Full-Scale Implementation of a Carbonic Acid Aeration System for Hydrogen Sulfide Removal

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The Econ Water Treatment Plant in Eastern Orange County, Florida, uses forced draft aeration for hydrogen sulfide removal. Pretreatment of the raw water for pH adjustment with sulfuric acid was included in the process to increase hydrogen sulfide removal efficiency. However, the use of sulfuric acid reduced the alkalinity buffering capacity in the finished water and increased the sulfate concentrations. These changes resulted in an increase of the corrosivity of the finished water in the distribution system.

In an effort to resolve the corrosion problems, the county implemented a pilot plant experimental program to evaluate the use of carbon dioxide in lieu of sulfuric acid for pretreatment. Based on results of the pilot study, the Econ WTP pretreatment chemical feed system was changed from sulfuric acid pretreatment to carbon dioxide pretreatment on October 27, 1994. Since the installation of the carbon dioxide system a water quality sampling program has been implemented to monitor the water quality in the distribution system including sampling at the customer faucet taps.

Background

The two-month pilot test at the Econ WTP compared the sulfuric acid used for pretreatment acidification to carbon dioxide gas for pH adjustment prior to aeration. When sulfuric acid was added for pH adjustment, the bicarbonate ions in the raw water were converted to carbon dioxide gas and then removed by the aeration process. This chemical reaction reduced the alkalinity values in the aerated water to range between of 20 to 60 mg/L as calcium carbonate. The Langelier Saturation Index (LSI) was aggressive, averaging -0.36. However, when carbon dioxide gas was injected into the raw water, the alkalinity of the aerated water remained essentially the same value as the raw water. The alkalinity of the raw water ranges between 120 to 150 mg/L as calcium carbonate. In addition, the calculated LSI was non-aggressive, ranging from 0.11 to 0.24. With the result of that pilot test the county implemented installation of a carbon dioxide chemical feed system for pretreatment prior to aeration.

Full Scale Implementation

The pilot plant utilized gaseous carbon dioxide for pH adjustment of the raw water due to its availability and ease of installation. During the bidding phase for the full scale installation of the carbon dioxide system, the capital costs presented for the liquid carbon dioxide system and the solution/gas carbon dioxide system were essentially the same. However, because the liquid system does not require a booster pump to dissolve the carbon dioxide gas into a carbonic acid solution, the liquid carbon dioxide system offered a lower operating power cost than the solution/gas system.

Installation of the liquid carbon dioxide consists of three main components, which include a carbon dioxide bulk storage tank, chemical feed panel and injector valves. Bulk storage is supported on two pier with no secondary containment required for chemical spills. The free standing control panel is adjacent to the chemical injection point. Chemical feed is controlled by a sequence of a valves—a pneumatic valve controls the dosage rate and a Praso valve controls the application rate.

The sulfuric acid feed system includes day and bulk storage tanks, and transfer and chemical metering pumps. Secondary containment is required for the chemical storage tanks and dual containment pipe is required from the chemical storage tanks to the point of injection. Chemical feed rate is controlled by metering pumps that must be calibrated on a continuing bases to maintain system reliability.

Operations

The operation of the carbon dioxide feed system offers several advantages over the sulfuric acid feed system. The carbon dioxide system requires only one controlling source to maintain a consistent carbon dioxide feed rate, whereas the sulfuric acid feed system requires a compound control loop to maintain a constant sulfuric acid feed rate.

Another operational advantage of the carbon dioxide feed system is that fewer mechanical components are required to operate the system when compared to the sulfuric acid feed system. In general, the carbon dioxide system would require less downtime because the equipment maintenance requirements for the system are not as extensive as the sulfuric acid feed system.

Finished Water Quality

The most important aspect of the implementation of the carbon dioxide feed system is the effects on finished water quality. Since the implementation of the carbon dioxide feed system, finished water quality has been closely monitored. Alkalinity is measured using a titration method and turbidity with a turbidimeter. Corrosivity of the water was evaluated by calculating the LSI using pH, alkalinity and hardness values of the finished water. The pH and alkalinity values of the finished water collected while operating the carbon dioxide feed system during January 1995 were compared to the alkalinity and pH values while operating the sulfuric acid feed system during June 1994.

The pH of the finished water ranged between 7.4 and 8.0. These values are similar to historic pH values from the plant. The alkalinity of the finished water ranged between 120 and 160 mg/L as calcium carbonate. These values represent a significant improvement over the alkalinity of the finished water while using sulfuric acid, which ranged between 20 and 60 mg/L as calcium carbonate.

The turbidity values of the finished water at times exceeded 1.0 NTU, whether sulfuric acid or carbonic acid was used for pretreatment. During high flow periods, the hydraulic capacity of the system is exceeded and the removal efficiency of the system is reduced, allowing turbidity forming elements to pass through the aerators and into the distribution system.

LSI values were calculated for the alkalinity range of the finished water using the carbon dioxide feed system and the sulfuric acid feed system. Pretreatment with sulfuric acid produced negative LSI values which is characteristic of a corrosive water. LSI values ranged from 0.11 to 0.24 for the carbon dioxide feed system, which suggests a non-aggressive potable water.

Cost Comparison

Capital and annual operation and maintenance costs were based on actual data collected for both chemical systems. These costs are presented based on a snap-shot of the data to show the present worth cost. Based on an average daily flow rate of 9 million gallons per day at the Econ WTP, both chemical feed systems were evaluated.

Capital costs for carbon dioxide and sulfuric acid are summarized in Table 1 and Table 2, respectively. Operation and maintenance cost estimates for each chemical feed system are presented in Tables 3 and 4, respectively. A present worth analysis was developed for each system using a 6 percent discount factor for a 15 year service life. The results of this analysis are presented in Table 5.

Conclusions

Based on the data that has been collected to date, it appears that the full scale carbon dioxide feed system provides a buffered and non-aggressive finished water throughout the entire treatment process. In addition to enhanced finished water quality, the overall operating cost is less than projected during the pilot testing and significantly less than the operating cost for the sulfuric acid system. Through further collection and analysis of carbon dioxide feed system operational data, the cost and operation of the Econ WTP will continue to be refined.

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Table 1. Sulfuric Acid Capital Cost

Description	Estimated Capital Cost
Sulfuric Acid Feed System	$60,900
Sodium Hydroxide Feed System	45,400
Sodium Bicarbonate System	43,600
Inhibitor System	15,500
Subtotal	**$165,400**
Contingency (15%)	24,800
Total Estimated Cost	**$190,200**

Table 2. Carbon Dioxide Capital Cost

Description	Estimated Capital Cost
Carbon dioxide Feed System	$202,000
Sodium Hydroxide Feed System	22,000
Inhibitor System	15,800
Subtotal	**$239,800**
Contingency (15%)	35,900
Total Estimated Cost	**$275,700**

Table 3. Estimated Annual O & M Costs—Sulfuric Acid System

Description	Average Annual Cost
Sulfuric Acid	$65,700
Sodium Hydroxide	301,000
Power	2,300
Labor	12,000
Total	**$381,000**
Cost per thousand gallons	**$0.12**

Table 4. Estimated Annual O & M Costs—Carbon Dioxide Feed System

Description	Average Annual Cost
Carbon Dioxide	$164,000
Sodium Hydroxide	36,500
Power	1,500
Labor	12,000
Total	**$214,000**
Cost per thousand gallons	**$0.07**

Table 5. Present Worth Analysis—Carbon Dioxide and Sulfuric Acid Feed Systems

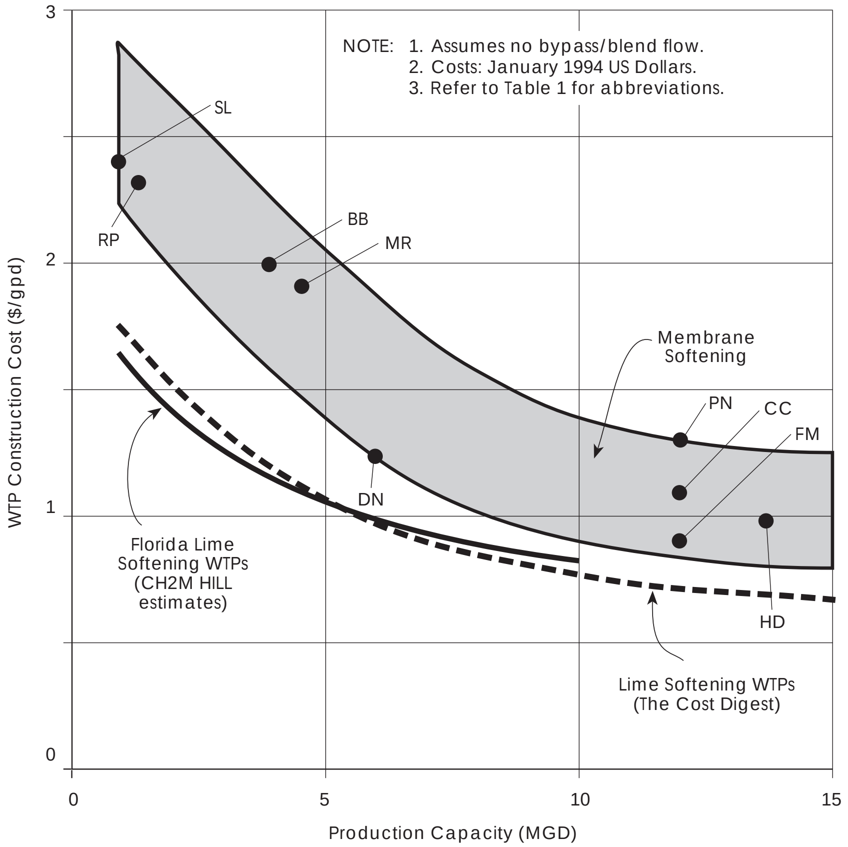
System	Present Worth
Carbon Dioxide	$2,600,000
Sulfuric Acid	$3,900,000

Comparative Economics of Membrane Softening and Lime Softening for Florida's Colored Groundwaters

Robert A. Bergman

New water softening plants in Florida have favored membrane softening over lime softening in treating colored groundwater because of the onset of more stringent drinking water standards, particularly for disinfectants and disinfectant byproducts, and the continuing development and improvements of nanofiltration membranes. This article explores the relative costs between lime softening and membrane softening for WTPs between 1 and 15 MGD.

Although construction and operation and maintenance (O&M) costs for lime softening generally are less than for membrane softening, membrane softening produces superior product water quality. If additional treatment processes are added to lime softening to produce comparable water quality, membrane softening can be less expensive. Furthermore, at some locations, a portion of the feedwater can be bypassed and blended with the membrane product, thereby meeting finished water quality goals at lower costs than lime softening.



NOTE: 1. Assumes no bypass/blend flow.
2. Costs: January 1994 US Dollars.
3. Refer to Table 1 for abbreviations.

Figure 1. Lime and Membrane Softening Water Treatment Plant Unit Construction Cost.

Data Sources

Construction and O&M cost data were obtained from eight of nine large membrane softening WTPs operating in Florida that responded to a cost survey sent in July and August, 1994. Three sources of data were used to determine lime softening construction costs for new plants in Florida: (1) construction cost of one new lime softening plant built in Florida, (2) The Cost Digest[1] data, and (3) in-house CH2M HILL cost estimate data for Florida lime softening plants.

Cost Analysis Methodology

Construction Costs—Construction costs and plant product water capacity data were used to develop unit construction costs. Costs were adjusted (from the date of startup) to January 1994 dollars using Engineering News Record construction cost index values (January 1994 ENR CCI is 5336).

Because many of the membrane softening plants were constructed with facilities sized for future membrane treatment capacity expansion, "adjusted" estimated unit construction cost per installed membrane capacity were determined. For example, where the treatment process building and other general plant facilities were sized for 8 MGD treatment capacity, but only 4 MGD of membranes were installed, the total treatment plant unit cost was calculated by adding the membrane system unit cost in dollars per gallon per day ($/gpd) permeate capacity (membrane system cost divided by 4 MGD) to the unit cost of the remainder of the plant's cost (the non-membrane system part of the construction cost divided by 8 MGD). In this manner, more comparable unit construction costs were determined for the various plants. Furthermore, the unit construction costs were based on the membrane permeate flow rate, i.e. not including any feed water that may bypass around the membranes for blending.

O&M Costs—O&M costs provided by the utilities were actual costs incurred while treating a given quantity of water or, in some cases, budgeted amounts. This is because either the actual cost data was not available or the plant recently had been commissioned and representative long-term costs were not yet available.

The ratio of actual production to plant capacity is called plant operating or utilization factor. Because the plants had varying plant operating factors, O&M costs per 1,000 gallons of membrane permeate were adjusted assuming 100 percent plant factor.

This was done by separating the O&M costs into two categories: (1) fixed costs, which generally are not dependent on plant operating factor (e.g., labor salaries and fringe benefits) and (2) variable costs, which vary proportionately with plant production output (e.g., chemicals, power, cartridge filter replacements).

Membrane replacement costs were considered fixed costs because the utilities reported annual budgeted costs, rather than membrane actual replacement costs; in some cases the

Figure 2. Lime and Membrane Softening Water Treatment Plant Unit Operation and Maintenance Cost.

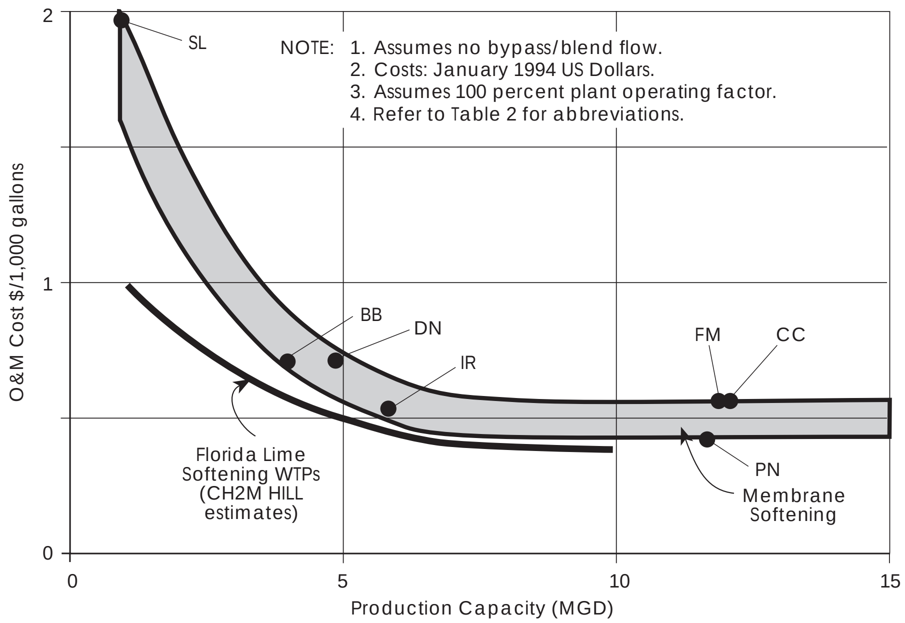


Table 1. Construction Costs—Membrane Softening Water Treatment Plants in Florida

Plant Name	Year Commissioned	Installed Membrane Capacity (MGD)	Treatment Plant Construction Cost ($ million)[a]	Estimated Unit Cost per Installed Membrane Capacity in 1994 Dollars ($/gpd)[b,c]		
				Treatment Plant	Disposal Well	Total
Operational						
Plantation (PN)	1991	12	12.2	1.13	0.23	1.36
Fort Myers (FM)	1992	12	14.2	0.88	NA	0.88
Collier Co. (CC)	1993	12	16.8	0.97	0.10	1.07
Indian River Co. South (IR)	1991 & 1994	6	11.4	1.90	NA	1.90
Dunedin (DN)	1993	4.8[d]	10.5	1.22	NA	1.22
Boynton Beach (BB)	1994	4	12.2	1.88	0.13	2.01
Village of Royal Palm Beach (RP)	1994	1.5	3.8	2.30	NA	2.30
St. Lucie West Development (SL)	1988	1	7	2.34	NA	2.34
In Construction						
Hollywood (HD)		14[e]	17.4	0.97	NA	0.97
Miramar (MR)		4.5	12.3	1.70	0.23	1.93

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[a] Does not include raw water supply and concentrate disposal well cost (if applicable).

[b] Estimated unit cost does not include raw water supply wells and piping and is expressed per gpd of membrane permeate capacity. Raw water supply costs typically vary from $0.03 to 0.24/gpd for the plants shown.

[c] Unit cost adjusted to account for component facilities with capacity greater than the installed membrane capacity. For example, if a building is sized for future membrane trains, the building cost is divided by the full-load membrane capacity.

[d] Six MGD of membrane capacity supplied in contract.

[e] Additionally, 4 MGD of RO capacity will be installed treating Floridan aquifer brackish groundwater.

NA = Not applicable.

Table 2. Operation and Maintenance Costs—Membrane Softening Water Treatment Plants in Florida

Plant Name	Membrane Permeate Capacity (MGD)	O&M Cost Assuming 100 Percent Operating Plant Factor without Membrane Bypass Blending ($/1,000 gallons)[a]					
		Labor[b]	Chemicals	Power	Membrane Replacement[c]	Other	Total
Plantation (PN)	12	0.12	0.10	0.13	0.08	0.02	0.45
Fort Myers (FM)	12	0.17	0.06	0.15	0.06	0.10	0.54
Collier Co. (CC)	12	0.15	0.08	0.14	0.08	0.09	0.54
Indian River Co. South (IR)	6	0.13	0.13	0.21	0.04	0.04	0.55
Dunedin (DN)	4.8	0.17	0.18	0.19	0.10	0.06	0.70
Boynton Beach (BB)	4	0.38	0.05	0.12	0.07	0.05	0.67
St. Lucie West Development (SL)	1	0.56	0.32	0.63	0.11	0.35	1.97

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[a] Membrane bypass (percent of finished water) for Indian River County South and Dunedin is approximately 30 percent and 50 percent, respectively. For other plants, all finished water is treated with membranes.

[b] Labor includes cost of fringe benefits.

[c] Budget values.

Costs given in January 1994 dollars.

utilities had not yet replaced any membranes or, in other cases, replacements were covered under warranty and the utilities did not incur costs.

To use the O&M cost data for estimating the costs at other locations, or for different production rates at the facilities surveyed, the actual plant operating factor would need to be considered. Under normal conditions, one can assume about 5 percent of the time should be adequate for membrane cleaning and other facility maintenance requiring full or partial plant shutdown. Therefore, a membrane softening plant operating as a base load plant could have costs vary similar to the costs presented.

Cost Comparisons

For lime and membrane softening WTPs from 1 to 15 MGD capacity, lime softening plant construction costs were lower than membrane softening throughout the entire range, but the relative difference in costs decreased with larger facilities (Table 1 and Figure 1). For 15 MGD plants, the lower range of the membrane softening plant costs is only about 10 percent greater than lime softening plant cost. If additional treatment processes are added to lime softening to produce water quality comparable to membrane softening permeate, membrane softening can have lower construction costs than lime softening.

For O&M costs (Table 2 and Figure 2), as in the case for construction costs, lime softening generally is less expensive than membrane softening, assuming no bypass/blending and a 100 percent plant operating factor. However, as plant capacities increase, the relative costs between the two processes decrease. Generally, membrane softening O&M costs are about 15 percent greater than lime softening O&M costs. Again, if additional treatment processes are added to lime softening to match the better membrane softening permeate water quality, or if some water can be bypassed around the membranes and blended to produce the finished water, membrane softening can be less expensive than lime softening.

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Evaluation of Color Removal Techniques for Highly Colored Groundwater

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Development of a new groundwater source in the southwestern region of Broward County could potentially change the quality of the finished water produced by the water treatment plants connected to the new source. The new supply has an average color of approximately 100 cu and an average total organic carbon (TOC) of approximately 19 mg/l. The existing supply has a color ranging from 50 to 60 cu and an average TOC of 9 mg/l.

As the new supply is utilized at increasing levels, it is anticipated that treatment modifications may be required to maintain the aesthetic quality (particularly low color) of the treated water while continuing to comply with anticipated disinfection by-product (DBP) regulations.

Due to the high level of TOC in the new supply, the use of chlorine beyond levels currently used as part of full scale treatment may increase the concentration of DBPs beyond the current maximum contaminant level (MCL). Moreover, promulgation of Stage 1 of the Disinfectant/Disinfection By-Product (D/DBP) rule with an anticipated trihalomethane (THM) MCL of 80 mg/L and newly regulated haloacetic acid (HAA) MCL of 60 mg/L will make it more difficult to conform to the D/DBP rule and achieve the target finished water color of 15 cu (or lower) with conventional lime softening and chloramination alone.

In response to the development of the highly colored water supply and in anticipation of the possible treatment modifications it may invoke, a literature review and bench-scale testing were conducted to identify practical treatment approaches which could potentially mitigate the elevated color level and DBP formation potential of the new water supply.

Through the literature review a number of treatment processes were identified as potentially viable alternatives: enhanced lime softening, lime softening preceded and/or followed by oxidation, ion exchange, and membrane softening (nanofiltration). The study did not include an examination of the effectiveness of either ion exchange or membrane softening. It is anticipated that each of the processes is capable of achieving the low color and DBP formation desired. However, the study focused on the more conventional treatment alternatives that are "easily" incorporated into the lime softening process.

The alternative treatment processes examined in this study are ranked below (from least effective to most effective) based on their ability to remove color from the new raw water source.

- Conventional lime softening;
- Potassium permanganate ($KMnO_4$) pretreatment followed by conventional lime softening;
- Ozonation followed by conventional lime softening;
- Ferric chloride ($FeCl_3$) enhanced lime softening;
- Potassium permanganate pretreatment with ferric chloride enhanced lime softening;
- Ozonation followed by ferric chloride enhanced lime softening;
- Conventional lime softening followed by ozonation;
- Ferric chloride enhanced lime softening followed by ozonation;

Throughout this article, the term "ferric chloride enhanced lime softening" is used. This term refers to the process of adding ferric chloride solution to the raw water during the rapid mixing stage. The ferric chloride promotes coagulant formation and destabilization, thereby enhancing the lime softening process's ability to remove contaminants through coagulation and sedimentation.

Raw Water Characteristics

Two sources (wells) of the new water supply were sampled and analyzed. The raw water analyses indicate that the quality of the new water supply is similar to that of existing supplies except that color, TOC, ammonia, and iron levels in the new supply are significantly higher. Table 1 summarizes the new water supply characteristics.

Table 1 New Water Supply Characteristics

Parameter	Well No. 1	Well No. 2
Color (cu)	100	100
pH	7.0	7.2
Total Iron, Fe (mg/L)	1.7	1.6
Ammonia Nitrogen (mg/L)	1.9	1.7
Hydrogen Sulfide (mg/L)	0.13	0.16
Total Alkalinity (mg/L)	244	252
Total Hardness (mg/L)	288	266
Total Organic Carbon (mg/L)	20	19
Total THM Formation Potential (mg/L)	—-	166
Total HAA Formation Potential (mg/L)	—-	254

Bench-Scale Testing

Bench-scale testing performed on the new raw water supply consisted of both jar tests and combination jar test/ozonation tests. In each test, an anionic polymer dose of 0.7 mg/l was used as a coagulant aid in the lime softening stage to improve color removal and sludge settleability. No chlorination or chloramination was performed during this study. Limited use of free chlorine residual can be expected to reduce color, however free chlorine residual will also form regulated DBPs and therefore achievement of a water color of 15 cu without the use of chlorine was selected as the goal for this study.

In general, it was observed that in each test, as the treatment dose increased, the settled water color decreased at a relatively steady rate. In testing employing the use of ferric chloride, as the dosage increased, a decrease in alkalinity and a concurrent increase in hardness in the settled water was observed. At high dosages of ferric chloride (typically above 25 to 30 mg/l), the alkalinity and hardness levels in the settled water reached undesirable levels.

Conventional and $FeCl_3$ Enhanced Lime Softening. The conventional lime softening process is very common is southeast Florida and therefore it was used as the baseline for this study. Conventional lime softening (with polymer addition) consistently reduced the color of the new supply to approximately 55 cu (a 45% reduction). As expected, $FeCl_3$ enhancement significantly improved the color removal performance of the lime softening process. The settled water color ranged from 47 to 17 cu as the applied $FeCl_3$ dose was increased from 10 to 60 mg/l.

$KMnO_4$ Pretreatment with Conventional and $FeCl_3$ Enhanced Lime Softening. Jar tests were performed to simulate $KMnO_4$ pretreatment followed by conventional and $FeCl_3$ enhanced lime softening. Potassium permanganate dosages of 5 and 10 mg/l and $FeCl_3$ dosages ranging from 0 (to simulate conventional lime softening) to 50 mg/l were used. Pretreatment contact times of 1 and 90 minutes were simulated.

Pretreatment with $KMnO_4$ followed by conventional lime softening was the least effective of the pretreatment options examined — resulting in a minimum settled water color of approximately 40 cu (a 60% reduction). Color removal improved significantly when $KMnO_4$ pretreatment was followed by $FeCl_3$ enhanced lime softening. As the $FeCl_3$ dose was increased from 10 to 50 mg/l, the settled water color decreased from 30 to 15 cu. The secondary standard of 15 cu was achieved with 5 mg/l of $KMnO_4$ at a contact time of 90 minutes followed by $FeCl_3$ enhanced lime softening at a $FeCl_3$ dose of 50 mg/l.

Test results indicate that a $KMnO_4$ dose of 10 mg/l was only slightly more effective than a 5 mg/l dose. On average, the 10 mg/l dose decreased color by only 1.5 cu over the 5 mg/l dose. Although it appears that increased $KMnO_4$ dosages (above 5 mg/l) do not significantly improve performance, results show that increased contact time does significantly improve color removal performance. The settled water color resulting from $FeCl_3$ enhanced lime softening with $KMnO_4$ pretreatment at a 90 minute contact time was consistently 5 cu lower than that achieved with a 1 minute contact time. When $KMnO_4$ pretreatment at the extended contact time was coupled with $FeCl_3$ enhanced lime softening, a settled water color of 20 cu was achieved with a $FeCl_3$ dose of 30 mg/l. To achieve the same level of color reduction with the short $KMnO_4$ contact time, a dose of 50 mg/l of $FeCl_3$ was required. When no $KMnO_4$ pretreatment was used, a $FeCl_3$ dose of approximately 55 mg/l was required to reduce the water color to 20 cu.

Ozonation Followed by Conventional and $FeCl_3$ Enhanced Lime Softening. Combination ozonation tests/jar tests were performed to simulate the pre-softening ozonation process. Raw water from the new supply was ozonated at a constant transferred dose of 18 mg/l and a contact time of 15 minutes. The water was subsequently subjected to either conventional or $FeCl_3$ enhanced lime softening. Ferric chloride dosages ranged from 10 to 60 mg/l.

Test results indicate that ozonation followed by $FeCl_3$ enhanced lime softening achieves a high level of color removal. As the $FeCl_3$ dose increased from 10 to 60 mg/l, the resulting settled water color decreased from 22 to 8 cu. The secondary color standard of 15 cu was achieved at an ozone dose of 18 mg/l and a $FeCl_3$ dose of approximately 22 mg/l.

The test results show that ozonation prior to conventional lime softening results in a settled water color of 52 cu. This indicates that ozonation at 18 mg/l provides only 3 cu of additional removal over the lime softening process alone. This does not appear reasonable. Extrapolating the color removal trend corresponding with ozonation and $FeCl_3$ addition back to a $FeCl_3$ dose of 0 mg/l (to simulate conventional lime softening), suggests that ozonation at 18 mg/l followed by conventional lime softening is capable of reducing the water color to approximately 30 cu. This seems to be more likely, however it has not yet been confirmed through additional testing.

Conventional Lime Softening Followed by Ozonation. Raw water from the new supply was lime softened and subsequently ozonated at dosages ranging from 3.5 to 15.2 mg/l and a contact time of 15 minutes. Post-softening ozonation improved color removal significantly over lime softening alone and performed better than pre-softening ozonation. As the ozone dose increased from 3.5 to 12.2 mg/l, the resulting filtered water color decreased from 30 to 12 cu, with a dose of approximately 11 mg/l corresponding to a color of 15 cu. Beyond a dose of 12.2 mg/l, no additional color reduction was observed. Post-softening ozonation with conventional lime softening resulted in up to 43% more color removal than did conventional lime softening alone, and approximately 10% more than pre-softening ozonation with conventional lime softening.

$FeCl_3$ Enhanced Lime Softening Followed by Ozonation. Raw water was treated by the $FeCl_3$ enhanced lime softening process at $FeCl_3$ doses of 10 and 40 mg/l. The settled water was then ozonated at dosages ranging from 2.9 to 15.0 mg/l and a contact time of 15 minutes.

The test results demonstrate that $FeCl_3$ enhanced lime softening followed by ozonation was the most effective color removal treatment examined. Excellent color removal results were achieved at moderate dosages of both $FeCl_3$ and ozone. At a $FeCl_3$ dose of 10 mg/l, as the ozone dose increased from 3.3 to 12.0 mg/l, the filtered water color decreased from 21 to 8 cu. At a $FeCl_3$ dose of 40 mg/l, as the ozone dose increased from 2.9 to 15.0 mg/l, the filtered water color decreased from 16 to 1 cu. The results indicate that a filtered water color of 15 cu can be achieved with approximately 10 mg/l of $FeCl_3$ and 7 mg/l of ozone or 40 mg/l of $FeCl_3$ and 3 mg/l of ozone.

Summary

Test results indicate that it is possible for the WTPs tributary to the new supply to reduce the water color from 100 to 15 cu without increasing chlorine. Table 2 contains the treatment processes found to achieve a color of 15 cu and the dosages required. It is possible that those treatment alternatives not listed in Table 2 are capable of reducing the color of the new supply to 15 cu, but that the dosages required to do so were not used in this study. Figures 1 and 2 summarize the performance of each treatment process evaluated in the study. Figure 1 includes conventional and enhanced lime softening alone, and with the pretreatment processes investigated. Figure 2 includes conventional and enhanced lime softening with post-softening ozonation at various dosages.

The intent of the study is to identify potential treatment modifications that can be made to the conventional lime softening process to effectively treat the new water supply. The results of the study are now being developed further by individual users of the new water supply to most cost effectively modify the existing treatment processes at their individual WTPs.

Treatment Alternatives	*Dosage Required to Achieve 15 cu*
$KMnO_4$ pretreatment with $FeCl_3$ enhanced LS	$KMnO_4$ dose of 5 mg/L at a contact time of 90 minutes and a $FeCl_3$ dose of 50 mg/L
Post-softening ozonation with no enhancement	Ozone dose of 11 mg/L
Post-softening ozonation with $FeCl_3$ enhancement	$FeCl_3$ dose of 10 mg/L and an ozone dose of 7 mg/L $FeCl_3$ dose of 40 mg/L and an ozone dose of 3 mg/L
Pre-softening ozonation with $FeCl_3$ enhancement	Ozone dose of 18 mg/L and a $FeCl_3$ dose of 22 mg/L

Table 2. Treatment Alternatives Found to Reduce Color to 15 cu

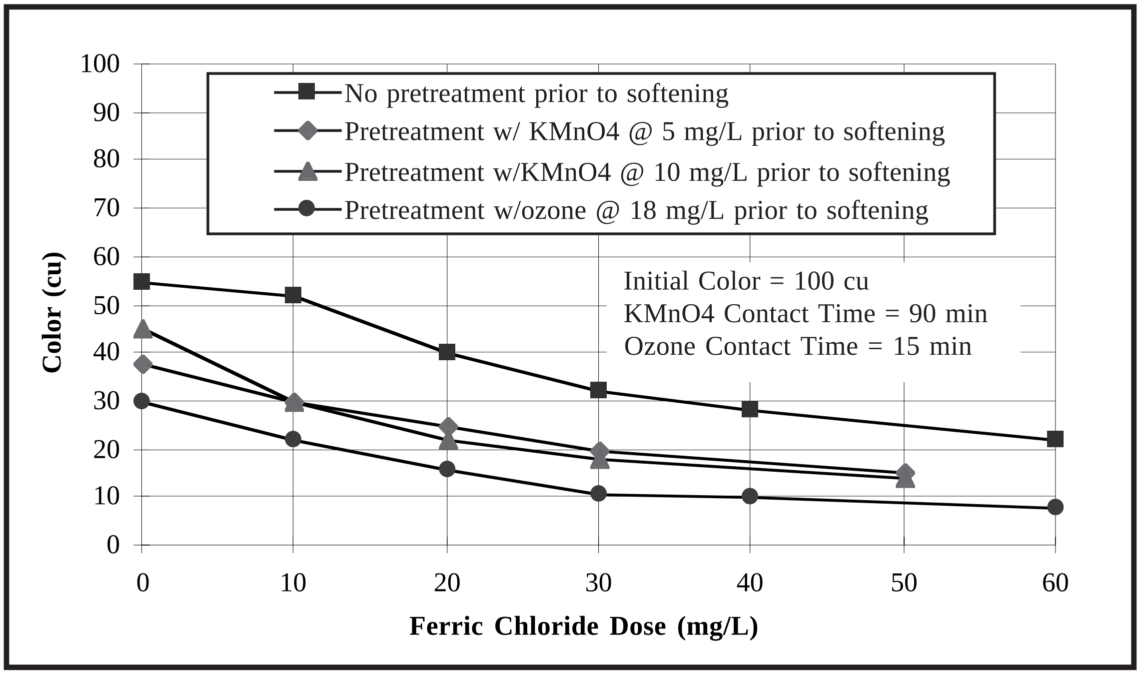

Figure 1. Pre-softening Oxidation

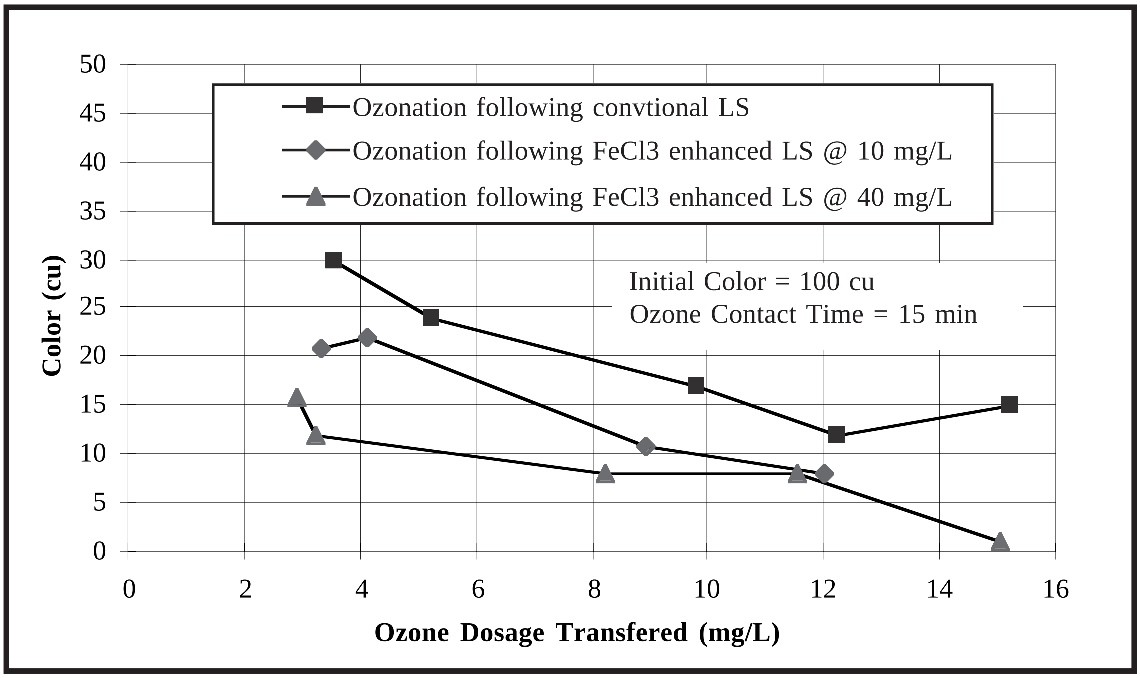

Figure 2. Post-softening Oxidation

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