASANT FLUORIDE REMOVAL PILOT TEST  
RUN 1 DATA SUMMARY

Cumulative Flow (gal)	(BV)	Fluoride Concentration		initial	pH final
		initial mg/l	final mg/l		
0		4.4	0.82	5.5	9.1
1,740	263	4.2	0.15	7.0	7.5
2,250	340	4.3	0.12	6.5	7.0
3,620	548	4.3	0.43	4.5	4.8
4,220	638	4.2	3.8	5.0	4.9

TABLE 3-2  
MT. PLEASANT FLUORIDE REMOVAL PILOT TEST  
RUN 2 DATA SUMMARY

Cumulative Flow (gal)	(BV)	Fluoride Concentration		initial	pH final
		initial mg/l	final mg/l		
0	0	4.3	0.95	5.5	9.5
570	86	4.4	0.24	5.7	5.4
1,920	290	4.3	0.23	5.5	5.3
2,340	354	4.3	0.23	5.7	5.3
3,840	581	4.3	0.25	5.9	5.8
4,490	679	4.2	0.28	5.5	5.2
5,700	862	4.2	0.28	5.8	5.6
6,230	943	4.3	0.35	5.9	5.8
7,660	1,159	4.2	0.32	6.5	6.4
8,070	1,221	4.2	0.48	5.7	5.8
9,070	1,372	4.3	2.4	5.8	5.7

TABLE 3-3  
CONWAY FLUORIDE REMOVAL PILOT TEST  
RUN 1 DATA SUMMARY

Cumulative Flow		Fluoride Concentration		pH	
(gal)	(BV)	initial mg/l	final mg/l	initial	final
0	0	2.6	0.85	5.5	9.3
1,540	233	2.2	0.15	6.0	7.9
1,670	253	2.3	0.10	4.9	6.5
3,680	557	2.0	0.11	5.5	6.0
5,380	814	2.2	0.10	6.0	5.8
5,540	838	2.1	0.20	5.5	5.7
7,260	1,098	2.3	0.91	5.5	5.6
7,420	1,123	2.3	1.40	5.5	5.7

TABLE 3-4  
CONWAY FLUORIDE REMOVAL PILOT TEST  
RUN 2 DATA SUMMARY

Cumulative Flow		Fluoride Concentration		pH	
(gal)	(BV)	initial mg/l	final mg/l	initial	final
0	0	2.3	0.43	5.5	8.3
1,390	210	2.2	0.24	5.5	5.2
1,630	247	2.0	0.25	5.5	5.1
3,450	522	2.1	0.17	5.4	5.8
3,620	548	2.2	0.24	5.6	5.7
5,290	800	2.2	0.91	7.3	7.1
5,650	855	2.2	0.60	5.6	5.8
6,920	1,047	2.2	1.80	5.7	5.7

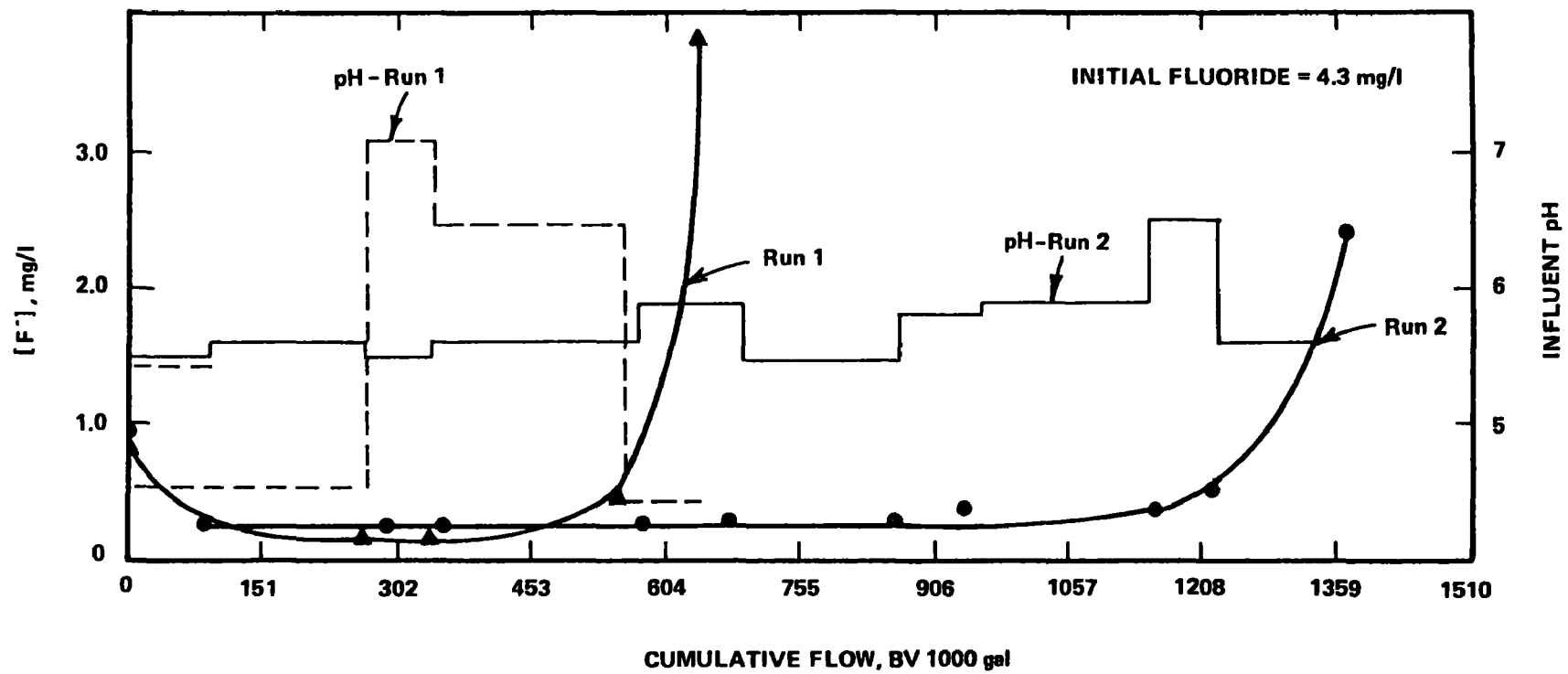


FIG. 3-1 . MT. PLEASANT PILOT PLANT RUN

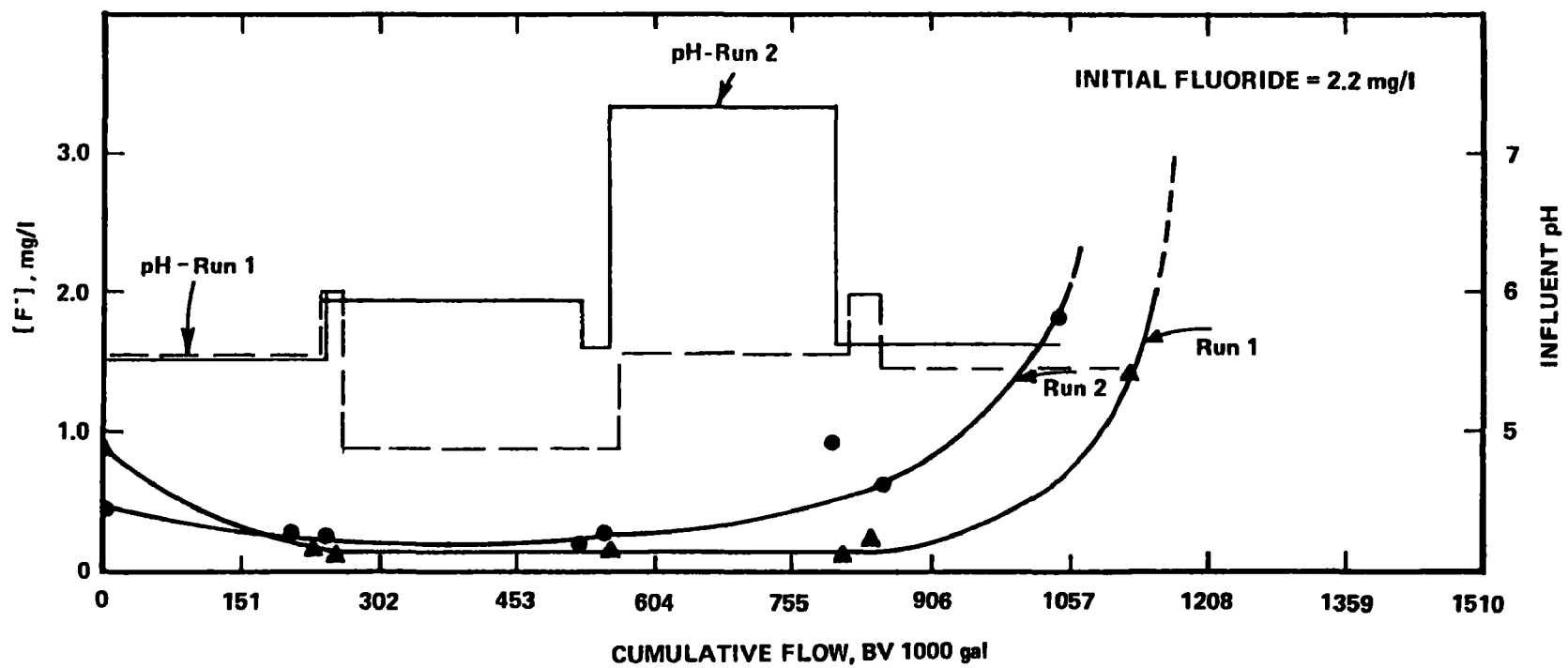


FIG. 3-2 . CONWAY PILOT RUN

## DISCUSSION OF RESULTS

The data indicate that influent pH control is of primary importance in the operation of the alumina column. AWARE had difficulty controlling pH during the initial runs at both Mt. Pleasant and Conway. The systems were equipped with a pulse type acid feed pump. During the initial runs pH was monitored utilizing grab samples. The pH was adjusted based on the grab sample analysis. Due to the pulsing of acid, this mode of operation was discontinued during the second run. During run 2 pH was monitored on a small composite sample. As a result of this modification, the run time was significantly greater at the Mt. Pleasant site during the second run. The Conway experimental data is somewhat erroneous. During the second run, acid feed was discontinued for a period of less than 24 hr. It is felt that this loss of acid feed was responsible for not obtaining the anticipated throughput of 10,000 gal/cu ft.



## CHAPTER 4

### BASIS OF DESIGN AND TREATMENT COSTS FOR ACTIVATED ALUMINA

Activated alumina fluoride removal systems were designed for systems in which this technology might be feasible. The process design summaries for a 100, 250, 500, and 1,000 gpm system are presented in Tables 4-1, 4-2, 4-3, and 4-4, respectively. The basis for design and associated capital and O&M costs for each system are described below.

#### BASIS OF DESIGN

Activated alumina column beds were designed based on maintenance of a 5 min bed detention time, 12 BV/hr. The detention time selected was based on the experience of operating activated alumina fluoride removal facilities and pilot studies conducted as part of this project. The column surface area was based on a hydraulic loading rate of 9 gpm/sq ft utilizing the design flowrate. The hydraulic loading rate of 9 gpm/sq ft was selected based on the flow required to expand the media bed during backwash. The column length was designed based on a 60 percent bed expansion volume.

Acid feed equipment would be required to adjust the column feed pH and adjust the alkaline byproducts generated during the regeneration and neutralization operating modes. No information on acid requirements was available during the investigation period. Experience of other facilities indicated that sulfuric acid usage was equivalent to 1 tank truck load (approximately 3,400 gal) per year for a 100 gpm plant operated 16 hr/day.

TABLE 4-2  
PROCESS DESIGN SUMMARY  
250 gpm System

Bed Depth in Single Column, ft	6.0
Length of Single Column, ft	9.6
Column Surface Area, sq ft	28.27
Initial Activated Alumina Inventory	
, cu ft	169.62
, lb	9,329.10
Time Between Regenerations, days	7.07
Operating Mode	Downflow
Flow Rate, BV/min	0.197
, gpm	250
, gpm/sq ft	8.84
Backwash Mode	Upflow
Backwash Rate, BV/min	0.197
, gpm	250
, gpm/sq ft	8.84
Backwash Duration, min	10
Backwash Volume, gal	2,500
Regeneration Mode	Upflow
Regeneration Rate, BV/min	0.06675
, gpm	84.7
, gpm/sq ft	3.0
Regeneration Duration, min	35
Regeneration Volume, gal	2,541
Rinse Mode	Upflow
Rinse Rate, BV/min	0.1335
, gpm	169.4
, gpm/sq ft	6.0
Rinse Duration, min	5.0
Rinse Volume, gal	847
Neutralization Mode	Downflow
Neutralization Rate, BV/min	0.197
, gpm	250
, gpm/sq ft	8.84
Neutralization Duration	240
Neutralization Volume	60,000

TABLE 4-3  
PROCESS DESIGN SUMMARY  
500 gpm System

Bed Depth in Single Column, ft	6.0
Length of Single Column, ft	9.6
Column Surface Area, sq ft	56.75
Initial Activated Alumina Inventory	
, cu ft	340.5
, lb	18,727.5
Time Between Regenerations, days	7.09
Operating Mode	Downflow
Flow Rate, BV/min	0.196
, gpm	500
, gpm/sq ft	8.81
Backwash Mode	Upflow
Backwash Rate, BV/min	0.196
, gpm	500
, gpm/sq ft	8.81
Backwash Duration, min	10
Backwash Volume, gal	5,000
Regeneration Mode	Upflow
Regeneration Rate, BV/min	0.06675
, gpm	170
, gpm/sq ft	3.0
Regeneration Duration, min	35
Regeneration Volume, gal	5,950
Rinse Mode	Upflow
Rinse Rate, BV/min	0.1335
, gpm	340
, gpm/sq ft	6.0
Rinse Duration, min	5.0
Rinse Volume, gal	1,700
Neutralization Mode	Downflow
Neutralization Rate, BV/min	0.196
, gpm	500
, gpm/sq ft	8.81
Neutralization Duration	240
Neutralization Volume	120,000

TABLE 4-4  
PROCESS DESIGN SUMMARY  
1,000gpm System

Bed Depth in Single Column, ft	6.0
Length of Single Column, ft	9.6
Column Surface Area, sq ft	113.5
Initial Activated Alumina Inventory	
, cu ft	681
, lb	37,455
Time Between Regenerations, days	7.09
Operating Mode	Downflow
Flow Rate, BV/min	0.196
, gpm	1,000
, gpm/sq ft	8.81
Backwash Mode	Upflow
Backwash Rate, BV/min	0.196
, gpm	1,000
, gpm/sq ft	8.81
Backwash Duration, min	10
Backwash Volume, gal	10,000
Regeneration Mode	Upflow
Regeneration Rate, BV/min	0.06675
, gpm	340
, gpm/sq ft	3.0
Regeneration Duration, min	35
Regeneration Volume, gal	11,900
Rinse Mode	Upflow
Rinse Rate, BV/min	0.1335
, gpm	680
, gpm/sq ft	6.0
Rinse Duration, min	5.0
Rinse Volume, gal	3,400
Neutralization Mode	Downflow
Neutralization Rate, BV/min	0.196
, gpm	1,000
, gpm/sq ft	8.81
Neutralization Duration	240
Neutralization Volume	240,000

The reviewed facilities did not require acid neutralization of spent regenerant. The acid holding tanks were sized based on the volume occupied by 40,000 lb of 92 to 96 percent  $H_2SO_4$  (approximately 3,500 gal).

Caustic storage was based on the use of 50 percent NaOH shipped in drums for the 100 gpm and 250 gpm systems. The vessels were sized based on the feed dilution of 1 percent caustic with sufficient volume for one regeneration cycle. The caustic storage tanks for the 500 gpm and 1,000 gpm systems were based on bulk delivery of 50 percent caustic. Caustic would be added in-line for dilution to 1 percent. The size of the caustic storage tank was based on the volume of 1 tank truck of 50 percent caustic, i.e., 4,000 gal. The volume is sufficient for 26 wk of 500 gpm system operation and 13 wk of 1,000 gpm system operation.

Raw water storage will be required for those systems which do not have sufficient storage to maintain pressure during bed regeneration. A survey of the systems which do not have sufficient storage indicated that the maximum system flow was 300 gpm. A raw water storage tank was sized and costs were estimated based on a 250 gpm system. The storage tank volume is 70,000 gal. The developed costs are included in the total treatment costs for those systems without storage.

A brine storage tank may be required at each site. The storage tank has three basic functions:

- Batch neutralization of the alkaline regenerant.
- Equalization for sewered systems.
- Holding vessel for contract disposal.

Costs were developed with and without brine storage. Brine storage may not be required in some of the larger sewered communities.

The brine storage tank volume was based on the anticipated quantity of regenerant, 52 BV.

Pump capacities were sized based on the required flows to backwash, regenerate, rinse and neutralize the bed. Those facilities with adequate water storage will utilize the well water pump to operate, backwash, regenerate, rinse, and neutralize the bed. This mode will require the use of a control valve and recycle loop to control flow.

Those systems which do not have adequate raw water storage will be supplied with a groundwater storage tank equipped with a variable speed pump to backwash, regenerate, rinse and neutralize the bed.

Operation of the facility requires strict pH control. Acid addition for maintenance of a feedwater pH of 5.5 will require pH controlled acid addition. Acid addition will be required to neutralize the bed. The neutralization acid feed system will be isolated to allow the continued raw water treatment in the second column. Acid addition will also be required for regenerant pH adjustment prior to disposal. The regenerant neutralization system will utilize the bed neutralization acid feed system on a batch basis.

The 100 and 250 gpm systems will not require in-line dilution for caustic regeneration. Caustic feed pumps have been based on a flow of 0.67 BV/min.

The 500 gpm system and 1,000 gpm system will utilize 50 percent caustic diluted in-line to 1 percent followed by a static mixer. The caustic feed pump sizes are based on flows required for dilution of 50 percent caustic to 1 percent. A summary of the basic equipment sizes and flowrates are presented in Table 4-5.

TABLE 4-5  
BASIC COLUMN INFORMATION

	100	250	System	500	1,000
Media Volume, cu ft	71.76	169.62		340.5	681
Column Surface Area, sq ft	11.04	28.27		56.75	113.5
Media Depth, ft	6.5	6.0		6.0	6.0
Column Height, ft	10.4	9.6		9.6	9.6
Backwash Flow, gpm	100	250		500	1,000
Regeneration Flow, gpm	40	85		170	340
Rinse Flow, gpm	72	170		340	680
Neutralization Flow, gpm	100	250		500	1,000
Brine Volume, Gal	28,000	66,000		133,000	265,000
Backwash Feed Tank, Gal	NA	70,000		NA	NA
Acid Storage Tank, Gal	5,000	5,000		5,000	5,000
Caustic Storage Tank, Gal	1,500	3,000		4,000	4,000

## ACTIVATED ALUMINA - COST ESTIMATES

Cost estimates were developed for 100, 250, 500, and 1,000 gpm activated alumina fluoride removal system models. Three systems were designed for each system capacity. The first system is a basic column design. The second system includes a brine storage or raw water storage vessel. The third system includes both raw water and brine storage facilities. The cost curves developed for the model systems are presented in Figure 4-1.

Capital costs were developed for each individual system based on the required system to achieve a 1.6 mg F<sup>-</sup>/l blended water supply. Annual capital costs are based on a 30 year amortization at an interest rate of 12 percent.

Operating costs are based on labor and chemical costs. Additional energy costs were assumed to be minimal. The anticipated labor requirements were based on the following:

- 100 gpm System            3 M-H per day
- 250 gpm System           5 M-H per day
- 500 gpm System           6 M-H per day
- 600 gpm System           8 M-H per day

Labor costs were assumed to be \$10/M-H. Chemical costs were based on the anticipated caustic and acid requirements FOB, Columbia, South Carolina.

Annual maintenance costs were based on 3 percent of the total capital cost. The developed annual O&M cost curve is presented in Figure 4-2.



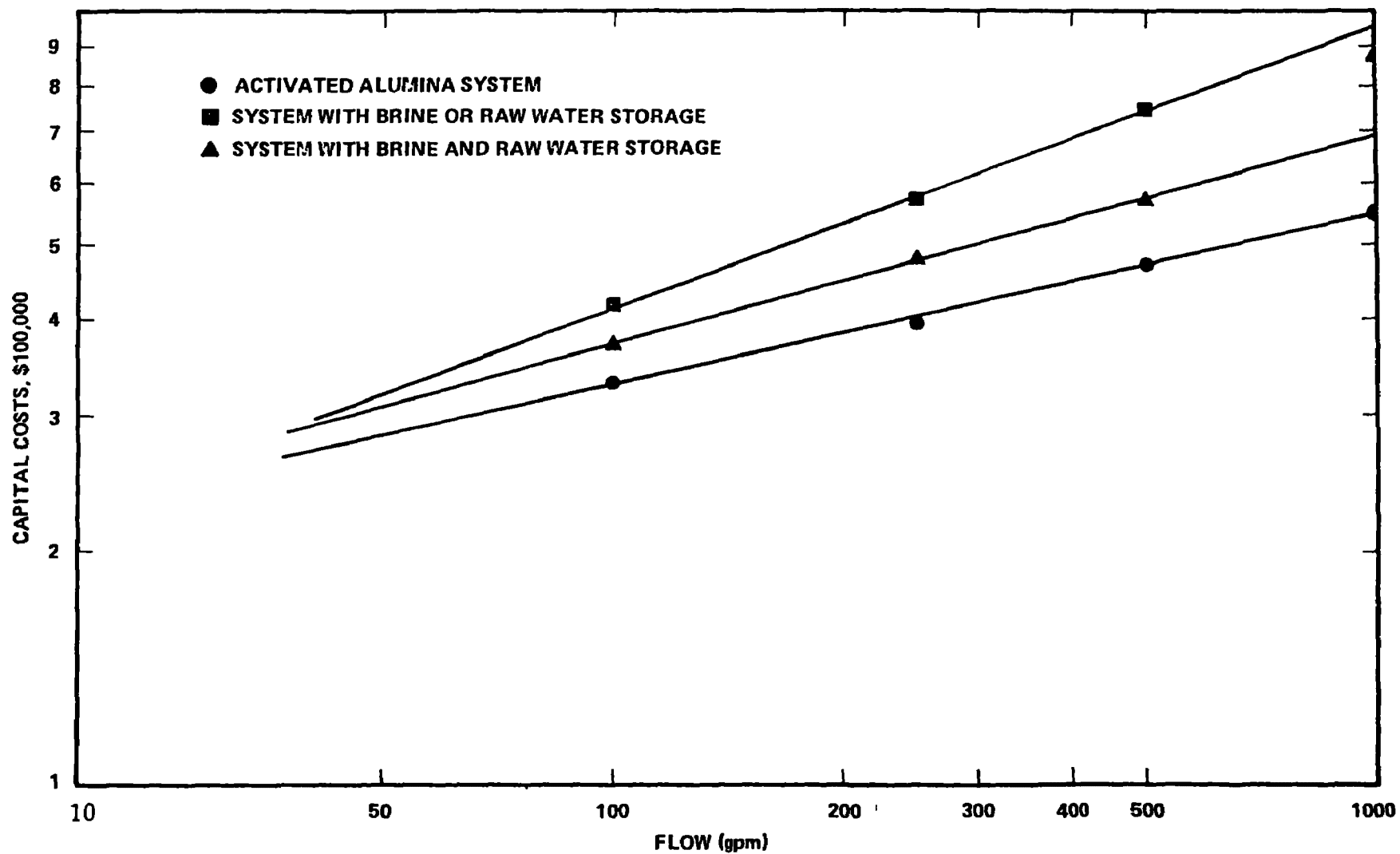


FIG. 4-1 . ACTIVATED ALUMINA SYSTEM – CAPITAL COST

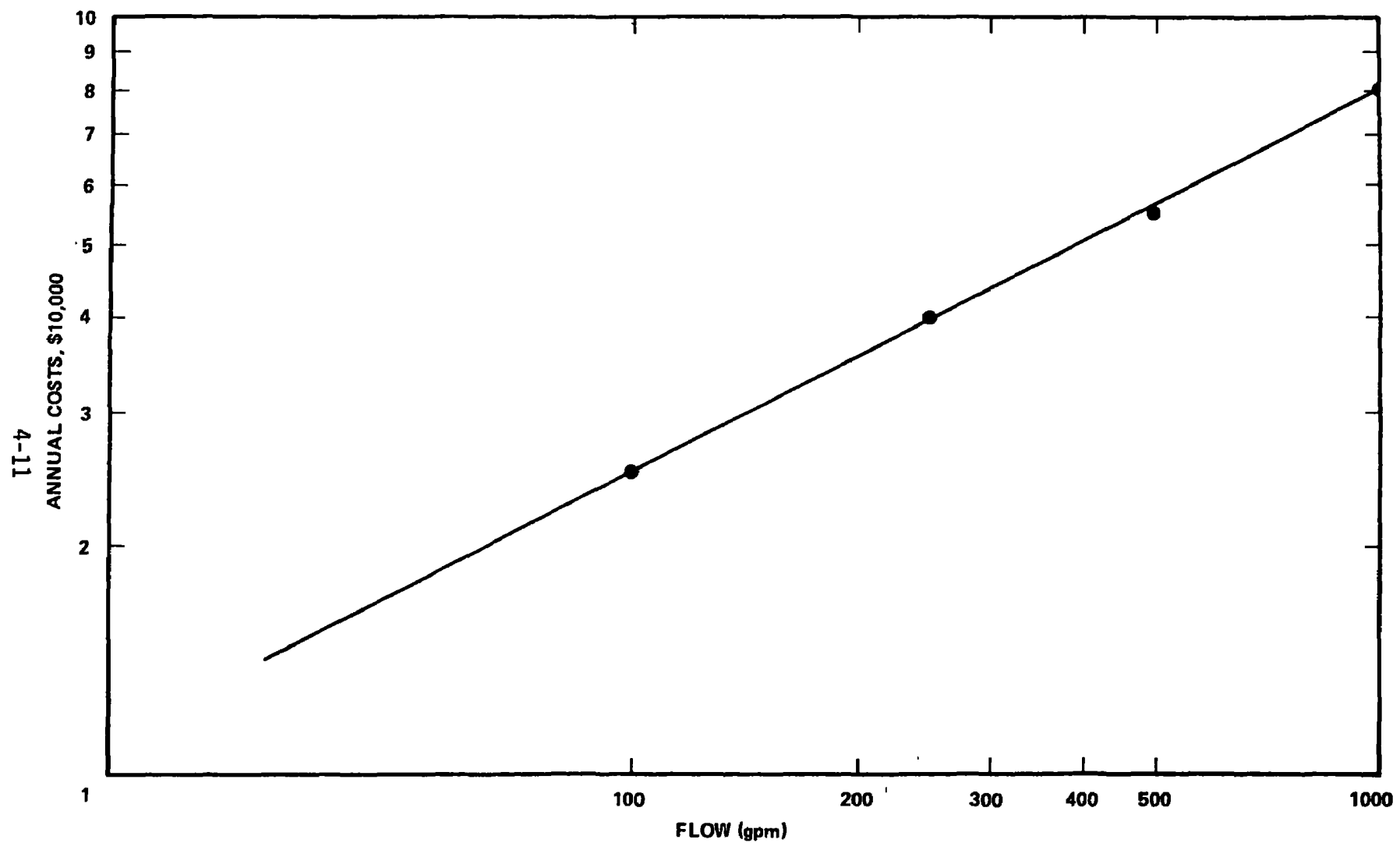


FIG. 4-2 . ACTIVATED ALUMINA SYSTEM – ANNUAL O & M COST

## PROCESS DESIGN - REGIONAL WATER TREATMENT FACILITY

A conventional surface water treatment facility was designed in order to develop capital and O&M costs. A process flow schematic is presented in Figure 4-3. The facility consisted of flash mixing, flocculation, sedimentation and filtration. Ancillary facilities consisted of gravity thickening and filter presses for sludge dewatering. A settling pond is also provided for sand filter backwash. A process design summary is presented in Table 4-6.

The estimated total capital cost for the treatment plant is \$26,250,000.

The estimated plant operating cost is \$509,000 per year. These costs include energy, chemical, and labor costs. The estimated maintenance cost for the facility is \$675,000 per year.

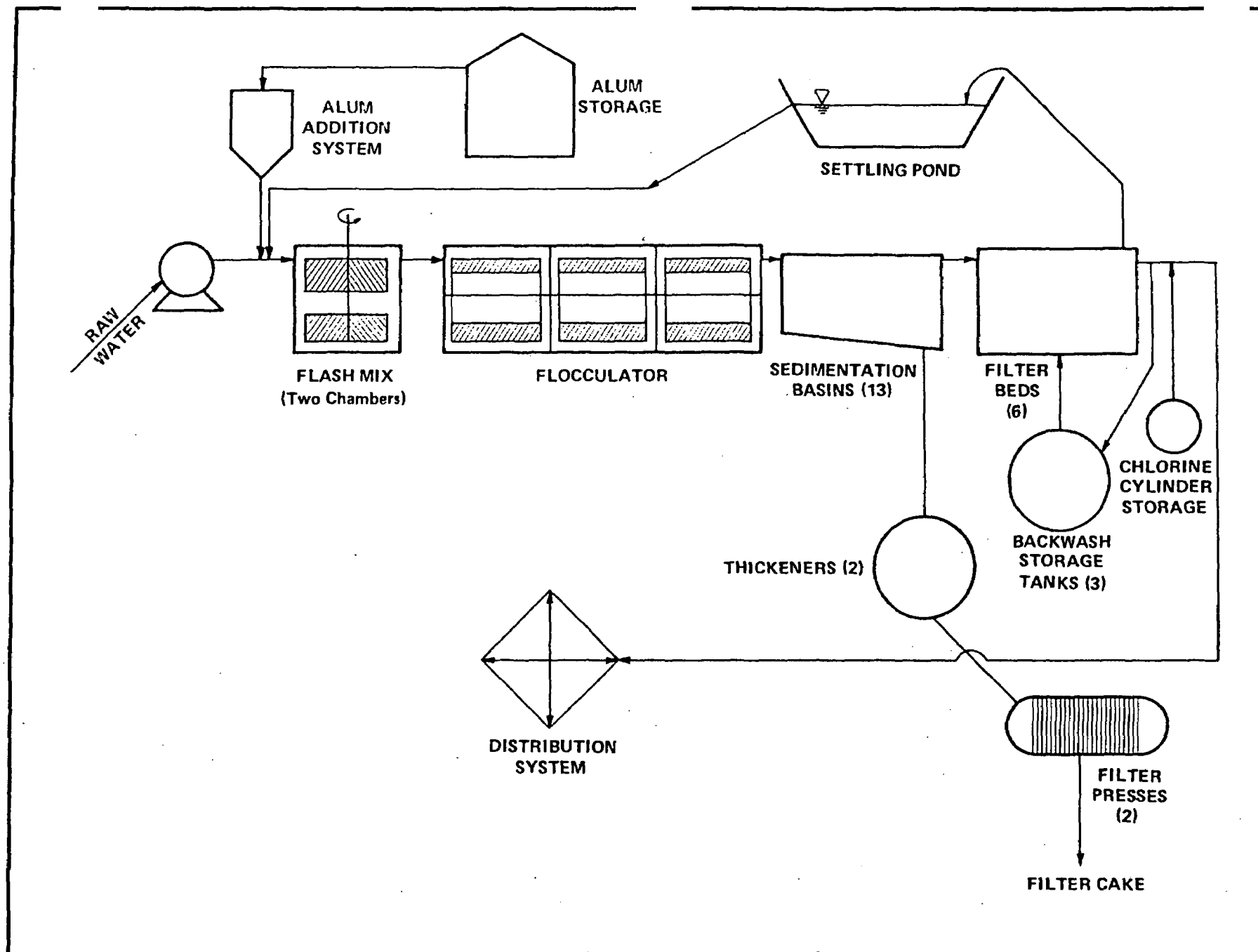


FIG. 4-3 . PROCESS FLOW SCHEMATIC – PEE-DEE RIVER REGIONAL FACILITY

TABLE 4-6  
PROCESS DESIGN SUMMARY

<u>Flash Mix Basins</u>	
Detention Time, sec	20
Number of Basins	2
Basin Volume, gal x 10 <sup>3</sup>	
Basin 1	5.4
Basin 2	5.4
Basin Dimensions, ea.	
Length, ft	8.5
Width, ft	8.5
Depth, ft	10
SWD, ft	12
Power Requirement, ea	
Basin 1, HP on line	40
Basin 2, HP on line	40
<u>Flocculation Basins</u>	
Number of Basins	4
Detention Time Total, min	40
Basin Volume, gal x 10 <sup>3</sup>	
Compartment 1	224
Compartment 2	224
Compartment 3	224
Basin Dimensions, ea	
Length	100
Width	30
Depth	10
SWD	12
Power Requirement, ea	
Compartment 1, HP on line	11
Compartment 2, HP on line	7
Compartment 3, HP on line	1
Paddle Area, sq ft	500

TABLE 4-6 (Cont.)  
PROCESS DESIGN SUMMARY

<u>Paddle Velocity</u>	
Compartment 1, fps	2.54
Compartment 2, fps	1.72
Compartment 3, fps	0.93
<u>Revolutions per minute</u>	
Paddle Compartment 1	6
Paddle Compartment 2	4
Paddle Compartment 3	2.25
<u>Alum Dosage Supply</u>	
Alum, ppm	40-100
Alum Dose, lb/day x 10 <sup>3</sup> , maximum	37.5
Alum Supply, days	30
30-day supply, gal x 10 <sup>3</sup>	208.5
Alum Storage Tanks	2
Volume, ea, gal x 10 <sup>3</sup>	108
Tank Dimensions	
height, ft	15
diameter, ft	35
<u>Sedimentation Basins</u>	
Surface Overflow Rate, gpm/sq ft	0.25
Detention Time, hr	5
Basin Volume Total, gal x 10 <sup>3</sup>	9.724
Number of Basins	13
Basin Dimensions	
Length, ft	200
Width, ft	50
Depth, ft	10
SWD, ft	12
Flow per basin, gpm	2,400
Flow Velocity, fpm	0.6
Weir Overflow Rate, gpm/ft	8
Weir Length Total, ft	3,900
Weir Length, per Basin, ft	300
<u>Thickener</u>	
Mass Dry Solids, lbs/mg	490
Volume Underflow Sludge, gal/mg	2,930
Thickener Surface Area, ft <sup>2</sup>	4,400
Number of Thickeners	2
Detention Time, days	3.9

TABLE 4-6 (Cont.)  
PROCESS DESIGN SUMMARY

<hr/>		
Thickener Dimensions		
Height, ft		15
Diameter, ft		55
Sludge Removal Rate, gal x 10 <sup>3</sup>		75
Sludge Mass Removed, lb/day x 10 <sup>3</sup>		21.9
 <u>Filter Beds</u>		
Filter Bed Components		
Coarse material	Anthracite	
Fine material	Sand	
Filter Bed Area Total, ft <sup>2</sup> x 10 <sup>3</sup>		12.6
Number of Filter Beds,		6
Surface Area, ft <sup>2</sup>		2,100
Filter Run Time		
Accepted Contemporary Maximum, hrs		60
Anticipated, hrs		48
Solids Accumulation Rate, lb/hr		6.6
Backwash Flow Rate, gpm	31,500	
Backwash Flow, gpm/ft <sup>2</sup>		15
Backwash Time, min	5-10	
Backwash Holding Tank		
Volume, ea, gal x 10 <sup>3</sup>		423
Volume, Total, gal x 10 <sup>3</sup>		1,269
Number of Tanks		3
Tank Dimensions		
height, ft		20
diameter, ft		60
 <u>Settling Pond</u>		
Number of Ponds		1
Volume for Settling, gal x 10 <sup>3</sup>	1,683	
Pond Dimensions		
Length, ft		300
Width, ft		75
Depth, ft		10
Detention Time, days		1.8

TABLE 4-6 (Cont.)  
PROCESS DESIGN SUMMARY

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Filter Press

Condition Solids, lb Lime/lb sludge solids	0.15
Coagulant Condition, lb/ton dry SS	5
Cake Solids, %	25
Cake Density, lb/ft <sup>3</sup>	70
Cake Thickness, in.	1.2
Filter Chamber Volume, ft <sup>3</sup>	1.88
Cycle Time, hr	2
Filter Pressure, psig	225
Dry Weight Solids, lb x 10 <sup>3</sup> /day	25.2
Cake Solids, lb/10 <sup>3</sup> /day	100.75
Filter Cake Volume, ft <sup>3</sup>	1,450
Number of Chambers/day	775
Cycles per Day	8
No. of Presses	2
No. of Chambers, ea press	50

Chlorination

Chlorine Dose, ppm <sup>-</sup>	3
Contact Basin	
Volume, gal x 10 <sup>3</sup>	945
Detention Time, min	30

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