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Pre-treatment optimisation of SWRO membrane desalination under tropical conditions

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ABSTRACT

Seawater quality and temperature under tropical conditions require particular attention to optimise the pre-treatment process and to minimise reverse osmosis (RO) membrane fouling in order to maximise plant performance. A skid-mounted pilot plant of 50 m^3d^{-1} was constructed to examine the impact of various pre-treatments on RO feed quality and to optimise the operation conditions for continuous seawater RO (SWRO) membrane desalination. The pilot plant consisted of five main sub-systems: (1) open seawater intake, (2) conventional pre-treatment, (3) one-stage RO membrane, (4) effluent discharge and (5) supervised control and data acquisition (SCADA). Such set-up allows pre-treatments of varying coagulants and their doses, hydraulic retention time, with and without chlorination, different filtration schemes of the dual-media filter, several methods of clean-in-place of the RO membrane and its service and flushing arrangement. The sub-systems were tested and the operation conditions optimised. The optimal operation conditions include intermittent chlorination (6 mg L^{-1} NaOCl) and dechlorination (6 mg L^{-1} $Na_2S_2O_5$), coagulant dose of 3 mg L^{-1} poly-aluminium chloride, 30 min hydraulic retention time in the clarifier, dual-media filter (DMF) operation cycle of 5 h service, 3 min backwashing and 1 min rinsing. Under these operation conditions, the pre-treatment continuously produced RO feed of high quality, with the silt density index always less than 5, most often around 3. The SWRO membrane performed as designed when being fed with the pre-treated high-quality water. The total salt rejection was greater than 99% with product water having total dissolved salt concentration less than 500 mg L^{-1}. All of the product water parameters, except boron, meet the WHO drinking water standards. After nearly 1 year of operation, a water recovery ≥35% could still be achieved when the RO membrane operation pressure was ≥60.5 bars, compared to 54.5 bars at the early stage of plant operation for the same rate of water recovery. The 10% increase in the operation pressure at the later stage is likely due to irreversible loss of the desalination capability of the SWRO membrane.

Keywords: SDI; RO permeate flux; Specific permeate flux; Membrane scaling; Fouling; Water recovery; Salt rejection

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1. Introduction

Two basic technologies dominate seawater desalination, namely, distillation and reverse osmosis. Depending on the technology used, the final product water is of high quality with a total dissolved solids concentration of below 500 mg L^{-1}. Globally, most of the large desalination plants are based on multi-stage flash distillation (MSF) [1,2] or reverse osmosis (RO) processes [2,3], with the remainder accounted for by multi-effect distillation (MED) and vapour compression [4], and electro-dialysis [5]. The modular SWRO membrane process is becoming the preferred choice for new desalination plants due to its energy efficiency and technology maturity [2,3,6], which makes it more attractive in many instances compared to MSF and MED [2,3,7]. The modular and compact nature of SWRO membrane desalination plants makes it possible to build distributed and medium-sized plants.

Several pre-treatment options are available for SWRO membrane desalination [8–10]. They include conventional chemical coagulation, settlement and filtration (CSF) [11–13]; membrane technologies such as various types of microfiltration and/or ultrafiltration [14–16]; and other processes such as nanofiltration and beach well extraction. Each of these pre-treatments has variations in terms of process configuration, chemical consumption and operating conditions. An effective and efficient pre-treatment process is measured according to cost, reliability and preparedness of the feed water for the SWRO membrane desalination. Under current situations, however, the CSF process remains a preferred choice because of its relatively low cost in construction, and its similarity to the front part of conventional water works and its simplicity of operation. The CSF process, if it is properly designed and optimised, can achieve a high-level pre-treatment to remove fouling and scaling substances in the feed water.

One breakthrough membrane technology is two membranes of nanofiltration (NF) [17]. The process has the potential of reducing one-third of energy consumption compared to the normal SWRO process. Another prosperous technology is forward osmosis (FO) with two steps: the first extracts water from the feed, the second separates water from the mixture of a draw solute and water [18]. The key to the effectiveness of the FO process in desalination is in the composition of the osmotic "draw" solution and the ease of the process in separating water from the draw solution [18,19]. The FO technology is currently still at its infant stage. All these, however, require pre-treatment of the feed water for optimum performance.

Several issues need special attention for SWRO membrane desalination under tropical conditions. The pre-treatment, a key to long-term success, should effectively and efficiently remove impurities (including oil and grease and colloidal materials) which might cause blockage of membrane pores, scaling, chemical and bio-fouling [3,6,20]. Use of chemicals to control fouling and scaling should be minimised for cost saving and environmental impact reduction. Optimisation of the system configuration and operation conditions leads to high efficiency in SWRO membrane desalination.

The main objective of this study was to optimize various pre-treatment operation conditions and to study their effect on the performance of the SWRO membrane desalination in terms of water recovery, permeate flux, salt rejection and permeate quality. This was achieved by removing suspended solids, minimising microbiological growth, removing oxidising compounds and colloidal materials and other fouling substances, adjusting pH and dosing anti-scalant to minimise scale formation on the SWRO membrane by cost-effective means.

2. Materials and methods

2.1. Pilot plant

The SWRO membrane desalination pilot plant was designed to produce 50 m^3d^{-1} using local coastal seawater as feed. It employed a conventional chemical coagulant, settlement and duo-media filtration prior to a one-stage SWRO membrane process. All instruments and components were housed in two 20-foot standard industrial containers and powered by a portable electric generator. The pilot plant was installed at the Tanah Merah ferry terminal (TMFT), Singapore, and tested for 1 year to study seasonal variations in the coastal seawater quality. The plant had the unit operations of open seawater intake, a conventional pre-treatment, SWRO membrane, and effluent discharge (Fig. 1). These unit operations were integrated and automatically controlled by the system control and data acquisition (SCADA).

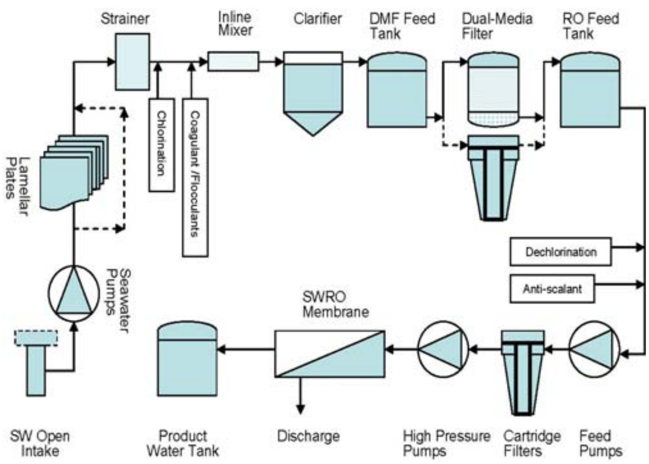

Fig. 1. Schematic flow diagram of the SWRO membrane desalination pilot plant using conventional pre-treatment.

2.1.1. Seawater intake

The intake unit operation drew raw seawater from the sea (E 103°59.2′ N 1°18.8′); the intake point was about 50 m from the shore, at least 1 m below the water surface and 1 m above the seabed. The system included a seawater pump (Yokota self-priming centrifugal pump, model UHN-0410), a coarse PBC mesh for removal of particles larger than 5 mm and a self-cleaning strainer (Amiad TAF, 2″ automatic electric filter) for separation of fine particles before the pre-treatment process. The pump has a capacity of 204 m^3d^{-1} with a total head of 30 m. The self-cleaning strainer has a maximum capacity of 600 m^3h^{-1} and 300 μm weave-wire screen, the strainer began the self-cleaning process when the due time of 15 min or the pressure differential pressure of 0.5 bar across the screen whichever came first. The flushing time was set at 10 s.

2.1.2. Pre-treatment

The pre-treatment system was designed to produce filtered seawater with silt density index (SDI) less than 5 to feed the RO sub-system. With a capacity of 165 m^3d^{-1}, the pre-treatment system consisted of a cylindrical clarifier preceded by a lamella plate settler, and two chemical dosing devices for chlorination and coagulation, a dual-media filter (DMF) and a 10 μm cartridge filter assembly. All equipment and components were mounted and housed inside one 20-foot container.

Clarifier. The clarifier (ID = 2 m and H = 1.7 m) was designed with a hydraulic retention time of 30 min and an effective volume of 4 m^3. The clarifier, made of fibre reinforced polymer (FRP), was circular in shape with a hopper tapering down to a manual valve for sediment removal. The hopper had a slope of 28° to the horizontal and elevated on 4 legs at 0.4 m above the floor. Seawater entered the clarifier through a centre port near the bottom and exit as overflow through weirs on upper part.

Chemical dosing systems. Two dosing systems, for chlorination and coagulant, were installed between the self-cleaning strainer and the clarifier. Sodium hypochlorite was used for chlorination and liquid polyaluminium chloride (PAC, pH 2.0) for coagulation. The dosing pumps (PULSAtron Series MP, model LMB2K2-WTCA-365) had a capacity of 0.8 L h^{-1} at 17 bar. The chemical tanks were made of heavy-density polyethylene (HDPE) with a capacity of 60 L. Calibrations were carried out to ensure accurate delivery of the chemicals by the dosing pumps.

Dual media filter (DMF). The DMF was designed to produce clean seawater of SDI less than 5 at 165 m^3d^{-1}; it was able to filter out effectively suspended solids greater than 20 μm when operated properly. The DMF feed pump (Grundfos multipurpose stainless steel pump, model CHI 8-10) was operated at 6.8 m^3h^{-1} at 1.4 bars. The rectangular DMF feed tank was constructed from FRP with a capacity of 4 m^3. The DMF pressure vessel (Park International) had a vinylester inner shell wound of fibreglass reinforced with high strength epoxy resin. The rated operating pressure was 10.2 bar (150 psi) at 49°C (120°F) with an effective volume of 0.725 m^3 and ID of 0.78 m (a surface area 0.292 m^2). The vessel was filled with 0.3 m of anthracite on the top of 0.6 m of sand with 0.7 m clearance at the top, and 0.2 m of gravel at the bottom. The particle size was 1.5–2.5 mm for anthracite, 0.4–0.55 mm for sand and 3–6 mm for gravel. The centrifugal pump (Grundfos multipurpose stainless steel pump, model CHI 8-15) was used to backwash the DMF using RO feed. In Auto mode, the filter backwash pump was interlocked with level switches in the RO feed tank and operated at 9.0 m^3h^{-1} at 1.6 bars.

10 μm cartridge filter assembly. The 10 μm cartridge filter assembly consisted of three 10 μm cartridge filters and served as by-pass to the DMF during the backwash mode. This enabled the system to continuously supply pre-treated water to the RO unit when the DMF was backwashed. The three 10 μm cartridge filters (Hydrosep 20″ polypropylene) were assembled in parallel and the assembly was installed parallel to the DMF. Under normal operating conditions, the cartridges were replaced once every 2 weeks.

2.1.3. Reserve osmosis membrane

The RO system was designed to produce 50 m^3d^{-1} of product water from the pre-treated water. The system included two chemical dosing systems, a 5 μm cartridge filter assembly, a high pressure pump and a RO membrane unit. All equipment and components were mounted and housed inside one 20-foot container.

Dechlorination and anti-scalant chemical dosing. Two dosing units were installed between the RO feed tank and the 5 μm cartridge filter assembly prior to the RO unit. Sodium metabisulphite was used for dechlorination and the Hypersperse MDC220 (GE-Betz) for anti-scaling. The dosing pumps and the chemical tanks were as those for pre-treatment.

5 μm cartridge filter assembly. The 5 μm cartridge filter assembly consisted of four 5 μm cartridge filters in parallel, serving as a safety guard and final filtration to ensure good feed to the RO membrane unit. The filters were 5 μm Hydrosep 20″ polypropylene cartridges.

RO feed pump. The RO feed pump (Grundfos multipurpose stainless steel pump, Model CHI 8-15) was installed to push seawater from the feed tank through the 5 μm cartridge filters to the high-pressure pump. The feed pump was operated at 6.5 m^3h^{-1} at 2.4 bars. The RO feed tank was made of HDPE with a capacity of 4 m^3. The

Table 1
Chemical, physical and biological characteristics of the raw seawater, the RO feed and permeate[a]

Water parameters	Unit	Raw seawater[b]		RO feed[b]		RO permeate[b]	
		Max	Min	Max	Min	Max	Min
pH	–	8.3	8	8.2	7.6	8.3	5.9
Fe	mg L^{-1}	0.743	0.15#	0.45	0.15#	0.009	0.0015#
Al	mg L^{-1}	2.33	0.95#	0.95#	0.95#	0.041	0.0095#
Na	mg L^{-1}	12,000	8,290	12,300	8,040	261	60.3
K	mg L^{-1}	593	355	570	361	7.97	2.71
Ca	mg L^{-1}	552	366	531	362	1.19	0.27
Mg	mg L^{-1}	1,510	1,060	1,480	1,120	4.19	1.29
Mn	mg L^{-1}	0.577	0.15#	0.15#	0.15#	0.0015#	0.0015#
Sr	mg L^{-1}	10.3	6.32	9.87	6.45	0.02	0.005
Ba	mg L^{-1}	0.515	0.2#	0.2#	0.2#	0.002#	0.002#
B	mg L^{-1}	7.44	2.63	7.12	1.63	2.35	1.06
Cl	mg L^{-1}	19,200	14,300	19200	13,800	298	73
F^-	mg L^{-1}	2	1#	2	1#	0.01#	0.01#
Ortho–P	mg L^{-1}	1.305#	1.305#	1.305#	1.305#	0.015#	0.015#
SO_4	mg L^{-1}	3,450	2,350	3,250	2,250	9.48	2.81
NO_3–N	mg L^{-1}	4.97	0.5#	7.22	0.5#	0.005#	0.005#
NH_4–N	mg L^{-1}	0.11	0.025#	0.09	0.025#	0.06	0.025#
Carbonate	mg L^{-1}	1#	1#	1#	1#	1#	1#
Bicarbonate	mg L^{-1}	113	99	111	92.6	5.79	2
SiO_2	mg L^{-1}	1.05	0.11	1.22	0.11	0.0315	0.0265
TOC	mg L^{-1}	4.22	0.25#	4.03	0.29	0.17	0.01
Total alkalinity	mg L^{-1}	115	101	112	93	5.8	2
Total hardness	mg L^{-1}	7,432	5,360	7,338	5,570	20.2	6.09
TDS	mg L^{-1}	35,700	30,600	34,400	30,700	478	188
UV abs (254 nm)	cm^{-1}	0.025#	0.025#	0.059	0.025#	0.025#	0.025#
TSS	mg L^{-1}	49.3	2.8	37.2	2.6	4.4	0.5
Turbidity	NTU	27.2	0.4	3.8	0.1	0.3	0.1
HPC	cfu mL^{-1}	8,600	10	1,450	0.15	13,100	1.84
Total coliform	cfu mL^{-1}	1,460	46	620	0.5#	0.5#	0.5#
Fecal coliform	cfu mL^{-1}	79	6	34	0.5#	0.5#	0.5#
Oil and grease	mg L^{-1}	1.50	0.30	—	—	—	

[a]Standard EPA and/or APHA methods were followed.
[b]Average of 43 samples taken during the period of May 2004–April 2005 for each type of water.
#, Value was half of the detection limit of the parameter.

cartridge filters were replaced at least once every week when the plant was operated under non-optimal conditions and once every 6–7 weeks under optimal operating conditions.

High-pressure pump. The high-pressure pump was installed to boost the RO feed pressure from 2.4 bar to the RO membrane unit operation pressure of 54–60 bars. The pump (Cat piston pump, Model 3537) was operated at 6.8 m^3h^{-1} at a discharge pressure of 70 bars. The pump was belt-driven by a 15 kW electric three-phase induction motor (Teco Electric & Machinery).

RO membrane unit. Four spiral-wound RO membrane elements with a 7.95" in diameter (Hydranautics SWC3) and in single-stage configuration was installed in serial in an RO pressure vessel (BEL 8-S-1500, Composite Industries). The maximum operation pressure is 102 bars at 49°C. The membrane is made of composite polyamide with a maximum applied pressure of 82.7 bars at 45°C. Each element had a nominal membrane area of 34.37 m^2. The anti-scalant (Hypersperse MDC220 from GE Betz) was dosed at 3 mg L^{-1} to minimise scale formation on the RO membrane. To minimise the polarisation impact on the RO membrane surface, an RO unit ope-rating cycle of 10 h service and 7 min flushing was set as recommended by the manufacturer.

Clean-in-place (CIP). The CIP pump (Grundfos multi-purpose stainless steel pump, model CHI 8-15) was operated at 6.5 m^3h^{-1} at 2.4 bars. Various CIP schemes were tested. In general, CIP was performed once every 1.5 to 2 months or when RO feed pressure increased by 15%.

Normally, the CIP protocols consisted of rinsing the membrane with RO feed, soaking (30 min) and re-circulating (60 min) the membrane with citric acid-EDTA solution at pH 2–4, draining and rinsing with RO feed, re-circulating (60 min)–soaking (30 min)–re-circulating (60 min) the membrane with caustic soda solution (pH 10) and draining and rinsing with RO feed.

2.1.4. Discharge

The discharge unit channelled all effluents, overflows, reject (brine), sediments and product water through a network of drainage into a sump for mixing. A discharge pump (Grundfos multipurpose stainless steel pump, model CHI 12-15) allowed the mixture to flow back into the sea. At the sea, the discharge point was 50 m away from the seawater intake point. The discharge pump installed was operated at 10.0 m^3h^{-1} at 2.4 bars. The sump constructed from FRP had a volume of 0.5 m^3.

2.1.5. Control and data acquisition

The control and data acquisition system automatically controlled all of the process sequences and acquired and stored all of the plant operation data and the performance parameters. The supervisory control and data acquisition (SCADA) was the Iconics SCADA software, Genesis 32. The programmable logic controller (PLC) was the Vision 280™, manufactured by Unitronics Industrial Automation Systems, containing a graphical 4.7" LCD screen and numeric keypad for easy human-machine (HMI) interaction. The VisiLogic™ Ladder software was used to create both the PLC ladder control program and the graphical HMI operator interface. A number of online instruments/sensors were installed for data acquisition, including pressure and flow rate, conductivity, pH and temperature.

2.2. Seawater characteristics

Seawater samples were collected from the TMFT site for laboratory analyses twice a week during the first 6 months of the plant test programme. The sampling frequency was reduced to once a week in the later half of the programme. The samples were analysed for a range of parameters shown in Table 1 using USEPA or APHA standard water analytical procedures. Based on these parameters, preliminary operation conditions were determined.

In addition to raw seawater samples, samples of clarifier effluent, DMF filtrate before dechlorination, RO feed after dechlorination and RO permeate were also taken regularly and characterised (Table 1). Selected samples were determined for silt density index following the recommended method [3,21].

3. Results and discussion

3.1. Optimum operational conditions

Each of the main sub-systems underwent an optimisation process. This includes the coagulation, chlorine dosing for disinfection, clarifier/sedimentation, DMF, RO membrane desalination, and CIP. Only detailed results from optimising coagulation and chlorination are presented here because of the length limit of the paper.

3.1.1. Coagulation

First polyaluminium chloride (PAC) and then ferric chloride ($FeCl_3$) were evaluated as a coagulant. Experiments were conducted to assess their effectiveness in removing suspended substances. Optimisation of the chemical dose was conducted both in the laboratory using the jar test and at the field using the pilot plant. PAC was shown to be a better coagulant than $FeCl_3$ although the latter is currently accepted in large scale SWRO membrane desalination plants [12,13]. The following only presents the results from the PAC study.

PAC doses of 3, 5, 7 and 9 mg L^{-1} were tested in the experiment for optimisation of the DMF cycle-service, backwash and rinsing. Results of the four PAC doses indicated that the lower the dose, the better the SDI values and the lower the differential pressure of the DMF (Fig. 2). The results showed that DMF permeate SDI_{15min} less than 3 was obtained for a period of about 5 h following the backwash when PAC was dosed at 3 mg L^{-1}. The SDI values were greater than 4 when PAC doses were 5, 7 and 9 mg L^{-1}. Use of PAC solution (pH 2–3) as coagulant eliminates the need of pH adjustment and polymer addition which simplifies the operation and reduces cost.

3.1.2. Chlorination and dechlorination

To inhibit microbial activities and protect the membrane from bio-fouling in the SWRO desalination process, chlorination was necessary [3]. Sodium hypochlorite (NaOCl) was used as the disinfectant. Dechlorination was undertaken prior to the RO membrane to reduce total chlorine concentration to less than 0.1 mg L^{-1}, a tolerable concentration of composite RO membranes [22–24]. Sodium metabisulphite ($Na_2S_2O_5$) was used for dechlorination and it reduces hypochlorite to soluble chloride which does not damage the RO membrane.

The dose of 5 mg L^{-1} sodium hypochlorite did not achieve the desired free chlorine concentration between 0.8 and 1.0 mg L^{-1} which is needed to suppress the growth of biofilm microorganisms in sub-systems prior to the RO membrane [25,26]. The dose was then increased to 6 mg L^{-1} and the free residual chlorine of 0.8–0.9 mg L^{-1} was achieved constantly (Fig. 3) at different sampling points in the treatment train.

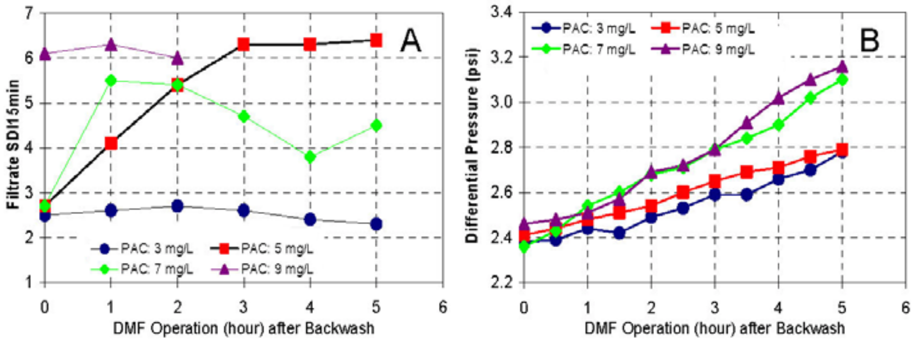

Fig. 2. SDI values of DMF filtrate (A) and DMF differential pressure (B) during operation with a PAC dose after backwash.

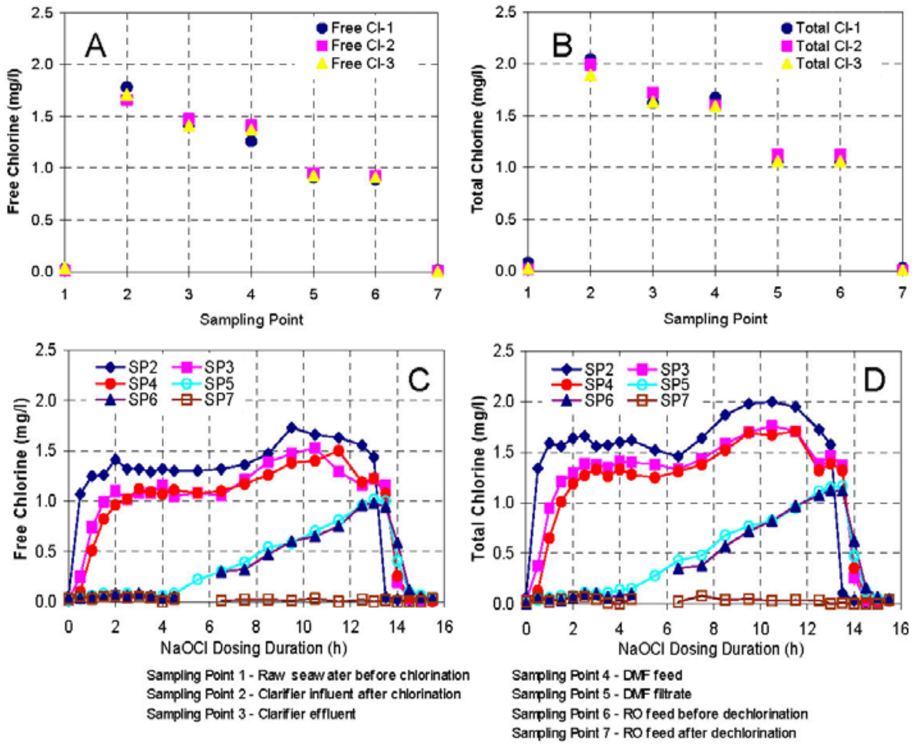

Fig. 3. Free (A) and total (B) chlorine concentrations at different sampling points in the treatment train, free (C) and total (D) chlorine concentrations measured over times at the NaOCl dose of 6 mg L^{-1}, which was dosed just before sampling point 2 and $Na_2S_2O_5$ was dosed at 3 mg L^{-1} right after sampling point 6.

Consequently, the dose of 6 mg L^{-1} NaOCl was fixed for the whole duration of the plant operation. Sodium metabisulphite dose was fixed at 6 mg L^{-1}, sufficient for removal of free chlorine in the RO feed (Fig. 3A) at sampling point 7 after dechlorination. Three experiments were conducted at different days to ensure repeatability of the chlorination and dechlorination results. Similar trend was observed for the total chlorine concentration (Fig. 3B).

To minimise the consumption of chemicals, intermittent chlorination was developed to eliminate biological growth while producing RO feed of high quality. Tests were conducted to study the effect of intermittent chlorine dosing on plant performance. Continuous chlorination for 4–6 h ensures thorough disinfection of the fluid up to the point of dechlorination. This was equivalent to at least three hydraulic detention times of the SWRO membrane desalination plant. At each of the sampling points, there was a period during which the residual free chlorine concentration was between 0.8 to 1.0 mg L^{-1}. According to the CT function (i.e. concentration in mg L^{-1} multiplies exposure time in min), this should eliminate biological growth in the system [25,26]. The parameters which were monitored included the residual free and total chlorine concentrations train (Fig. 3).

In conducting the test, all of the other operating conditions were unchanged except the chlorination and dechlorination scheme. Once a free chlorine concentration of between 0.8 and 0.9 mg L^{-1} was achieved, the chlorine dosing pump was turned off and the DMF filtrate was measured periodically for SDI. Dechlorination continued for an additional 3 h after chlorination was stopped to ensure complete removal of residual free chlorine. When SDI value was greater than 4, the chlorine dosing pump was turned on again to provide a dose of 6 mg L^{-1}, and at the same time, the dechlorination was resumed.

The chlorine dosing pump was idled for about 312 h (13 days) during which the SDI value of DMF filtrate was less than 4. Immediately after the chlorination was resumed, free and total chlorine concentrations at each of the six sampling points were closely monitored. Free chlorine concentration reached a value more than 0.8 mg L^{-1} in 1.5 h for sampling points prior to the DMF, including the point right after the clarifier (Fig. 3). It took, however, about 12 h for the DMF filtrate to reach the desired level of 0.8–1.0 mg L^{-1} free chlorine (Fig. 3C). Similar patterns were observed for the total chlorine concentration (Fig. 3D).

Intermittent chlorination and dechlorination were implemented at the plant from 19 October 2004. The cycle of chlorination was 12 h of 6 mg L^{-1} NaOCl dosing followed by 324 h of non-dosing. Dechlorination was activated whenever chlorination was switched on and continued for an additional 3 h after chlorination was stopped.

3.2. RO membrane performance

The RO membrane unit was routinely operated after the pre-treatment system was tuned up for optimum operation conditions. Water recovery, permeate flux and salt rejection are discussed in the following sections.

3.2.1. Water recovery

The designed water recovery rate was up to 35% for the one-stage RO with four membrane elements in one vessel. During the early stage (April 2004), a water recovery rate of 30% to 34% was achieved under a feed pressure between 54 and 59 bars (Fig. 4). After 1 year, however, the feed pressure needed to be between 54.5 and 60 bars to achieve the same recovery rate. The RO feed pressure needed to be at least 61 bars to achieve a recovery rate of 35% (Fig. 4).

3.2.2. Specific permeate flux

A pressure cycle started right after the RO membrane was flushed or cleaned. At each pressure cycle, the RO membrane feed pressure increased and the water recovery rate decreased as operation continued within the cycle (data not shown). For selected pressure cycles, both flux

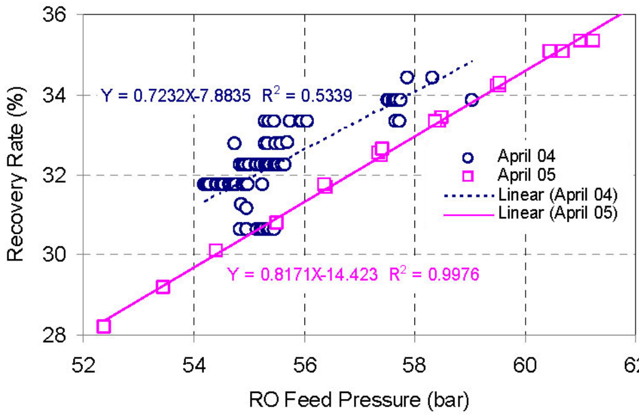

Fig. 4. Water recovery rate as a function of RO feed pressure at early and later stages of plant operation.

(F) and specific flux (F_S) were calculated at the start and end of the cycle using the corresponding RO membrane permeate flow rate according to the following equations:

$$F = \frac{Q}{A} \tag{1}$$

$$F_S = \frac{Q}{A \times P_{TM}} \tag{2}$$

where Q is the RO membrane flow rate (m^3 h^{-1}), A the working RO membrane surface area (m^2), and P_{TM} the trans-membrane pressure of the RO membrane, which is calculated using the following equation:

$$P_{TM} = \frac{P_{Feed} + P_{Reject}}{2} - P_{Permeate}$$

in which P_{Feed} = RO membrane feed pressure (bar), P_{Reject} the RO membrane reject pressure (bar); and $P_{Permeate}$ is the RO membrane permeate pressure (bar).

Permeate flux and specific permeate flux were plotted against operation time (Fig. 5). The results showed that flux both at the early and the late stage of the pressure cycle decreased linearly with operation time and the decreases were in similar pace. Specific permeate flux (F_s) decreased faster than permeate flux during the whole pressure cycle because of the increases in transmembrane pressure needed to sustain the water recovery rate. The decreases in the specific permeate flux were associated with concentration polarization [27,28], scale formation [29] and blockage of the membrane pores by organic and bio-fouling materials [3,30,31]. The concentration polarization can be reduced by periodically flushing the membrane with RO permeate while the scale and the bio-fouled-substances could only be removed by an effective

CIP process. The progressive decreases in specific flux at the early period of a cycle indicated that part of the lost flux could not be recovered by the flushing. The CIP (8 August 2004) was able to recover a portion of the lost flux. However, such recovered flux was quickly lost again (Fig. 5).

3.2.3. Salt rejection and permeate quality

The minimum and maximum values of the water characteristics are presented in Table 1 and the averages in Table 2 for the raw seawater, RO feed and RO permeate. The rejections of the ions and the physical and the biological parameters are presented in Table 2.

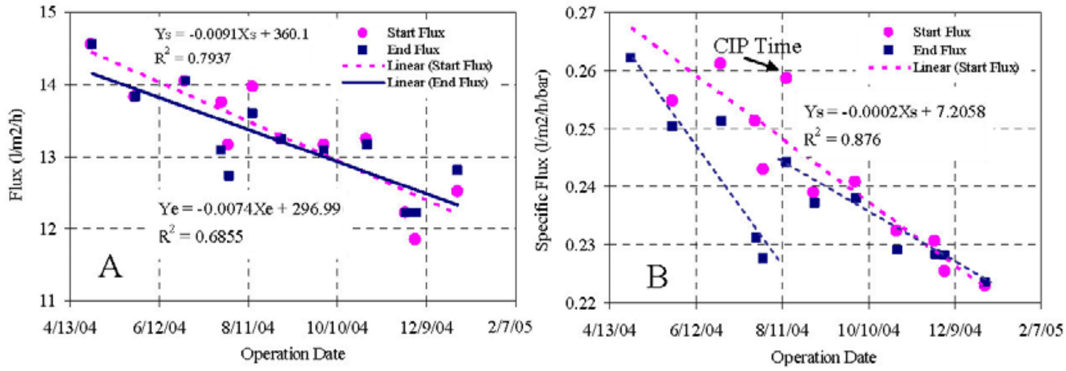

Fig. 5. Changes in RO membrane permeate flux (A) and specific permeate flux (B) at the beginning (Start Flux) and at the end (End Flux) of point selected pressure cycles (each point represents a pressure cycle).

Table 2
Average concentrations and removal of ions and selected water parameters during the pre-treatment and the combined pre-treatment and SWRO membrane desalination process

Parameter	Concentration average[a], mg L^{-1}			Percent removal, %	
	Raw seawater	RO feed	RO permeate	Pre-treatment	Pre-treatment + RO
Fe	0.52	0.35	0.005	14.0	99.0
Al	2.33	0.95	0.022	59.2	99.1
Na	10,189	10,151	121	0.4	98.8
K	444	446	4.59	—	99.0
Ca	427	423	0.51	0.9	99.9
Mg	1,288	1259	1.80	2.3	99.9
Mn	0.52	0.15	0.0015	71.1	99.7
Sr	7.67	7.60	0.01	0.9	99.9
Ba	0.48	0.2	0.002	58.1	99.6
B	5.24	4.78	1.64	8.8	68.7
Cl	16,719	16,919	161	—	99.0
F	2	1	0.01	50.0	99.5
Ortho-P	1.305	1.305	0.015	—	98.9
SO_4	2,705	2,690	3.91	0.5	99.9
NO_3-N	2.83	3.12	0.005	—	99.8
NH_4-N	0.08	0.07	0.043	16.7	46.9
Bicarbonate	107	102	3.29	4.5	96.9
SiO_2	0.54	0.53	0.029	1.5	94.6
TOC	1.98	1.87	0.03	5.4	98.3
Total alkalinity	108	103	3.27	5.0	97.0
Total hardness	6,366	6242	8.68	2.0	99.9
TDS (drying method)	32,484	32,436	287	—	99.1
Total suspended solids	19.9	14.01	2.75	29.6	86.2
Turbidity (NTU)	6.1	0.51	0.15	91.7	97.6

[a]Average of 43 samples taken during the period May 2004–April 2005 for each type of water.

The RO membrane functioned well in terms of ion rejection except boron. The concentrations of all tested parameters of the RO permeate (Tables 1 and 2) meet the WHO drinking guidelines, except boron. The average salt (TDS) rejection was 99.1%. The average removal was 97.0% for alkalinity, 86.26% for TSS and 97.6% for turbidity. Silica and total organic carbon rejections were 94.6% and 98.3%, respectively. Ammonium-N concentrations were below the detection limit (0.5 mg L^{-1}) for most of the analyses, leading to a low calculated rejection (Table 2). Boron concentrations varied between 1.63 and 7.12 mg L^{-1} in the feed with the average of 4.8 mg L^{-1}. The average RO membrane boron rejection was 68.7%. The relatively high B concentration in the RO permeate (average 1.64 mg L^{-1}) was due to (1) the inability of the ordinary SWRO membrane in rejecting non-dissociated neutral boron molecule which is small [3,32,33], and (2) there was no pH adjustment for B rejection by the SWRO membrane desalination process.

The heterotrophic plate count (HPC) was up to 13,100 cfu mL^{-1} for samples from the RO permeate (Table 2). This was due to zero free residual chlorine after dechlorination of the RO feed (Fig. 3) and the time lapsed in analysing the samples might have permitted biological growth. With chlorination at a low concentration (1 to 2 mg L^{-1}), the HPC should be well under control. Both total coliform and faecal coliform were below the detection limit of 1 cfu mL^{-1} in the RO permeate.

5. Conclusions

The effective pre-treatment train for SWRO membrane desalination consisted of chemical dosing, chlorination, coagulation, clarification, and filtration. The optimum operating conditions for the pre-treatment were a coagulant–liquid PAC dose of 3 mg L^{-1}, coagulation and sedimentation hydraulic retention time 30 min, and DMF operation cycle of 5 h, 3 min backwash and 1 min flush. When low pH liquid coagulant (such as PAC) was used, it was not necessary to adjust the feed water pH. Intermittent chlorination of the raw seawater was as effective as continuous chlorination in production of high-quality DMF filtrate. The intermittent chlorination cycle was 12 h continuous dosing of 6 mg L^{-1} NaOCl followed by 324 h non-dosing. A high-quality RO feed (SDI <5, most often SDI≈3) was produced when the raw seawater was pre-treated under the optimal conditions.

A water recovery rate of 35% or higher was achievable after 1 year if the RO feed was of high quality and the RO feed pressure >60.5 bars. Salt rejection of the SWRO membrane was greater than 99% and the total dissolved solids in the product water was less than 500 mg L^{-1}. Without specific process for boron removal, the boron concentration in the product water ranged from 1.06 to 2.35 mg L^{-1} after a cumulative 68.7% removal in the pre-treatment and the RO processes. CIP with citric acid at pH ~2.5 and the addition of EDTA to the cleaning solution substantially improved the effectiveness of cleaning. Part of the RO membrane flux loss, however, was irreversible and the CIP performed was not able completely to strip off the fouling/scaling materials.

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