Increased Water Treatment Plant Capacity at a Mexican Power Station Using Micro Media Filters Ahead of RO and Replacing EDI with Ion Exchange

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ABSTRACT: This paper describes the process and equipment used to treat a Mexican secondary water source in order to supply boiler feed make-up to a combined cycle power plant. RO pretreatment is provided by a highly efficient micro media filter that produces a filtrate with a turbidity of 0.1 – 0.2 NTU and an SDI value of 3 – 5. Following the RO, a compressed-bed ion exchange system was supplied to replace an existing electrodeionization polisher. The final product water produced has a conductivity less than 0.06 µS/cm.

INTRODUCTION

Recovery of wastewater for industrial use presents many challenges for the design and operation of water treatment equipment. The term wastewater can apply to a range of water sources from municipal sewage to storm water runoff to industrial discharge water. As such, the water characteristics may have great variability and equipment used to treat the water must be robust and forgiving. Wastewater typically has high levels of dissolved solids, higher metals content than in natural water sources and potentially high levels of organic matter, both natural and man-made.

This paper describes a water treatment system fed with secondary water used in the production of boiler water makeup water. This is a gas-fired combined cycle plant with a generating capacity of 600 MW and has been in operation since 2002. The plant obtains its water (for both cooling and boiler feed makeup) from municipal primary treatment lagoons/settling ponds. The main source of water entering the lagoons is municipal sewage. Minor sources include storm water runoff and industrial discharge water (both process and sewage).

Secondary treatment of the water is provided at the power plant itself. The sewage treatment plant uses a biological treatment process to first oxidize organic matter and NH_3 in an aerobic step and then remove nitrates formed by NH_3 oxidation by bacterial action under anaerobic conditions. This eliminates biological contaminants and reduces other contaminants in the water.

The water is then treated with lime, which raises the pH of the water, causes the precipitation of dissolved minerals such as calcium and magnesium and reduces the overall dissolved solids content. The lime addition takes place in a clarifier where the precipitates are removed. The water is then treated with sulfuric acid to bring the water to a neutral pH and disinfected by chlorine addition. The precipitated sludge is thickened and dehydrated on a belt press to produce a non-hazardous solid, which can be sent to a landfill.

The clarified water is then suitable for use as makeup water for the cooling towers. A portion of the water requires further treatment in order to obtain water that is suitable for use in a boiler and for other plant water users. The original water treatment plant consisted of sand filtration, cartridge filtration, two-pass reverse osmosis and electrodeionization.

After three years of operation the plant needed additional capacity. Also, the plant was experiencing operational difficulties. Reverse osmosis units fouled rapidly and required frequent cleanings. The EDI system was not consistently producing water of sufficient quality.

Burns & McDonnell was hired to assist in evaluating improvement of the pre-filtration ahead of the RO, increased RO capacity and improved product water quality.

The clarified feed to the plant was found to have the following characteristics:

Table 1: Feed Water Composition

Parameter	**Units**	**Design Value or Range**
TSS	mg/L	5 – 10
Silica	mg/L as SiO_2	10 – 20
TDS	mg/L as $CaCO_3$	720 - 1030
pH	---	7 – 7.5
Alkalinity	mg/L as $CaCO_3$	140 – 200
Conductivity	μS/cm	1800 – 2400
Phosphate	mg/L as PO_4	1 – 3
Hardness	mg/L as $CaCO_3$	220 – 300
Calcium	mg/L as $CaCO_3$	100 – 180
Potassium	mg/L as K	8.5
Fluoride	mg/L as F	0.43
Nitrate	mg/L as NO_3	16.9
Sodium	mg/L as Na	230 – 330
Chloride	mg/L as Cl	250 – 350
Manganese	mg/L as Mn	0.2 – 1.0
Iron	mg/L as Fe	0.2 – 0.5

After extensive evaluation the system shown in Figure 1 was supplied by Eco-Tec and consisted of these main elements:

- **Pretreatment** using a single high efficiency micro media filter to replace 4 existing sand filters.
- Upgrade of existing **Reverse Osmosis** capacity with new membrane elements, valving and instruments.
- Replacement of EDI system with a **compressed short bed IX demineralizer**.

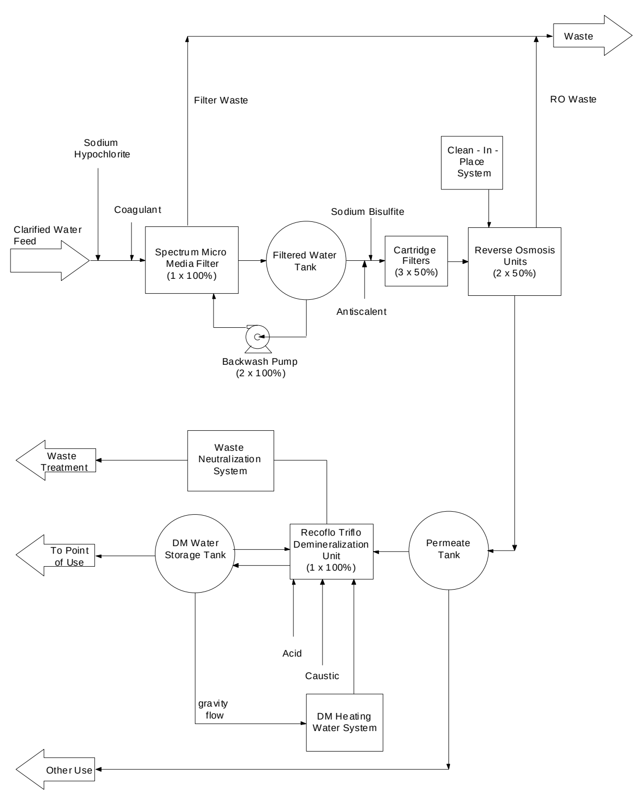

Figure 1 - Schematic showing proposed main system elements

PRETREATMENT

Secondary water sources like this cannot be used directly as feed to RO systems because they contain relatively high levels of suspended solids, organics, colloids and biological material even after clarification. A tertiary treatment process of some sort is required to prevent fouling of the RO membranes. Amongst the most commonly recommended techniques for pretreatment are: conventional depth media filtration; microfiltration (MF); and ultrafiltration (UF) [1,2].

A major advantage of depth media filtration is that the cost of media replacement is relatively low, since service life is long and the media itself is inexpensive. It is also not prone to fouling and generally is quite robust. Together with a reasonable initial capital cost, this explains the prominent place that this pretreatment method has held for many years. Nevertheless, MF and UF membrane methods do provide a barrier for suspended solids and ensure consistently low levels of TSS to downstream processes. The downside is that these processes tend to have higher initial capital costs and operating costs for membrane replacement are appreciable. Also, in some cases, membrane pretreatment may just move the fouling problem upstream from the RO.

The original system for tertiary treatment of the secondary water supply consisted of two trains of two-stage media filters (Figure 2). The primary sand filters had an effective media size of 0.85 mm, and the secondary filters had an effective media size of 0.65 mm. Coagulant (cationic polymer) was added to the feed stream before the primary filters. Sodium hypochlorite also was dosed into the feed for disinfectant purposes. The filter vessels had diameters of 4.6 ft and generally were operated at a service flow rate of ~6 gal/min/ft^2. Backwashing of the existing system took ~2 hours to complete and was only done on one train at any given time to ensure a sufficient supply of RO feed. Performance of this filtration system was poor and there were frequent spikes in the solids content of the filtrate. Cartridge filters which followed the media filters were overtaxed and solids leaked through to the RO. This had an adverse effect on RO performance, and increased the required frequency of membrane cleaning.

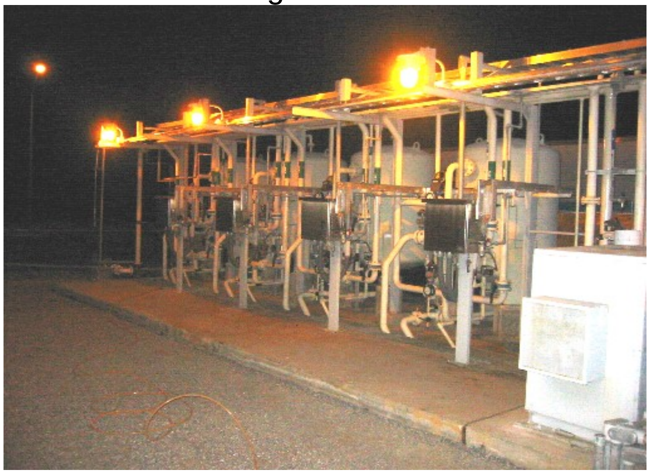

Figure 2 - Previous Two Stage Media Filter Trains

High Efficiency Micro Media Filter

The two-train system of primary and secondary media filters has been replaced with a single "micro media" dual-media filter (Figure 3), which differs in a few key characteristics from typical dual media designs.

Figure 3 - Current Single Micro Media Filter

While the top media layer consists of coarse anthracite (effective size ~0.7 mm) similar to that used in many conventional dual media filters, the lower "micro media" layer is a high density media with an effective size of less than 0.10 mm. This is significantly finer than the fine media used in most designs which generally has an effective size of about 0.35 mm.

Since the lower layer removes the residual quantity of suspended solids not retained in the upper layer it effectively defines the filtration efficiency. The use of finer media allows operation at a higher service flow rate while maintaining adequate filtrate quality. While the higher service flow and finer media result in larger initial *(i.e.* clean) pressure drops across the filter, the pressure drops are manageable (typically ~22 psi compared to ~5 psi for a conventional filter).

As a result of the higher flow rate provided by the micro media filter, the diameter of the filter vessel required to treat a given flow of water can be reduced. The single micro media vessel which has replaced the previously used two-train system is 5.5 ft in diameter and has a design service flow rate of 15 gal/min/ft^2.

Unlike membrane filters, the performance of a depth media filter will vary over the duration of a service cycle. The previous sand filters exhibited the classical ripening effect typical of media filters at the beginning of the cycle where the initial filtrate quality was poor and gradually improved as the cycle proceeded. To ensure that only high quality filtrate was passed downstream to the RO, it was necessary to bleed the initial ~10 column volumes of filtrate to waste over the first 2 hours of the service cycle. The micro media filter, however has a much shorter ripening period: suitable filtrate quality is obtained after 2 column volumes (<10 minutes) of the process cycle.

To maximize the cleaning efficiency of the backwash, the micro media filter operates with a simultaneous air scour/backwash cleaning cycle. Water is first drained to the top of the media. Air and water are then passed simultaneously up through the media bed. The water flow expands the media and allows the air to agitate the media much more violently and uniformly than with water alone. When the vessel has been refilled with water, the air is turned off and a water backwash flushes dirt from the filter in the usual manner. The cleaning cycle for the micro media filter is approximately 15 minutes long.

Filter Performance

Prior to installation of the full-scale micro media filter, a performance comparison was carried out between the existing sand filter trains and a pilot-scale micro media filter (6 inch diameter column). The feed to the pilot system was the same as that to the sand filters, the feed stream being split after the point of coagulant addition.

Figure 4 shows effluent turbidities over a three day period for which the inconsistent performance of the sand filters is apparent. The initial point of upset for the sand filters (about midday on January 18) coincided roughly with the time when one filter was brought back into service after being backwashed. Unfortunately, no SDI measurements were available during the period of comparison. It is also important to note that the coagulant dosage was not necessarily at the point required for optimal performance. This is the probable reason for the filtrate turbidities never falling below ~0.2 NTU.

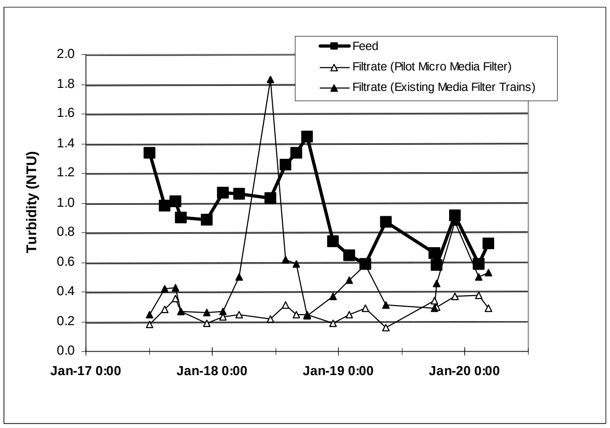

Figure 4 - Filtrate Turbidities for Sand Filters and Pilot Micro Media Filter.

Operating data for a service cycle for the full-scale micro media filter is shown in Figure 5. The service cycles are typically ~8 hours long. Filtrate turbidities of ~0.1 NTU are achieved and $SDI_{1,5}$ values are <3. The unit was being operated at a flow rate below its design value of 15 gal/min/ft^2 due to flow limitations of the secondary water source.

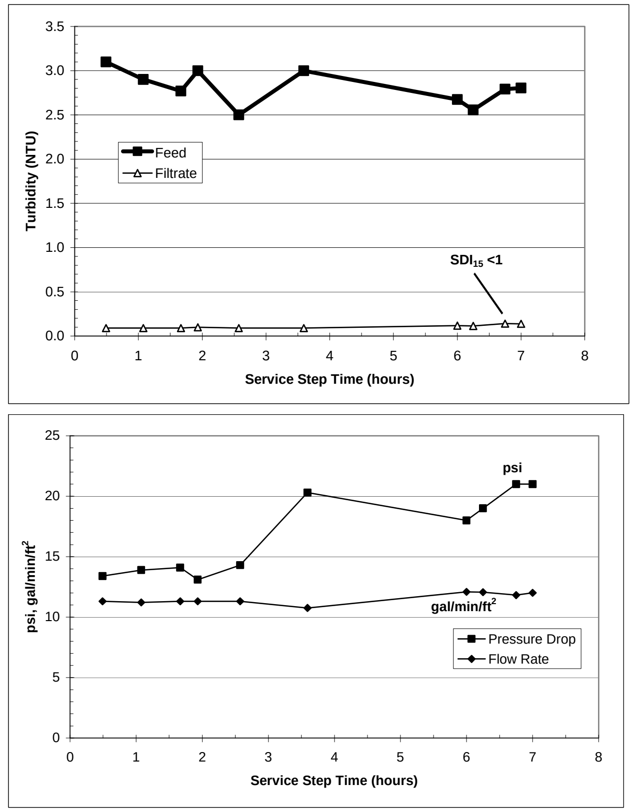

Figure 5 - Operating Data for Full-scale Micro Media Filter.

REVERSE OSMOSIS SYSTEM

Prior to the upgrade, the existing reverse osmosis system was configured into two trains. Each train contained three stages. Three vessels with five elements in the first stage, two vessels with five elements in the second and one vessel with five elements in the final stage. Each train produced 14.2 m^3/hr. A sample of permeate from the RO taken just prior to the upgrade indicated a conductivity of 392 µS/cm. At the time of the analysis the RO membranes were approximately three years old. The poor permeate quality was attributed in part to leaky seals between the membrane elements and membrane degradation due to fouling of the membrane.

The RO system upgrade involved replacing the existing membranes with higher surface area membrane elements. This enabled the

existing arrangement of vessels and piping to be used to produce higher flow capacity. The permeate flowrate per train was increased to 28.4 m^3/hr per train. In addition some existing valving and instruments were replaced.
The original membrane configuration used membrane elements with a 7% lower surface area and a flux rate (permeate flow per membrane surface area) of 7.5 gal/ft^2/day. The upgraded RO system operates at a flux of 14 gal/ft^2/day. The use of a design flux rate of 14 gal/ft^2/day is based upon the RO element manufacturer's guidelines for wastewater pre-treated with multi-media filtration or ultrafiltration. Despite the increase in fluxes the membranes have experienced less fouling since commissioning. Once commissioned the upgraded RO system produced permeate with a conductivity ranging from 30 - 70 µS/cm.
Just less than half of the RO permeate is passed to the ion exchange process. The balance is used elsewhere in the plant.

COMPRESSED SHORT BED ION EXCHANGE SYSTEM

The original system utilized EDI stacks. This well established technology is typically used to treat double pass RO permeate. A detailed description of EDI with a comparison to ion exchange can be found in reference (3). The only concern expressed about the installed EDI equipment, beyond the requirement for additional capacity was the slow and inconsistent restart after shutdown.
The principal benefit of EDI is usually identified as operation without the consumption and storage of regenerant chemicals and the lack of a waste stream. Indeed, with the exception of some make-up salt, no regenerant chemicals are consumed by the EDI portion of the water treatment plant, but chemicals are still required and stored onsite for such things as cleaning RO membranes, feed water pH adjustment, anti-scalants, coagulants, and lime softening.
Since EDI uses no regenerant chemicals they are not added into plant waste, however the salts removed into the effluent stream from the concentrating compartment of the stack technically would be classified as a waste and presumably would be identified as such under most environmental operating permits. Other concerns related to this technology include hardness scaling, leaks and the inability to field service the stack assemblies.
The plant decided to replace the existing EDI cells and expand capacity using a packed bed ion exchange system. Aside from the more favorable economics, the principal technical considerations in arriving at this decision were the rapid start-up after shut down provided by the packed bed system, elimination of the second pass RO requirement, and operating and maintaining a single type of polishing system.

Packed and Compressed Bed Ion Exchange

In polishing applications resins are normally used in mixed bed polishers that are either regenerated on site or are shipped to centralized off-site locations. For economic reasons off site regeneration is limited to feed streams with very low TDS. In this case a design TDS in excess of 80 mg/L as $CaCO_3$ makes this an impractical option.
Standard onsite mixed bed systems have a long and proven history of successful performance. However, relative to newer EDI and packed bed technologies operation and particularly regeneration can be much more complicated, particularly when resins age. Many of these problems are related to resin separation and uniform re-mixing

before and after regeneration[4]. Another disadvantage of conventional mixed bed technology is the relatively high regenerant consumption and waste water production.

Packed resin bed technology reduces these problems by using separate beds of cation and anion resin that are counter-currently regenerated. Separating the resins eliminates the separation and mixing problem. Counter-current regeneration minimizes regenerant consumption and allows production of high quality water by ensuring the highest degree of resin regeneration at the bottom of the column. Thus, the cleanest resin is the last to contact the product water ensuring the lowest contaminant levels.

A further refinement to packed bed technology is the compressed short bed system[5]. Its most obvious feature is the short bed height of 3 or 6 inches. Other features include the use of a much smaller resin bead and complete elimination of internal freeboard. The smaller bead size greatly improves the kinetics of the exchange process, which allows operation at higher flowrates and reduces the depth of the active exchange zone (permitting the use of such short columns). Eliminating freeboard and operating the resin in a slightly compressed state ensures that flow is uniformly distributed (crucial for such short beds) and that the resin position is maintained during regeneration in order to obtain full benefit of the counter-current operation.

One of the main concerns when using packed bed systems is resin fouling. Since resin is not backwashed during regeneration, any accumulated dirt will not be removed. Hence the need for highly efficient pretreatment. With this in mind treatment of RO permeate by packed bed systems is an ideal application.

Demineralizer Design Basis

The bid specification identified the composition of the RO permeate being fed to the demineralizer as follows:

Table 2: Demineralizer Feed Composition

Parameter	Unit	Value
Calcium	ppm as $CaCO_3$	4
Chloride	ppm as Cl	32
Silica	ppm as SiO_2	3
pH		7.15
Temperature	Celsius	16 - 30
Bicarbonate	ppm as $CaCO_3$	30
Nitrate	ppm as NO_3	5.9
Sodium	ppm as Na	35.91
Sulfate	ppm as SO_4	7

Note that the average operating TDS is 30 mg/L as $CaCO_3$. The system was designed to give a net product flowrate of just under 100 gpm. The target product water quality is shown in the following table.

Table 3: Product Water Quality Target

Parameter	Units	Value
Conductivity	µS/cm	$\leq$ 0.06
Total Silica as SiO_2	ppb	$\leq$ 5
Copper as Cu	ppb	$\leq$ 2
Sodium as Na	ppb	$\leq$ 5
Sulphate as SO_4	µg/L	$\leq$ 3
Chloride as Cl	µg/L	$\leq$ 3
TOC	µg/L	$\leq$ 20
pH		6 - 7

System Description

When using separate beds of cation and anion resin the limiting factor with regards to product water quality is generally sodium leakage from the cation bed or from residual caustic left on the anion resin after regeneration. This becomes more problematic when trying to produce high quality water (<0.1 µmho/cm) and with higher feed TDS (>25 mg/L).

Given the feed composition and product water quality target a cation/anion bed combination is not sufficient even with a compressed short bed system. In this case a third polishing bed of cation resin is added to remove low levels of sodium. This low level allows the use of a bed only 3" in depth that is operated at an even higher flux with very infrequent regeneration.

A single train skid-mounted compressed bed system with cation/anion/polishing cation beds was supplied and is shown in Figure 6. The main cation and anion beds are only 30" in diameter and 6" in depth. The final polishing cation bed is 24" in diameter and 3" deep. Net product flow is 100 gpm with a total system cycle time of 21 minutes. This includes the 5 minute fully automated regeneration sequence. Given the low sodium load to the polishing bed regeneration is required only after 100 cycles of the primary beds.

Figure 6 – Single train, skid-mounted, compressed bed system.

To ensure < 5 ppb silica in the product the anion bed is regenerated with warm caustic at 60ºC. Since, the caustic regenerant volume is less than 10 gallons a small 3 ft diameter electrically heated tank is used to supply warm water for regeneration.

The only other significant auxiliaries are two small 4 ft diameter tanks for combining the regen wastes and trimming pH before discharge. Day tanks for premixing of regenerants are not required. The concentrated regenerants are supplied to metering pumps on the skid and injected inline to the regen water. Regenerant consumption and waste volumes are given below.

Chemical	Consumption
Sulfuric Acid *	0.53 lbs/1000 gallons product
Sodium Hydroxide *	0.45 lbs/1000 gallons product
Waste Volumes	17 gals/1000 gallons product

* - denotes 100% weight basis

These values are based upon treatment of the 30mg- $CaCO_3$ /l average feed TDS. Even though this is a relatively high TDS for a polishing application, regenerant consumption is small. Based on the above numbers and assuming continuous system operation standard caustic and acid supply totes would last 1 and 2 months respectively. This helps to reduce chemical volumes stored on site and also reduces the hazards associated with chemical transfer and handling.

System Performance

The system was commissioned in August of 2006 and all performance objectives were achieved. The equipment has since been in continuous service without any problems. Figure 7 shows the onstream conductivity profiles after the primary cation and anion beds and then the final product water profile after the polishing bed, which confirms that the product water was <= 0.06 µmho/cm target value. Feed conductivity was 50 – 76 µmho/cm. Product SiO_2 levels were measured at less than 2ppb.

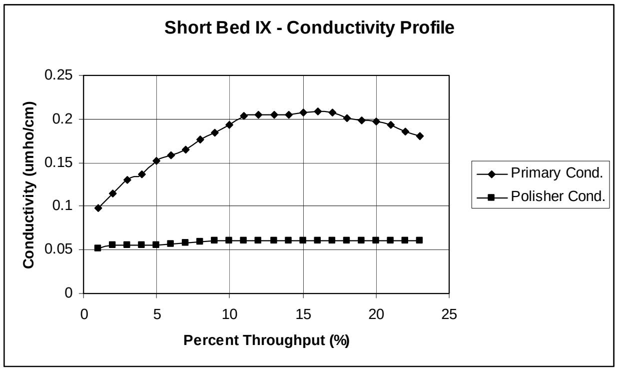

Figure 7 – Onstream conductivity profiles

SUMMARY

An upgraded and expanded water treatment plant was supplied to treat a secondary waste water stream from a Mexican municipal waste water treatment plant. A high purity product water with a conductivity <= 0.6 µmho/cm was produced and supplied as boiler make-up water to a 600 MW combined cycle power plant.

Tertiary treatment is provided by a high efficiency media filter that replaced existing sand filters and was found to be more economic than ultrafiltration. This filter typically produced filtrate with a turbidity 0.1 NTU and SDI values of <3.

Primary demineralization is provided by a single pass RO system using low fouling composite membranes. Permeate quality ranges from 30 – 75 µmho/cm when treating feed water with a TDS of 700 – 1,000 mg/L.

Polishing is provided by a compressed short bed ion exchange system, which replaced existing EDI cells. The new system is able to operate on a feed with a much higher TDS eliminating the need for the existing second pass RO. The problem of slow EDI start-up after shut down was also addressed. The system has now been in service for over a year without any problems.

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