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Ultrafiltration pretreatment to reverse osmosis for seawater desalination — three years field experience in the Wangtan Datang power plant

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ABSTRACT

Systems composed of ultrafiltration (UF) pretreatment for seawater reverse osmosis (SWRO) desalination are often termed "integrated membrane system" or "dual membrane system". These systems promise to offer reliable handling of very difficult waters, also in regions which have traditionally experienced feed water and pretreatment problems such as Middle East and United States. Nevertheless, data sets presenting long-term integrated operation from a UF and SWRO point of view at a larger scale level, and hence validating the potential cost benefit of UF pretreatment, are very scarce. The dual membrane seawater desalination system at the Wang Tan Power Plant has been operating for three years and is an ideal case to present learnings. Despite some gaps in the data set, some operational problems, and an unconventional low flux and low chemicals operation approach, the authors believe that transparent sharing of this data set can significantly contribute to a better industry understanding of integrated operation. The data shows that UF system operation is possible using a low flux (25 L/m²/h), low chemical approach. This approach totally eliminates the need for coagulation or chemical enhanced backwash, and uses only yearly clean in place operations. This requires higher upfront capital investment, but results in lower chemicals cost, lower sludge and chemical brine disposal, better ease of operation and higher safety level. This approach has allowed reliable water production in the DOW™ UF and FILMTEC™ SWRO unit for 3 years and should be interesting for very environmentally aware regions with difficult waters, such as Australia or United States. Based on limited data, turbidity removal rate was 98–99.5% and outlet SDI typically <2.5. On a water with very high temperature fluctuation, this enabled SWRO operation with slow pressure drop increase and normalized flux loss, hence resulting in low cleaning frequency of around yearly clean in place operation, and low replacement rate of 1%/a. The data also shows that care should be taken that chlorine employed in ultrafiltration backwash operations does not attack SWRO membranes — improved modes eliminating these problems are available and discussed within the paper. Ultimately, a one-year pilot trial in the SWRO plant shows that the combination of ultrafiltration and internally staged design, employing high productivity elements such as FILMTEC™ SW30ULE-400i is synergistic and can enable unprecedented SWRO vessel productivity of 5 m³/h and flux rate in the range of 25 L/m²/h, while achieving excellent water quality in the range of below 500 μS/cm.

Keywords: Ultrafiltration; Reverse osmosis; Seawater; Desalination; Integrated membrane system; Dual membrane system

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1. Introduction to dual membrane systems

In the last ten years, ultrafiltration (UF) or microfiltration (MF) pretreatment has gained widespread attention as potential pretreatment to seawater desalination by seawater reverse osmosis (SWRO). While in the period until 2002, mostly pilot studies were undertaken, in recent years there have been about 10-15 seawater reverse osmosis (SWRO) plants implemented using ultrafiltration pretreatment. Due to the little difference in TMP between UF and MF, but the large impact that MF or UF outlet quality can have on the pressure in the seawater stage, it is generally believed that UF is the preferred option. Therefore the Dow application development work as well as this paper focus on UF technology.

In the late 1990s and early 2000s, UF technology was perceived as a higher cost solution as last resort to deal with very difficult waters. More recently the claim has been made that UF could be equivalent or lower cost compared to conventional pretreatment (CPT) in certain situations. This would obviously be the case, when very difficult waters are to be handled, and therefore the geographies that have had difficulties with their introduction to seawater reverse osmosis due to feed water/ pretreatment problems are (notably Middle East and United States) focusing on UF technology. Hence, the main argument still is more reliable treatment of difficult water, and higher plant availability, less energy consumption and less chemicals use, especially in the SWRO stage.

There are however other drivers for considering ultrafiltration as pretreatment:

- Higher chemical doses and sludge quantities created by conventional pretreatment can not be tolerated due to environmental reasons –this trend apparently starts to develop in some of the Australian projects
- Desalination plant footprint would merit from a UF solution, and site specific conditions show high cost of building or site area
- High value is given to reliably reach maximum plant capacity, e.g. in industrial plants, where the cost of down time due to lack of water is much larger than the water production cost.
- Ease of design and operation: Ultrafiltration provides more stable water quality than a multi media filtration system, because there is no conditioning at the beginning of the cycle and no break through phase at the end of the operation cycle. In addition the membrane supplier takes most ownership of process design. Therefore process design and control is much easier than with conventional pretreatment, and this enables a wider market group access to reliable SWRO operation, especially the participants with less know-how in multi-media filtration.

Based on the above arguments, cost models have been established to compare UF and CPT. Some authors are basing their cost comparison of UF vs. CPT solely on capital and operation cost of UF and CPT operations. Although the assumptions with regards to the various very different UF system designs (submerged or pressurized? outside-in or inside out?), and on the unit operations in CPT (only media filters, or dual media filtration or sedimentation/dissolved air flotation included before, and what is the cartridge filter replacement rate downstream?) tend to be very different, this provides a very transparent comparison.

The main argument, though, is more reliable SWRO operation. Many cost models use improved operating conditions in the SWRO stage in their models (lower energy consumption, lower membrane replacement rate, higher flux and recovery operation, less chemicals use due to lower cleaning frequency) to demonstrate that UF technology saves cost in the integrated desalination system. Some of the advocates of ultrafiltration propose remarkably high performance increases (e.g. up to 80% higher SWRO flux, 40% higher recovery, [1], or that all of the advantages on the SWRO can be combined. Obviously, a realistic assessment is needed to objectively evaluate the potential cost advantage of the dual membrane system. Such an objective assessment must be based on a thorough data evaluation of RO performance after UF pretreatment, ideally side-by-side to a CPT system. The RO data for such an assessment should include including all the parameters mentioned above, energy consumption (or pressure), membrane replacement, flux, recovery, cleaning frequency.

Large scale plant experience is limited, and most knowledge is based on pilot trials. A look at the literature reveals that, pilot studies were often carried out and evaluated by independent institutions, which neutrally evaluated various pretreatment technologies side by side. In most cases, the reported work focused on pretreatment with little attention given to the SWRO process or integrated technologies.

For example, a limited look at the recent 2007 International Desalination Association conference's ultrafiltration papers [2–16] reveals that there are a few studies that have reported downstream process information and are site specific [6,13]. The vast majority of papers focus only on the UF process performance, and not the integrated system with RO data.

The lack of SWRO section performance after UF systems, especially in larger scale plants, presents a true bottleneck in accelerated development of ultrafiltration pretreatment to SWRO. With three years operational experience of the dual UF–SWRO system, Wang Tan power plant is one of the longest running SWRO plants with ultrafiltration pretreatment. It therefore offers the unique opportunity to learn about the integrated system, considering both UF and SWRO performance. This paper complements our previously published paper about the plant [17] and provides a detailed review of three years of operation of the dual UF-SWRO system at WangTan facility. The paper aims at presenting all, positive and negative, aspects of the combined system in an objective and open manner, in order to help accelerating the development of this application.

2. The dual membrane system at Wang Tan power plant

2.1. Introduction to the WangTan DaTang power plant

The WangTan DaTang power plant is a fossil power plant with 2 × 600 MW capacity. A picture of the plant in shown in Fig. 1.

The plant is located in JingTang port in HeBei province, in the Northern part of China, next to TangShan city. It supplies HeBei region and TangShan city with energy, but during the Olympic Games also supplied BeiJing.

The plant uses industrial and boiler make-up water and construction phase 1 which has been implemented is designed to provide 300 m³/h SWRO permeate (part of it being used as industrial water) and 210 m³/h BWRO permeate, which is being converted by ion exchange (IX) to boiler make-up water.

Fig. 1. Wang Tan Da Tang power plant.

City water is one of the supplies to the plant, however water resources are limited, and it was not possible to satisfy the water need of the power station exclusively with city water. Therefore the Wang Tang power plant decided to divert seawater from the Bohai Sea, for use as the raw water source. The geographic location of the plant and its raw seawater supply are shown in Fig. 2.

The water is captured from the Bohai Sea via a channel and pond system. The intake situation is shown via satellite picture in Fig. 3. A larger channel (roughly 500 m width, see top right in Fig. 3) drives the seawater to a smaller channel (roughly 150 m width), from where the water is capture in a lagoon (center of Fig. 3).

Land photos showing the intake situation are shown in Fig. 4.

2.2. History of the plant

At the time of building the plant, it became clear that the only option to supply sufficient water at appropriate quality was a multiple pass seawater desalination system. However, the raw water quality posed a challenging condition for desalination treatment. The sea has high turbidity that fluctuates greatly with wind and tides.

Additionally, the raw seawater temperature ranges from 0°C to 27°C for winter and summer respectively. The raw water temperature range made design of the membrane system and optimizing operations more difficult. To reduce the temperature range, it was decided to operate the desalination plant on seawater that was first passing through the condenser, and hence heating up by roughly 5°C. The temperature curves of the raw

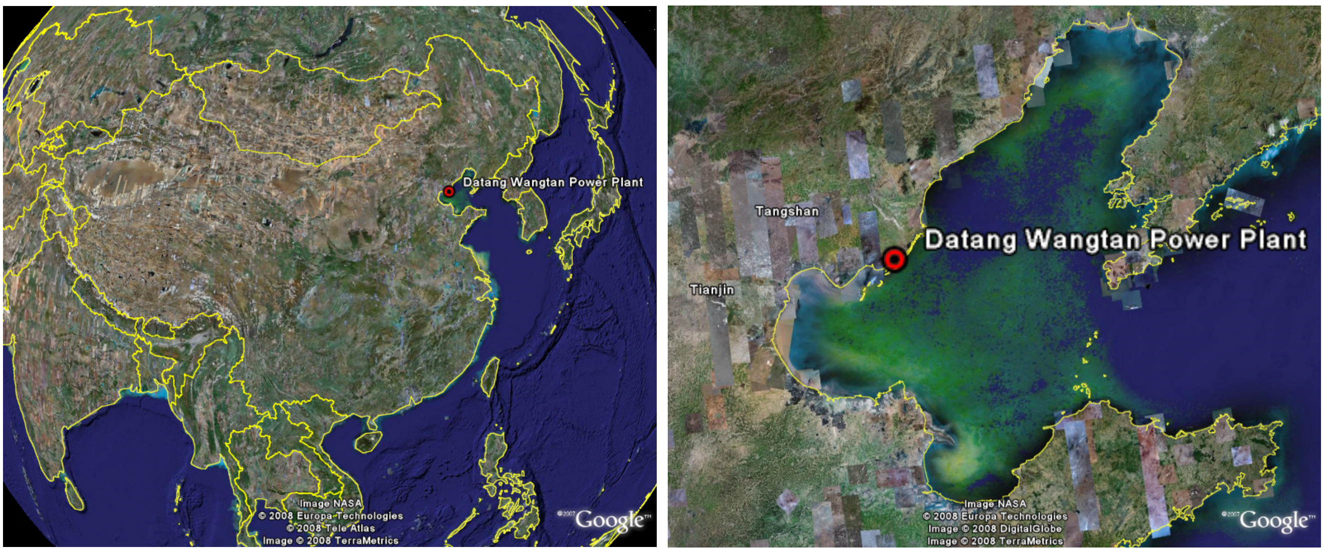

Fig. 2. Geographic location of WangTan DaTang power plant.

Fig. 3. Satellite picture of intake channel and pond.

Fig. 4. Land picture showing the intake situation.

Bohai Sea water and the condenser heated seawater are shown in Fig. 5

Therefore DOW™ UF membranes were pilot tested in parallel to conventional media filtration, to assess performance and potential advantages of using ultrafiltration membranes for sea water desalination pretreatment and to aid design and identify operating parameters.

UF membranes were proven to significantly reduce the turbidity despite the fluctuation in feed water quality. During the seven month testing period the turbidity ranged from 20 to 100 NTU. The feed water turbidity could vary as much as 20 NTU during one day of testing. The DOW UF permeate was consistently below 0.3 NTU and a Silt Density Index (SDI) of 3.0.

In addition to the pilot results, the main arguments for using ultrafiltration pretreatment in SWRO desalination have been cost savings of the dual system of UF and SWRO (given high land costs) and consistent filtrate quality compared to conventional media filtration systems.

Based on these arguments, Dow™ UF technology from Dow Water Solutions was selected as pretreatment to the SWRO operation. Downstream of the SWRO, a BWRO and IX using FILMTEC™ membranes, also from Dow Water Solutions.

2.3. *Treatment process design*

The plant was originally planned to treat 1,200 m^3/h

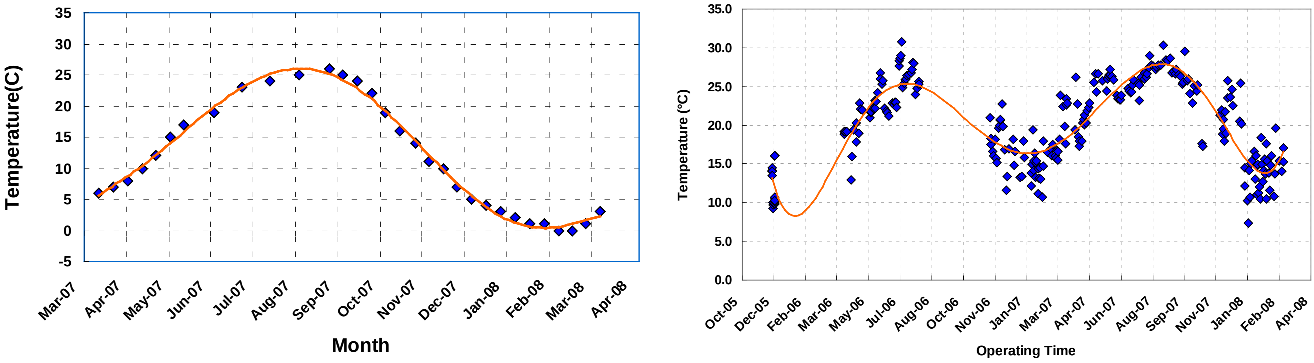

Fig. 5. Temperature of raw seawater (left) and after condenser adjustment (right).

of raw seawater, installed in two construction phases, a first one of 800 and a second phase of 400 m³/h. Based on this input, it could produce 450 m³/h SWRO permeate and 315 m³/h of BWRO permeate. BWRO permeate is fed to an IX unit, which provides high purity water for boiler make up supply.

The plant uses the separation component technology from Dow Water Solutions in the most critical unit operations of UF, SWRO and BWRO. The unit process design information is shown in Table 1.

Construction phase 1 was implemented in 2005. In 2006, after successful commissioning of phase 1, it was determined that it was possible and cost-effective to substitute half of the capacity by the lower cost city water, of which excess capacity became available.

Therefore the plant only produces roughly 400 m³/h of ultrafiltrate, and 130 m³/h of SWRO permeate. Due to this reason, the operation mode that was chosen strongly differs from the design: typically only one of the SWRO trains operates, while all UF units continue to operate at half the flux rate. Obviously, since there is currently no need to identify additional water resources and significant spare capacity is available, the expansion has been canceled.

A detailed process flow schematic is shown in Fig. 6.

Chlorine (NaOCl) is added from an electrolysis bath at a free chlorine concentration of 1 ppm, to control biofouling in the disc filter and ultrafiltration. Then the water is fed to a disc filter, which is used to filter large particles and prevent irreversible damage to the UF membrane. The disc filter (supplier: Amiad), with a 150 μm pore size, using automatic wash at an hour intervals and a design turbidity of 15 NTU.

Despite the high turbidity and chemical oxygen demand (COD) levels, a coagulation step is not used for pretreatment. A coagulation unit was designed and built, however it is not being used by the end-user. The reasons for avoiding this process step are:

1. Despite not using chemically enhanced backwash (CEB), and despite very infrequent (yearly) clean-in-place (CIP) operations in the UF unit, coagulation is not needed to maintain the (50% lower than designed) UF capacity.
2. The low chemicals approach (no coagulation, no CEB, very infrequent CIP) reduces chemicals cost and chemical waste and sludge disposal problems and provides ease of operation and a higher safety level.

The ultrafiltration system includes seven skids each with sixty DOW SFP 2660 UF modules. After the UF process, a break tank is used to collect UF permeate water for backwash supply to the UF system and balance flow to the RO system. More details on the UF unit are described in the following section.

Before the SWRO unit operation, sodium metabisulfite (SMBS, chemical formula $NaHSO_3$) is used to remove re-

Table 1
System information on unit operations in Wang Tan

Unit operations	Total capacity (m³/h)	Capacity per skid (m³/h)	Number of skids	Component installed
Disc filter	800 (+400)	400	2 (+1)	
UF	840 (+360)	120	7 (+3)	DOW SFP-2660 UF Module
1st pass SWRO	300 (+150)	150	2 (+1)	FILMTEC SW30HR LE-400i
2nd BWRO	210 (+105)	105	2 (+1)	FILMTEC BW30-400

Numbers in brackets indicate a possible future expansion in construction phase 2

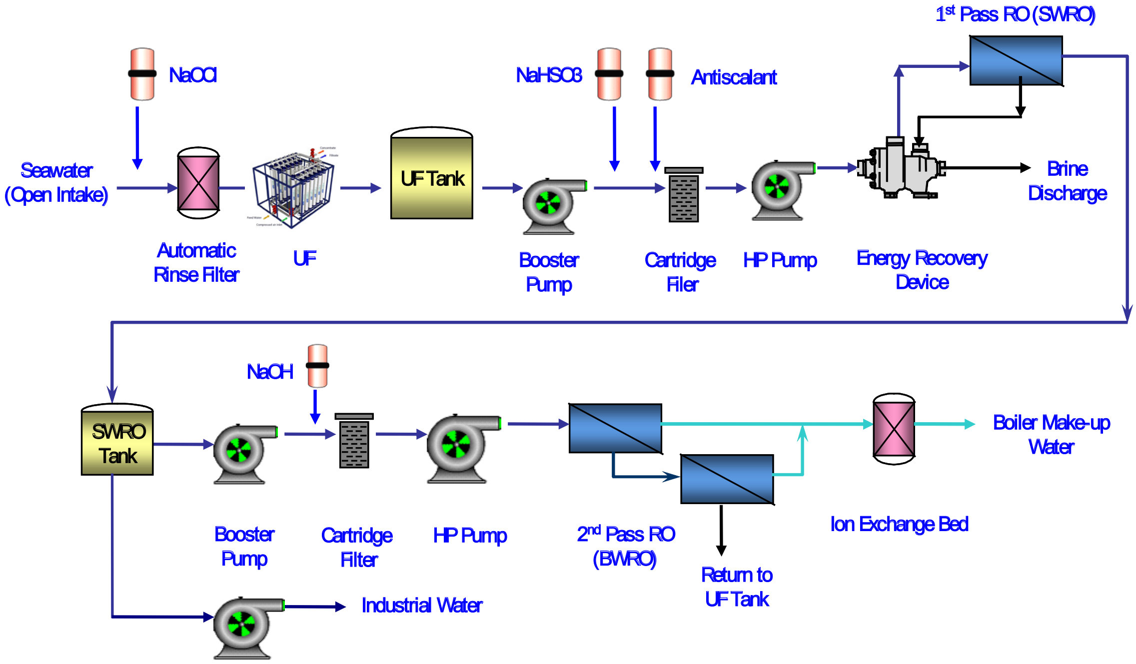

Fig. 6. Treatment system process flow schematic.

sidual NaOCl in the UF filtrate water. An anti-scalant (CP 101) is added to prevent RO membrane scaling. Caustic soda is added to control pH in the RO feed.

The cartridge filters (5 μm polypropylene filters with 60 m^3/h capacity each) provide additional protection to the SWRO high pressure pumps and membranes.

In the seawater section, turbo chargers from PEI are used, model HTC AT1800, with an efficiency in the range of 70–75%.

A portion of the 1st pass SWRO water is used for other industrial purposes. The remaining water goes through a 2nd pass BWRO and IX treatment for boiling water make.

A photo of the main process room shows the 7 UF and 2 SWRO trains that were built in construction phase 1 (see Fig. 9: UF (left) and SWRO trains (right).

3. Ultrafiltration operations and performance

3.1. UF feed and product quality characterization

Water samples were taken at different times within the 2 years of operation and performed by various labs including the Wang Tan power plant laboratory, Dow Biocides Shanghai, and the Nalco Analytical Lab, in Jurong Singapore. Data was routinely collected during the first year of operations and less frequently after that.

Consistent long-term feed and product quality characterization proved to be challenging, due to the lack of local certified laboratories, and the inconvenience of bringing water samples from the remote location to a water lab in BeiJing or ShangHai. Nevertheless an attempt was made to better describe the feed and product water quality.

Fig. 7. Disc filter and cartridge filter.

Fig. 8. Feed pump and turbocharger.

Fig. 9. UF (left) and SWRO trains (right).

3.1.1. UF feed quality

The UF system was designed for a maximum feed water turbidity of 50 NTU. The average water quality conditions for the raw water diverted from the bay area of the Bohai Sea and fed to this plant, estimated from a sampling program in the early phase of operation (late 2005, early 2006), are shown in Table 2.

It can be seen that turbidity, iron, silica and phosphate do not seem too high, but total suspended solids reading is high, so this means a relatively high colloidal fouling tendency. The measurement of organics (oil, TOC, COD) indicate high potential for organic fouling. The COD/TOC ratio* of 1.6 is relatively low, compared to a typical ratio of 2.5–3.3 in most waters [18], which means that organics in this water display high degree of oxidation. A water sample taken in January 2008 showed turbidity of 35 NTU and a TOC of 7.1 mg/L. TOC seems very reasonable, but turbidity seems much higher than the sampling form the early operation period.

In addition to the early sampling program and the 2008 sample, an online turbidity instrument was used. Turbidity is monitored on the feed flow to the UF skids and on the composite from the operating skids. UF permeate turbidity is not measured from individual skids. Turbidity is used routinely to monitor the UF system performance. The data is shown in Fig. 10. In the left chart a longer period is shown, when an online instrument was feeding data to the plant PLC. In the right chart, a period of 15 days at the end of the operation period is shown, in which a service engineer collected data with a hand held turbidity meter.

* This parameter describes the degree of oxidation, or oxygen content of organic matter. Lowest COD/TOC ratio is 0.7 for oxalic acid, and highest ratio is 5.4 for methane.

Table 2
Raw water analysis (average)

Parameter	Avg. raw water	Maximum
Turbidity, NTU	4.3	6.8
TSS, mg/L	60	220
Temperature, °C	14	25
pH	8.1	8.4
Fe^{2+}, μg/L	20	40
Fe^{3+}, μg/L	60	160
Total silica, SiO_2, mg/L	22	64
Total phosphorous, P, μg/L	20	90
TDS, mg/L	36200	36800
Oil, mg/L	1.2	3.2
TOC, mg/L	3.6	7.3
COD ($K_2Cr_2O_7$), mg/L	5.7	10.0

In the left chart, it can be seen that during the first year of operation (November 2005–August 2006), the online instrument indicates a wide scatter (20–100 NTU), while in the second year (August 2006–August 2007), the instrument shows a stable value of 100 NTU. It is unclear if the data indicates a wide variation of feed water composition or if this is a sign of unreliable instrument. For the 2nd year, the instrument is calibrated to a maximum feed turbidity of 100 NTU and readings of >100 NTU are reported as 100 NTU. Overall it seems that the data in the left chart is not as reliable.

In the right chart, data was collected in a 15 day period at the end of the operation period. Turbidity ranges between 3 and 10 and the average is 6 NTU.

If the data for the original sampling program from 2005/2006, the 2008 sample, and the online instrument

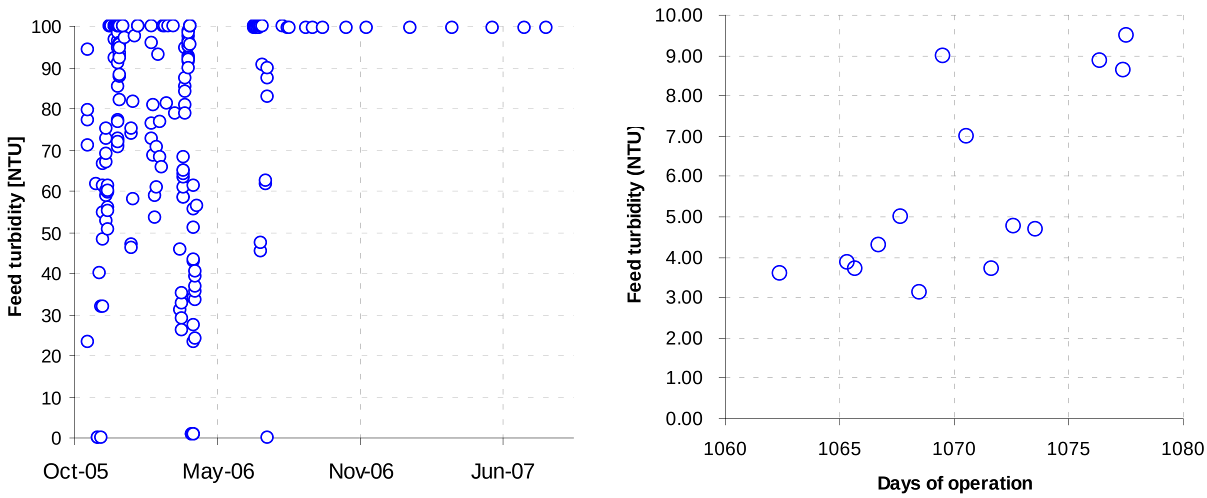

Fig. 10. Feed turbidity.

data are combined, then it seems likely that feed turbidity is higher than the originally assumed 4 NTU.

3.1.2. UF product quality

The water quality requirements from the UF system were turbidity <1 NTU and SDI <3.0. Product turbidity results are shown in Fig. 11. In the chart to the left, UF product turbidity was measured with the same online instrument that showed the above described inconsistent results for the feed turbidity. The data shown in the graph to the right were collected by a service engineer at the end of the operation period using a well calibrated hand instrument.

During start-up and commissioning some very high product water turbidity levels were observed, which then trended down to a still relatively high level of 0–1.5 NTU. Since the same instrument was used, that yielded the very high feed turbidity measurement, the UF product turbidity data shown to the left, should be handled with care. In the graph to the right, the data is more reliable. Except for train 4 (which must have some broken fibers and turbidity was 0.3–0.6), turbidity in the remaining trains 1–3 and 5–7 was in the range of 0.1–0.2.

Regardless of the online turbidity online instrument problems and high readings on UF feed and product with the online turbidity instrument, the turbidity passage rate can be calculated, which compensates for calibration problems (Fig. 12). Again, to the left the online instrument in the beginning period, and to the right the hand calibrated instrument operated by the service engineer in October 2008.

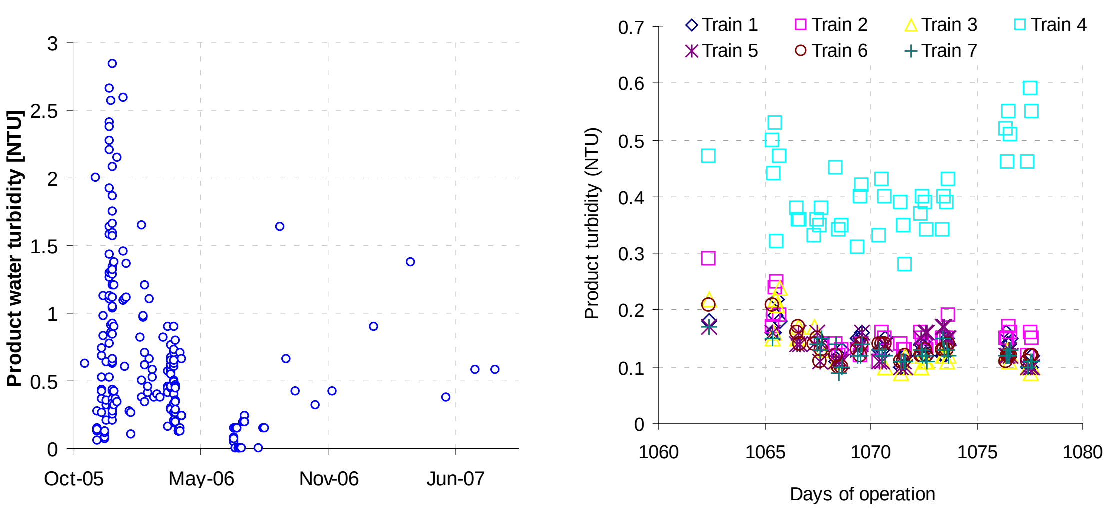

Fig. 11. UF product turbidity by online instrument.

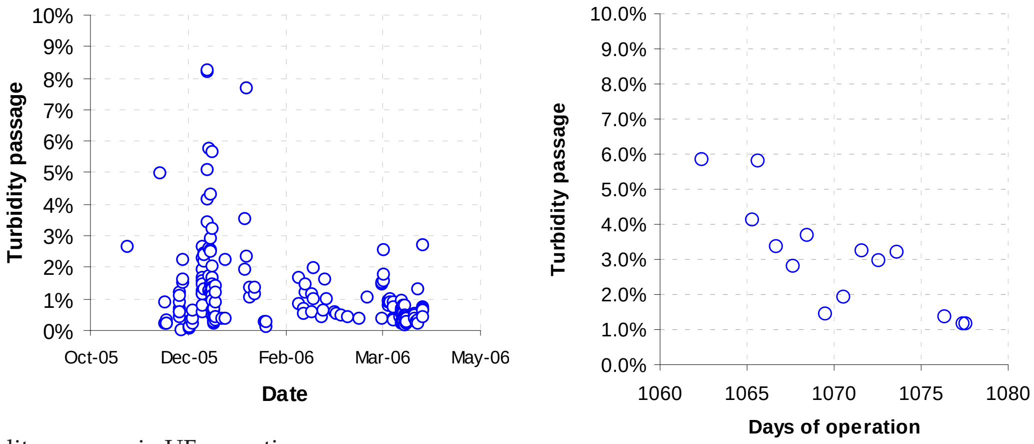

Fig. 12. Turbidity passage in UF operation.

From the turbidity passage rate it can be assessed that the turbidity removal by UF was consistently high, in the range of typically 96–99.5%.

It should also be noted that the turbidity could be significantly improved, if coagulation would be used. Coagulation would agglomerate the smaller suspended solids, which are still passing through the ultrafiltration. It has been widely proven that coagulation with Fe or Al can not only significantly improve outlet quality, but that it can also improve operation performance in terms of lower cleaning frequency. However, due to the good performance of the SWRO, the preference is to follow the "low chemicals" strategy, and it was chosen to avoid the coagulation.

UF product SDI was tracked with some intensity during two periods in the early operation time range, during December 2005 and March 2006, as well as at the end of the operation range. This is shown in the graphs to the left and the right in Fig. 13.

It can be seen that as with UF product turbidity, a strong drop was observed in the initial period. Then, values below 2.5 were observed. After 3 years of operation, SDI was in the range of 2.5–3.5.

In summary it can be seen that, despite not using coagulation, the DOW UF module was able to effectively deal with the wide ranging feed water, and produced a stable RO feed water.

3.2. UF productivity

The UF system had originally been designed to operate at a flux of 60 L/h/m² with a 95 % recovery. Initial operations showed that the skids operated at the design product flow. However, there was an additional source, and water supply was available from the city. Therefore only one of the two available SWRO trains was operated and only half of the UF production capacity was required. Instead of shutting down selected skids and operating only a part of them at full flux rate, the plant operator elected to keep all skids operational and reduce the flux rate on most skids.

As had been mentioned already in the previous chapter, after successful commissioning of the project at full capacity, it was then decided to reduce the desalination system's production to about 50%, due to lower demand. Hence, each train had been designed to produce 120 m³/h,

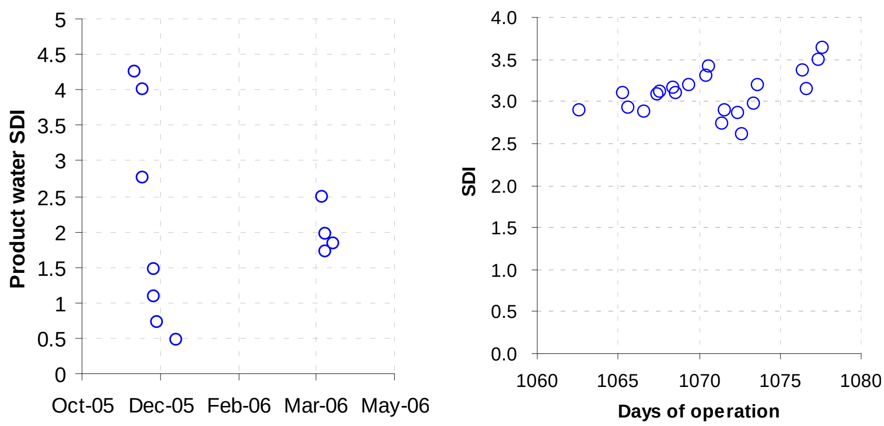

Fig. 13. UF product SDI

but operated at this level only in the first month, before the lower production route was chosen, and production adjusted to 50 m^3/h per train.

In October 2008, a service engineer returned to the site and performed a short test aiming at running at the full capacity for 2 days. The test was successful and the skids produced between 100–120 m^3/h at a TMP of 0.8–1.0 bar. The scatter shown in Fig. 14 comes from more tests performed in the same period in October, which were operated at the plant's typical operation condition of about 50 m^3/h per skid.

The UF production can also be expressed from a membrane flux point of view (Fig. 15).

It can be seen that the UF unit only worked for a

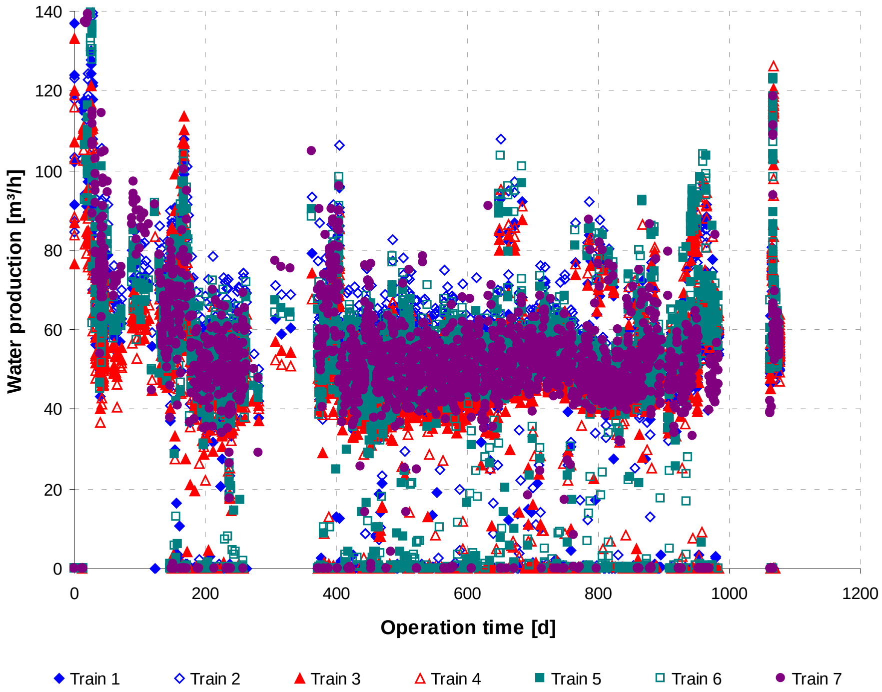

Fig. 14. Water production per UF skid.

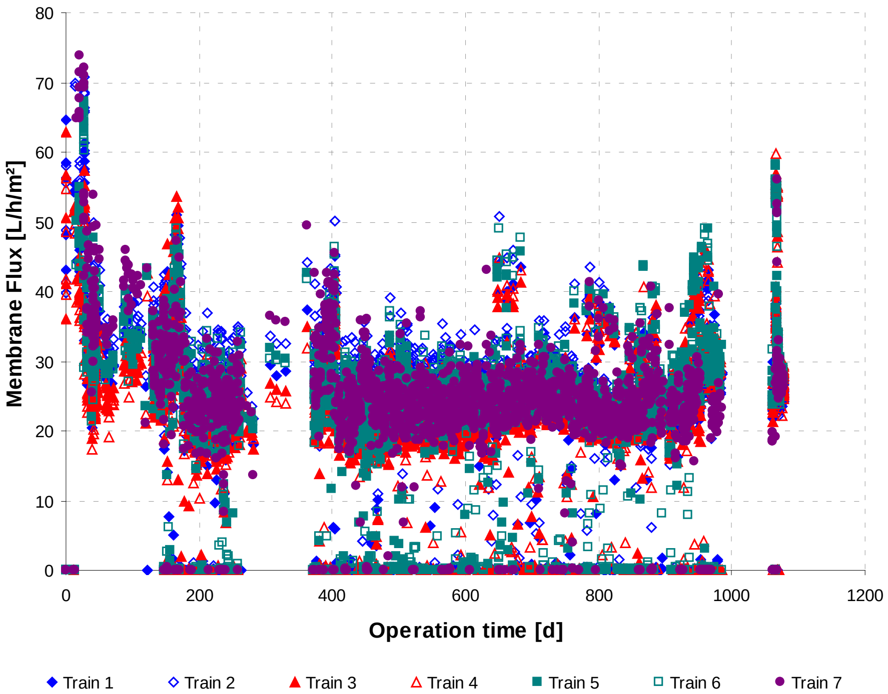

Fig. 15. UF flux.

short time period at the design flux of 60 L/m²/h. For the remainder of the time, it worked at roughly 25 L/m²/h, except for the 2nd test period in October 2008, when it operated again briefly at 50–60 L/h/m².

3.3. Cleaning protocols

Based on the lowered production, the coagulation system was not used, CEB was eliminated and CIP was done very infrequently. As what regards the design of the backwash cycle, the frequency and duration of each operating cycle is shown in Table 3. Chemical usage during each cycle is included.

3 CIP operations were executed, the first in October 2006 (300 d), the second in October 2007 (~650 d) and the third in April 2008 (850 d).

3.4. Transmembrane pressure and permeability

Fig. 16 shows feed, product and transmembrane pressure, which is the difference between the two. In a previous publication, we had shown the feed and product pressure for different trains [17], which was maintained at surprisingly similar levels. This can be explained by the design of the trains, which are fed by one single feed pump from the pump room, via a common feed header to the trains, and product is collected via a common header from all trains to be fed to the common UF product buffer tank. Therefore only one feed, product and transmembrane pressure is shown for all trains. It should be noted that this figure includes two types of pressure measurements: some are made by service engineers on site and they are recorded directly in front of the modules, within the skid (especially in the start period of 0–150 d, and in the October 2008 test at 1080 d). Most of the measurements (150–1000 d) were taken from the main header's feed and product measurements, which are automatically recorded on PLC. The pressures inside the headers includes entrance and exit pressure drop inside the skids.

It can be seen that initially feed and product pressure (range of 1–2 bar) were far higher than the required transmembrane pressure (0.3 bar). It was probably not possible to throttle the feed pump to lower levels, since no frequency transformer was installed. This could have led to running the UF at a larger flux, causing faster foul-

Table 3
UF operating process — backwash conditions

	Filtration	Air scour	Backwash	Forward flush	CEB	CIP
Frequency	56 min	56 min	56 min	56 min	None	8–12 months
Duration	56 min	40-60s	4 min	60 s	None	6 h
Chemical consumption	NaOCl, 0.5 ppm residual		15 ppm NaOCl			Alkaline: 0.05% NaOH, 0.2% NaOCl. Acid: 0.36% HCl

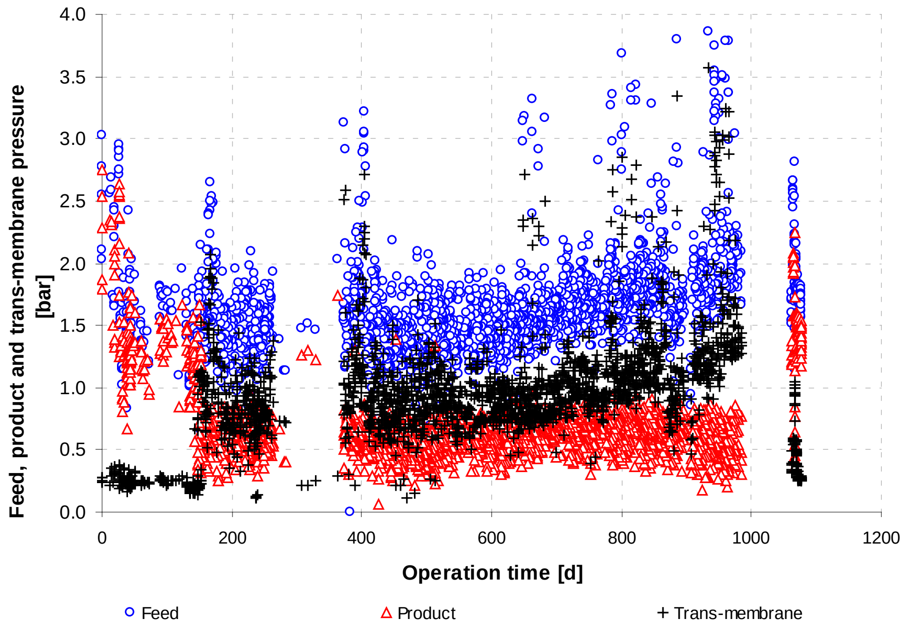

Fig. 16. Feed, product and transmembrane pressure.

ing, hence larger trans-membrane pressure and lower permeability.

The feed water pressures to each skid was measured and along with the skid permeate pressure and was used to calculate the trans-membrane pressure. On the average feed pressure was approximately 1.5 bar. The initial TMP after cleaning was typically in the range of 0.2–0.4 bar. At the point where the measurement of pressures is changed from manual measurement in the skid to automatic measurement in the headers, a TMP increase from ~0.3 to 1.0 bar appears to occur. This is probably due to a higher error in the measurement and to entry and exit losses in the skids. The entry and exit losses to the skids appear to contribute about 0.7 bar to TMP in the period between 150 and 1000 d.

After 8 months of operation and prior to a clean in place cycle, the TMP appeared to rise to 1.0 bar. After the CIP cycle the TMP returned to the values recorded during start up indicating that the membranes were well cleaned and that no irreversible fouling had occurred. Since the time period without CEB and CIP is far longer than the short time periods after CIP, the impact does not last for a long time and usually disappears relatively fast again. It is also possible that, due to the low frequency, the CIP cleanings are not effective anymore. This is often also observed with RO membrane systems, when cleaning is done too late or too infrequently: various harsh cleaning operations would be required to restore permeability. This was not done in this plant, and therefore the permeability remained low even after the occasional CIP cleans.

It is worth noting that the feed pressure was maintained at the same level all the time (around 1.5 bar), and a change in TMP mainly affected product pressure, which was reduced, when TMP increased. Based on the design of the buffer tank, the variation in product pressure did not cause any inconvenience. This is far larger than the theoretically required transmembrane pressure of 0.2–0.3 bar. Nevertheless, designing a UF feed pump at 0.5 bar instead of 2.0 bar would unduly reduce the safety margin, while not providing significant saving, compared to the pressure requirement of the cartridge filters, especially the SWRO unit, but also some of the other transfer operations.

In October 2008 (1080 d of operation) a service engineer visited the site, performed cleanings of the skids and recorded data within the skids. It can be seen that the TMP required was only 0.5 bar, except for the period when the unit operated at 50–60 L/h/m² — in that period TMP required was 1.0 bar.

Despite feed pressure and TMP and hence membrane permeability being of lower importance, a look at the permeability shall be taken in Fig. 17, to assess the impact of the low chemicals approach. The permeability of the membranes on each train was calculated using the flux, membrane area, and trans-membrane pressure. Despite the low temperatures observed in this plant (down to 10°C), the permeability was not corrected for temperature.

Fig. 17 shows as a permeability loss in the UF operation, from initially around 100-250 L/m²/h/bar (average in the range of 150), to the range of 25 L/h/m²/bar, and

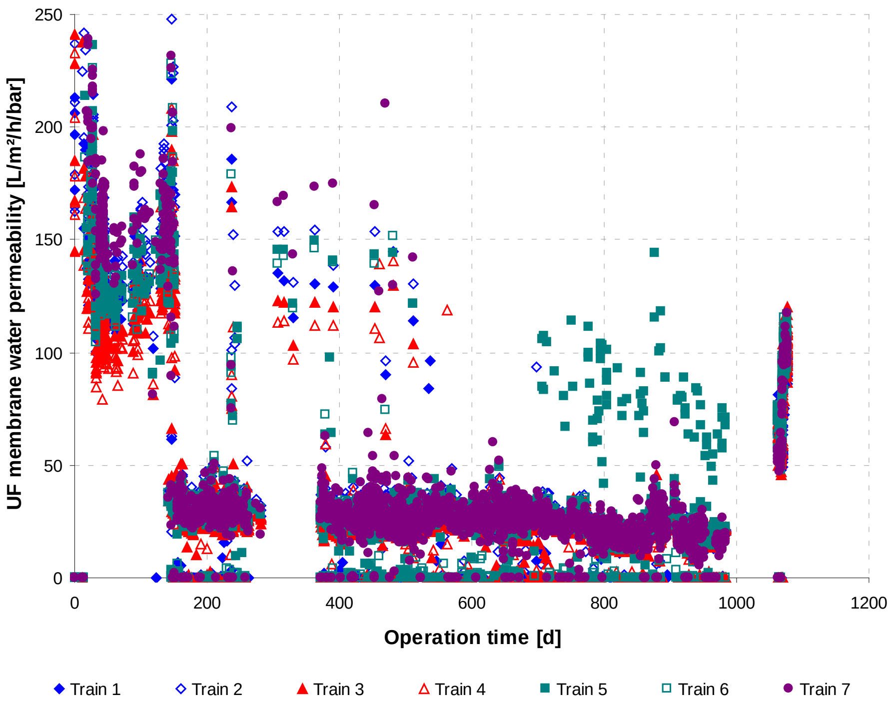

Fig. 17. Membrane permeability UF section.

at the end again a permeability increase to the range of 50–120 L/h/m²/bar.

This loss can be explained as follows:

- A large portion comes from the way that pressures were recorded. The low permeability period (150–1000 d) is equivalent to the period when pressure was recorded in the headers, and not in the skids. This means that entry and exit losses cause much of the low permeability, which is evident, considering they roughly tripled the TMP measurement. This means that it would be more appropriate to consider the triple permeability for the period of 150–1000 d, hence roughly 75 L/h/m²/bar. This is inline with the data recorded in the end period, of 50–120 L/h/m²/bar.
- Most of the reduction from average 150 to average 75 L/h/m²/bar likely comes from the lack of CEB and CIP over long time periods.
- The series resistance model can be used to assess the increase in resistance by fouling. Resistance is defined as the reciprocal permeability. Resistance increases from an initial resistance of 6.7 mbar/(L/h/m²), to 13.3 mbar/(L/h/m²). Hence, the cake layer, which remains on the membrane permanently due to the lack of CEB and CIP operations, is in the range of 6.6 mbar/(L/h/m²) resistance, and contributes most of the resistance.

It can be concluded that the low chemicals approach definitely causes a permeability reduction from 150 to 75 L/h/m²/bar, as well in the range of 0.3 bar additional pressure. An economical calculation should be made to assess the economics of lower chemical consumption compared to the somewhat larger energy consumption (and larger required capital investment for low flux operation).

3.5. Returned modules

During the October 2008 visit, two modules were uninstalled before the cleanings and returned from Wang Tan power plant to the Huzhou Dow laboratory. One of the modules was autopsied as is. Photos are shown in Fig. 18.

Table 4
Foulant analysis

Item	Value
Total ferric (%,as Fe)	5.29
Manganese (%,as Mn)	2.09
Alum (%,as Al)	8.79
Calcium (%,as Ca)	0.57
Magnesium (%,as Mg)	2.29
Total silicon (%,as Si)	2.02
Total organic substance (%)	19.4

It can be seen that there was a significant amount of fouling on the module, and the fouling appeared to be of brown to red color. Foulant was collected from the module, dried (12 h at 105 °C), then burned (680°C, 2 h) and then an inorganic element analysis was carried out. Results are shown in Table 4

The color of the fouling can be explained by the presence of high organic and iron levels. In addition the alum level is very high.

For the other module, an initial permeability test was carried out, which indicated a permeability of 60 L/h/m², which is inline with the observation from the plant. Various cleaning protocols were carried out in order to define the most appropriate cleaning condition (chemical, temperature, duration).

The most appropriate cleaning condition was as follows: oxalic acid 2%, temperature 35°C, circulation time 2 h, then soaking 3 h, then backwashing (air scrubbing 10 Nm³/h, 60 s, backwash flow rate 90 L/h/m² 60 s, forward flushing 60 s). A permeability of 215 L/h/m², which is inline with the permeability of new fibers. The module was then autopsied as well and a photograph is shown in Fig. 19.

Both from the permeability data as well as from the picture it can be seen that the module can be very well cleaned again and the full permeability can be restored.

Fig. 18. Returned module from Wang Tan — autopsy before cleaning.

Fig. 19. Cleaned module after 3 years operation.

4. SWRO unit performance

4.1. SWRO train with FILMTEC™ SW30HR LE-400i

The operational data of one of the 2 SWRO trains was logged, normalized and charted, using the FTNorm normalization program available from Dow Water Solutions. The permeate flow from these trains is shown in Fig. 20.

It can be seen that there were 2 periods in which the SWRO train 2 did not operate. This was due to the reduced water demand, which has previously been described. In the case of the RO system only one train was operated.

Fig. 21 shows the flux in the SWRO section. It can be seen that flux was between 9.5 and 16.5 L/h/m², and data is a bit skewed due to a lower flux at start-up. The median flux was 14.5 L/h/m².

Based on an evaluation with the ROSA (Reverse Osmosis System Analysis) program, it seems that fouling factor after cleanings was 1.0, while on average it was at 0.8. This means that even the very rare cleanings were capable of restoring new membrane performance on this difficult water. This can be contributed to the use of ultrafiltration as a pretreatment to the SWRO membranes, to the reliable performance of DOW ultrafiltration mem-

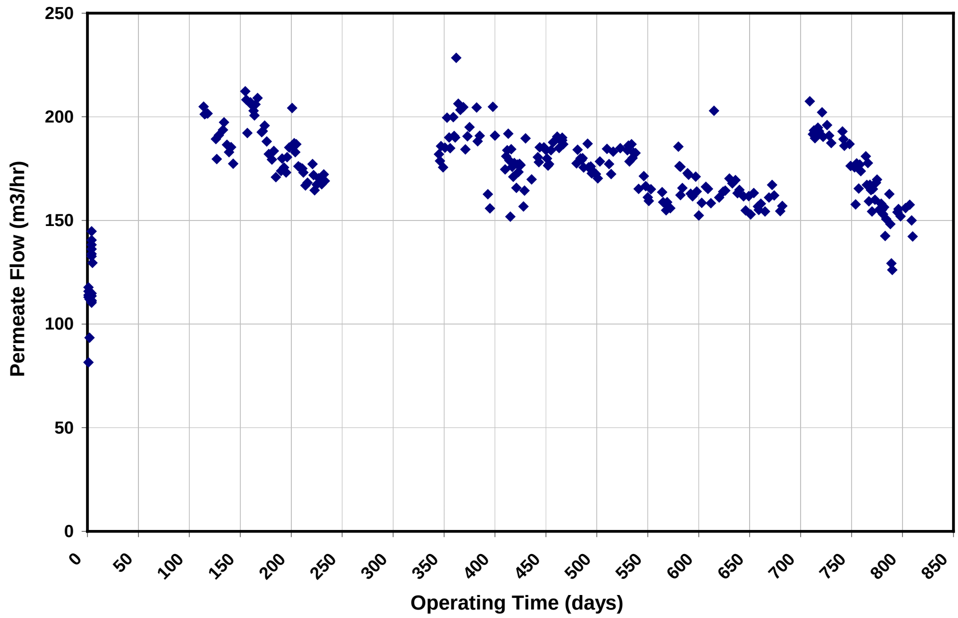

Fig. 20. RO permeate flow over time.

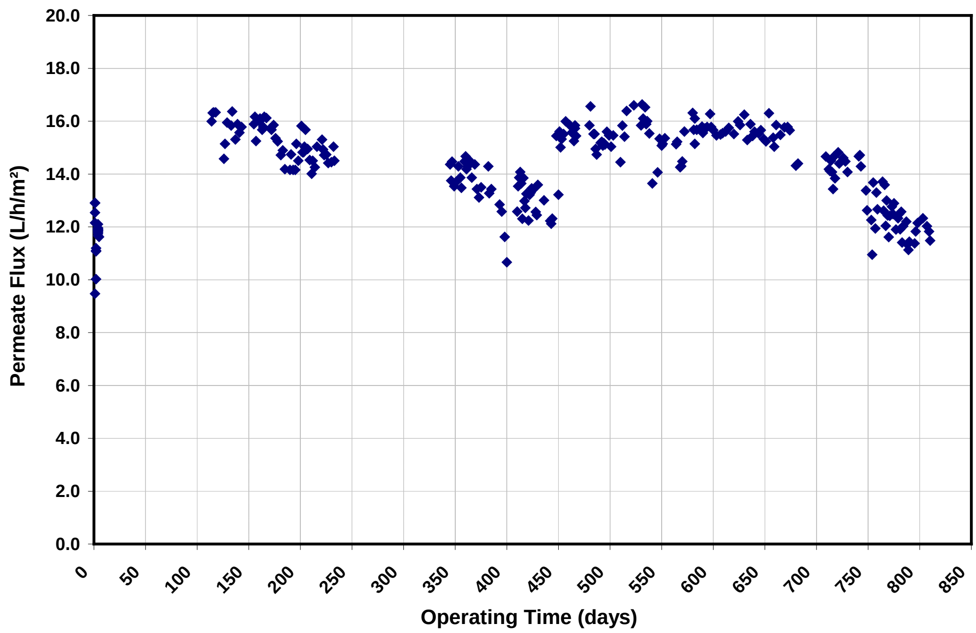

Fig. 21. Permeate flux in full scale plant.

branes, and to the reliable design and performance with FILMTEC™ SWRO membranes.

Pressure drop is shown in Fig. 22. It can be seen that pressure drop after cleanings was in the range of 1–1.5 bar, while at maximum and before cleaning it reached 2.5–3.0 bar. Average pressure drop was 1.7 bar. ROSA predicted pressure drop is 1.6–1.9 bar (depending on feed flow and temperature). This means after cleanings pressure drop was lower than expected, while shortly before cleanings it was higher. On the average, pressure drop was within the expected interval.

Permeate TDS was on average 180 ppm as compared to predicted 145 ppm in the average conditions, hence roughly 25% higher than projected by ROSA. This is still within expectation of the plant operators, but is contrary to the observation that a lot of systems using FILMTEC™ SWRO membranes show better permeate salinity than projected.

In the Wang Tan case, advanced analysis of the SWRO elements was carried out, to improve the understanding of the effect of ultrafiltration pretreatment. Tests and autopsies indicated that the returned SWRO elements were indeed relatively clean, compared to a conventional pretreatment of open intake feed. Gas chromatography–mass spectrometry analysis after extraction, and X-ray fluorescence indicated signs of limited oxidative damage. A potential damage might explain the increase in SWRO permeate salinity, which is unexpectedly 30% higher than predicted by ROSA.

It was unclear, if a potential oxidative damage might have occurred close to start-up, when chlorine was still used in the chemically enhanced backwash (CEB), or if it had occurred due to an upset in the chlorination / de-chlorination system, or due to chlorine use in the backwash. In theory, the SMBS dosing point before the SWRO should have mitigated moderate chlorine doses in the feed. This observation suggests that more research should be done on the potential downstream effect of chlorine used in UF operation, and on identifying options to mitigate or eliminate this need. Vial et al. [19] had presented a backwash protocol, where chlorine was dosed only in the first half of the backwash, which ensured a significantly lower chlorine concentration in the UF fiber product section before return to operation (and hence feeding the RO). As described previously, it was decided eliminate the chemically enhanced backwash (and related chlorine use) and replace this cleaning step with a clean in place (CIP) operation every 8 months. This reduced the risk of a potential negative downstream effect of CEB operations.

In the three years of operation studied, normalized flow was stable between 150 and 200 m^3/h. To sustain the stable productivity, only 3 cleanings were necessary in the SWRO, one after a start-up problem, after 350 days and after 700 days. Only 2 vessels (14 elements of the installed 504) have been replaced (based on Dow wish, in order to do element analysis of returned elements in the Dow labs, and to free up the vessel for pilot testing) within the three years since start-up, hence the replacement rate amounts to a very low 1% per year.

4.2. Pilot trial with internally staged design (ISD) using FILMTEC™ SW30ULE-400i

Pilot trials were started mid 2007, using Dow's 11,000 gpd high productivity seawater membrane FILMTEC™ SW30ULE-400i in internally staged design (ISD) configuration (1 element SW30HR LE-400i, 1 SW30XLE-400i, 5 SW30ULE-400i). This concept relies on enabling higher average flux by a better balance of

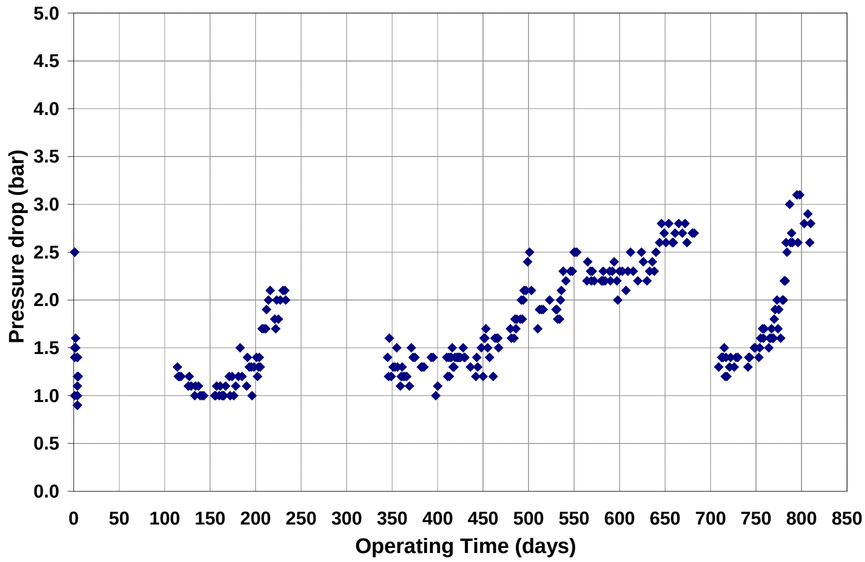

Fig. 22. RO Pressure drop vs. operating time.

individual element flux and recovery. This controls concentration polarization and minimizes fouling despite higher flux operation.

On 20 June 2007, the test vessel with internally staged design was started up. Between 20 June 2007 and 19 March 2008, 6 plant visits were made and data from the test vessel collected. This data is shown in Fig. 23.

The flux rate of the vessel with 7 elements is shown in the following Fig. 24.

The performance in the ISD test vessel, with 4.5–6 m³/h permeate flow and 220–400 μS/cm permeate conductivity compares favorably to the performance of the average SW30HR LE-400i vessel which produces on average 3.5 m³/h at 350–400 μS/cm. Unfortunately it was not possible to measure feed flow or concentrate flow in the ISD test vessel, or to take a concentrate sample. Therefore it was not possible to assess recovery in this vessel, which limits the analysis by ROSA. Due to the lack of recovery data, a detailed analysis of the ISD vessel is not possible and only a range estimation can be made. The fouling factor is definitely larger than 1, and the conductivity is at ROSA prodiction or slightly better.

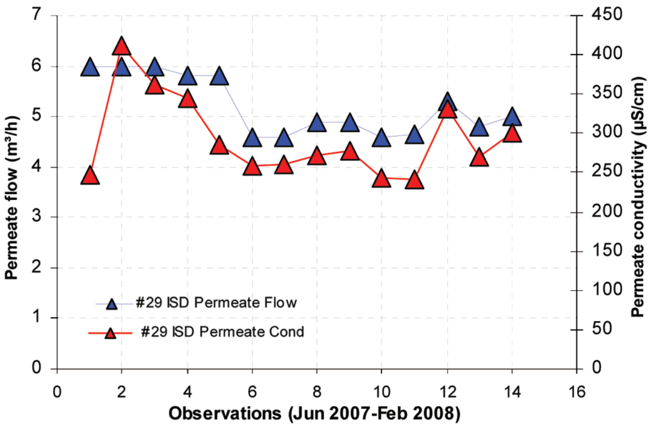

Fig. 23. ISD vessel performance in Wang Tan.

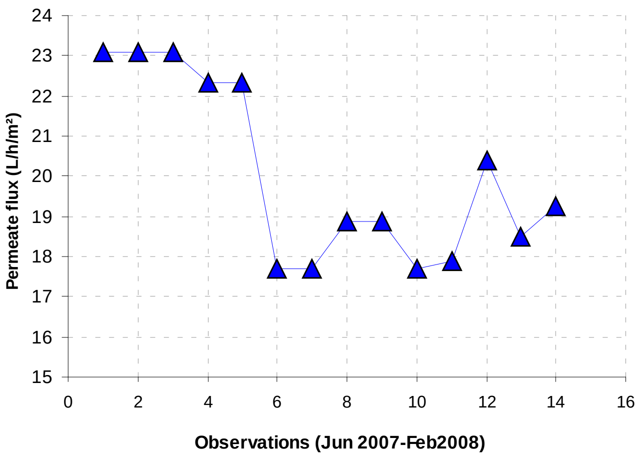

Fig. 24. Flux in the ISD test vessel.

It can be said that the ISD design was definitely performing very stable over the operation period, at very high flux of 18–23 and on average 20 L/h/m². After 9 months of operation, the test vessel has been producing consistently 5–6 m³/h permeate at 300–400 μS/cm, hence roughly 150–200 mg/L TDS. These preliminary results indicate that operation at 20 L/h/m² and possibly beyond is sustainable, when ultrafiltration pretreatment and internally staged design approach are used in combination.

This is very promising and suggests the potential for significant cost savings downstream of a pretreatment system using DOW™ ultrafiltration membranes. Compared to a conventional design of an open intake seawater feed, 10% higher flux could be obtained in the SWRO from the ISD concept, and 20% more flux when ultrafiltration pretreatment is applied. The synergistic nature of both concepts could cause a drastic reduction of SWRO unit cost in the future. However, the pilot tests with the ISD design will need to be continued and evaluated in more detail, to confirm these very positive preliminary results.

5. Summary and conclusions

Three years of operation data for both UF and SWRO section of an industrial scale plant are presented. This enables a thorough assessment of the long term gains of UF pretreatment technology

Despite the high and fluctuating feed turbidity and water temperature, and despite low chemical operation approach (no coagulation, no CEB), the DOW Ultrafiltration Modules were able to manage these challenges and produce desirable feed water for the Wang Tan power plant. No UF module replacement has been done to date. The only pretreatment required for the DOW UF modules is a disc filter. This enables the elimination of coagulation, sedimentation or media filtration unit operations, which is a tribute to the outside-in flow configuration of the UF module.

For the UF operation, a low flux, low chemical, low maintenance approach was used and coagulation and chemically enhanced backwash eliminated, CIP done only yearly. After 3 years of operation, it was still possible to restore full permeability of returned modules in the lab, which was further confirmed by optical inspection of the cleaned fibers after module autopsy. While running at low flux requires more UF modules, and the very low CIP frequency requires slightly higher energy consumption due to the estimated 0.3 bar higher TMP, this approach minimizes chemicals use, sludge and spent solutions disposal, maintenance and safety problems. Therefore this concept represents a somewhat unique but possibly interesting approach.

The SWRO industrial system has been running stable (with slow pressure drop increase and slow permeability loss between CIP operations in the RO) on a very difficult water and required only yearly cleaning and a low replacement rate of 1%/a. Some symptoms of oxidative attack were seen in autopsies, which suggest that chlorine use should be reviewed in the UF operation (and only the first half of the backwash should be done using chlorine, to eliminate chlorine peaks to the RO). It can concluded from the operation that the UF pretreatment has definitely provided a very reliable and safe pretreated water suitable for smooth RO desalination operations.

The combination of ultrafiltration with high productivity FILMTEC elements, especially in ISD configuration promises unprecedented high and stable productivity levels. This could allow capital savings in the range of 20–30% in the SWRO stage but more research in this area is required.

Ultrafiltration has kept the promise of allowing reliable operation of an industrial scale SWRO system on a difficult water for three years, and it has allowed unprecedented productivity at a flux rate of 17–23 L/h/m^2 for 9 month when combined with the ISD approach and high productivity SW30ULE-400i elements. In conclusion, DOW UF modules provide an economical and effective solution for sea water desalination and are able to produce water with acceptable water quality for the downstream operation.

Acknowledgements

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Appendix

1. ROSA projection for train 2

Feed flow to Stage 1, m³/h	302.22	Pass 1 permeate flow, m³/h	135.98	Osmotic pressure:	
Raw water flow to system, m³/h	302.22	Pass 1 recovery, %	44.99	Feed, bar	22.91
Feed pressure, bar	51.75	Feed temperature, °C	20.3	Concentrate, bar	42.59
Fouling factor	1.00	Feed TDS, mg/l	33092.19	Average, bar	32.75
Chem. dose (100% H_2SO_4), mg/l	0.00	Number of elements	252	Average NDP, bar	17.98
Total active area, m²	9364.32	Average pass 1 flux, lmh	14.52	Power, kW	543.14
Water classification: seawater (open intake) SDI < 5				Specific energy, kWh/m³	3.99

Stage	Element	#PV	#Ele	Feed flow (m³/h)	Feed press (bar)	Recirc flow (m³/h)	Conc flow (m³/h)	Conc press (bar)	Perm flow (m³/h)	Avg flux (lmh)	Perm press (bar)	Boost press (bar)	Perm TDS (mg/l)
1	SW30HRLE-400i	36	7	302.22	51.40	0.00	166.24	50.07	135.98	14.52	1.00	0.00	145.07

Pass streams (mg/l as ion)					
Name	Feed	Adjusted feed	Concentrate	Permeate	
			Stage 1	Stage 1	Total
K	380.00	380.00	688.99	2.25	2.25
Na	10000.00	10000.00	18137.40	52.04	52.04
Mg	1300.00	1300.00	2362.08	1.61	1.61
Ca	410.00	410.00	744.97	0.50	0.50
Sr	7.20	7.20	13.08	0.01	0.01
CO_3	6.38	6.38	13.57	0.00	0.00
HCO_3	130.00	130.00	231.63	0.95	0.95
Cl	18000.00	18358.61	33305.11	86.52	86.52
SO_4	2500.00	2500.00	4544.02	1.19	1.19
CO_2	1.84	1.83	3.05	2.05	2.05
TDS	32733.58	33092.19	60040.85	145.07	145.07
pH	7.60	7.60	7.65	5.85	5.85

Permeate Flux reported by ROSA is calculated based on ACTIVE membrane area. DISCLAIMER: NO WARRANTY, EXPRESSED OR IMPLIED,AND NO WARRANTY OF MERCHANTABILITY OR FITNESS, IS GIVEN. Neither FilmTec Corporation nor The Dow Chemical Company assume liability for results obtained or damages incurred from the application of this information. FilmTec Corporation and The Dow Chemical Company assume no liability, if, as a result of customer's use of the ROSA membrane design software, the customer should be sued for alleged infringement of any patent not owned or controlled by the FilmTec Corporation nor The Dow Chemical Company.

Design warnings

-None-

Solubility warnings

Langelier Saturation Index > 0
Antiscalants may be required. Consult your antiscalant manufacturer for dosing and maximum allowable system recovery.

Stage details

Stage 1	Element	Recovery	Perm flow (m^3/h)	Perm TDS (mg/l)	Feed flow (m^3/h)	Feed TDS (mg/l)	Feed press (bar)
	1	0.11	0.92	66.88	8.40	33092.19	51.40
	2	0.10	0.78	87.11	7.47	37178.23	51.13
	3	0.10	0.64	115.55	6.69	41504.80	50.89
	4	0.08	0.51	155.85	6.05	45885.77	50.69
	5	0.07	0.40	213.06	5.54	50099.91	50.51
	6	0.06	0.30	294.35	5.14	53943.88	50.35
	7	0.05	0.22	408.34	4.84	57274.72	50.21

Scaling calculations

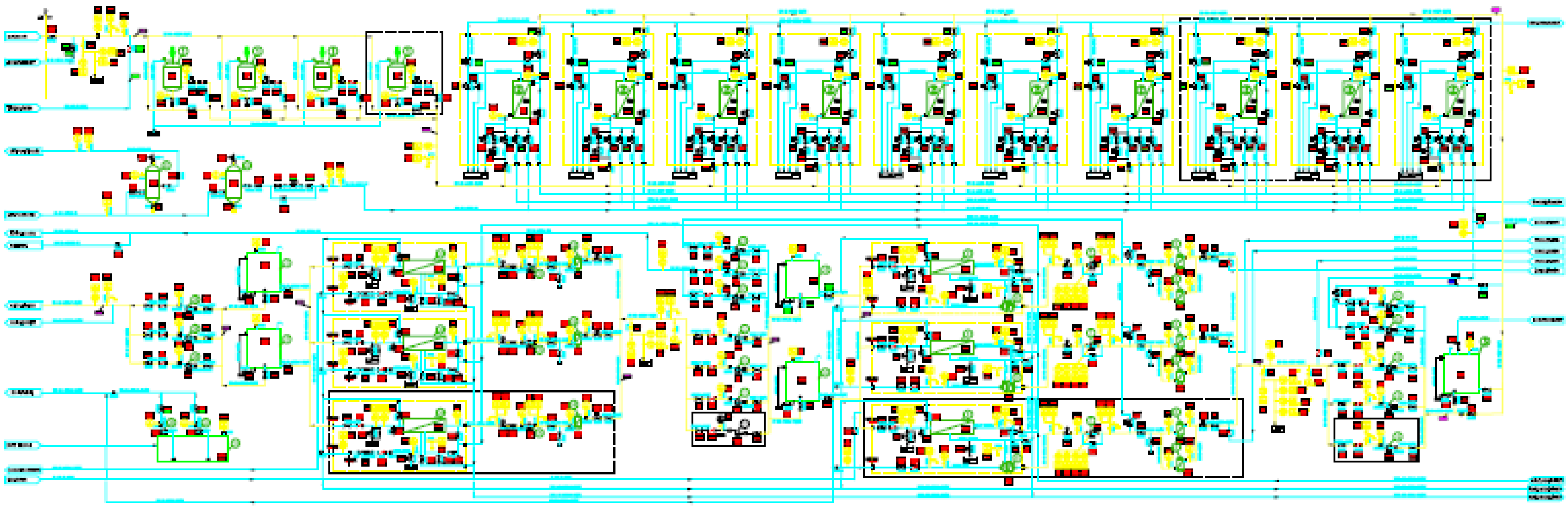
	Raw water	Adjusted feed	Concentrate
pH	7.60	7.60	7.65
Langelier saturation index	0.52	0.52	1.07
Stiff & Davis stability index	–0.43	-0.43	–0.12
Ionic strength (Molal)	0.68	0.68	1.28
TDS (mg/l)	32733.58	33092.19	60040.85
HCO_3	130.00	130.00	231.63
CO_2	1.84	1.84	3.05
CO_3	6.38	6.38	13.57
$CaSO_4$ (% saturation)	19.17	19.17	39.30
$BaSO_4$ (% saturation)	0.00	0.00	0.00
$SrSO_4$ (% saturation)	13.11	13.11	29.76
CaF_2 (% saturation)	0.00	0.00	0.00
SiO_2 (% saturation)	0.00	0.00	0.00
$Mg(OH)_2$ (% saturation)	0.07	0.07	0.16

To balance: 358.61 mg/l Cl added to feed.

2. P&ID

The P&ID of the plant is shown in the following section. Since the plant was designed, built and operated in China, the language in the P&ID is Chinese. It was not possible to translate the P&ID in the framework of this publication.

2.1. Entire plant, overview



Explanations:

- Top left, pretreatment to UF
- Top right, the 7 UF units plus 3 for future expansion
- Bottom left, 2 trains SWRO, plus 1 for future expansion
- Bottom right, 2 trains BWRO, plus 1 for future expansion