

Dow  
Liquid Separations



# **FILMTEC**

## **Membranes**

### **An Economic Comparison of Demineralization with Reverse Osmosis and Ion Exchange Technology**

**Power-Gen Americas '94  
December 7-9, 1994**

by: Steven D. Coker - The Dow Chemical Company  
Scott S. Beardsley - The Dow Chemical Company  
Sharon S. Whipple - The Dow Chemical Company

# An Economic Comparison of Demineralization with Reverse Osmosis and Ion Exchange Technology

## Introduction

In this paper we will consider in detail the factors impacting the decision to retrofit reverse osmosis (RO) ahead of an existing ion exchange (IX) system. The break-even point in total dissolved solids above which it is more economical to use one of these technologies over the others depends on a number of factors which will be addressed in this paper. The economic factors affecting the break-even point and project payback time include chemicals, resins, membranes, energy, operating labor, maintenance, and capital-related items (1-9).

In some instances technical considerations will outweigh economics in evaluating the possibility of retrofitting RO in front of an existing IX system. For example, retrofitting RO may insure compliance with environmental regulations which require a reduction in the volume of regenerant waste from IX. Or if the water source has a varying dissolved solids level, retrofitting RO will greatly minimize the impact on product water quality.

It should be noted that The Dow Chemical Company markets both reverse osmosis elements and ion exchange resins. It is our intention to provide an objective study utilizing conservative economic analysis without bias for one technology over another. It is also our intention to show the impact of the latest advances in membrane and resin technology on the total cost to produce water. This paper is a follow-up to the one written in 1987 (6) and is a companion paper to the one recently presented at WATERTECH '94 (9) on the economics of reverse osmosis and ion exchange in new systems.

In recent years the cost of chemicals and energy has not changed significantly. Higher active surface area elements (400 active square feet) and higher rejection of salt using

FILMTEC® FT30 thin film composite membranes have lowered the total cost to produce water from a RO system. The cost of both IX resin and RO elements are lower than in previous economic models which reduces their periodic replacement costs.

## Design Basis

The assumptions used in this economic evaluation are listed in Tables A, C and D. The water treatment systems were sized to produce three different quantities of mixed-bed quality water. The flow rates were two hundred fifty thousand (250 Mgalpd), five hundred thousand (500 Mgalpd), and one million gallons per day (1000 Mgalpd). Identical storage facilities for product water were assumed for each product water flow rate and for all cases studied. One train was used for the 250 Mgalpd system size, two trains for the 500 Mgalpd size, and four trains were used for the 1000 Mgalpd size.

Four feed water qualities, varying only in the quantity of total dissolved solids (TDS) were utilized in the study. The TDS levels were 80, 160, 320, and 480 ppm (as CaCO<sub>3</sub>). The quality of the feed water can vary significantly depending on the geographic location, and can affect any system design as well as the need for pretreatment systems, especially where reverse osmosis is contemplated. Surface water sources typically require more pretreatment while ground water sources typically need less. The feed waters used in this study have a high hardness ratio, high alkalinity, and no problems with organics, colloidal particles, or turbidity. Capital was included in the study for the additional pretreatment required by the RO system.

A raw water inlet temperature was assumed to be 55°F, a national average which may vary depending on the geographic location. Unlike

**Table A**

**Bases and Assumptions for Cost Analysis  
Water Analysis, ppm as Calcium Carbonate**

**Case 1:**

Ca	31.0	HCO <sub>3</sub>	55.5	SiO <sub>2</sub> (as SiO <sub>2</sub> )	5.0
Mg	32.5	SO <sub>4</sub>	11.8	Temperature	55°F
Na	15.8	Cl	10.5	pH	7.6
		NO <sub>3</sub>	1.5		
TDS	79.3	TDS	79.3		

**Case 2:**

Case 1 x 2

TDS

160

**Case 3:**

Case 1 x 4

TDS

320

**Case 4:**

Case 1 x 6

TDS

480

**Costs:**

Energy	\$0.05/KWH
Steam	\$1.75/1,000 lbs
Caustic Soda	\$0.16/lb
Sulfuric Acid	\$0.038/lb (100% basis)
Scale Inhibitor	\$1.45/lb
Lime	\$0.02/lb
Feed Water	\$0.05/1,000 gallons
Waste Disposal	\$0.05/1,000 gallons

**Depreciation of Capital:**

10 years, SL

**System Sizes:**

250,000 gpd

500,000 gpd

1,000,000 gpd

360 days/year

**System Operating Rate:**

**Product Water Purity:**

Mixed-bed polished water	< 0.01 ppm sodium
	< 0.01 ppm silica
	> 10 megohm-cm

previous studies, we did not heat the feed water for this analysis since there are many systems that do not utilize preheating to decrease the feed pressure required for reverse osmosis systems.

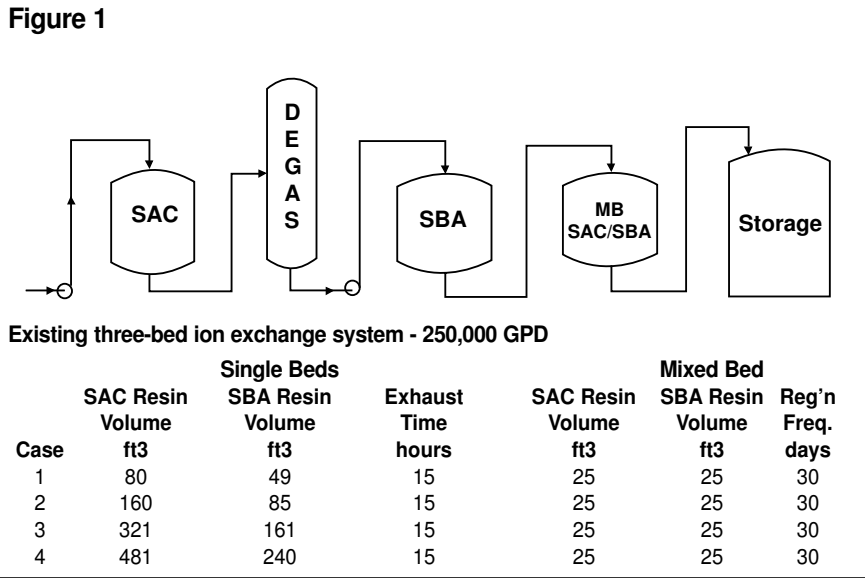
The costs of the most dominant operating factors, energy and caustic, were set at \$0.05/KWH and \$0.16/lb (100%), respectively. The caustic price reflects a high purity grade specification. The cost of feed water to the RO/IX or straight IX system and the cost of waste disposal have been considered at \$0.05/1000 gallons each. Labor and maintenance costs were also considered in the evaluation. Operating labor was considered minimal at one-eighth to one-quarter man per shift for the relatively continuous RO operations depending on system size. For straight IX, with more batch-type operations, the operating labor was doubled. Maintenance costs were set at 5% of equipment costs.

The initial direct fixed capital (DFC) costs were estimated by obtaining equipment cost estimates from two water treatment system manufacturers, based on defined system criteria provided by the authors. The estimates were then factored to represent a reasonable installed capital cost, which includes piping, instrumentation, and auxiliaries. It was assumed that the land and building were existing. The base estimates used in this study are listed in Table B for purchased, preassembled (not installed) equipment, and membranes. Capital for pretreatment is not included in these totals but was estimated separately. A 10-year straight-line depreciation based on total direct fixed capital, and taxes and insurance at 2% of DFC were assumed.

**Table B**  
**Purchased Equipment (Preassembled) Capital Estimates \***

	Case 1	Case 2	\$M Case 3	Case 4
<b>Feed TDS, ppm as CaCO<sub>3</sub></b>	80	160	320	480
<b>Straight ion exchange (IX)</b>				
250,000 gpd	\$0.24	\$0.29	\$0.33	\$0.40
500,000 gpd	\$0.45	\$0.53	\$0.62	\$0.74
1,000,000 gpd	\$0.83	\$1.00	\$1.16	\$1.38
<b>Reverse osmosis/ion exchange (RO/IX)</b>				
<b>Thin film composite</b>				
250,000 gpd	\$0.31	\$0.31	\$0.31	\$0.31
500,000 gpd	\$0.55	\$0.55	\$0.55	\$0.55
1,000,000 gpd	\$1.00	\$1.00	\$1.00	\$1.00

\*Estimates are the average of figures provided by Glegg Water Conditioning and U.S. Filter/Illinois Water Treatment



**Existing Three-Bed Ion Exchange System**

A three-bed ion exchange system utilizing a strong-acid gel cation bed, a vacuum degasifier, a strong-base gel anion bed, and a mixed-resin

polishing bed used in the design comparisons was assumed to be existing. All of the IX resins were assumed to have a gaussian particle size distribution. A flow diagram in Figure 1 depicts the ion exchange system utilized in this study.

Table C summarizes the bases and assumptions for the ion exchange computer projections and the subsequent cost analysis.

The degasifier was used to remove the carbon dioxide from the acidic cation effluent in order to reduce the quantity of anion resin and also the amount of caustic regenerant. Inclusion of the degasifier is logical due to the high level of alkalinity in the feed water.

In order to size the ion exchange demineralizer it is necessary to provide water for regeneration and rinse requirements as well as account for outages associated with regeneration cycles. Thus the average feed water flow was 209 gpm for each train which was designed to yield 250 Mgal. The exhaustion times of the cation and anion beds ranged from 20 to 21 hours when the resins were new. It was assumed that an operating capacity decline of 50% would occur during the life of the anion resins due resin degradation and organic fouling. In order to estimate an average cost to produce water by IX, an exhaustion time of 15 hours was utilized. The regeneration cycle was approximately 4 hours. In all cases the mixed-resin polishers in the three-bed ion exchange system were regenerated every 30 days rather than upon exhaustion.

The computer projections of the primary beds used co-current regeneration as this is the predominant regeneration scheme used in the United States. The proposed system utilized realistic regenerant levels and produced water low in sodium and silica with a resistivity of approximately 10 megohm-cm.

**Table C**

**Bases and Assumptions for Cost Analysis Existing Three-Bed Ion Exchange System**

<i>Operation Sequence</i>	<i>Specification</i>
<b>Pretreatment</b>	
Sand filters	
Carbon beds	
<b>Demineralized Water Train</b>	
Cation resin bed	Strong acid cation - gel
Degasifier	Removes carbon dioxide
Anion resin bed	Strong base anion - gel
Mixed resin bed	Strong-acid cation - gel Strong-base anion - gel
<b>Demineralized Water Storage</b>	
<b>Waste Neutralization</b>	
(Waste IX regenerants)	Neutralize to pH 7.0
250,000 gpd - One (1) train	209 gpm
500,000 gpd - Two (2) trains	417 gpm
1,000,000 gpd - Four (4) trains	834 gpm
Operating efficiency	85%
Regeneration	Co-current
Cation regenerant	H <sub>2</sub> SO <sub>4</sub>
Anion regenerant	NaOH, 120° F
Regenerations	As needed
Time	4 hours
Resin life	
Cation	5 years
Anion	4 years
Mixed Bed	6 years

Sulfuric acid was used to regenerate the cation resins. The primary cation beds were regenerated with 6 pounds per cubic foot of resin which was applied at concentrations of 2%, 4% and 8% acid in a stepwise fashion. The anion beds were regenerated with 5 pounds of high purity caustic soda per cubic foot of resin applied at a temperature of 120°F to maintain a low level of silica leakage. The elevated temperature regeneration sequence included pre-heat, regeneration, and slow rinse on the anion bed. The mixed bed resins were regenerated with 8 pounds of regenerant per cubic foot of resin.

## New Reverse Osmosis System/Existing Ion Exchange System

The new RO treatment system is shown schematically in Figure 2. Since a RO system is a continuous operation, the average inlet RO flow rate is 232 gpm per train with the outlet flow to the IX being 174 gpm or 250 Mgd of permeate. The feed flow rate increases to 463 gpm for the 500 Mgd and 926 gpm for the 1,000 Mgd systems. A 5-micron cartridge filter is required ahead of the RO system as a polishing filter. This RO system was assumed to be a retrofit ahead of and used in conjunction with the existing three-bed IX system described above.

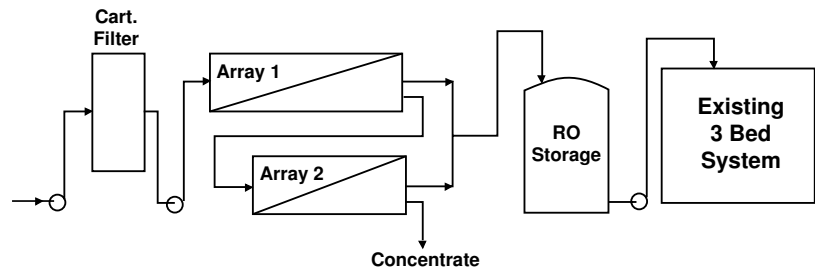
The thin film composite (TF) reverse osmosis section consisted of a 5-2 array utilizing 42 high surface area, spiral wound, low pressure RO elements to produce 250 Mgd (see Figure 2). The 500 Mgd and 1000 Mgd system sizes used twice and four times the number of RO elements and pressure vessels, respectively. Feed pressures of 241 to 248 psig were required for operation in the TDS range of 80 to 480 ppm as CaCO<sub>3</sub>. The addition of a high quality antiscalant ahead of each RO system was used to control the formation of calcium carbonate and calcium sulfate scale in all four cases. Acid addition was used only in Case 4 in order to keep the Langelier Saturation Index (LSI) of the RO concentrate below + 1.5.

A system recovery of 75% was used in all cases. Higher recovery levels are theoretically possible at lower TDS levels, however, in order to optimize the total cost to produce water recoveries of 70 to 80% are typically utilized.

Table D summarizes design parameters and assumptions that were used for the computer projections of the new RO followed by the existing three-bed IX system. These projections of product quality and flow rate

**Figure 2**

New reverse osmosis ahead of existing ion exchange system - 250,000 GPD



Case	Reverse Osmosis System				Feed Pressure (psig)
	Array 1 PV	Total Elements	Array 2 PV	Total Elements	
1	5	30	2	12	241
2	5	30	2	12	242
3	5	30	2	12	245
4	5	30	2	12	248

Above equipment plus necessary pretreatment equipment.

**Table D**

### Bases and Assumptions for Cost Analysis

#### Reverse Osmosis/Existing Ion Exchange System

##### Operation Sequence

##### Specification

##### Pretreatment

Flocculation clarifier  
Sand filters

##### Reverse Osmosis System

Pretreatment  
Acid addition  
Antiscalant addition  
5-Micron cartridge filter

##### Membranes

Type  
Thin film composite, spiral wound  
Life  
Three years  
Recovery  
75% in two stages  
Feed pressure  
248 psig (Thin film)  
Temperature  
55°F

##### Ion Exchange System

Cation resin bed  
Strong-acid cation - gel  
Degasifier  
Removes carbon dioxide  
Anion resin bed  
Strong-base anion - gel  
Mixed-bed  
Strong-acid cation - gel  
Strong-base anion - gel

##### Demineralized Water Storage

##### Waste Neutralization

Waste IX regenerants  
Neutralize to pH 7.0

##### Resin life

Cation  
8 years  
Anion  
6 years  
Mixed Bed  
6 years

were run using current reverse osmosis and ion exchange computer design programs.

### Discussion of Results

The results of this study are summarized in Figures 3 through 10 showing first the base cases and then the effect of caustic pricing and power pricing. We need to point out that the cost curves presented here only apply when using the set of assumptions as listed. The payback time period and break-even points are likely to change when significant changes in the assumptions occur. For example we assumed that the land and building already existed for the new RO system retrofit. If these needed to be purchased, then the payback period would be extended.

### Three-Bed Ion Exchange System

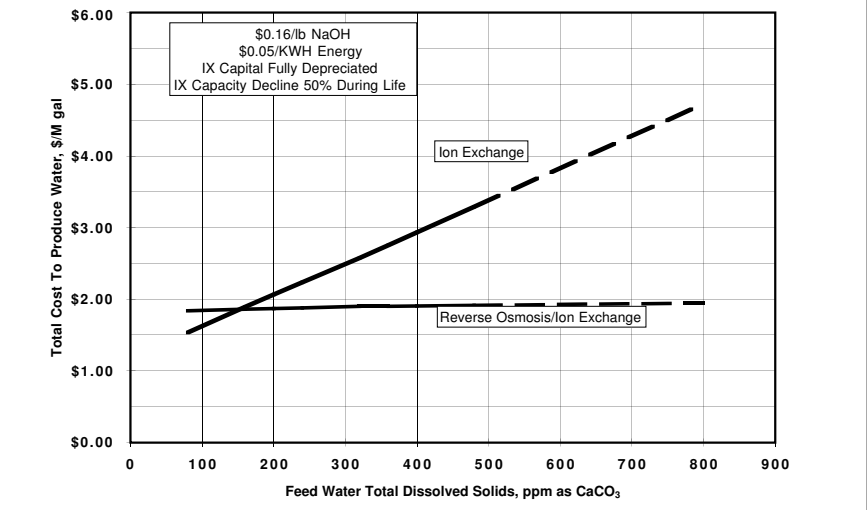
The base case results for the three system sizes are shown in Figures 3 through 5. As would be expected, the total cost to produce water for straight ion exchange increases with increasing feed TDS. For the 250 Mgalpd case the cost increases from \$1.50 at 80 ppm as CaCO<sub>3</sub> to \$3.35 per 1000 gallons of product water at 480 ppm as CaCO<sub>3</sub>.

The effect of increasing system size is to lower the total cost to produce water. The total cost to produce water decreases from \$1.50 to \$3.35 per 1000 gallons for the 250 Mgalpd to \$0.90 to \$2.50 per 1000 gallons for a system size of 1000 Mgalpd.

The total cost to produce water is lower than projected for new ion exchange equipment due to our assumption that capital is fully depreciated with respect to the ion exchange vessels, pretreatment equipment, neutralization and storage facilities. (9) Therefore the cost of depreciation is minimal for the IX system in the calculation of the total cost to produce water. However, since the equipment and resins are

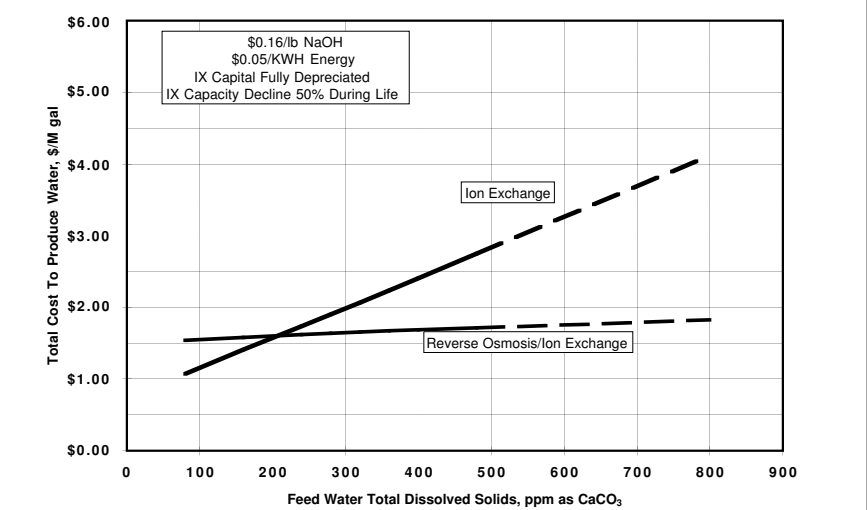
**Figure 3**

**Impact of Retrofitting Reverse Osmosis Ahead of Existing Ion Exchange - 250,000 gpd**



**Figure 4**

**Impact of Retrofitting Reverse Osmosis Ahead of Existing Ion Exchange - 500,000 gpd**



not new, the operating capacity of the IX system has been reduced to account for resin degradation and organic fouling of the anion resins. This has resulted in increased chemical costs with respect to regeneration and increased waste disposal costs.

### New Reverse Osmosis/Existing Ion Exchange System

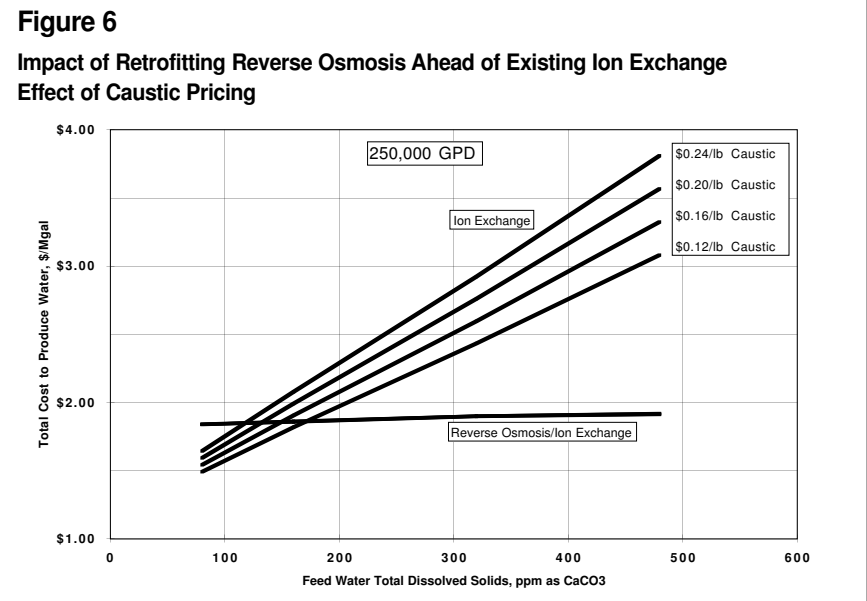
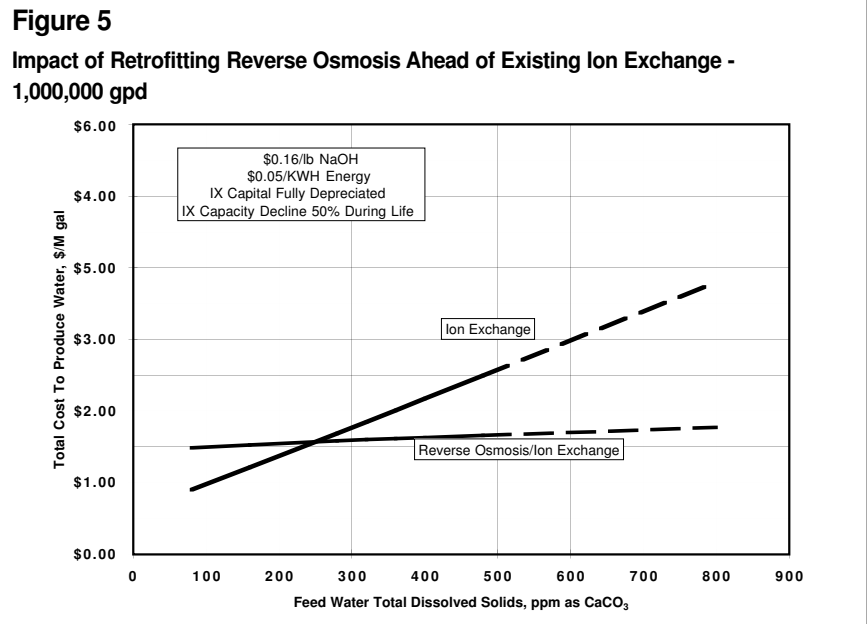
Figures 3 through 5 show the RO/IX system water production costs increase only slightly compared to the existing IX system which shows

a much greater increase in total cost to produce water with increasing feed water TDS. This is because the costs associated with the ion exchange system (caustic, sulfuric acid and resin replacement) are greatly reduced by retrofitting RO ahead of the existing IX system. Also, the capital costs associated with the new RO are relatively unaffected by increasing TDS.

The break-even point in TDS as CaCO<sub>3</sub> above which the total cost to produce water is more economical

using RO ahead of existing IX versus straight IX is 150 ppm for the 250 Mgpd system size. This break-even point moves up to 200 ppm at 500 Mgpd and to 250 ppm at 1000 Mgpd capacity. Above these break-even points it is possible to calculate a payback period for retrofitting RO ahead of IX.

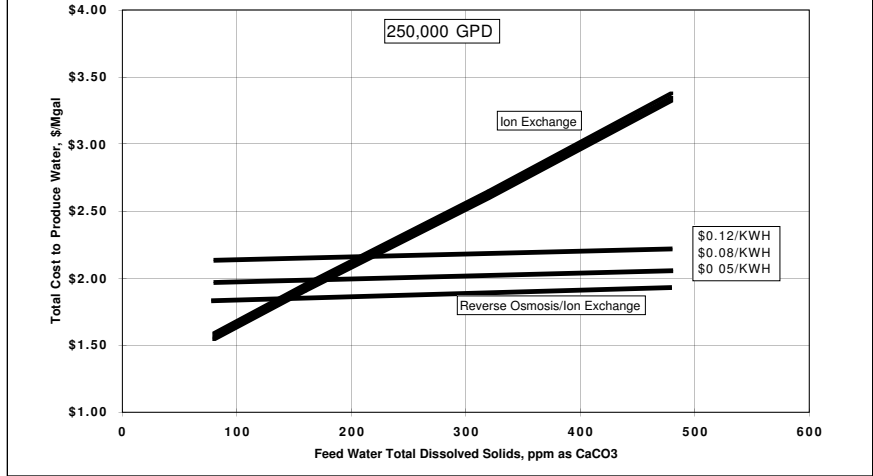
Figure 6 shows the impact of caustic pricing on the break-even point and total cost of water for a 250 Mgpd system. Increasing caustic pricing affects the IX system much more than the retrofit of RO ahead of IX when you look at the total cost to produce water. Also the break-even point moves from 120 ppm TDS at a caustic price of \$0.24/lb to 180 ppm TDS at a caustic price of \$0.12/lb which is significant considering the current trend of increasing caustic pricing.



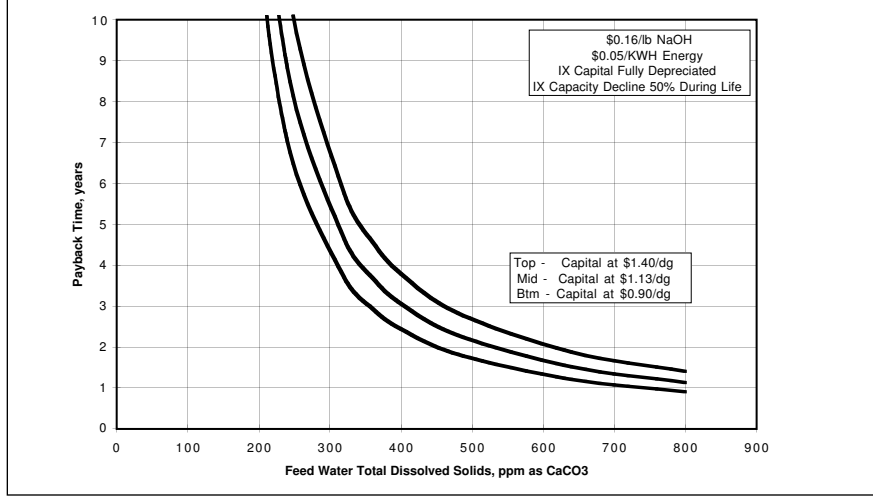
The effect of power costs is shown in Figure 7 for a 250 Mgalpd system. Here the increase in power costs affects the RO/IX system much more than the straight IX system. The break-even point moves from 150 ppm at \$0.05/KWH to 220 ppm at \$0.12/KWH.

The payback period in years for retrofitting RO ahead of an existing IX system is shown in Figures 8 through 10 for system sizes of 250 Mgalpd, 500 Mgalpd and 1000 Mgalpd. The base case is shown as the middle line and reflects the capital estimate average obtained from two system suppliers. The top and bottom lines show the impact on payback if the equipment cost increases or decreases by about 25%. If you assume that a 5 year payback is acceptable then you can justify retrofitting RO ahead of IX if your feed water TDS is in the range of 280 to 340 ppm for a system capacity of 250 Mgalpd. As the system capacity increases the TDS range for justifying a RO retrofit increases to 320 to 390 ppm for 500 Mgalpd capacity and to 360 to 430 ppm at 1000 Mgalpd capacity. This interpretation assumes that the IX capital is fully depreciated and that the IX operating capacity declines by 50% over the resin lifetime.

**Figure 7**  
Impact of Retrofitting Reverse Osmosis Ahead of Existing Ion Exchange  
Effect of Power Cost

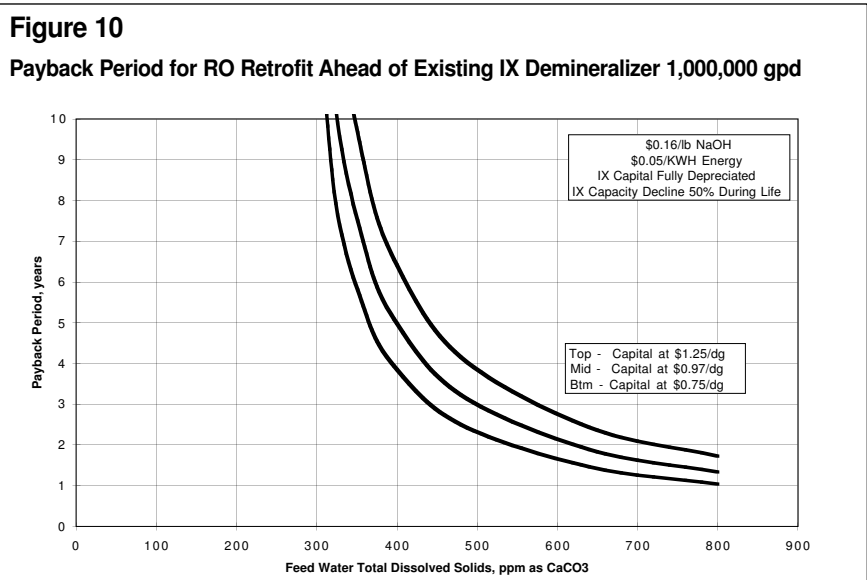
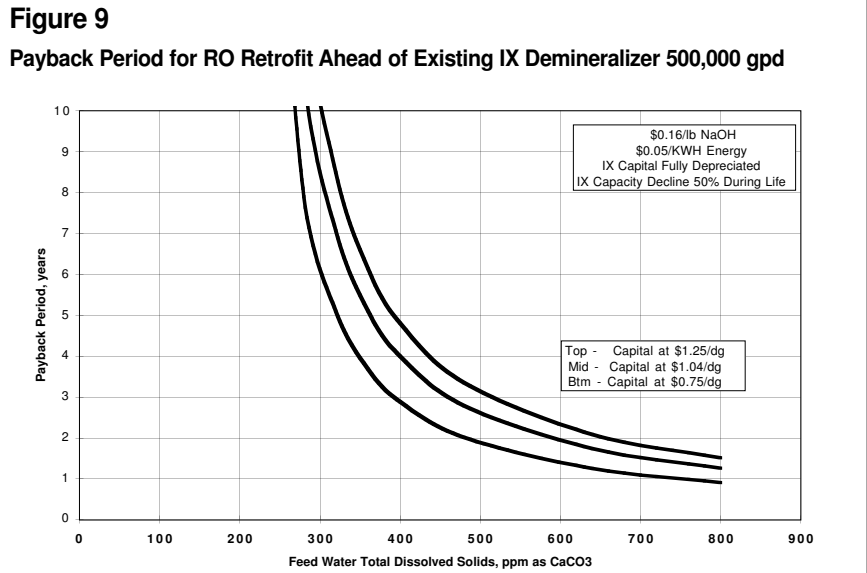


**Figure 8**  
Payback Period for RO Retrofit Ahead of Existing IX Demineralizer 250,000 gpd



**Conclusions**

1. The break-even point above which the total cost to produce water is more economical using a new RO ahead of an existing IX system rather than an existing IX system is about 150 ppm as CaCO<sub>3</sub>.
2. Increasing system size increases the break-even point from 150 ppm for a 250 Mgalpd system to 250 ppm for a 1000 Mgalpd system.
3. A retrofit of RO ahead of an existing IX system may be economically feasible above 320 ppm assuming an acceptable payback is 5 years.
4. The payback period at a given TDS level becomes longer as system size increases from 250 Mgalpd to 1000 Mgalpd.
5. Increasing caustic pricing favors the economics of retrofitting RO ahead of IX.
6. Increasing power pricing favors staying with existing IX rather than retrofitting RO ahead of IX.



---

## References

1. Lefevre, L., "Water Demineralization Using Reverse Osmosis and Ion Exchange," Technical Data Sheet Volume 8, Number 2, Dow Chemical USA, 1978.
2. Coulter, B. and Jones, G.D., "The Application of Reverse Osmosis to Mexican Waters," paper presented at the First Mexican Conference, Mexico City, Mexico, February 20-22, 1980.
3. Printz, J. and Wainwright, R., "Comparing Ion Exchange and Reverse Osmosis in the Electric Utility Industry," paper presented at the American Power Conference, Chicago, IL, April 26-28, 1982.
4. Little, D. and Lefevre, L., "The Economics of Reverse Osmosis and Ion Exchange," paper presented at the Water Supply Improvement Association, 10th Annual Conference, Honolulu, HI, July 27, 1982.
5. Pittner, G., Levander, R., and Bossler, J., "Unique Double-Pass Reverse Osmosis System Eliminates Ion Exchange for Many Deionization Applications," ULTRAPURE WATER, September/October 1986.
6. Whipple, S., Ebach, E. and Beardsley, S., "The Economics of Reverse Osmosis and Ion Exchange," paper presented at the Ultrapure Water Conference and Exposition, Philadelphia, PA, April 13-15, 1987.
7. Smith, B. and Whipple, S., "RO Pretreatment Eases Strain on Demin System," POWER, March 1994.
8. Hamann, H., Buyok, W., and Whipple, S., "Reverse Osmosis Pretreatment of Boiler Water for Cost Savings and Waste Reductions," paper presented at the International Water Conference, Pittsburgh, PA, October 31-November 2, 1994.
9. Beardsley, S., Coker, S., and Whipple, S., "The Economics of Reverse Osmosis and Ion Exchange," paper presented at WATERTECH '94, November 9-11, 1994.

## Acknowledgments

The authors greatly appreciate the cooperation of Christine T. Wilson of Glegg Water Conditioning, Inc., and Robert D. Gornal of Illinois Water Treatment Company in providing capital equipment estimates.

**Dow Liquid Separations Offices.**

**For more information call Dow Liquid Separations:**

**Dow Europe**

Dow Information Centre  
Liquid Separations  
Schurenbergweg 5  
1105 AP Amsterdam Zuidoost  
P.O. Box 12121  
1100 AC Amsterdam  
The Netherlands  
Tel. +31-20-691-6268  
Fax +31-20-691-6418

**Dow Pacific**

Dow Chemical Japan Ltd.  
Liquid Separations  
Tennoz Central Tower  
2-24, Higashikanagawa 2-chome  
Shinagawa-ku, TOKYO 140  
Japan  
Tel. (813) 5460 2100  
Fax (813) 5460 6246

**Dow Pacific**

Dow Chemical Australia Ltd.  
Liquid Separations  
26 Rodborough Road  
French's Forest  
New South Wales 2086  
Australia  
Tel. 61-2-9776-3226  
Fax 61-2-9776-3299

**Dow Latin America**

Dow Quimica S.A.  
Liquid Separations  
Rua Alexandre Dumas, 1671  
Sao Paulo - SP - Brazil  
CEP 04717-903  
Tel. 55-11-5188 9345  
Fax 55-11-5188 9919

**Dow North America**

The Dow Chemical Company  
Liquid Separations  
Customer Information Group  
P.O. Box 1206  
Midland, MI 48641-1206  
USA  
Tel. 1-800-447-4369  
Fax (517) 832-1465

**Internet**

<http://www.dow.com/liquidseps>

**Notice:** No freedom from any patent owned by Seller or others is to be inferred. Because use conditions and applicable laws may differ from one location to another and may change with time, Customer is responsible for determining whether products and the information in this document are appropriate for Customer's use and for ensuring that Customer's workplace and disposal practices are in compliance with applicable laws and other governmental enactments. Seller assumes no obligation or liability for the information in this document. NO WARRANTIES ARE GIVEN; ALL IMPLIED WARRANTIES OF MERCHANTABILITY OR FITNESS FOR A PARTICULAR PURPOSE ARE EXPRESSLY EXCLUDED.

Published April 1998.



\*Trademark of The Dow Chemical Company

Form No. 609-00183-498QRP  
CH 172-163-E-498