State-of-the-art of reverse osmosis desalination

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Abstract
Throughout the world, water scarcity is being recognised as a present or future threat to human activity and as a consequence, a definite trend to develop alternative water resources such as desalination can be observed. The most commonly used desalination technologies are reverse osmosis (RO) and thermal processes such as multi-stage flash (MSF) and multi-effect distillation (MED). In Europe, reverse osmosis, due to its lower energy consumption has gained much wider acceptance than its thermal alternatives. This review summarises the current state-of-the art of reverse osmosis desalination, dealing not only with the reverse osmosis stage, but with the entire process from raw water intake to post treatment of product water. The discussion of process fundamentals, membranes and membrane modules and of current and future developments in membrane technology is accompanied by an analysis of operational issues as fouling and scaling and of measures for their prevention such as adequate cleaning procedures and antiscalant use. Special focus is placed on pre-treatment of raw water and post-treatment of brine as well as of product water to meet drinking and irrigation water standards, including evaluation of current boron removal options. Energy requirements of reverse osmosis plants as well as currently applied energy recovery systems for reduction of energy consumption are described and cost and cost structure of reverse osmosis desalination are outlined. Finally, current practices of waste management and disposal as well as new trends such as the use of hybrid plants, i.e. combining reverse osmosis with thermal processes and/or power generation are addressed.

Keywords: Desalination; Reverse osmosis

1. Introduction
Throughout the world a trend to intensified use of desalination as a means to reduce current or future water scarcity can be observed. Water scarcity, which occurs not only in arid regions, may be characterised as a mismatch between water supply and water demand: Pollution and exploitation of groundwater aquifers and surface water have led to a decrease of quantity and/or quality of available natural water resources in many regions. The ongoing growth of population,
industry and agriculture further increases water demand. In addition, higher living standards, especially in industrial countries, result in higher per capita water consumption and in intensified water scarcity.

The ratio of the average amount of withdrawal to the amount of long term available freshwater resources is called “water stress index”. The intensity of water resources usage in European countries and Israel has been illustrated using this water stress index (Fig. 1). A value of 40% indicates acute water scarcity and a value of 10% is considered as the lower limit of water scarcity [1]. There are, however, a number of factors that are not adequately reflected in the global information of the water stress index. Due to seasonal and local variations, some countries with low water scarcity index, nevertheless, exhibit conditions of serious water stress. An example for this situation is the UK, where in spite of overall ample water availability, (index <0.1) serious summer drought conditions tend to occur in the southeast. On the other hand, a low contamination use such as cooling water taken from rivers, which contributes highly to water abstraction figures in industrialised countries such as Germany, may create much less water stress than agricultural or urban use.

Frequently, part of the water is less accessible or of lower quality. As soon as the convenient water resources are utilised, a hard competition for water arises between agriculture, industry and the public water users. This competition may lead to higher water prices, constricted economic development and social problems in regions with limited water access. As a result, the general welfare of a country under water stress conditions is threatened.

Exploitation of natural fresh water resources combined with higher water demand has led to an increased demand for alternative fresh water resources. Desalination provides such an alternative source, offering water otherwise not accessible for irrigational, industrial and municipal use.

Desalination technologies can be classified by their separation mechanism into thermal and membrane based desalination. Thermal desalination separates salt from water by evaporation.

![Fig. 1. Water stress index for European countries and Israel [1].](image-url)
Fig. 2. Desalination capacities of European countries according to [2]. Only plants with a capacity >700 m³/d are considered.

Table 1

<table>
<thead>
<tr>
<th>Thermal desalination technologies</th>
<th>Membrane based desalination technologies</th>
</tr>
</thead>
<tbody>
<tr>
<td>Multi-stage flash distillation (MSF)</td>
<td>Reverse osmosis (RO)</td>
</tr>
<tr>
<td>Multi-effect distillation (MED)</td>
<td>Nanofiltration (NF)</td>
</tr>
<tr>
<td>Vapour compression distillation (VCD)</td>
<td>Electrodialysis (ED)</td>
</tr>
</tbody>
</table>

and condensation, whereas in membrane desalination water diffuses through a membrane, while salts are almost completely retained. An overview of available desalination techniques is given in Table 1.

Reverse osmosis and multi-stage flash are the techniques that are most widely used. The decision for a certain desalination technology is influenced by feed water salinity, required product quality as well as by site-specific factors such as labour cost, available area, energy cost and local demand for electricity.

A great share of the world’s desalination capacity is installed in the Middle East, and although RO is rapidly gaining market share, thermal processes still dominate the Middle Eastern market due to the low cost of fossil fuel-based energy in this region and due to their suitability to combination with generation of electric energy (cogeneration of steam and electricity). Arid and semi-arid countries in Europe also use desalination as a way to overcome regional water scarcity. As shown in Fig. 2, Spain is the European country which possesses by far the highest desalination capacity. In Europe almost all recently installed desalination plants use reverse osmosis.

2. Desalination overview

2.1. Expected demand for desalination

Desalination markets have expanded in the last decades. The market with the greatest installed capacity is the Gulf Region (Middle East),
where low cost of fossil fuel led to preferred application of thermal desalination processes. The Mediterranean market follows, ahead of the American and Asian markets [3].

The Gulf region will continue to be the greatest market for new desalination equipment because of a rapid growth in population and necessary replacement of over-aged plants. A doubling in capacity until 2015 is expected. The countries around the Mediterranean Sea will experience the largest growth rate with a smaller expansion expected in Asia until 2015 [3].

Asia will become a fast growing market in the long run, due to its enormous population and economic growth that will most likely lead to a water demand that cannot be satisfied with conventional water sources. Fig. 3 shows a forecast of desalination capacity until 2015 in different regions.

2.2. Desalination in Europe

In Europe, desalination capacities are concentrated around the Mediterranean Sea in Spain and Italy (cf. Fig. 2) where desalination is used to overcome water shortage in regions with limited water resources that suffer from intense water demand from tourism and agriculture. For example, on islands like Mallorca, water consumption is very high due to tourism and agriculture, while natural water resources are scarce. Water had and has to be shipped to the island, resulting in prohibitive cost. By building a large desalination plant near Palma, the region was able to reduce
its water price and its dependence on water shipments from the mainland [4].

Water scarcity constrains a society in terms of economic development and thus restricts living standards. This interaction between development and water availability is valid for all sectors, but is most apparent in the tourism industry, which strongly relies on water availability, e.g., on
Lanzarote, between 1986 and 2002 the number of tourists visiting the island has risen from around 10,000 to 1.9 million. This increase in visitors would not have been possible without the drastic increase in water availability due to desalination capacities that were expanded [5] from 450,000 m$^3$/y to 17,210,000 m$^3$/y. Further growth strongly depends upon a continued expansion of desalination.

Seawater desalination capacities in Europe are illustrated in Figs. 4 and 5. Thermal desalination dominated from 1960 till 1985. Then membrane processes took over and have been the technology of choice in most plants constructed until today. Fig. 5 shows currently installed desalination capacities in Europe. A definite trend to membrane based desalination and to larger facilities is apparent. Most plants are installed in coastal regions or on islands of Spain and Italy. Table 2 presents an overview of recently installed RO desalination plants across Europe and of typical capacities.

The trend to RO can be exemplified by the Canary Islands (see Figs 6 and 7). From 1960 to 1985 only a few thermal desalination plants were installed on two islands, while since 1985 the majority of desalination capacities are membrane based and extend to four islands.

Newly installed capacities in Europe display a trend to larger plants as can be seen in Fig. 8. Numerous plants with an individual capacity higher than 50,000 m$^3$/d have been installed during the last years.

Most desalination plants in Europe have been installed in Spain and the overall capacity there will continue to increase during the next years. This tendency is related to a change of policy:

Under the National Hydrological Plan NHP Spain originally intended to use inter basin water transfer to overcome regional water scarcity. In addition to the already existing water transfers and other minor hydraulic works, the plan intended the creation of a new water transfer of 1,050 cubic hectometres per year from the Ebro River to the areas of Catalonia (190 hm$^3$), Comunidad Valenciana (315 hm$^3$), Murcia (450 hm$^3$) and Almería (95 hm$^3$). Except for the 190 hm$^3$ to be used for supplying fresh water to the urban area of Barcelona, the transferred volumes were reserved for irrigation and to improve the ecological quality of water ecosystems subject to severe degradation in the southwest of Spain. Along with the framework for regulation of the public hydraulic domain, the NHP established the management of aquifers, the improvement of the water use for urban supply, the development of awareness-raising activities towards an efficient use of resources and the investment of public money in R&D activities related to water resource management [14].

After a change of government in 2005, Spain stopped the Ebro inter-basin water transfer and instead pushes increased usage of desalination to overcome water shortages. The objectives of the Spanish AGUA program are [15]:

- Rehabilitation of some of the most important Mediterranean ecosystems
- Improvement and modernisation of numerous supply systems in urban areas and irrigation
- Provision of more than 1000 million m$^3$/a of new water resources guaranteeing high water quality.

### Table 2
Recently installed RO desalination plants in Europe

<table>
<thead>
<tr>
<th>Site</th>
<th>Capacity, m$^3$/d</th>
<th>Constr. year</th>
<th>Ref.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Larnaca (Cyprus)</td>
<td>56,000</td>
<td>2001</td>
<td>[6]</td>
</tr>
<tr>
<td>Las Palmas</td>
<td>78,000</td>
<td>1969–2004</td>
<td>[7]</td>
</tr>
<tr>
<td>Carboneras</td>
<td>123,000</td>
<td>2002</td>
<td>[4]</td>
</tr>
<tr>
<td>Bahia de Palma</td>
<td>67,500</td>
<td>2001</td>
<td></td>
</tr>
<tr>
<td>Marbella</td>
<td>55,000</td>
<td>1996</td>
<td>[8]</td>
</tr>
<tr>
<td>Telde</td>
<td>35,000</td>
<td>2001</td>
<td>[9]</td>
</tr>
<tr>
<td>Lanzarote</td>
<td>20,000</td>
<td>2000</td>
<td>[10]</td>
</tr>
<tr>
<td>Canal de Alicante</td>
<td>63,000</td>
<td>2006</td>
<td>[11]</td>
</tr>
<tr>
<td>Cartagena</td>
<td>140,000</td>
<td>2006</td>
<td>[12]</td>
</tr>
<tr>
<td>Dhekelia</td>
<td>40,000</td>
<td>1996</td>
<td></td>
</tr>
</tbody>
</table>
Fig. 6. Seawater desalination plants constructed from 1960 to 1985. (Only plants with a capacity of at least 700 m³/d were considered). Modified from [2].

Fig. 7. Seawater desalination plants constructed since 1985 to the present. (Only plants with a capacity of at least 700 m³/d were considered). Modified from [2].

Fig. 8. New installed capacity and plant size in Europe adapted from [13].
Table 3
Desalination plants in the Spanish Mediterranean basin [15]

<table>
<thead>
<tr>
<th>No. of plants</th>
<th>Total capacity, Mm³/y</th>
<th>Plant capacity range, Mm³/y</th>
</tr>
</thead>
<tbody>
<tr>
<td>Working since 2005</td>
<td>5</td>
<td>152</td>
</tr>
<tr>
<td>Under construction</td>
<td>4</td>
<td>118</td>
</tr>
<tr>
<td>In tender processes</td>
<td>10</td>
<td>263</td>
</tr>
<tr>
<td>April–June 2006</td>
<td></td>
<td></td>
</tr>
<tr>
<td>In public information</td>
<td>4</td>
<td>49</td>
</tr>
<tr>
<td>Drafting stage</td>
<td>3</td>
<td>50</td>
</tr>
</tbody>
</table>

While 15% of the water will come from modernisation of irrigation systems, 20% from water reuse, an additional 15% from better use of superficial and subterranean water as well as improvements in water management, the greatest part will be provided by desalination of brackish water and seawater.

Overall, more than 100 projects will be implemented with a total investment of more than €4,000 million. Various desalination plants are under construction, in tender or drafting stage (Table 3) [15].

All desalination plants will use reverse osmosis and will be powered from the existing electric network. Increased gas emissions will be compensated by increased usage of renewable energies.

2.3. Desalination technology overview

Desalination technologies can be divided into two groups (cf. Table 1). Multi-stage flash distillation (MSF) is the most frequently applied thermal desalination technology and is (still) preferred in the Middle East. Reverse osmosis is the most common membrane based desalination option in seawater and brackish water desalination, dominating in the area around the Mediterranean Sea. Thermal desalination is more energy intensive than membrane based desalination, but can better deal with more saline water and delivers even higher permeate quality [2,3]. Table 4 illustrates key operational data for thermal and membrane based desalination options.

Table 4
Comparison of key operational data of thermal and membrane based desalination technologies [2,16]

<table>
<thead>
<tr>
<th></th>
<th>MSF</th>
<th>RO</th>
<th>Electrodialysis</th>
</tr>
</thead>
<tbody>
<tr>
<td>Thermal energy consumption [kWh/m³]</td>
<td>12</td>
<td>—</td>
<td>—</td>
</tr>
<tr>
<td>Electrical energy [kWh/m³]</td>
<td>35</td>
<td>0.4–7</td>
<td>1</td>
</tr>
<tr>
<td>Typical salt content of raw water</td>
<td>30,000–100,000</td>
<td>1,000–45,000</td>
<td>100–3,000</td>
</tr>
<tr>
<td>Product water quality (ppm TDS)</td>
<td>&lt;10</td>
<td>&lt;500</td>
<td>&lt;500</td>
</tr>
</tbody>
</table>

2.3.1. Reverse osmosis

Reverse osmosis is by far the most widespread type of membrane based desalination process. It is capable of rejecting nearly all colloidal or dissolved matter from an aqueous solution, producing a concentrate brine and a permeate which consists of almost pure water. Although reverse osmosis has also been used to concentrate organic substances, its most frequent use lies in seawater desalination applications.

Reverse osmosis is based on a property of certain polymers called semi-permeability. While they are very permeable for water, their permeability for dissolved substances is low. By applying a pressure difference across the membrane the water contained in the feed is forced to permeate through the membrane. In order to overcome the feed side osmotic pressure, fairly high feed
pressure is required. In seawater desalination it commonly ranges from 55 to 68 bar [17,18]. Operating pressures for the purification of brackish water are lower due to the lower osmotic pressure caused by lower feed water salinity.

A flow sheet of a reverse osmosis based desalination plant is shown in Fig. 9.

The process includes the following stages:
- Water abstraction
- Pre-treatment
- Pumping system
- Membrane separation unit
- Energy recovery system
- Post-treatment
- Control-system

The abstraction (1) of feed water can be realised either through coast- and beach wells or through open seawater intake systems. Coast- and beach wells provide better quality water with less turbidity, algae and total dissolved solids than open seawater intakes [6,19], but require more space. In brackish water desalination, wells are used to abstract feed water. Pre-treatment (2) includes all activities to adjust the intake water in constitution and pH-value. Particulate matter is removed from the feed-water and chemicals are added to prevent scaling and fouling. The pumping system (3) is required to overcome height differences within the distribution chain and to apply the necessary pressure to the feed. The membrane is capable of separating salt from water with a rejection of 98–99.5%, depending on the membranes in use [6]. The energy recovery system is responsible for the transfer of potential energy from the concentrate to the feed. Current energy recovery systems such as work exchangers operate with efficiencies of up to 96% [17]. In post-treatment (6) permeate is re-mineralised, re-hardened, disinfected by chlorination and adjusted to drinking water standards. A control system maintains a continuous and reliable production. Fig. 9 shows a simplified reverse osmosis scheme with energy recovery system and open seawater intake. A typical assembly of the RO stage is shown in Fig. 10.
The main technologies competing with RO desalination are electrodialysis and thermal evaporation techniques such as the multi-stage flash process. These techniques will be shortly presented in the following.

2.3.2. Electrodialysis

For brackish water applications, electrodialysis (ED) is a membrane technology competing with reverse osmosis. It can be used for concentration or removal of charged species in aqueous solutions. ED has been used on an industrial scale since the 1960s.

The process is based on the movement of charged species in an electrical field. Dissolved anions e.g. \( \text{Cl}^- \), \( \text{NO}_3^- \) move towards the anode, while cations (e.g. \( \text{K}^+ \) or \( \text{Na}^+ \)) are attracted by the cathode. The movement of the ions is controlled by ion-selective membranes between the anode and cathode. Anion-exchange membranes are permeable for anions, while cations are held back. Cation-exchange membranes show the opposite behaviour (cf. Fig. 11).

The general principle of electrodialysis as shown in Fig. 12 is known since the 1940s. The electrodialysis stack is divided into several cells by anion (AEM) and cation (CEM) exchange membranes in an alternating sequence. Thus, the concentration of ionic species is reduced in the so called diluate compartments and increased in the concentrate compartments. The basic unit of a stack consists of a pair of a diluate and a concentrate compartment [20].

In the first and last cells of the electrodialysis stack the electric circuit is closed. Depending on the electrolyte and operation conditions, the following electrode reactions can be observed [21]:

- \( \text{NaCl} \) solution, \( \text{pH} < 7 \): Chlorine gas formation
Fig. 11. Ion movement by current [20].

\[
\text{Cl}^- + 2e^- \rightarrow \text{Cl}_2
\]

- Hydrogen gas usually evolves at the cathode:

\[
2\text{H}_2\text{O} + 2e^- \rightarrow \text{H}_2 + 2\text{OH}^-
\]

Fig. 12. ED process principle [20].

Fig. 13 shows a schematic illustration of the specific total process cost over the feed salinity for distillation, reverse osmosis and electrodialysis. For low ion concentration, electrodialysis can offer advantages over RO and is used in the chemical and petrochemical industries to desalinate dilute aqueous or organo-aqueous solutions. Even for brackish waters with up to 3000 ppm salt, electrodialysis is competitive. Due to the higher energy consumption at higher concentration, ED it is rarely used for seawater desalination.

2.3.3. Multi-stage-flash distillation

The most important thermal desalination option is the multi-stage-flash distillation process. MSF is frequently applied in the Middle East where it benefits from low available energy price and high feed water salinity. Multi-stage-flash distillation includes the following stages:

- Water abstraction
- Pre-treatment
- Flash and heat recovery section
- Heating section
• Post-treatment and
• Brine outlet and product water delivery

P-treatment in thermal desalination processes mainly consists of addition of antiscalant and anti-corrosion additives. In the flash and heat recovery sections (3), the feed water, originally at ambient temperature, is led through the different stages, where it is used as coolant, heating up in each stage until it reaches the brine heater (4).

Here it is contacted with steam from the boiler that transfers energy to the preheated water sufficient for partial evaporation in the first pressure vessel, which is under the highest pressure. Non-evaporated feed water passes to the next stage where lower pressure decreases the boiling point. The pressure decreases with each stage as well as the amount of water carried on. The vapor condenses at the feed water pipe, which runs through each vessel, and is collected. After the last stage the brine is discharged. Fig. 14 shows a multi-stage-flash distillation process scheme.

2.4. Application of desalination processes

The proper choice of a desalination technology will depend on the feed water quality, which is mainly characterised by its total dissolved solids content (TDS value). Different feed water qualities and their corresponding salt content are given in Table 5.

Reverse osmosis is the dominant desalination option in Europe. In seawater desalination as well as in brackish water desalination, reverse osmosis makes up about three quarter of total desalination capacity. In seawater desalination, one quarter of total production is processed by thermal desalination technologies such as multi-stage flash distillation and multi-effect evaporators (cf. Fig. 15). In the smaller market of brackish water, membrane based desalination technology is even more prevalent (cf. Fig. 16) [2].

Energy consumption of reverse osmosis is the lowest among all options for seawater desalination, making it most cost efficient in regions with high energy cost [6]. Especially in brackish water desalination, reverse osmosis offers great advantages over thermal desalination technologies due to its much lower energy consumption at low salt concentration [3]. The variable cost of thermal desalination plants is almost independent of feed water salinity, while membrane process variable cost is nearly proportional to the feed water salinity and therefore lower in brackish water than in seawater desalination, making reverse osmosis and electrodialysis the most economic processes [3].

Recovery rates in brackish water reverse osmosis (BWRO) applications are limited by the risk of scale precipitation and are typically in the

<table>
<thead>
<tr>
<th>Feedwater</th>
<th>Minimum salinity TDS [ppm]</th>
<th>Maximum salinity TDS [ppm]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Seawater</td>
<td>15,000</td>
<td>50,000</td>
</tr>
<tr>
<td>Brackish water</td>
<td>1,500</td>
<td>15,000</td>
</tr>
<tr>
<td>River water</td>
<td>500</td>
<td>1,500</td>
</tr>
<tr>
<td>Pure water</td>
<td>0</td>
<td>500</td>
</tr>
</tbody>
</table>

Table 5 Feedwater characterisation by salt content [3]
range of 75%–80%. In seawater reverse osmosis systems (SWRO) recovery rates may at most reach 60% due to limited feed pressure and increased energy consumption at elevated salt concentrations. Feed pressure typically reaches 55–65 bar in SWRO, BWRO desalination uses moderate feed pressures of 10–15 bar [6].

Brackish water desalination is assumed to grow at higher rates than seawater desalination in the near future. Delivery of fresh water from seawater desalination plants demands piping and pumping systems to transport product water from coastal regions to residential areas, which increase cost. High availability of brackish water in residential areas makes expensive delivery piping and pumping unnecessary. One restricting problem in brackish water desalination is the discharge of brine, because disposal options are limited, disposal is associated with high additional costs and environmental damage has to be expected. Nevertheless, brackish water desalination will see a rapid overall growth during the next years [3].

A relatively new field of application of reverse osmosis is its combination with low pressure membrane technologies such as ultrafiltration (UF) or microfiltration (MF) for secondary wastewater effluent purification. By combining RO with UF/MF, water of almost any quality can be produced, as was demonstrated in the Newater
plant in Singapore. So far, water from effluent purification is barely used for direct production of drinking water. Instead, it is generally used in aquifer recharge for indirect drinking water production [22].

The size and number of RO plants for secondary effluent purification are increasing significantly. Some examples of recent or future applications are Sydney Olympic park, Kuwait’s Sulaibiya project with a production capacity of 375,000 m³/d, which uses ultrafiltration and reverse osmosis, the enlarged plant in Orange County, California, with 272,000 m³/d or Singapore, where in 2010, 25% of fresh water demand will be supplied by seawater desalination and water reclamation using reverse osmosis in combination with other membrane technologies [23].

This report will focus mainly on RO applications in sea- and brackish water desalination.

2.5. Basic cost comparison of thermal and reverse osmosis technologies

For all desalination technologies, costs have steadily decreased in the last decades. Generally, thermal desalination is more cost intensive than reverse osmosis desalination. Data on desalination cost of thermal desalination plants is very limited. Zhou and Tol [24] predicted unit production costs by use of plant capital costs, which are more publicly available.

Typical production costs in 2005 for thermal desalination are about 0.65 to 0.90 $/m³. Recent examples for MSF facilities are the Abu Dhabi’s Taweelah desalination plants constructed in 2005 having a water production cost of 0.7 $/m³ and the Shuweihat plant with a capacity of 63,000 m³/d producing water at 0.73 $/m³ [3]. Cost specifications of thermal desalination plants are mostly given on a rudimentary basis and it is not clear if prices for fuel have been included in calculation of water cost.

The strongest decrease in water production cost was achieved for operation of reverse osmosis processes. This strong decrease was due to technological improvements of membranes, economy of scale, improvement of pre-treatment options and the application of energy recovery systems. Still, there is no generally preferable option [4]. Costs for recently installed RO desalination plants are shown in Table 6.

The world’s largest reverse osmosis desalination plant in Ashkelon, Israel, achieves a product water price of 0.53 US$/m³ and water production costs for the Tuas plant, Singapore, for the first time have been below 0.5 US$/m³.

The fact that thermal desalination processes offer advantages in terms of ease of operation, can deal with more saline water and produce better quality water has to be considered in the decision for one or the other option. The cost to produce fresh water from brackish water sources is in the range of 0.2–0.3 $/m³ [6].

3. Introduction to desalination by RO

Reverse osmosis is a (differential) pressure driven separation technique. By applying a pressure difference, the permeating component(s), in most applications nearly exclusively water, are forced through the membrane. Besides its application for production of drinking water, reverse osmosis is also applied in the treatment of effluent water and separation of organic and inorganic compounds from aqueous solution for industrial applications.
Fig. 17 shows separation capabilities of RO and other pressure driven membrane separation processes used in water treatment, namely nanofiltration, ultrafiltration and microfiltration. While microfiltration (MF) and ultrafiltration (UF) membranes are used to remove fine colloidal particles (MF/UF), bacteria (MF/UF), viruses (UF) and larger molecules (UF) such as proteins, nanofiltration and reverse osmosis are able to remove even smaller compounds such as dissolved salts.

Nanofiltration membranes typically reject molecules with a molecular weight higher than 200 g/mole and in reverse osmosis nearly all dissolved compounds are removed from water. The US Environmental Protection Agency has designated RO the best technology for removal of various harmful compounds such as arsenic, barium and nitrate [25].

Pressures applied in reverse osmosis applications vary between 15 bar (brackish water desalination) and 60 to 80 bar in seawater desalination. However, for special applications such as in landfill leachate treatment even pressures of up to 200 bar have been used [20].

3.1. Pressure driven desalination — RO process fundamentals

Osmosis occurs when a semi-permeable membrane (permeable to water and not to the solute) separates two aqueous solutions of different concentration (Fig. 18). At equal pressure and temperature on both sides of the membrane, water will diffuse (“permeate”) through the membrane resulting in a net flow from the dilute to the more concentrated solution until the concentrations on both sides of the membrane become equal. This process will also take place if the pressures on both sides are different, as long as the pressure difference Δp between the concentrated side and the dilute side is not larger than a certain value that depends upon the difference of the respective concentrations and is called the osmotic pressure difference ΔΠ. If the differential pressure Δp is larger than ΔΠ, the direction of flow is reversed and water flows from the concentrate to the dilute side. This process is called RO. In water desalination, the feed side is operated under elevated pressure and the solute concentration on the permeate (dilute) side is negligible compared to the feed concentration. In this case, permeate
flow is observed, as long as the differential pressure exceeds the osmotic pressure $\Pi_F$ of the feed solution.

3.1.1. Osmotic pressure

Thermodynamically, the osmotic pressure is defined as:

$$\pi = \frac{R \cdot T}{V_b} \ln(x_W)$$

with the osmotic pressure $\pi$, the molar volume of water $V_b$, the mole fraction of water $x_W$ and the ideal gas constant $R$.

In dilute solutions, the osmotic pressure can be estimated using van t’Hoff’s law, which is of the same form as the ideal gas law:

$$\pi = -\frac{n_s}{V}RT \quad \text{or} \quad \pi = CRT$$

with the total amount of solutes in solution $n_s$ [moles], total concentration of solutes $C$ [moles/L] and the volume of solvent $V$.

Taking into account non-ideality and dissociation of the ions in solution, van t’Hoff’s law can be rewritten as

$$\pi = i \phi CRT$$

with $i$ representing the dissociation parameter, which is equal to the number of ions and molecules per mole of solute produced by dissolution of the solute and $\Phi$ representing a correction factor that takes into account nonidealities.

For a rough estimate the osmotic pressure of sea water can be calculated assuming an NaCl solution of equal total dissolved solids concentration (TDS). For NaCl at low concentrations the osmotic pressure is approximately equal to [20]:

$$\pi = 8 \text{ bar / wt}\%_{\text{NaCl}}$$

The actual osmotic pressure of seawater is approximately 10% lower than that of an NaCl water solution of equal TDS concentration, due to the presence of higher molar mass species [25].

3.1.2. Models for Description of water and salt flow

It is not the intention of the authors to present a general list of mass transfer models, which have been developed in the course of RO membrane development and application. These are presented elsewhere [20,25]. Instead, only the most commonly used model for prediction of water and salt flows across the membrane, the so called solution diffusion model (SDM) will be shortly presented. For a detailed description we refer to the literature [20].

The SDM is based on the following assumptions:
The active membrane layer is a dense membrane without pores. Permeating components dissolve in the membrane phase.

At all times there is chemical equilibrium at the phase interface between membrane and feed/permeate side.

Salt and water flux are independent of each other.

Salt flux results solely from concentration gradient, but not from pressure.

Due to membrane swelling, water concentration and water diffusion coefficient across the membrane are constant.

According to the SDM, the driving force for permeation of each component can be split into two terms, the concentration or activity difference and the pressure difference between the feed and the permeate sides.

At relatively low salt concentrations, the pressure driving force for permeating salt components is negligible, while, due to the assumption of constant water concentration in the membrane, solely the applied pressure difference $\Delta p$ causes water flux across the membrane.

With these simplifications the salt flux $J^*_{SW}$ and the water flux $J^*_{WW}$ can be estimated:

- water flux:
  \[ J^*_{WW} = A \cdot (\Delta p - \Delta \pi_w) \]

- salt flux:
  \[ J^*_{SW} = B \cdot (w_{s,F} - w_{s,P}) \]

For the process design of an RO plant, usually salt rejection $R_i$ instead of salt flux is used according to

\[ R_i = 1 - \frac{w_{i,P}}{w_{i,F}} = \frac{w_{i,F} - w_{i,P}}{w_{i,F}} \]

where $w_i$ is the salt concentration in the feed (index $F$) and permeate (index $P$). Mass transfer model parameters $A$ and $B$ need to be determined by experiments.

For a GE Osmonics RO membrane, permeate flux and rejection according to the solution diffusion are shown in Fig. 19. Rejection increases with feed pressure and is relatively constant up to a certain value. At higher concentrations rejection strongly decreases and overall efficiency will go down. However, the presented form of the solution diffusion model is not capable of quantitatively describing mass transfer at high concentrations since no interactions between the permeating components are taken into account.

If necessary the extended solution diffusion model (ESDM), which accounts for these interactions, should be used. The ESDM is presented elsewhere [20].

Besides the solution diffusion model, many other models have been developed and applied to
predict mass flows across the membrane. The most important ones are [25]:

- Pore flow models, which assume a porous RO membrane with pore sizes smaller than 1 nm.
- Preferential sorption-capillary flow models, where ions are repelled from the membrane surface and a water layer is formed. Absorbed water flows through capillary pores, while repelled salt ions are rejected and stay in the feed solution.
- Coupled models, which combine different modelling approaches.

3.1.3. Concentration polarisation

Rejection of dissolved and particulate matter by the membrane leads to accumulation of these substances in front of the membrane, with highest concentrations directly at the membrane surface. This phenomenon is called concentration polarisation. A typical concentration profile for the rejected component is shown in Fig. 20.

Depending on particle size, which will determine diffusive back-transport from the membrane, concentration polarisation is more or less distinct. Although concentration polarisation can also be found on the permeate side as indicated in Fig. 20, it is usually neglected in RO since it is much less pronounced than feed side polarisation.

Concentration polarisation has several negative effects on RO performance:

- Rejection decreases due to higher salt flux because of increased salt concentrations at the membrane surface.
- Especially for divalent ions solubility limits can be exceeded, leading to a precipitation layer on the membrane surface, which negatively influences mass transfer.
- Water flux is reduced due to higher osmotic pressure associated with higher salt concentration at the feed side membrane surface.
- Particles are accumulated at the membrane which can lead to cake formation on the surface.

3.2. Limiting factors (fouling, scaling, membrane deterioration)

Real RO processes are not only limited by increasing osmotic pressure due to concentration polarisation and rising overall concentrations along the membrane, but by other factors reduc-
3.2.1. Membrane deterioration

Various chemicals can harm the active layer of the membrane, leading to irreversible damage associated with reduced rejection capability and even destruction of the membrane. Oxidants used in pre-treatment of the reverse osmosis feed water or as cleaning chemicals are the most important group of chemicals responsible for membrane deterioration. Presence of even trace amounts of these compounds may oxidise the membrane surface and damage the active membrane layer. Membrane suppliers therefore give restrictions on exposure to oxidants. In addition, polymeric membranes are more or less susceptible to very low or high pH values. Therefore pH adjustment and control is necessary to ensure stable operation.

Membrane deterioration is addressed in more detail in Section 4.1, Reverse osmosis membranes.

3.2.2. Blocking mechanisms

During operation of a reverse osmosis plant, care has to be taken that no dissolved, colloidal or biologic matter accumulate at the membrane surface, building a continuous layer that reduces or inhibits mass transfer across the membrane. Depending on the mechanisms we distinguish scaling, which denotes precipitation of inorganic material on the surface of the membrane and fouling due to transport of particulate matter to the surface or biological growth on the surface.

**Scaling:** Scaling of the membrane is caused by super-saturation of inorganic compounds concentrated on the feed side. Super-saturated salts can precipitate on the membrane surface building a thin layer, which hinders mass transfer through the membrane. Scaling always occurs at the membrane surface because of the increased salt concentration near the membrane caused by concentration polarisation. Some of the most important scaling substances are CaCO₃, CaSO₄, BaSO₄ and silica. Scaling can dramatically reduce permeate flux, and has to be avoided by all means. Most susceptible to scaling is the downstream part of the RO stage where concentration in the feed solution is the highest.

Even though scaling for some compounds can be removed by flushing the membrane with acid, in practice it is often not possible to transport the crystalline mud out of the module, especially in spiral wound modules. Therefore, pre-treatment is used for stabilisation of substances that could cause scaling. By pH adjustment and use of so called antiscalants precipitation can be inhibited.

Crystal growth is usually divided into three stages as shown in Fig. 22. Antiscalants inhibit one or more of these building stages [20]. Detailed information on compounds responsible for scale precipitation and information on proper chemical treatment can be found in Sections 5.1.1 and 6.1.

**Fouling:** Membrane fouling is caused either by convective and diffusive transport of suspended or colloidal matter or by biological growth, the so called bio-fouling. An existing fouling layer adds to the overall resistance to mass transfer of the membrane and overall
performance decreases significantly. In addition, membrane fouling also increases pressure loss along the membrane, while rejection is decreased.

Particulate fouling in current practice is inhibited by mechanical pre-treatment of the RO feed water by use of screens, sand filtration and cartridge filters or membrane pre-treatment. Biological fouling, caused by microorganisms sticking to the membrane producing a gel like layer, is a serious problem to operation of a RO plant and has to be prevented by chlorination in pre-treatment prior to the actual RO stage.

Fouling can never fully be prevented even with optimised pre-treatment. Therefore, periodical membrane cleaning has to be performed (cf. Section 4.2.2, Membrane cleaning). Complete removal is not possible and fouling has to be tolerated up to a decrease of mass flux down to 75% of original flux [20].

Good operating practice calls for chemical cleaning of the membranes if either normalised permeate flow decreases by 10%, feed channel pressure loss increases by 15% or normalised salt rejection decreases by 10% from initial conditions during the first 48 h of plant operation [27].

Overviews over the different foulants, pretreatment measures for fouling prevention and fouling prediction measures are given in Sections 5.1, Foulants; 5.3, Analytic methods for water characterisation and fouling prediction; and Section 6, Pre-treatment. Cleaning chemicals, frequency of cleanings and cleaning procedure are described in Section 4.2.1, Membrane cleaning.

4. Implementation

4.1. Reverse osmosis membranes

For efficient desalination with reverse osmosis membranes, membranes should in general display high flux and high rejection. High permeability requires very thin membranes, since the flux is inversely proportional to the membrane thickness. Extremely thin membranes today consist of a very thin active non-porous layer and a porous supporting layer for mechanical stability. The support layer protects the membrane from ripping or breaking, while the active layer is responsible for almost all resistance to mass transport and the selectivity of the membrane. Membranes featuring this combination of active layer and supporting structure are called asymmetric membrane. Two asymmetric membranes are shown in Fig. 23.

In the early 1960s the first asymmetric reverse osmosis membranes were produced by Loeb and Sourirajan [29]. These membranes showed up to 100 times higher flux than any symmetric membranes known. This development paved the way
for the commercial success of reverse osmosis. The first commercially available RO membranes introduced into the market in the early 1970s were cellulose-acetate (CA) membranes.

Cellulose acetate membranes are produced by phase inversion, using the miscibility gap of a polymer, a precipitant and a solvent. A film of dissolved polymer (cellulose acetate and acetone) is immersed into a precipitant (water). Acetone is replaced by water and the polymer rich phase starts to precipitate. After complete solidification the membrane is tempered in hot water to remove existing surface defects.

One of the major drawbacks of cellulose acetate membranes is the possibility of membrane deterioration by hydrolysis. In contact with water the acetate membrane becomes damaged according to the reaction:

\[
\begin{align*}
R_1\text{C}=O\bar{\text{OH}} + R_2\text{HO} &\rightarrow R_1\text{C}=O\bar{\text{O}}\cdot R_2 + \text{H}_2\text{O} \\
\text{Acid} &\rightarrow \text{Condensation} & \text{Hydrolysis reaction} &\rightarrow \text{Ether}
\end{align*}
\]

This reaction is strongly dependent on pH and is faster under acidic or alkaline conditions. Fig. 24 shows the dependence of the hydrolysis reaction constant on the pH of the feed water. For pH 4–5 membrane lifetimes can easily reach up to 4 years while for pH 1 and 9 membrane lifetime is in the range of only days. But at pH 6, membranes need to be replaced after 2.5 years. Application of cellulose acetate membranes therefore needs careful pH adjustment and control. In addition, CA membranes under high pressure tend to strongly compact and flux as well as overall performance decreases.

Although cellulose acetate membranes are still in use and commercially available, they are more and more replaced by composite membranes.
Table 7
Influence of pH value on the lifetime of CA membranes [20]

<table>
<thead>
<tr>
<th>pH value</th>
<th>Membrane life time</th>
</tr>
</thead>
<tbody>
<tr>
<td>4–5</td>
<td>4 years</td>
</tr>
<tr>
<td>6</td>
<td>2.5 years</td>
</tr>
<tr>
<td>1 and 9</td>
<td>Couple of days</td>
</tr>
</tbody>
</table>

Composite membranes are made of an active layer from polyamide and a porous support of different material, which offers an additional degree of freedom to design a suitable membrane for each application type. RO composite membranes are produced by interfacial polymerisation, a method developed by Cadotte [30] that enables production of polymer layers below 50 nm thickness.

The supporting layer is usually a micro- or ultrafiltration membrane made of polysulphone. The supporting layer is wetted by a hydrophilic monomer (e.g. 0.7% polyethyleneimide, PEI) and immersed into a hydrophobic monomer (e.g. 0.5% toluenediisocyanate). At the interface the polymerisation takes place until a thin membrane film poses a diffusion barrier and the reaction is terminated.

Composite membranes are chemically and physically more stable, display a strong resistance to bacterial degradation, do not hydrolyse, are less influenced by membrane compaction and are stable in a wider range of feed pH (3–11). However, composite membranes are less hydrophilic and therefore have a stronger tendency for fouling than CA membranes and are deteriorated by very small amounts of free chlorine in the feed stream (Table 7).

All membrane materials available on the market have only limited stability to oxidants used in pre-treatment such as free chlorine or ozone. Residual free chlorine therefore has to be reduced by addition of sodium metabisulfite or contacting with activated carbon (cf. Section 6.1.5).

In separation of organics from water, membrane deterioration can also occur due to the presence of solvents as acetonitrile, vinyl acetate, dioxane, DMF, NMP and a number of chlorinated solvents, substances that usually are not present in drinking water feed streams. Table 8 summarises chemical resistances of the available membrane materials.

Over the last decades, membrane performance has significantly increased with respect to both permeability and salt rejection. In 1981 Cadotte [30] reported flux values of 43 L/(m² day bar) for composite membranes at standard test conditions (32,000 mg/L NaCl, 55 bar, 25°C, 8% recovery), while today’s membranes reach flux values of up to 201 L/(m² day bar) [31]. These values are achieved with brackish water of low salinity, while typical flux values for seawater membranes are significantly lower, even though advances in permeability and salt rejection have also been achieved there. From 1996 to now the rejection of typical seawater membranes increased from 99.6% to 99.8%, while flux increased from 43 L/(m² day bar) to 69 L/(m² day bar) [31]. The historical development of seawater RO membranes is shown in Fig. 25.

4.2 Modules and module staging

4.2.1 Module elements

Early RO desalination plants often used hollow fibre reverse osmosis membranes due to
the optimal membrane area to volume ratio. Feed water passes at the outside and is forced through the membrane, while permeate is processed inside the lumen. Originally, hollow fibre RO membranes were produced by DuPont, who stopped their commitment in RO membrane development and production. Today, this type of module is still available from the Japanese company Toyobo, which also has supplied several recently installed SW desalination plants with their capillary membranes [e.g., Fukukoa (Japan) 2005, Ad Dur (Bahrain) 2005].

However, most membranes to be installed in RO membrane desalination use flat sheet membranes in a spiral wound module (SWM) configuration in spite of its smaller packing density of $<1000 \text{ m}^2/\text{m}^3$. SWMs offer a good balance in terms of permeability, packing density, fouling control and ease of operation.

A summary of advantages and disadvantages of spiral wound modules is shown in Table 9. The basic assembly of an RO spiral wound module (SWM) is shown in Fig. 26. The SWM consists of several flat sheet membranes that are glued together pair-wise on three sides with a permeate spacer in between to form a membrane pocket. Each pocket is connected to the permeate collector tube with its open end. The membrane pockets are rolled around the tube with feed spacers between each pocket. Thus alternating feed and permeate channels are created. The feed enters the module and is partly forced through the membrane. Permeate is collected in the permeate

Fig. 25. Historical development of typical RO membrane elements modified from [31].

Table 9
Spiral wound modules — advantages and disadvantages [20]

<table>
<thead>
<tr>
<th>Advantages</th>
<th>Disadvantages</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cheap and relatively simple production</td>
<td>High feed side pressure loss</td>
</tr>
<tr>
<td>High packing density $&lt;1000 \text{ m}^2/\text{m}^3$</td>
<td>Susceptible to fouling</td>
</tr>
<tr>
<td>High mass transfer rates due to feed spacers</td>
<td>Hard to clean</td>
</tr>
</tbody>
</table>
Fig. 26. Flow through a spiral wound module adapted from [20].

Tube, while retentate leaves the membrane elements on the opposite side of the feed inlet. Typically, single RO elements are operated with a recovery of 5 to 15% [25].

Besides providing a flow path for the feed along the membrane leaf, feed spacers also create eddies, which reduces concentration polarisation and thus increases mass flow through the membrane. By reducing concentration polarisation, feed spacers significantly reduce fouling potential and it was found that spacers may enhance critical flux by a factor of two [31].

However, feed spacers inevitably increase feed channel pressure drop and for feed channels below 0.6 mm, excessive loss of productivity has been found due to a strong decrease in transmembrane pressure difference. On the other hand, lower feed channels increase total packing density. An optimal feed channel height was found to be between 0.6 and 1.5 mm [31].

Flow inside the permeate channel generally is smaller than in the feed side, which calls for smaller permeate channel height. Optimising pressure loss and packing density led to a permeate channel height of 0.25 to 0.5 mm, depending on the width of the membrane leaves. For maximised production rate of one single element an optimum ratio of feed channel to permeate channel height of 0.5 to 1 was found [31].

The current industrial standard SWM element measures 8" in diameter, but 4" and 2.5" versions for laboratories are also available. To minimise piping and reduce the number of pipe connections, 4 to 8 SWM elements are placed in series inside one pressure vessel with a connected permeate collector tube. Standard pressure vessels are commercially available for system pressure up to 65–80 bar.

The 8" standard diameter was more or less arbitrarily chosen. The selection criterion was that one element should be as large as possible, yet small enough to be handled and installed by a single individual.

Due to the modular construction of membrane plants, the current 8" modules offer only a limited economy of scale. Bartels et al. [32] determined the influence of module diameter on cost. Savings from increased module diameter are expected via a reduction of system footprint, number of housings, piping interconnections and seals between the modules.

The operating cost has been calculated for various plant capacities and three different water sources, seawater (TDS 38,000), brackish groundwater (TDS 2,200) and effluent water (TDS 930). Cost reduction in terms of overall life cycle cost of a desalination plant is shown in Table 10. Savings are smallest for seawater and
are expected to amount to 4–7%. For all water sources a major cost reduction was achieved by increasing the element diameter from 8” to 16”. A further diameter increase did not result in substantial cost reduction.

As a result of these studies, a new standard element diameter of 16” was recommended. Unlike 8” membrane elements, 16” diameter module elements with a membrane area of 2800 ft$^2$ cannot be loaded into the pressure vessel by a single worker. Instead, auxiliary mechanical loading tools are needed. However, loading times in terms of membrane area are actually shorter than with traditional 8” elements.

Koch Membrane Systems (KMS) in 2004 introduced the first commercially available large diameter (MegaMagnum™) element with a nominal diameter of 18”. Permeability and rejection of the new element are only slightly lower than of the traditional 8” elements [33]. So far, KMS does not offer its large diameter element for seawater applications due to increased complexity of housing an 18” element at pressures of 69–83 bar [32]. However, for BW applications cost reductions can be expected by reduction of equipment cost, plant footprint and of installation time.

4.2.2. Membrane cleaning

Membrane fouling cannot totally be avoided, even if effective pre-treatment of RO feed water is conducted. Fouling occurs due to mineral scaling, colloidal particles or bio-foulants. The deposition of any fouling layer is accompanied by increased feed channel pressure drop and by decreased permeate flux and reduced salt rejection.

To restore performance and to avoid permanent membrane damage, any existing fouling layer has to be removed by membrane cleaning in defined intervals. Good operation practice calls for chemical cleaning if either normalised permeate flow decreases by 10%, feed channel pressure loss increases by 15% or normalised salt rejection increases by 10% from initial conditions during the first 48 h of plant operation [27]. The typical design of cleaning equipment is shown in Fig. 27.

A typical cleaning procedure consists of the following steps [27]:

- Make-up of the cleaning solution. Depending on system size, this can be done either automatically or by manual preparation.
- Low flow pumping: Pre-heated cleaning solution is processed at minimum pressure head to avoid re-deposition of dirt on the membrane. Almost no permeate is produced. Remaining concentrate from normal plant operation is pushed out of the system and disposed.
- Recycling of cleaning solution: Cleaning solution is recycled to achieve a stable cleaning temperature. If necessary, the pH is adjusted.
- Soaking: Pumps are turned off and depending on the type and intensity of fouling, RO membranes are soaked with the cleaning solution for 1 to 15 hours.
- High flow operation: Foulants are flushed out of the system.
Fig. 27. Membrane cleaning equipment, modified from [27].

- **Flush-out.** Any cleaning solution present in the system has to be flushed out of the system with permeate. Application of raw water might lead to a reaction of its constituents with the cleaning solution.

  The choice of adequate cleaning chemicals and of the pH of the cleaning solution for efficient membrane cleaning is strongly dependent on the fouling conditions. A summary of cleaning solutions for different foulants is given in Table 11.

  **Cleaning by direct osmosis** — A relatively novel procedure for RO membrane cleaning is direct osmosis by feeding of a high salinity solution (DO-HS) to the RO train. A short injection of feed water with increased salt concentrations (25% NaCl solution) with an associated osmotic pressure of 200 bar overcomes feed pump gauge pressure and reverse osmosis shifts to direct osmosis, leading to a permeate backwash stream through the reverse osmosis membrane [34].

  Liberman et al.[34,35] identify four effects that take place within a few seconds after injection of high salinity solution, which together account for effective membrane cleaning:
  - Lifting of an existing fouling layer due to backflow of permeate through the membrane.
  - Sweeping and stripping of the lifted fouling layer due to increased velocities inside the feed channel leading to reduced pressure drop along the pressure vessel.
  - Separation of biomass from the membrane surface due to bio-osmotic shock. Water is sucked out from cytoplasm such as bacteria, fungi and algae, cell membranes shrink and detach from the membrane surface.
  - Dissolving of micro-crystals due to the ionic strength of the DO-HS.

  DO-HS has been implemented at two BWRO plants with new as well as old, silica scaled membranes. DO-HS has been applied once a day, five days a week.

  Performance of old membranes has been significantly improved. Pressure drop was reduced from 6.5 to 3.4 bars, permeate conductivity decreased from 815 to 437 µm and about 4–5 kg
Table 11
Cleaning solutions and their applications modified from [27] and [28]

<table>
<thead>
<tr>
<th>Cleaner</th>
<th>0.10%</th>
<th>1.00%</th>
<th>0.10%</th>
<th>0.03%</th>
<th>0.20%</th>
<th>1.00%</th>
<th>0.50%</th>
<th>1.00%</th>
</tr>
</thead>
<tbody>
<tr>
<td>NaOH</td>
<td>pH 12</td>
<td>pH 12</td>
<td>pH 12</td>
<td>pH 1–2</td>
<td>pH 5</td>
<td>pH 1–2</td>
<td>pH 3–4</td>
<td></td>
</tr>
<tr>
<td>pH</td>
<td>max. 35°C</td>
<td>max. 35°C</td>
<td>max. 35°C</td>
<td>max. 35°C</td>
<td>25°C</td>
<td>25°C</td>
<td>25°C</td>
<td>25°C</td>
</tr>
<tr>
<td>Foulant</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Inorganic salts</td>
<td>++</td>
<td>+</td>
<td>+</td>
<td>+</td>
<td>+</td>
<td>+</td>
<td>+</td>
<td></td>
</tr>
<tr>
<td>(e.g. CaCO₃)</td>
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<td></td>
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<tr>
<td>Sulphate scales</td>
<td>o</td>
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<td></td>
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<tr>
<td>(CaSO₄, BaSO₄)</td>
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<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Metal oxides</td>
<td>++</td>
<td>+</td>
<td>+</td>
<td>+</td>
<td></td>
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<td></td>
<td></td>
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<tr>
<td>(e.g. iron)</td>
<td></td>
<td></td>
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<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Inorganic colloids</td>
<td>+++</td>
<td>+</td>
<td>+</td>
<td>+</td>
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<td></td>
<td></td>
<td></td>
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<tr>
<td>(silt)</td>
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<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Silica</td>
<td>++</td>
<td>+</td>
<td>+</td>
<td>+</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Biofilms</td>
<td>+</td>
<td>++</td>
<td>+</td>
<td>++</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Organic</td>
<td>+</td>
<td>++</td>
<td>+</td>
<td>++</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Na₂EDTA</td>
<td>2-ethyl-acetate</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Na-DSS</td>
<td>Sodium-lauryl-sulphate</td>
<td>+</td>
<td></td>
<td>Very good results</td>
<td></td>
<td></td>
<td></td>
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<tr>
<td></td>
<td></td>
<td></td>
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</tr>
</tbody>
</table>

of fouling debris could be removed from each membrane element. Membrane cleaning by DO-HS of the new membranes ensured stable operation with consistently low pressure drop and permeate conductivity.

Application of DO-HS offers several advantages [34,35]:
- On-line technique without interruption of the operation.
- High effectiveness.
- Low cost.
- DO-HS can be easily implemented at existing facilities and in design of new plants without great changes to plant equipment.
- Simplified disposal due to a reduction of cleaning chemicals.
- Low membrane replacement rate.

4.2.3. Module staging and process design

Typically, six to eight identical spiral wound membrane elements are loaded into one single pressure vessel (PV). The overall plant performance of the plant relies very much on the maintenance of the single membrane elements. By measuring process parameters, normalising data, membrane autopsies and visual inspections, membrane elements can be identified that do not perform adequately and require immediate attention [37]. Three options are available to improve or restore performance of a single pressure vessel to keep design limits of a plant like maximum TDS value:
- Replacement of old membrane elements with new ones.
- Interchange badly performing membrane elements with membranes which have been chemically cleaned.
- Inter-arrange membranes changing the location of the elements inside one pressure vessel [37].

**Internal staged design** — The essential problem with the current plant design which uses
Table 12
Approximate critical fluxes and dependence on particle size adapted from [26]

<table>
<thead>
<tr>
<th>Particle size (µm)</th>
<th>Approximate critical flux (L/m² h)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>39,342</td>
</tr>
<tr>
<td>0.5</td>
<td>39,179</td>
</tr>
<tr>
<td>0.2</td>
<td>39,083</td>
</tr>
</tbody>
</table>

similar membrane elements in the pressure vessels is the difference in production of the membrane elements depending on their location inside the PV. Critical for the design and choice of a membrane is the so-called critical flux design criterion. Critical flux defines the flux at which concentration polarisation leads to severe fouling [36]. Obviously this is a strong function of particle size due to differences in diffusivity and shear stress causing different back-transport of particles of different size. An estimation of critical flux for different submicron size particles is shown in Table 12.

Highest flux along a pressure vessel occurs in the first element due to minimum osmotic pressure. Thus, this first element is most prone to bio and colloidal fouling and critical flux must not be exceeded there. However, flux along the membrane decreases because of increasing osmotic pressure due to increase in salt concentration. Using equal membrane elements causes a reduction of productivity from pressure vessel inlet to outlet.

Recently, a new design approach has been suggested by Mickols et al. [31]. Using a seawater membrane element as a first element and high flux membranes towards the end of a pressure vessel can lead to significantly higher recoveries at the same feed pressure and to large energy savings. Overall, water production cost has been estimated to be reduced by 9–15% on average. However, in this study new experimental high flux membranes have been used, which so far are not commercially available.

Fig. 28 outlines the novel approach. By combining seawater membrane elements with high flux membranes the drawbacks of each type of element could be at least diminished.

**Array configuration** — Desalination plants based on RO membrane technology are usually multiple stage processes. There are three basic plant designs for straight-through operation (cf. Fig. 29), and a selection of a proper design will depend on plant capacity and production requirements [26]. The simplest plant design uses the series array configuration. Several SWM elements are connected in series; as mentioned before, usually 6–8 elements are loaded into one housing. The series array design is limited by the feed fouling potential and restrictions on pressure head loss, which defines the maximum housing length.

For higher plant through-put, multiple housings are used in parallel. If feed side flow rates are significantly reduced by permeation and fall below minimum requirements, the tapered array configuration can be applied to reduce cross-sectional membrane area proportional to decreasing flow rates.

Pressure loss along the housings and increasing concentration on the feed side reduce net-driving force and interstage pumps, so called booster pumps, are needed. The number of parallel housings of a specific stage and element number per housing is mainly based upon the maximum allowed pressure, defined by the membrane manufacturer, the maximum and minimum flow rate through an SWM element and the targeted overall recovery.

Maximum pressure is defined to avoid membrane damage. Membrane compaction is not an issue, since flux in today’s membranes is not affected by membrane compaction below typical operational pressures.

The maximum flow rate determines the number of parallel lines. Higher flow rate would result in excessive pressure loss along the modules.
Fig. 28. Internal staged design adapted from [31].

Fig. 29. Reverse osmosis plant configurations adapted from [26]. Top: series array. Middle: parallel array. Bottom: tapered array.
Finally, a restriction on flow rate in terms of minimum flow is necessary to avoid excessive concentration polarisation, which would cause permeate flux reductions, increased fouling potential and lower rejection rates [26].

Several Windows-based software tools for plant design of RO systems exist. The most commonly used software tools are available from the major RO membrane suppliers and are listed in Table 13. These programs offer the possibility for calculation of one and two staged RO plants after defining feed composition, flow rates as well as plant structure. In addition, scaling and fouling potential can also be estimated. All software tools offered by the membrane suppliers only evaluate plant design using their own membrane elements.

### Table 13

<table>
<thead>
<tr>
<th>Company/institution</th>
<th>Software</th>
</tr>
</thead>
<tbody>
<tr>
<td>Dow</td>
<td>Rosa</td>
</tr>
<tr>
<td>Department of Chemical Engineering, RWTH Aachen University</td>
<td>Rodesign 5.0 (Fig. 30)</td>
</tr>
<tr>
<td>GE Osmonics</td>
<td>Winflows 2.0</td>
</tr>
<tr>
<td>Hydranautics</td>
<td>IMS Design</td>
</tr>
<tr>
<td>Koch Membrane Systems</td>
<td>ROPRO</td>
</tr>
<tr>
<td>Saehan Industries</td>
<td>CSMPRO v2.0</td>
</tr>
</tbody>
</table>

4.3. Energy recovery systems

The application and development of energy recovery systems are a major reason for the decreasing cost for seawater desalination. In general, energy recovery devices (ERD) use the remaining energy of the brine, which otherwise...
Fig. 31. Principle sketch of pressure exchangers [20].

would be wasted, to apply part of the necessary pressure to the feed. Depending on overall recovery and efficiencies of ERD and pumps, this can substantially reduce energy requirements of an RO plant. The energy recovery systems used in reverse osmosis applications can be divided into two groups: Pressure exchangers often referred to as work exchangers, which directly transfer pressure from the brine to part of the feed water, and turbine systems, which mostly are either Pelton wheel or turbocharger systems, which convert potential energy from the brine to mechanical energy either supplied to the feed pump as auxiliary power supply or directly to the feed water. Turbine systems are the older option of the two types of ERD and work at efficiencies of up to 90%.

Pressure exchangers directly transfer pressure from the brine to the feed achieving efficiencies of around 96%-98% [17]. The general principle of pressure exchanger systems is shown in Fig. 31. Feed water is led into a duct, which is then closed by a valve. Another valve opens and gives way to brine at elevated pressure entering the duct, this way pressurising the feed. Feed at elevated pressure exits the duct, mixes with feed from the high pressure pump and is led to the RO stage. A pressure exchanger consists of multiple ducts, which operate in parallel. While the principle is similar for all pressure exchangers, commercially available systems vary in executions and design. The process flowsheet of the RO desalination process using a pressure exchanger systems is shown in Fig. 32. Using pressure exchanger systems, only part of the overall feed needs to be pressurised in the high pressure pump. Due to pressure loss in the RO systems and piping, feed leaving the pressure exchanger needs additional pumping prior to the RO stage.

For turbine systems, the high pressure pump has to deal with the entire feed flow. Turbine ERDs are either the Pelton wheel or the turbocharger system. In Pelton wheel type ERDs the high-pressure concentrate enters the turbine through the inlet nozzle. The high pressure water stream drives the rotor which then produces rotating power to a shaft connecting turbine and high pressure pump, thus assisting the main electric motor in driving the high-pressure pump (cf. Fig. 33). Brine is discharged at atmospheric pressure.

Turbochargers consist of a pump and a turbine section combined in one housing. Both pump and turbine sections contain a single stage impeller or rotor. Hydraulic energy from the brine stream is converted to mechanical energy by the turbine rotor. The pumping section re-converts the mechanical energy back to pressure energy supplied to the feed stream. The process scheme is given in Fig. 34. The process pumping section consists of two steps. At first, all feed is pressurised by high pressure pumps driven by an electric motor to an intermediate pressure level. The feed pressure is then further increased by the turbocharger to the RO stage inlet pressure.

Turbochargers are the dominant technology despite the fact that pressure exchangers offer significant advantages in terms of efficiency.
However there is a general lack of reliable figures on installed capacities for the competing technologies and leading manufacturers of pressure exchanger and turbine systems both claim that their technology has outperformed the competitor in terms of new installed capacities in 2005 [39, 40].

Pressure exchangers require additional auxiliary equipment such as high-pressure circulation pumps. Thus, equipment and maintenance costs of pressure exchangers exceed those of a turbine systems [17]. Other disadvantages of the pressure exchanger systems are increased salinity of the feed due to mixing of brine and feed water, causing higher osmotic pressures in the RO stage.

According to Oklejas et al., published data of ERI, a leading supplier of pressure exchanger mentions increases in salinity by 3 to 5%, which is similar to an increase in osmotic pressure of up to 2 bar depending on overall feed salinity [39].

On the other hand, pressure exchangers maintain a very high efficiency even if membrane recovery is changed or if changes occur due to aging, fouling, or seasonal variation of temperature and salinity, which are often seen with beach well intake systems [39]. Turbine systems suffer stronger reductions in efficiency if operated outside the actual design point.

Higher water recoveries and lower system pressures in brackish water applications limit applicability of energy recovery systems in BWRO. However, even in BW desalination the installation of energy recovery systems might lead to reductions in overall water production cost [41].

The application of energy recovery systems has led to achievable energy consumption as low as 2–4 kWh/m³ in seawater desalination and <1 kWh/m³ in brackish water desalination [6].
5. Raw water characterisation

Beyond the rough distinction between different water qualities shown in Table 5, a more detailed analysis of the feed water is necessary to adjust pre-treatment and operating parameters. As an example of a typical seawater composition, Table 14 exemplarily shows characteristic parameters of Mediterranean seawater for Cyprus and the Canary Islands.

Feed water composition varies depending on local industries’ discharges, water depth, water temperature, ocean currents, algae growth and many more parameters. The salt content in the Mediterranean Sea varies from 3.6% to 3.9% increasing from west to east. However, total salt content is not the only important parameter for membrane desalination. Biomass, dissolved atmospheric gases, concentration of single salts, heavy metals and chemical discharge from industries are relevant to the operation of RO plants. By means of different key parameters the water quality can be characterised. Nonetheless, a detailed analysis with respect to existing foulants of any feedwater is necessary for optimal pre-treatment design.

Table 14
Water characterisation of feed water from the Mediterranean Sea [42]

<table>
<thead>
<tr>
<th>Analysis</th>
<th>Dekhelia, Cyprus [mg/L]</th>
<th>Canary Islands [mg/L]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ca$^{2+}$</td>
<td>450.0</td>
<td>962</td>
</tr>
<tr>
<td>Mg$^{2+}$</td>
<td>1,452.4</td>
<td>1,021</td>
</tr>
<tr>
<td>Na$^{+}$</td>
<td>12,480.0</td>
<td>11,781</td>
</tr>
<tr>
<td>K$^+$</td>
<td>450.0</td>
<td>514</td>
</tr>
<tr>
<td>NH$_4^+$</td>
<td>0.0</td>
<td>0.004</td>
</tr>
<tr>
<td>HCO$_3^-$</td>
<td>160.0</td>
<td>195</td>
</tr>
<tr>
<td>CO$_3^{2-}$</td>
<td>0.2</td>
<td>0</td>
</tr>
<tr>
<td>SO$_4^{2-}$</td>
<td>3,406.0</td>
<td>3,162</td>
</tr>
<tr>
<td>Cl$^-$</td>
<td>22,099.0</td>
<td>21,312</td>
</tr>
<tr>
<td>F$^-$</td>
<td>0.0</td>
<td>1.5</td>
</tr>
<tr>
<td>NO$_3^-$</td>
<td>0.0</td>
<td>2.6</td>
</tr>
<tr>
<td>PO$_4^{3-}$</td>
<td>n/a</td>
<td>0.08</td>
</tr>
<tr>
<td>NO$_2^-$</td>
<td>n/a</td>
<td>0.03</td>
</tr>
<tr>
<td>Total hardness in CaCO$_3$</td>
<td>n/a</td>
<td>6,600</td>
</tr>
<tr>
<td>Total salinity (TDS)</td>
<td>40,498.2</td>
<td>38,951</td>
</tr>
<tr>
<td>Fe$^{2+}$</td>
<td>n/a</td>
<td>0.04</td>
</tr>
<tr>
<td>Al$^{3+}$</td>
<td>n/a</td>
<td>0.001</td>
</tr>
<tr>
<td>pH</td>
<td>8.1</td>
<td>6.33</td>
</tr>
<tr>
<td>Conductivity, $\mu$S</td>
<td>n/a</td>
<td>46,200</td>
</tr>
</tbody>
</table>

5.1. Foulants

Rejected constituents by the RO membrane pose a general fouling risk to plant operation. Foulants can be classified into four categories [43]:

- Chemical foulants, which cause scaling.
- Physical foulants or particulate matter, which are related to deposition of particles on the membrane surface.
- Biological foulants, which can either deteriorate the membrane or form a biofilm layer, which inhibits flux across the membrane.
- Organic foulants, which can interact with the membrane.

5.1.1. Chemical foulants

Scaling of a reverse osmosis membrane occurs if concentrations of sparingly soluble salts, i.e. divalent and multivalent ions exceed their solubility level. Concentrations in the feed channel inside a module increase, and with increasing recovery, the risk of scaling grows. However, solubility levels only define the minimum concentration level at which scaling might occur. In practical operation, even at higher concentrations scaling may not occur due to the long induction times of crystallisation. However, it is common practice not to exceed solubility limits.

Dissolved inorganics most likely to cause scaling are Ca$^{2+}$, Mg$^{2+}$, CO$_3^{2-}$, SO$_4^{2-}$, silica and iron [44]. If solubility limits are exceeded, CaCO$_3$, sulphates of calcium, strontium and barium, CaF$_2$ and various silica compounds are the most likely compounds found as scaling on the membrane.
surface. Hydroxides of Al, Fe and Mn are normally precipitated before contact with the membrane. Most natural surface and groundwaters display high CaCO$_3$ concentrations close to saturation. The scaling tendency of a given feed water is therefore often evaluated using the Langelier saturation index (LSI) for brackish waters and the Stiff and Davis Stability index (S&DSI) for seawaters.

LSI and S&DSI are defined as [45,46]:

\[
\text{LSI} = \text{pH} - \text{pH}_S \quad \text{(TDS <10,000 mg/L)} \\
\text{pH}_S = \text{pCa} + \text{pAlk} + pK_2 - pK_S \\
\text{S&DSI} = \text{pH} - \text{pH}_S \quad \text{(TDS >10,000 mg/L)} \\
\text{pH}_S = \text{pCa} + \text{pAlk} + K
\]

where \( \text{pH}_S \) = pH level at which the water is in equilibrium with calcium carbonate; \( \text{pCa} \) = negative log$_{10}$ of calcium concentration [mol/l]; \( \text{pAlk} \) = negative log$_{10}$ of total alkalinity [mol/l]; \( pK_2 \) = negative log$_{10}$ of ionization constant of HCO$_3$; \( pK_S \) = negative log$_{10}$ of the solubility product of calcium carbonate; and \( K \) = the ionic strength constant at 25°C.

Carbonate scaling can be prevented by pH adjustment. At acidic pH values the equilibrium according to

\[
\text{Ca}^{2+} + \text{HCO}_3^- \leftrightarrow \text{H}^+ + \text{CaCO}_3
\]

is shifted more to the left-hand side and the solubility of calcium ions is increased. Therefore, precipitation of carbonate scaling is usually avoided by maintaining pH of the RO feed at values of 4–6. The acid used for pH adjustment should be of food grade quality [47]. Most common acids for pH adjustment are hydrochloric acid and sulphuric acid. The latter is more easily handled and often more readily available, but poses the risk of increased sulphate scaling due to higher overall sulphate concentration. To control calcium carbonate scaling by addition of acids without additional antiscalants, LSI or S&DSI should be smaller than 2 to 2.5 in order to reduce acid consumption.

Carbonate, sulphate and calcium fluoride scaling can be avoided by addition of antiscalants such as organic polymers, surface active agents, organic phosphonates and phosphates, e.g. polyhexametaphosphate (Calgon), which interfere with the kinetics of crystal nucleation, formation and/or growth. The presence of silica greatly complicates an RO desalting process. Threshold limits of silica scale precipitation are difficult to predict as they are influenced by a large number of parameters. Another difficulty is the lack of a silica anti-scale that can be confidently used to extend water recovery limits. Moreover, silica scales deposited on a membrane are difficult and costly to remove. In the presence of silica it is customary to restrict the recovery limits below the silica saturation limit of about 120 mg/L. Antiscalants may allow operation to a silica concentration of at most 220 mg/L [43].

### 5.1.2. Particulate fouling

Particulate matter in natural waters can be classified according to Potts et. al. [44] into four different categories depending on particle size:

- Settable solids \( > 100 \mu\text{m} \)
- Supra-colloidal solids \( 1–100 \mu\text{m} \)
- Colloidal solids \( 0.001–1 \mu\text{m} \)
- Dissolved solids \( < 10 \text{ Å} \)

Particles larger than >25 μm can be easily removed by various treatment options such as screens, cartridge filters, dual-media filters etc.

The most common inorganic particles are aluminium silicate clays (0.3–1 μm) and colloids of iron, e.g. iron oxide, aluminium oxide and silica. The most problematic feeds are those containing colloidal particles not easily removed by granular beds either because of their minute size or because of electrostatic repulsion effects of the media. In such cases it is necessary to add a coagulant or flocculating agent. The most
frequently used coagulants are ferric chloride, alum and cationic polymers. The polymeric coagulants are effective in very small dosages but can cause membrane fouling difficulties [43].

The presence of suspended solids can be monitored by the silt density index (SDI) test, turbidity analysis, zeta potential measurement and particle counting. For prevention of colloidal fouling, membrane manufacturers require a turbidity NTU <0.2, zeta potential > -30 mV and SDI <3–5 [43]. Waters from wells are much less loaded with colloidal material and often no further reduction of colloidal content is needed.

5.1.3. Biofouling

All raw waters contain microorganisms such as bacteria, fungi, algae, viruses and higher organisms such as protozoa, living or dead, or biotic debris such as bacterial cell wall fragments. At the large membrane surface dissolved organic nutrients of the water are concentrated due to concentration polarisation. Microorganisms entering a RO system therefore find ideal growth conditions resulting in possible formation of a biofilm [47].

Biofilm formation consists of three stages [36]:

- transport to the membrane surface
- attachment to the surface and
- biofilm growth

Biological fouling can influence the performance of an RO system, resulting in increased pressure loss along the feed channel and significant flux decline. Biofouling cannot be easily removed because microorganisms are protected from shear forces and disinfectants by a gel like layer. If the gel layer is partly removed by disinfection, e.g. chlorination, dead cellular matter serves as additional nutrient leading to intensive biologic re-growth. It is therefore essential for good RO performance to reduce biological fouling to a minimum by effective pre-treatment [48].

The potential for biofouling is higher with open intake water than with well water with its higher water quality. Pre-treatment is generally required for surface water to prevent biofouling. For a bacterial count higher than $10^6$ CFU/ml significant biofouling problems have to be expected [48].

5.1.4. Organic foulants

Degradation of organic matter such as plants produces a matrix of macromolecules called humic acids. Organics in natural waters are usually humic substances in concentrations between 0.5 and 20 mg/L in BW and up to 100 mg/L in surface seawater TOC [48].

Humic acids are of polymeric phenolic structure and have the ability to form chelates with metal ions, most importantly with iron ions; a gel like fouling layer is formed by complexation of multivalent ions. The adsorption of these organics on the membrane surface results in permeability decline, which even can be an irreversible process. At high pH values (>9) fouling can be prevented since membrane and organic substances assume a negative charge. The resulting repulsion is used for cleaning of the membranes [48].

Irreversible fouling is mostly due to complexation of calcium. Calcium complexes form a highly compactable floc-like structure which also causes the highest flux decline compared to other chelates.

It was found that mainly the hydrophobic humic substances are deposited on the membrane surface [50] and that the adsorption process is favoured with positively charged, high molecular mass compounds. Similarly, the most hydrophilic membranes have been found less prone to fouling by organic colloids, i.e. humic acids [49].

In RO operation it is recommended that humic acids are removed prior to filter pre-treatment by flocculation, coagulation with hydroxide flocs, ultrafiltration or adsorption on activated carbon. Other organic foulants in natural waters are oil and grease droplets.
5.2. Key parameters

Scaling and fouling are the main problems in seawater and brackish water desalination and the potential for scaling and fouling will strongly depend on the water composition. The abundance of water constituents can never be fully integrated in a process designing exercise. However, water quality can be characterised by certain key parameters which are used to gain information about the tendency of the feed water to cause scaling and fouling.

Parameters that are commonly used to characterise the feed water are:

- The SDI Index (Silt Density Index), which describes the fouling potential of the feedwater and is determined in filtration tests with the feed or raw water using porous microfiltration membranes. The test arrangement and procedure are further described in Section 5.3.1.
- The content of Ca\(^{2+}\) and Mg\(^{2+}\) in the feedwater gives information about the hardness of the water, which causes calcinations of pipes. Total hardness is defined by the content of calcium, magnesium, barium and strontium in solution. However, in general only the amount of calcium and magnesium are used to characterise the hardness of a feedwater.
- The solubility product, which gives information about the limiting salts and the potential for scaling. The solubility product depends on the mole fraction of the salt in water in saturation conditions. In solution the salt dissolves into its ions following the general equation [51]:

\[ A_n B_m \rightleftharpoons n \cdot A^p + m \cdot B^q \]

The solubility product changes with the temperature and is defined by equation [51]:

\[ SP = C_a^n \cdot C_b^m \]

where \( SP \) is the solubility product, \( C_a \) [mole/l] the concentration of the negative ion in saturation condition and \( C_b \) [mole/l] the concentration of the positive ion under saturated conditions. The comparison of ion product of the feed water and the solubility products of ions present allows the determination of the limiting salt and the adjustment of scaling control. The solubility product of selected salt in water at 298° K is given in Table 15.
- The conductivity is directly proportional to the content of dissolved salts in the water and is also used to determine the amount of dissolved salts in the feed. Conductivity of sea water depends strongly on temperature. Therefore, if salinity is to be calculated from conductivity, temperature must be measured at the same time as conductivity, to eliminate the temperature effect and obtain salinity.
- The TDS (total dissolved solids) characterises the content of dissolved solids in the feed, but does not differentiate between different salts.

5.3. Analytic methods for water characterisation and fouling prediction

The most common and accepted tools for prediction of colloidal fouling are the Silt Density Index (SDI) and the Modified Fouling Index (MFI). Both are based on cake filtration on a nominal 0.45 µm Millipore™ microfiltration membrane at constant pressure.

5.3.1. Silt Density Index

The test arrangement and parameters are
Table 16
Definitions and conditions for SDI determination [20]

<table>
<thead>
<tr>
<th>Definition of the Silt Density Index</th>
<th>Condition: $V_1 = V_2$</th>
</tr>
</thead>
<tbody>
<tr>
<td>$SDI_{15(5)} = \left(1 - \frac{t_1}{t_2}\right) \cdot 100 \quad % \min^{-1}$</td>
<td>$V_{1,2} = 500 \text{ ml for } T = 15 \text{ min}$</td>
</tr>
<tr>
<td>$t_1$ Filtration time for Volume 1</td>
<td>$V_{1,2} = 100 \text{ ml for } T = 5 \text{ min}$</td>
</tr>
<tr>
<td>$t_2$ Filtration time for Volume 2</td>
<td></td>
</tr>
<tr>
<td>$(1-t_1/t_2) \cdot 100$ Percentage blocking</td>
<td></td>
</tr>
<tr>
<td>$T_{15(5)}$ Time for testing: Measured from the start of filtration 1 to start of filtration 2: $T_{15} = 15 \text{ min}; T_5 = 5 \text{ min}$</td>
<td></td>
</tr>
</tbody>
</table>

Pressure difference: $p = 2 \text{ bar } \pm 5\%$, Filter: $0.45 \mu m$ pore diameter.
Membrane area: $1350 \text{ mm}^2$.

Fig. 35. Test arrangements for the determination of the SDI [20].

shown in Fig. 35 and Table 16. From a feed tank the raw water is fed through a microporous membrane with a defined membrane area at a constant transmembrane pressure (cf. Table 16). The SDI is determined by comparison of filtration times $t_1$ and $t_2$ of two filtration measurements at standardised intervals $T_{15(5)}$ of 5, respectively 15 min, for a defined filtration volume.

$$SDI_{15(5)} = \left(1 - \frac{t_1}{t_2}\right) \cdot 100 \quad \% \min^{-1}$$

The SDI is used to estimate the potential of fouling caused by fine suspended organic or inorganic colloids. Bio-polymers causing biofouling are not sufficiently held back by the membrane and thus, the bio fouling potential can not be estimated by the SDI test procedure.

From SDI measurements recommendations for pre-treatment can be developed. Membrane manufacturers often demand for stringent SDI values of 2 to 4.

Although the SDI method is widely used in the industry, SDI prediction of fouling is seen controversially. Moody et al. [52] reported fouling problems with an SDI less than 1. On the other hand, Potts et al. [44] reported cases of economic viability with a feed water SDI greater than 5 and Lipp et al. [53] found no clear correlation between the SDI and the fouling behaviour.

According to Yiantsios et al. [54], the SDI is an empirical parameter and a very poor indication of actual RO conditions. Reasons are the permeation rate, which is more than 1000 times higher than at RO membranes and the limited rejection of particles smaller than 0.45 $\mu m$, considering that particles $<0.05 \mu m$ are largely responsible for flux decline in RO membranes. Furthermore, the SDI does not show a linear relationship to the concentration of colloidal and suspended matter.

It must be concluded that a mere SDI measurement is not sufficient to either characterise fouling potential or the quality of pre-
treatment, but that classes of substances that contribute to fouling must be identified and their fate in pre-treatment has to be studied.

### 5.3.2. Modified Fouling Index (MFI)

The Modified Fouling Index (MFI) was developed by Schippers et al. [55]. The intention of MFI was to develop a fouling index which increases linearly with foulant concentration.

The MFI index is measured with the same equipment as the SDI measurement. It is based on the filtered volume, measured every 30 seconds over a maximum time of 20 minutes.

$$MFI = \frac{\eta_{20} \cdot \Delta p}{\eta_r \cdot \Delta p_0} \cdot \tan \alpha$$

where $\eta_{20}$ is the viscosity at 20°C; $\eta_r$ is viscosity at water temperature; $p$ is transmembrane pressure (kPa); $p_0$ is transmembrane pressure as reference at 20°C (kPa); and $\tan \alpha$ is the slope from the linear part of the plot, equal to $d(t/V)/dV$. The ratio of $t/V$ is plotted versus the total filtered volume $V$ as shown in Fig. 36. The linear part of the filtration curve describes the cake or gel filtration and is used for the determination of the MFI as given in the formula.

The MFI index recommendations for acceptable RO/NF operation range from 0–2 s/L$^2$ for RO and 0–10 s/L$^2$ for nanofiltration application [56]. Boerlage et al. [57] found only a poor correlation of the SDI and MFI with colloidal fouling observed at RO and NF installations, and attributed this problem to particles smaller than 0.45 µm. Sung et al. [58] also state only a slightly better, but not very strong correlation with the quality parameters colour, turbidity and TOC for the MFI compared to the SDI. Thus, the MFI displays similar limitations in fouling prediction as the SDI.

### 5.3.3. MFI–UF

Limitations of the SDI and MFI have led to the development of the MFI–UF test. As for the MFI tests, the approach is to establish cake filtration conditions during the test and obtain representative values of the fouling resistance of the cake formed on the membrane [54]. The MFI–UF has been developed to give a more accurate colloidal fouling prediction tool by using ultrafiltration membranes instead of a 0.45 µm microfiltration, thereby retaining smaller particles. Disadvantages are the lower flux and the higher cost of UF membranes compared to microfiltration membranes. Hence, the time needed for the test is significantly increased.

Various membranes have been evaluated for the MFI–UF test. Boerlage et al. [59] identified polyacrylonitrile membranes with pores 1000 times smaller than the MFI microfiltration membrane as adequate for the MFI–UF test. Run times of the MFI–UF test are between 1–5 h and MFI–UF is defined similar to the modified fouling index.

$$MFI - UF = \frac{\eta_{20} \cdot \Delta p}{\eta_r \cdot \Delta p_0} \left( \frac{A}{A_0} \right)^2 \frac{d(t/V)}{dV}$$
where $A$ is the membrane surface area ($m^2$) and $A_0$ is the reference surface area of the MFI membrane with a 0.45 $\mu$m pore size ($m^2$).

MFI–UF values are significantly higher than corresponding MFI values, typically ranging from 2000 to 13,300 s/L$^2$ compared to 1–10 s/L$^2$ for the MFI.

5.3.4. Biofouling prediction

The prediction of biofouling is still at its infancy. Current practice for prevention is a combination of pre-treatment by chlorination and membrane cleaning. In addition, possible biofouling formation is detected in an early stage on test surfaces in the feed stream. These surfaces are periodically examined for attached bacteria. Furthermore, inspections of the cartridge filters and the interior of the feed side piping are performed on a regular basis [48].

The biological fouling potential is related to the concentration of microorganisms in the raw water, feed stream and concentrate channel. In addition, concentration and type of nutrients also determine biological growth of a biofilm.

Vrouwenvelder et al. [60] have evaluated different biomass and growth potential parameters for prediction of biofouling in RO membranes. Adenosinetriphosphate ATP concentration, i.e. a measure for the amount of active biomass, and microscopic total direct cell count TDC, which does not differentiate between active and non-active cellular matter have been used as parameters for the concentration of microorganisms. Parameters analysed for assessment of the concentration of growth promoting substances in water are the concentration of easily assimilable organic carbon AOC and the biofilm formation rate BFR.

The authors showed that severe biofouling was found if BFR values temporarily exceeded 120 pg ATP/(cm$^2$ d) and/or the AOC value exceeded 80 $\mu$g Ac-C/L. For BFR values lower than 1 pg ATP/(cm$^2$ d), no biofouling was detected. However, so far no standardised test exists for biofouling prediction.

6. Pre-treatment

Due to the fouling sensitivity of reverse osmosis units, high quality feedwater is required to ensure stable, long term performance. A sufficient pre-treatment supplying high quality feedwater, regardless of fluctuation of raw water quality, is therefore essential for plant operation. Pre-treatment serves to reduce fouling potential, increase reverse osmosis membrane life, maintain performance level and to minimize scaling on the membrane surface [61]. To characterize the fouling potential of a feed stream, the SDI is generally used. Membrane suppliers recommend SDI$_{15}$ values <3 of the RO feedwater to avoid increased pressure loss in the modules and to minimize fouling. An unreliable pre-treatment system results in high rates of membrane fouling, frequent membrane cleaning, lower recovery rate, reduced membrane life, higher operational pressure and poor product quality. All these effects have a direct influence on operational cost. Therefore, careful consideration has to be given to the pre-treatment system when designing a desalination plant.

Pre-treatment can be divided into two groups: the physical pre-treatment and the chemical treatment. The first is responsible for mechanical filtration through screening, cartridge filters, sand filters or membrane filtration. Chemical pre-treatment includes the addition of scale inhibitors, coagulants, disinfectants and polyelectrolyte [62].

In the past, most reverse osmosis plants used conventional pre-treatment, which is defined as chemical and physical pre-treatment without the use of membrane technologies. Conventional pre-treatment generally uses flocculation, settling, sand filtration and a following cartridge filtration as physical pre-treatment. With declining raw water quality and decreasing membrane costs, in
more projects the use of membrane pre-treatment prior to the reverse osmosis stage is being considered as an alternative to conventional pre-treatment [63,64]. Micro- and ultrafiltration membranes are considerable alternative options and it is estimated that membrane pre-treatment will rapidly grow in the coming years [71].

The extent of pre-treatment depends on the feedwater quality which varies with the location of the plant and the intake system. While for feedwater from well sources cartridge filtration is usually sufficient, feedwater from open seawater intakes demand more extensive pre-treatment. In RO wastewater treatment, membranes already are the state of the art solution in pre-treatment due to advantages in area demand, turbidity fluctuation and fouling control. The extent of membranes in pre-treatment to SWRO is expected to increase [71] because it offers a more efficient pre-treatment at small footprint and decreasing costs [63] (cf. Section 6.4).

The physical pre-treatment usually consists of flocculation and multimedia filtration followed by cartridge filtration, but different set-ups such as laminar settlers, dissolved air flotation or membranes are found in plants around the world. A simplified pre-treatment process scheme is given in Fig. 37. The chemical pre-treatment depends on the physical pre-treatment in use.

6.1. Chemical pre-treatment

Chemical pre-treatment includes any addition of chemicals upstream of the reverse osmosis stage. The kind of physical pre-treatment significantly influences the amount of chemicals that have to be added into the process stream. Membranes in pre-treatment usually require less chemical addition than conventional pre-treatment, which is characterised by a rather high consumption of chemicals [61].

6.1.1. Chlorination

Chlorination is necessary independent of the applied physical pre-treatment to disinfect the water and prevent biological growth which causes fouling of filters and membranes and reduces treatment performance. Chlorine is added to the
raw water as sodium hypochlorite NaOCl or chlorine gas Cl₂, which in water hydrolyses to hypochlorous acid:

\[
\text{Cl}_2 + \text{H}_2\text{O} \rightarrow \text{HOCl} + \text{HCl}
\]

NaOCl + H₂O → HOCl + NaOH

In water hypochlorous acid dissociates to hydrogen and hypochlorite ions:

\[
\text{HOCl} \leftrightarrow \text{H}^+ + \text{OCl}^-
\]

The sum of Cl₂, NaOCl, HOCl and OCl⁻ is referred to as free residual chlorine. For continuous chlorination at the intake point, a free residual chloride concentration of 0.5–1.0 mg/L should be maintained along the pre-treatment line to prevent biofouling [65]. Dechlorination upstream of the RO membrane is necessary to prevent the membrane from oxidation and/or hydrolysis. Continuous chlorination has been the industrial standard for years, but biofouling downstream of the dechlorination point is common. Chlorine is believed to react with organic matter and cut it down to smaller fragments which serve as nutrients for enhanced biological growth at the RO membrane surface where no chlorine is present to prevent this growth. To overcome this problem, periodical shock injection of chlorine with off-line RO stage is conducted. After shock injection all chlorinated feed water has to be rinsed out before operation start-up, ensuring that no chlorinated water reaches the RO membrane.

The germicidal efficiency is depending on the concentration of un-dissociated HOCl, which is 100 times more efficient than the dissociation product OCl⁻. The dissociation equilibrium of HOCl in water changes with pH value, temperature and salinity as shown in Table 17.

Table 17
Concentration of HOCl depending on the pH value, temperature and salinity [65]

<table>
<thead>
<tr>
<th>pH value</th>
<th>Temp., °C</th>
<th>TDS, mg/L</th>
<th>HOCl, in % of residual chlorine</th>
</tr>
</thead>
<tbody>
<tr>
<td>7.5</td>
<td>25</td>
<td>40</td>
<td>50</td>
</tr>
<tr>
<td>6.5</td>
<td>25</td>
<td>40</td>
<td>90</td>
</tr>
<tr>
<td>7.5</td>
<td>25</td>
<td>40,000</td>
<td>30</td>
</tr>
<tr>
<td>7.5</td>
<td>5</td>
<td>40</td>
<td>62</td>
</tr>
</tbody>
</table>

The predominantly created biocide in seawater is therefore HOBr compared to HOCl in brackish water. Since hypobromous acid is a weaker acid than hydrochlorous acid, i.e. it dissociates to a smaller extent, pre-treatment can be performed at higher pH values than in brackish water pre-treatment because more HOBr is present at higher pH values.

In shock dosing operation, the concentration of chlorine and the frequency of dosing seems to depend strongly on the process and the site. Dosing frequencies have been reported to vary from once to 4 times a day with a chlorine concentration of 5 ppm [62]. Another report speaks of ideal shock dosing rates of 2–3 ppm during 2 h a week [61].

As an alternative to chlorination, ultraviolet (UV) radiation is able to disinfect the raw water, but so far is seldomly applied and must be seen problematic due to lack of depot effect.

6.1.2. Coagulation and flocculation agents

Coagulation and flocculation agents are added to cause dissolved matter to adsorb on hydroxides
formed and colloidal matter to agglomerate. In conventional pre-treatment a subsequent sedimentation and sand filtration remove those agglomerates from the feedwater. To increase the agglomerate size, a selection of proper chemicals and dosage is necessary. Iron or aluminium salts are used, sometimes in combination with polymers. Most commonly used coagulants are ferric chloride FeCl$_3$, ferric salts Fe$_2$(SO$_4$)$_3$, or aluminium sulphate Al$_2$(SO$_4$)$_3$. During coagulation alkalinity is reduced and CO$_2$ is produced [62]. The addition of coagulants to the raw water prior to a membrane pre-treatment helps to reduce the fouling potential in the membrane pre-treatment and to provide better feedwater quality to the RO membrane.

6.1.3. pH adjustment

In reverse osmosis applications the pH value is generally shifted to lower values where reverse osmosis membranes show better performance and where cellulose acetate membranes show less hydrolysis. Acids such as sulphuric acid are used to achieve the targeted pH-value. pH adjustment is also necessary to prevent CaCO$_3$ scaling (cf. Section 5.1.1).

6.1.4. Antiscaling agents

Scaling is the precipitation of salts on the membrane surface caused by super saturation. Scaling reduces membrane productivity and water recovery. Various salts can cause scaling and the limiting salt has to be determined using the solubility product (cf. Section 5.2). Depending upon the limiting salt, different scale inhibitors are used. For calcium carbonate scaling the addition of sulphuric acid is usually sufficient. Scale inhibitors may control scaling caused by sulphates, carbonates and calcium fluoride. Due to lower recovery rates in SWRO compared to BWRO, scaling here is less of a problem. Generally, the addition of antiscaling agents is advisable for SWRO systems that work with recoveries greater than 35% [62]. Anti scaling agents are necessary independent of the physical treatment options selected.

Sodiumhexametaphosphate SHMP was commonly used as an antiscalant, but has been widely replaced by polymeric compounds due to eutrophicating properties of SHMP and associated disposal problems.

6.1.5. Dechlorination

Dechlorination has to be performed prior to the RO stage because residual chlorine in the feedwater to the reverse osmosis element may damage the membrane by oxidation. The resistance to chlorine varies depending on the membrane material. Leading membrane manufacturers expect degradation of the membrane after an exposure of 200–1,000 h at 1 mg/L of free chlorine for composite membranes. The rate of chlorine damage to the membrane also depends on the pH value. Degradation is faster in alkaline water than in neutral or acidic water. Higher temperatures also speed up membrane degradation by oxidation.

Commonly, sodium metabisulphite is used for dechlorination due its high cost effectiveness [63,66]. In water it reacts to sodium bisulphite:

\[
Na_2S_2O_5 + H_2O \rightarrow 2HaHSO_3
\]

Sodium bisulphite then reduces hypochlorous acid:

\[
2NaHSO_3 + 2HOC1 \rightarrow H_2SO_4 + 2HCl + Na_2SO_4
\]

In practice, 3.0 mg of sodium metabisulphite is typically used to remove 1.0 mg of free chlorine, in contrast to a theoretically necessary 1.34 mg metabisulphite per 1.0 mg chlorine. Besides sodium metabisulphite, activated carbon is very effective to reduce residual free chlorine. In an activated carbon bed water reacts with carbon and chlorine:
\[ C + 2\text{Cl}_2 + 2\text{H}_2\text{O} \rightarrow 4\text{HCl} + \text{CO}_2 \]

The dechlorination dosing point in the pre-treatment chain has an influence on membrane performance. A study in a desalination plant at Al-Birk on the Red Sea coast observed the influence of the dechlorination dosing point in the pre-treatment chain on bacterial growth and biofouling potential [66]. The pre-treatment of the plant includes sand filtration and cartridge filtration as physical pre-treatment as well as chlorination and the addition of sodium metabisulphite to remove residual chlorine as chemical treatment. Dechlorination prior to the cartridge filtration resulted in clogging of the filters. A shift of the dosing point closer to the RO membrane showed higher biofouling potential of the water but lower clogging of the cartridge filter. The longer chlorine remains in the water, the more nutrients are created by breaking down larger molecules into smaller ones, which enhances the growth of biofouling. Variation of residual chlorine in the range of 0–1 mg/L upstream of the RO membrane showed no significant change in biofouling potential. A residual chlorine content of >0.5 mg/L in the cartridge filtration stage is required to prevent biofilm build-up [66].

6.2. Conventional pre-treatment

While chemical pre-treatment is responsible for pH adjustment, increasing the solubility of salts and disinfection, the physical pre-treatment is responsible for the separation of dispersed particles from the feed water to prevent blocking, fouling and flux decreases in the membrane. Conventional and membrane pre-treatment are the two currently applied pre-treatment options.

Conventional pre-treatment is still the predominantly used pre-treatment option. Conventional pre-treatment demands chemical addition and generally contains flocculation, sedimentation and filtration to mechanically remove colloidal particles and algae. Conventional pre-treatment measures for an SWRO plant usually include the following steps [63]:
- rotating screens for coarse pre-filtration
- chlorination
- acid addition
- coagulation
- flocculants
- single- or double-stage sand filtration
- addition of sodium bisulphite (to remove residual chlorine) and antiscalants
- cartridge filtration (mesh size 5–10 µm)

Conventional filter systems are backwashed with filtered water and air at least once a day. The filter replacement rate varies, depending on the raw water quality and ranges between every two to eight weeks [68].

A single stage sand filtration or even a simple cartridge filtration is able to achieve SDI values well below 3 if the system is fed by water from beach well sources which provide good quality raw water [68]. But when open seawater intakes are used, poor raw water quality during storms or algae bloom can cause problems even in a properly tuned conventional pre-treatment system. Additional disadvantages of conventional pre-treatment systems for the operation of reverse osmosis membranes have been observed [7,63]:
- fluctuations of feed water quality to the RO membrane
- difficulties to supply a constant SDI < 3.0
- difficulties to remove particles smaller 10–15 microns
- large footprint due to slow filtration velocities
- coagulant influences membrane performance

The footprint of a conventional pre-treatment is about 35–40 m²/1000 m³/day permeate [6]. Conventional pre-treatment is generally complex, labour intensive and space consuming [69].

Besides the most commonly used mechanical separation steps of multimedia filtration and cartridge filtration, various other combinations of mechanical separation technologies are known
and applied worldwide. Plate or tube settlers combined with multimedia filtration for physical pre-treatment have reportedly achieved average filtrate SDI$_{15}$ values of 3.0 [67].

Dissolved air flotation (DAF) combined with filtration as physical pre-treatment is another alternative. In this process about 10% of the raw water is taken from the raw water stream, pressurised and saturated with air, before it is released back through nozzles or valves into the raw water leaving the flocculation chamber. The sudden pressure release forms micro bubbles of about 60 microns in size onto which preformed flocs and particulate matter attach and are carried by the bubbles to the surface. DAF can achieve effluent turbidity <0.5 NTU, effectively removes high concentration of algae and shows advantages in treating very cold raw water [67].

In a study on pre-treatment of the reverse osmosis desalination section of the hybrid plant at Al Fujairah (UAE), conventional pre-treatment showed a reduction of SDI$_{15}$ values from 12–19 in the raw water to SDI$_{15}$ 2–4.5 after pretreatment and a turbidity reduction of up to 75%. This was achieved through chemical disinfection and flocculation, multimedia sand filtration and cartridge filtration [61].

In the RO stage about 170,000 m$^3$/d of permeate is produced at a recovery of 40%. The chemical dosing rate and pre-treatment effects on the filtrate quality are presented in Tables 18 and 19.

To prevent biological growth, sodium hypochlorite (NaOCl) was shock injected once a week at 2–3 ppm for two hours. Coagulation agents in the form of ferric chloride were added at a dose ranging from 3 to 5 ppm [61].

Costs for chemicals used in pre-treatment are reported to be in the range of 4.8–5.7% of total O&M costs. Chemical costs are 0.03 $/m$^3$ permeate, i.e., in agreement with reported values. Energy consumption for conventional pre-treatment in this case study amounts to 0.416 kWh/m$^3$ filtrate. Assuming a recovery rate of 40%, the energy consumption is 1.04 kWh/m$^3$ permeate, representing 20.2% of the total energy consumption of the reverse osmosis section of the plant [61].

### 6.3. Membrane pre-treatment

Microfiltration and ultrafiltration membranes have been successfully applied in pre-treatment of much more difficult raw waters than seawater, such as in industrial and municipal wastewater for many years. Both membrane types form a barrier against suspended particles, colloidal materials and bacteria. Therefore they guarantee a low SDI value of the RO feed water; even with strong fluctuation of raw water quality, enabling operation with a high and stable permeate flux even in long term operation [7,63,64].

In a process using membranes in pre-treatment, the raw water is usually roughly pre-filtered by a mechanical screen before it is fed to

### Table 18

**Chemical dosing [61]**

<table>
<thead>
<tr>
<th>Chemicals</th>
<th>Concentration, ppm</th>
</tr>
</thead>
<tbody>
<tr>
<td>Chlorine</td>
<td>3</td>
</tr>
<tr>
<td>Ferric chloride</td>
<td>3</td>
</tr>
<tr>
<td>Cationic coagulant</td>
<td>0.85</td>
</tr>
<tr>
<td>Sulphuric acid</td>
<td>25</td>
</tr>
<tr>
<td>Antiscalant</td>
<td>1.05</td>
</tr>
<tr>
<td>Sodium bisulfite</td>
<td>6</td>
</tr>
</tbody>
</table>

### Table 19

**Parameters to evaluate the efficiency of pre-treatment [61]**

<table>
<thead>
<tr>
<th>Characteristics</th>
<th>Before pre-treatment</th>
<th>After pre-treatment</th>
</tr>
</thead>
<tbody>
<tr>
<td>Turbidity</td>
<td>0.368</td>
<td>0.157</td>
</tr>
<tr>
<td>SDI</td>
<td>14.6</td>
<td>3.4</td>
</tr>
<tr>
<td>Iron, ppm</td>
<td>0.058</td>
<td>0.009</td>
</tr>
<tr>
<td>Total petroleum hydrocarbons, ppb</td>
<td>0.005</td>
<td>Not detected</td>
</tr>
</tbody>
</table>
the membrane. Chemical dosing in membrane pre-treatment is significantly reduced compared to conventional pre-treatment [63]. Commercially available modules are: immersed plate-, pressure driven capillary-, pressure driven spiral wound and immersed hollow fibre modules. Most commonly, hollow fibre modules are applied for pre-treatment [68].

Fouling that occurs on the surface of the membrane has to be removed. Back flushing with permeate or back flushing with chlorinated permeate combined with air sparging has been reported to be very efficient in removing particles that deposit on the membrane surface [63].

Kumar et al. [70] compared microfiltration and ultrafiltration membranes in pre-treatment to determine differences in filtrate quality. In the test runs, feed to the membrane was prefiltered by a 1 µm filter to create similar feed water quality in all tests. 0.1 µm microfiltration as well as 100 kDa and 20 kDa ultrafiltration membranes were tested. The tests showed no significant difference between microfiltration and 100 kDa ultrafiltration membranes regarding flux decrease in the RO element, suggesting equal fouling potential of the filtrate. The 20 kDa ultrafiltration membrane resulted in an reduced flux decline in the RO element, suggesting less membrane fouling. However, higher pressures had to be applied to achieve similar flux [70].

Experiences with membrane pre-treatment — Membrane pre-treatment offers advantages in terms of RO feed water quality and has been successfully applied at various desalination sites. Vial et al. [71] implemented 0.1 µm hollow fibre membranes for pre-treatment of Mediterranean seawater with a raw water quality as reported in Table 20.

Storms caused significant peaks of turbidity and SDI of the raw water and minor increases in filtrate turbidity and SDI. Intensive chemical cleaning was able to restore normal permeate flux and SDI [71]. Membrane pre-treatment provided high quality feedwater to the RO membrane with an SDI consistently below 1.8, allowing operation at higher recovery rates reducing total system running cost.

Pearce et al. [7] analysed membrane pre-treatment to reverse osmosis desalination at Port Jedda, Saudi Arabia, as an alternative to its conventional pre-treatment facility, which could not meet targeted feedwater quality during algal bloom and storms. The feed to the UF pre-treatment membranes was taken from the stream optimised for conventional treatment, i.e. it was acidified and dosed with ferric chloride.

Membrane pre-treatment with daily air enhanced backwash achieved an average filtrate SDI of 2.2 with all values below 3 even during storms and algal bloom. Compared to the existing conventional pre-treatment system, this has been an improvement of RO feed water quality in terms of SDI by two units. Higher RO feed water quality resulted in reduced fouling of the reverse osmosis element by 75% and thus will avoid downtime or production losses [7].

Problems with the pre-treatment system at Tampa Bay, Florida, combined with operational difficulties have significantly increased cost for the desalination process. The plant capacity is 94,000 m³/d and originally featured a 2-stage dual media sand filtration pre-treatment system. However, pre-treatment has been insufficient and could not meet targeted level of feed water SDI values, which caused severe fouling of the RO elements. The resulting higher chemical usage and energy consumption as well as more frequent RO replacement significantly increased operational costs.

Table 20
Feedwater quality [17]

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed temp., °C</td>
<td>24–31</td>
</tr>
<tr>
<td>pH</td>
<td>7.8–8.2</td>
</tr>
<tr>
<td>Turbidity on average (peaks), NTU</td>
<td>2.9 (24)</td>
</tr>
<tr>
<td>TDS</td>
<td>35,000–43,000</td>
</tr>
<tr>
<td>SDI₁₅ (median)</td>
<td>13–27 (17.8)</td>
</tr>
</tbody>
</table>
Application of immersed hollow fibre ultrafiltration membranes substituting the conventional pre-treatment facility ensured stable operation by producing an RO feed water with a SDI$_{15}$ of less than 1.0 regardless of feed water turbidity and algal bloom [63]. In addition, implementation of the UF membrane pre-treatment system allowed the reverse osmosis plant to be operated at much higher flux and recovery rates [7].

A comparative overview of the influence of both membrane and conventional pre-treatment according to [63] is given in Table 21. Membrane pre-treatment increases permeate flux, supplies better quality feedwater to the RO stage and often makes the use of cartridge filters unnecessary. On the other hand, ultrafiltration membranes display shorter life times than sand filters as their conventional counterpart. Therefore, higher costs for membrane replacement come along with benefits in feedwater quality to the reverse osmosis stage.

### 6.4. Cost comparison of membrane and conventional pre-treatment

While operational benefits of membrane pre-treatment have been described, a cost comparison is necessary to evaluate the overall benefit of each option. The economic benefit of one option is depending on many site specific factors such as raw water quality, membrane replacement costs, utility cost and energy cost. Wolf and Siverns [63] analysed cost for RO desalination with conventional and membrane pre-treatment based on the assumptions presented in Table 22 [63,68].

Membrane pre-treatment displays higher investment cost for the pre-treatment facility (cf. Table 23), which is balanced by higher investment costs for the RO stage of the desalination plant using conventional pre-treatment. Thus, overall investment cost of the two systems is comparable.

Overall operating costs, similar to investment costs, are nearly independent of pre-treatment technology (cf. Table 24). RO membrane replacement and overall maintenance cost using membrane pre-treatment decrease due to better RO feedwater quality; fewer chemicals and less manpower are required due to highly automated processes, which also reduces cost. However, advantages of membrane pre-treatment are balanced by higher investment and operational costs of the pre-treatment step [63].

Côte et al. [72] compared conventional dual-
media filtration and ultrafiltration membrane pre-treatment of seawater for a plant capacity of 75,000 m³/d. They concluded that membrane pre-treatment is slightly more expensive than conventional dual media filtration. It is further suggested that ultrafiltration pre-treatment is capable to reduce overall costs of SWRO desalination if operational benefits such as [72] higher RO flux, smaller footprint, longer RO-membrane life, and increased plant availability are taken into account.

A higher RO feed water quality allows the process to run at a higher RO flux without the danger of increased fouling. With increased flux, less membrane area is required for the same permeate production rate, permeate salinity is decreased, but energy consumption increases due to higher feed side pressures. In addition, longer membrane life caused by better protection from particulate fouling by application of membrane pre-treatment decreases overall life-cycle costs [72].

An answer to the question if membrane pre-treatment or conventional pre-treatment is economically more feasible strongly depends on site-specific factors. Fluctuations in the feed water quality in terms of turbidity and TDS as well as algae bloom can cause problems for conventional pre-treatment, which might result in additional cost for the process. Membrane pre-treatment is a lot less sensitive to fluctuations of feed water quality and supplies the RO stage of the plant with superior water quality for long-term ope-

<table>
<thead>
<tr>
<th>Table 22</th>
<th>System assumptions</th>
</tr>
</thead>
<tbody>
<tr>
<td>SWRO desalination plant capacity</td>
<td>74,000 m³/day</td>
</tr>
<tr>
<td>Seawater TDS</td>
<td>35,000 ppm</td>
</tr>
<tr>
<td>Raw water quality</td>
<td>Poor, high/variable turbidity, SDI immeasurable</td>
</tr>
<tr>
<td>RO flux</td>
<td>17 L/m²h after UF, 13.6 L/m²h after sand filtration</td>
</tr>
<tr>
<td>Power cost</td>
<td>US $0.045/kWh</td>
</tr>
<tr>
<td>Interest rate</td>
<td>6.5%</td>
</tr>
<tr>
<td>Plant life</td>
<td>25 years</td>
</tr>
<tr>
<td>Second pass RO</td>
<td>For boron removal as needed</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Table 23</th>
<th>Specific investment cost [63]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Component</td>
<td>UF + 2-stage SWRO + 2-stage sand filtration + 2-stage SWRO [US$/d]</td>
</tr>
<tr>
<td>Infrastructure</td>
<td>205</td>
</tr>
<tr>
<td>Pre-treatment system</td>
<td>202</td>
</tr>
<tr>
<td>Desalination system</td>
<td>572</td>
</tr>
<tr>
<td>Total investment cost</td>
<td>979</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Table 24</th>
<th>Operational cost [63]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Component</td>
<td>In line coagulation + 2-stage sand filtration + 2-stage SWRO [US$/m³]</td>
</tr>
<tr>
<td>Investment cost</td>
<td>0.2377</td>
</tr>
<tr>
<td>Replacement for UF cartridges/ sand filtration material + cartridges</td>
<td>0.0234</td>
</tr>
<tr>
<td>Replacement RO membranes</td>
<td>0.0161</td>
</tr>
<tr>
<td>Process and cleaning chemicals</td>
<td>0.0411</td>
</tr>
<tr>
<td>Power consumption</td>
<td>0.1773</td>
</tr>
<tr>
<td>Spare parts</td>
<td>0.0382</td>
</tr>
<tr>
<td>Manpower – O&amp;M</td>
<td>0.0286</td>
</tr>
<tr>
<td>Overhead</td>
<td>0.0196</td>
</tr>
<tr>
<td>Total water cost</td>
<td>0.5819</td>
</tr>
</tbody>
</table>

*Not considering any penalties for alternative water supply in case of plant under-performance caused by pre-treatment.

*Excluding cost for land acquisition.
ration. Membrane pre-treatment therefore might be able to increase RO membrane life and ensure stable operation even under adverse conditions and could thus lead to overall cost reductions [63,64].

7. Post-treatment

Untreated permeate from sea- or brackish water reverse osmosis plants does not conform with drinking water standards such as the World Health Organisation (WHO) drinking water guidelines [73] or the Gulf Drinking Water quality standards [74]. Drinking water requirements according to different drinking water legislation are listed in Table 25.

Due to the low TDS values RO permeate water can be unpalatable, corrosive and unhealthy. Post-treatment to meet drinking and irrigation water standards is therefore an essential part of most reverse osmosis plants.

Permeate has to be re-hardened in order to prevent corrosion of pipes in the distribution network, pH value and CO₂ content need to be adjusted for scaling prevention and permeate water needs further disinfection. In addition, restrictions on boron content exist, which poses a separation problem on RO desalination due to limited boron rejection of existing RO membranes. Therefore special measures have to be taken to comply with boron limits.

7.1. Recarbonation and remineralisation

Corrosiveness of water is often characterised by its Langelier saturation index (LSI). “Corrosive water” is defined as water having a negative LSI.

The LSI is defined as the measured pH of the water minus the equilibrium pH value (pHₐ) of the system if saturated with CaCO₃ at the measured alkalinity and calcium values. If the LSI is negative, the water is corrosive to calcium carbonate. For LSI equal to zero, the water is non-aggressive and no deposition layer will be found. Positive LSI will result in deposition of calcium carbonate.

The aim of any re-hardening of RO permeate is to produce a slightly positive LSI to prevent pipe corrosion and to have a fine precipitation layer for additional protection. Achievement of minimum TDS values is not an issue in reverse osmosis desalination like it is for thermal desalination processes.

There exist several methods for re-hardening of permeate water summarised by Withers [75] and Hasson and Bendrihem [76].

7.1.1. Dissolution of lime by carbon dioxide

A commonly used method to add alkalinity to water to make it non-aggressive and/or non-corrosive [75] is the dissolution of lime by carbon dioxide according to

\[ 2\text{CO}_2 + \text{Ca(OH)}_2 \rightarrow \text{Ca(HCO}_3\text{)}_2 \]

CO₂ acidified desalinated water is treated with a milk of lime, which is prepared in a lime saturator (cf. Fig. 38).

Use of hydrated lime may result in an increased total product water turbidity (NTU) and use of hydrated lime solution with <96% might raise NTU to values higher than 5, violating WHO drinking water guidelines. Lime slurry dosing using hydrated lime of 98% purity is therefore recommended [75].

7.1.2. Dissolution of limestone by carbon dioxide

The most widely applied method [76] for remineralisation is the dissolution of limestone by carbon dioxide. CO₂ acidified desalinated water is contacted with limestone and the water is mineralised according to

\[ \text{CO}_2 + \text{CaCO}_3 + \text{H}_2\text{O} \rightarrow \text{Ca(HCO}_3\text{)}_2 \]
### Table 25
Drinking water requirements

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Gulf drinking water guidelines [74]</th>
<th>WHO drinking water guidelines [73]</th>
<th>European Standard (98/83/CE of3)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Colour (Pt-Co scale)</td>
<td>15</td>
<td>15</td>
<td></td>
</tr>
<tr>
<td>Turbidity, NTU</td>
<td>5</td>
<td>&lt;1</td>
<td></td>
</tr>
<tr>
<td>Taste</td>
<td>Acceptable</td>
<td>Acceptable</td>
<td></td>
</tr>
<tr>
<td>Odour</td>
<td>Acceptable</td>
<td>Acceptable</td>
<td></td>
</tr>
<tr>
<td>pH</td>
<td>6.5–8.5</td>
<td>8</td>
<td>6.5–9.5</td>
</tr>
<tr>
<td>TDS [mg/L]</td>
<td>100–1000</td>
<td>1000</td>
<td></td>
</tr>
<tr>
<td>Conductivity [µS/cm]</td>
<td>160–1000</td>
<td>2500</td>
<td></td>
</tr>
<tr>
<td>Total hardness [mg/L CaCO₃]</td>
<td>500</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Calcium [mg/L]</td>
<td>200</td>
<td>200</td>
<td>200</td>
</tr>
<tr>
<td>Magnesium [mg/L]</td>
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<td></td>
</tr>
<tr>
<td>Sodium [mg/L]</td>
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<td>200</td>
<td>200</td>
</tr>
<tr>
<td>Chloride [mg/L]</td>
<td>250</td>
<td>250</td>
<td>250</td>
</tr>
<tr>
<td>Aluminium [mg/L]</td>
<td>0.2</td>
<td>0.2</td>
<td></td>
</tr>
<tr>
<td>Iron [mg/L]</td>
<td>0.3</td>
<td>0.3</td>
<td></td>
</tr>
<tr>
<td>Copper [mg/L]</td>
<td>1.0</td>
<td>1.0</td>
<td></td>
</tr>
<tr>
<td>Zinc [mg/L]</td>
<td>5.0</td>
<td>3.0</td>
<td></td>
</tr>
<tr>
<td>Manganese [mg/L]</td>
<td>0.1</td>
<td>0.1</td>
<td></td>
</tr>
<tr>
<td>Residual chlorine [mg/L]</td>
<td>0.2–0.5</td>
<td>0.6–1.0</td>
<td></td>
</tr>
<tr>
<td>Sulphates [mg/L]</td>
<td>250</td>
<td>250</td>
<td></td>
</tr>
<tr>
<td>Boron [mg/L]</td>
<td>0.5</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Fig. 38. Dissolution of lime by carbon dioxide, modified from [75].
This method will produce water with a pH equal to \( \text{pH}_s \). Theoretically only half of the carbon dioxide needed for the lime process has to be added because of the carbonate content of limestone. However, due to the slow reaction rate, complete conversion will not be achieved and \( \text{CO}_2 \) will be only reduced by 15–35\% [75].

Residual \( \text{CO}_2 \) has to be neutralised by addition of \( \text{NaOH} \) or \( \text{Na}_2\text{CO}_3 \). For larger plants, \( \text{CO}_2 \) degassing with \( \text{CO}_2 \) recovery might be more economical than neutralisation. However, residuals of carbon dioxide still need to be neutralised for control of the desired pH and alkalinity (cf. Fig. 39). Major advantages are the economic benefits using limestone instead of lime, and reduction of \( \text{CO}_2 \) usage. In addition, the equipment for handling limestone is cheaper compared to the equipment used for preparing and dosing of lime slurries.

The process displays several disadvantages compared to usage of hydrated lime:
- need for limestone dissolver units
- need for \( \text{CO}_2 \) desorption towers
- excess \( \text{CO}_2 \) has to be neutralised by addition of lime, caustic or sodium carbonate
- hydrated lime requirement is usually about 70–80\% of the mass equivalent limestone usage, resulting in lower transport costs.

7.1.3. Alternative re-hardening measures

Other methods for re-hardening usually use dosage of a chemical solution based on calcium chloride or bicarbonate. Due to high preparation and dosing costs, these methods can only be applied for small scale plants.

Dosage of hydrated lime and sodium carbonate — Hydrated lime and sodium carbonate are mixed with permeate. The method is usually applied to natural water containing some alkalinity and some free \( \text{CO}_2 \). Under these conditions predominantly a non-adherent calcium carbonate is crystallised according to

\[
\text{Ca(OH)}_2 + \text{Na}_2\text{CO}_3 + \text{H}_2\text{O} \rightarrow \text{CaCO}_3 + 2\text{NaOH}
\]

The method is not used in large desalination
plants due to high operating costs associated with sodium carbonate [75].

Usage of calcium chloride and sodium bicarbonate — Anhydrous calcium chloride and sodium bicarbonate are added to the product water according to:

\[
\text{CaCl}_2 + 2\text{NaHCO}_3 \rightarrow \text{Ca(HCO}_3\text{)}_2 + \text{NaCl}
\]

Plants, which require a minimum TDS value, could benefit from the increase in chloride concentration, but could also render the water corrosive. In addition, further pH adjustment is required.

The method is only applied for industrial water treatment but not for municipal water production [75].

Blending of RO desalinated water with treated water from a saline water source — Product water from thermal desalination processes is sometimes blended with mineral rich waters such as BW or SW to improve taste and aroma properties of the water and TDS values. However, blending is never used with permeate from reverse osmosis plants [75].

When permeate is blended with mineral-rich water, re-hardening is insufficient, since mainly sodium and chloride are added. Only very small amounts of sweater can be added, otherwise TDS values would exceed plant requirements. In addition, blending does not sufficiently increase alkalinity.

7.2. Disinfection of produced desalinated water

Although the permeate from a reverse osmosis plant is of high quality, free of bacteria and viruses, disinfection is still necessary to protect consumers from pollution introduced during distribution, further treatment or storage. Methods for disinfection include: treatment with chlorine, on-site sodium hypochlorite generation and treatment with bulk hypochlorite [75].

The choice of disinfection method will depend on availability of chemicals, their cost and on the safety issues related to chlorine use. Chlorine disinfection is usually the cheapest choice calculated for the life-time of a desalination plant. Depending on required quantity, either liquid or gaseous chlorine is preferred. Liquid chlorine is generally used for withdrawal rates greater than 28 kg/h [77].

Bulk delivery of sodium hypochlorite is another option for disinfection, but application will greatly depend on temperature conditions of the site. Over the life-time of a plant it may be the most costly option due to decomposition of the disinfectant. Cooling of storage facilities and during transport will reduce decomposition but will increase overall whole life cost. Typical cost therefore exceeds the cost of chlorination with liquid or gaseous chlorine by 45–50% [75].

The cost of sodium hypochlorite application can be reduced by on-site electrolytic generation. For control of chlorate concentration, which is limited by drinking water regulations, cooling of the electrolytic feed solution to 15–25°C [75] is necessary, which can significantly add to the overall cost. Although cost is decreased by about 10% compared to bulk delivery of sodium hypochlorite, it is still 40% higher than in liquid or gas chlorination.

7.3. Boron removal

Boron is found in any natural water source. However, boron at high concentration in drinking water is suspected to cause birth defects, fetal abnormalities and to disrupt normal fetal development. The WHO therefore limits boron content in drinking water to 0.5 mg/L, while the EU suggests concentrations below 1.0 mg/L.

In addition, boron at elevated concentrations may be harmful to crops when desalinated water is used for irrigation purposes. Although boron as a trace element is vital for plant growth, it can lead to foliage damage, reduction of fruit yield
and premature ripening of sensitive fruits such as citrus or kiwis at concentrations above 0.3 mg/L [78]. Typical tender documents therefore display boron limits in the RO permeate that lie between 0.3 and 1 mg/L. Typical boron concentrations in seawater by far exceed required values and can be as high as 7 mg/L in the Arabian Gulf and usually are about 4.5 mg/L.

The problem of high boron concentration was observed after commissioning a SWRO plant in Eilat, Israel, in 1997. Farmers using the post-treated product water for irrigation noticed poisoning of crops and partly discoloured leaves. Later, boron was identified as the toxin responsible for these effects. Since then, several post-treatment methods have been developed for boron removal in SWRO desalination.

In seawater boron is usually present as boric acid $\text{H}_3\text{BO}_3$ and the following equilibrium reaction takes place in seawater [78], which leads to a pH-dependent distribution of boron containing species, with a $\text{pK}_a$-value around 9.3:

$$
\text{B(OH)}_3 + \text{OH}^- \leftrightarrow \text{B(OH)}_4^-
$$

Typical RO membranes have a high rejection of the charged form, dominating at pH-values above $\text{pK}_a$, whereas rejection is low for the uncharged species, which dominates at lower pH.

Under standard test conditions (32 g/L NaCl, 8% recovery, 55 bar feed pressure), SWRO high rejection membranes display a boron rejection between 88 and 91% [79]; BWRO membranes with associated high permeabilities reject between 30% and 80% of the uncharged boron compound [78]. In general, boron rejection decreases with decreasing membrane permeability.

Actual rejection not only depends on pH, but on various parameters such as temperature and salt concentration. At higher pH rejection
strongly increases due to a shift to the charged form. A shift to pH 10 elevates rejection of SW membranes to about 99% and of BW membranes to 93%, pH 11 to 99% for BW and 99.5% for SW membranes [79]. Boron rejection of current RO membranes is shown in Fig. 40.

Removal of boron with RO membranes therefore requires elevated pH values. In a single pass RO operation, high pH is however problematic due to high alkalinity resulting in an excessive consumption of caustic and high hardness which could cause precipitation of scaling layers. Increased pH is therefore used primarily in double pass operation at the second RO pass. Alternatively boron selective resins can be used instead of a second RO stage.

The main options for boron removal are:

- single pass RO with high boron rejection membranes
- SWRO followed by BWRO
- SWRO followed by boron selective ion exchange resin (BSR)
- SWRO followed by a hybrid process of BSR and BWRO
- SWRO followed by electro dialysis reversal (EDR)

The options are illustrated in Fig. 41.

7.3.1. Single pass SWRO

To date, most systems use multiple stages for boron removal. Development of high rejection RO membranes however could lead to the use of single pass RO systems. In a recent paper by Tabiguchi et al. [81] from Toray Ind. the authors have evaluated the application of high boron rejection membranes in single pass operation and compared their results with results from two- or multiple pass operation with and without boron selective resins.

The high rejection membrane developed by TorayTM displays a boron rejection of 94–96% at pH 8, which is approximately a decrease of boron passage by two-thirds compared to other seawater RO membranes. Cost for desalination including boron removal has been evaluated with two different feed waters for all options. The authors conclude that for high boron concentrations (>0.9 mg/L for Asian seawater (ASW) and >1.4 mg/L for Middle East seawater (MESW)) costs are minimised using a standard RO membrane in a single pass mode. Application of the new high rejection membrane leads to lower cost for permeate boron concentration higher than 0.45 mg/L ASW and 1 mg/L MESW. Lower concentrations require double pass systems.

Single pass operation poses a greater threat not to keep required boron levels. Fluctuations in feed salt content and temperature might change boron rejection significantly and it is questionable if under any conditions boron concentrations will comply with stringent boron requirements.

7.3.2. SWRO followed by BWRO

Many recently built RO desalination plants, e.g. Ashkelon, use a combination of SWRO (first pass) and BWRO (second pass) for achieving minimum boron concentrations. Typical single-stage boron permeate concentrations are in the range of 0.8 to 1.3 mg/L using seawater RO membranes with limited rejection and feed pH of 7.6 [79]. These high concentrations show that the stringent requirement of 0.4 mg/L cannot be achieved using single pass RO operation.

For boron removal, permeate is collected at different locations of the first pass RO train. Permeate close to the feed entry displays lower boron concentrations and is used for blending. Rear-end permeate with higher boron concentration is treated in a second pass.

The second pass uses high flux BW membranes, which generally have low boron rejection. Therefore pH is elevated by addition of caustic soda. The second pass can be realised as a conventional BW unit with recoveries around 85%, but is usually designed for higher recoveries for an efficient usage of first pass permeate.
Fig. 41. Boron removal options adapted from [79] (single pass RO top left, 2nd pass BWRO top right, 2nd pass BSR bottom left, hybrid BWRO/BSR process bottom right).
Brackish water membranes are employed due to their low energy consumption in desalination. Busch et al. [79] give a reduction in energy consumption of 0.29 kWh/m$^3$ using BWRO membranes instead of SWRO membranes.

High recovery and elevated pH values result in high scaling potential and measures against scaling have to be taken. High recovery units with recovery up to 90% can be designed using antiscalants. With additional intermediate acid dosage, the recovery of the second pass can even reach 95% [79].

The boron removal section of the Ashkelon plant uses a sequence of four RO stages including first pass RO (Fig. 42). After the first pass, part of the permeate is send to a 2nd stage with elevated pH after dosage of caustic. Brine leaving the 2nd stage is re-softened for scaling prevention and fed to the third stage. In a last step, the pH of the permeate of the 3rd stage is raised by addition of caustic soda, and fed to the 4th RO stage. Permeate of the 1st, 2nd and 4th stages are mixed and the resulting product water displays a boron concentration below 0.4 mg/L.

7.3.3. Use of SWRO followed by boron selective ion exchange resin (BSR)

In general, ion exchange membranes are used to remove ions from a solution. The active ion exchange membrane consists of a solid phase of naturally occurring minerals or a synthetic resin having a mobile ion attached to an immobile functional acid or base group [82]. In the exchange process, the mobile ions are exchanged with solute ions having a stronger affinity to the functional group. The process can be highly selective and achieves a reduction of the boron concentration of 99 to 99.99% using a boron selective exchange resin [79], while other ion concentrations are practically constant.

An ion exchange membrane consists of a column filled with ion exchange resin pellets and a pellet strainer system for pellet retention. Advantages of the BSR system are [79,82]:

- Unlike a two stage RO system the process is highly boron selective.
- Very low boron concentration allows blending with first pass permeate.
- Higher recovery of first pass permeate.
- Temperature independent operation.
- Simplicity of operation.
- Low water loss.
- Low power consumption.

Disadvantages can be summarised as [79,82]:

- Storage of chemicals is required for regenerating BSRs.
- High expense is incurred for the regeneration of BSRs.
• The units are sensitive to fouling and enhance bacterial growth.
• BSR cleaning solutions for regeneration are highly concentrated and require expensive and careful disposal.
• Two different technologies are employed.

Regeneration of the BSR is usually achieved in co-current flow. The selective resin is first treated with HCl or sulphuric acid, then rinsed with water and finally regenerated with caustic followed by a final rinsing step.

Typically, boron removal is nearly complete within one stage of BSR treatment and boron is removed by 99 to 99.99%. A challenge when designing a BSR system is the swelling and shrinking behaviour of the resin with changes in pH. The resin shrinks with protonation and swells when regenerated [79].

7.3.4. SWRO followed by a hybrid process of BSR and BWRO

The advantage of the BSR process is the almost complete removal of boron. However, because the ion-exchange (IX) process displays no rejection for salt ions, the required salinity must be achieved in the first RO stage. Alternatively, the BSR process can be combined with a 2nd BWRO stage for reduction of boron and salinity. This hybrid process can be especially beneficial for old SWRO plants.

Actual design of the system will depend on many parameters such as feed boron and salt concentrations as well as feed water temperature. A typical design of a hybrid SWRO/BWRO/BSR process is given in Fig. 43.

The partial second pass BWRO process produces 50 to 60% of the final blend with a typical boron concentration of 0.2 to 0.35 mg/L. The BSR stage contributes between 10 to 25% with a
Table 26
Comparison of alternative boron removal techniques [79,83]

<table>
<thead>
<tr>
<th>Boron removal</th>
<th>Feed boron concentration [mg/L]</th>
<th>Boron concentration requirement [mg/L]</th>
<th>Temperature range [°C]</th>
<th>Unit costs [$/m^3]</th>
<th>Additional specifications</th>
</tr>
</thead>
<tbody>
<tr>
<td>1-pass RO</td>
<td>4–5</td>
<td>0.8–1</td>
<td>18–26</td>
<td>0.38–0.52</td>
<td></td>
</tr>
<tr>
<td>2–pass RO</td>
<td>4–6.3</td>
<td>0.4–0.5</td>
<td>&lt;34</td>
<td>0.45–0.55</td>
<td></td>
</tr>
<tr>
<td>RO–BSR</td>
<td>n/a</td>
<td>&lt;0.4</td>
<td>n/a</td>
<td>0.48</td>
<td>Low TDS</td>
</tr>
<tr>
<td>Hybrid RO/BSR</td>
<td>4.5–6.0</td>
<td>&lt;0.4</td>
<td>n/a</td>
<td>0.5–0.55</td>
<td>Low TDS</td>
</tr>
<tr>
<td>2–pass with pH change</td>
<td>4.8–6.3</td>
<td>&lt;0.4</td>
<td>n/a</td>
<td>0.47–0.52</td>
<td>Low chlorides</td>
</tr>
</tbody>
</table>

Total boron concentration of about 0.1 mg/L. Mixed with the feed side end permeate from the first pass SWRO unit (0.5–0.6 mg/L) a total boron concentration in the water leaving the desalination plant of less than 0.4 mg/L can be achieved.

7.4. Boron removal cost comparison

Several authors have investigated operating costs of the different boron removal options. Busch et al. [79] have analysed post-treatment cost for double pass desalination systems with currently available BWRO membranes and boron selective resins.

By looking only at the cost for post-treatment without looking at the combined process of first pass RO and secondary boron removal, further excluding civil works, mechanical and electrical installation work, labour, overhead and insurance, the two options are comparable with 4.5 (BWRO) to 4.35 US¢/m³ (BSR) (cf. Fig. 44). Higher cost of the BWRO boron removal system mostly stems from the much higher energy cost, both from second stage operation and from water losses [80]. However, if the first stage is included, the difference increases. The BSR unit treats less water, and due to the low boron concentration, higher amounts of first pass permeate can be blended without violating boron restrictions.

As a result, Busch et al. predict total water production costs of 48.0 US¢/m³ using the boron selective resin compared to 53.6 US¢/m³ with the double pass RO unit. This means an increase due to post-treatment in unit production cost of 3.5 ¢/m³ for treatment with BSR compared to 9.1 US¢/m³ by use of a double pass RO.

Total cost is shown in Fig. 45 together with actual cost without boron post-treatment.

Table 26 summarises typical applications for the different post-treatment options. It can be clearly seen that the low boron concentrations required today can only be achieved with modern double pass systems. Costs for double pass RO, BSR and hybrid systems are comparable and a choice will greatly depend on costs of electricity, chemicals etc.
8. Waste management and environmental impact

Desalination plants supply water for municipal, agricultural and industrial use. Furthermore, they preserve natural water resources from exploitation and provide water for recreational areas and forests [84]. Besides this environmental protection effect, they have several disadvantages concerning their impact on the environment. Noise is emitted, energy is consumed and highly concentrated brine as well as waste membranes have to be discharged. Leaks in the feed water distribution system may affect aquifers, the intake as well as the outfall systems interfere with the marine environment and the location of a desalination plant in certain areas reduces the recreational value of the area, which might reduce public acceptance and add to environmental damage (Fig. 46).

Special attention has to be paid to the way brine is discharged and to plant efficiency in terms of energy consumption e.g. by using renewable energies to make a desalination project environmentally sound. Negative influences may not only damage the environment or reduce public acceptance, but can also result in financial penalties if toxicity standards are not met. In addition, fishing industry as well as tourism may suffer through the installation of a desalination plant in an improper location.

8.1. Air emissions

Since desalination plants of any kind consume energy, every process at least indirectly emits greenhouse gases into the atmosphere. In terms of primary energy, reverse osmosis requires 5–6 times less energy than thermal processes [85]. With the use of energy recovery systems together with high efficiency pumps and state-of-the-art RO membranes, reverse osmosis processes are able to produce fresh water at an energy consumption of 2.0 kWh/m³ [86]. However, most RO facilities have higher energy consumption. With increasing use of desalination worldwide, an energy consumption close to 3.0 kWh/m³ should be the near future goal to reduce impacts on the atmosphere [85]. Most energy for desalination processes is derived from thermo-electric power generation. The typical amount of greenhouse gases associated with the production of one cubic meter of pure water in seawater desalination is presented in Table 27 [85].

The production of 1 m³ of pure water through SWRO demands about 1 kg of oil per cubic metre of fresh water, assuming an energy consumption of 3.0–4.5 kWh/m³ [85,87]. The ongoing construction of new desalination facilities all over the world will continue to require fossil fuels as energy resource to produce potable water. This makes the water production highly dependent upon energy prices, which will most likely increase in the future and therefore put pressure on water prices.

To mitigate the environmental influence of desalination on the atmosphere and to reduce the dependency on oil as a primary energy source, the application of renewable energies should be considered in the future.

8.2. Acoustic disturbances

The high pressure pumps, energy recovery systems and turbines operating in a desalination plant generate noise, may be as high as 90 dB (A) [88]. The noise generation does not allow the
operation of a desalination plant in the vicinity of population centres without the use of means to reduce noise emission. Technological means to reduce noise may be housings over pumps and an appropriate acoustical planning of the plant [87]. Hearing damage starts at a noise level of 85 dB (A). Workers who work in this or higher noise levels have to use ear protectors in order to prevent hearing damage.

In a SWRO desalination plant in Ghalilah (UAE) a 15 dB sound reduction enclosure is built around a pressure exchanger ERD to ensure a 85 dB maximum noise requirement. Without the sound enclosure the pressure exchanger devices emit noise at a level of approximately 87 dB measured at a distance of 1 m [89].

8.3. Water quality and marine life

Water quality and marine life are affected by desalination plants in various ways. The intake system may effect the marine environment by creating currents, which can disturb marine organisms and suck large amounts of organisms into the intake. The brine disposal has the strongest effect on water quality and marine life.

The higher salt concentration and the chemical composition of the brine are creating disturbances in the vicinity of the outfall.

8.3.1. Brine composition

There are different options for the disposal of brine. In seawater desalination discharge into the open sea is considered to be the least expensive option. The brine typically contains [90]:
- backwash water from physical pre-treatment (high loads of solids, containing biological, mineral and organic matter),
- saline concentrate from the reverse osmosis separation unit, often containing anti scalants and
- membrane cleaning solutions

The backwash water (1) contains a high load of biological, mineral and organic matter and usually amounts to 2–10% of the feed for conventional pre-treatment and to 5–10% and a lower suspended solids load for membrane pre-treatment. Solids in the backwash water are sometimes treated by flocculation before discharge into the sea, according to the site’s operational constraints. Sludge is thickened by the addition of
polymers and dewatered by a centrifuge or filter press and then discharged to spread or landfill [90].

The concentrate (2) is characterised by a high salt content, which depends on the recovery rate of the plant and feed water salinity. Concentrate salinity and concentration are given by

\[
TDS_C = TDS_P \cdot \left( \frac{1}{1 - Y} \right) - \left( \frac{Y}{1 - Y} \right) \cdot TDS_P
\]

\[
Q_C = (1 - Y) \cdot Q_f
\]

where \(C\) is the concentrate, \(F\) th feed, \(P\) the permeate and \(Y\) the recovery = permeate flow/ feed flow. Recovery \(Y\) in seawater RO desalination usually varies from 40 to 60%.

The brine density is higher than that of seawater, which causes the brine to sink towards the seabed where it influences the marine biota, which is used to conditions of stable salinity. The problem of hotter brine in thermal desalination is absent in membrane desalination. Instead, high osmotic stress associated with the brine has the highest impact on the benthic biota [85]. The magnitude of the environmental effect depends on the chemical composition, natural hydrodynamics at the discharge point and the kind of marine life that inhabits the discharge area [90]. The concentrate needs pH adjustment in the form of addition of an alkalising agent if the discharge may reach sensitive areas such as coral reefs [90].

The chemical cleaning solution (3) contains high loads of dissolved solids and can be either quite acid or alkaline. The amount of chemical cleaning solutions is small compared to the other effluents [90]. Chemical cleaning is usually done by acid cleaning prior to alkaline cleaning. Both solutions should be mixed for neutralisation in a buffer tank before discharge. Chemical cleaning can be minimised by efficient pre-treatment. Chlorination may be replaced by ultraviolet radiation and conventional pre-treatment by membrane pre-treatment to reduce the environmental impact of chemicals.

A summary of chemicals added in pre- and post treatment and their further processing is given in Table 28. Table 29 compares feedwater composition and brine composition for a SWRO desalination plant in Dekhelia, Cyprus.

8.3.2. Discharge and its effects on marine life

The environmental impact during installation of pipes for intake and outfall systems are temporary but may also be significant to the marine environment. The significance of this impact depends on the level of disturbance, the specific nature of the marine habitat and the specific communities [87].

The brine discharge certainly has the greatest impact on marine environment. Especially the trend to larger desalination plants, with the associated larger concentrate discharge, poses a threat to the environment. Besides the volume of discharged brine, other factors determine the actual impact. The way brine is discharged and hydrological features such as waves, currents and water depth can reduce the environmental impact due to improved brine dispersion. Other relevant factors are

- the length of the outfall pipe,
- its distance to the shore,
- its level from the seafloor and
- the existence of diffusers.

Osmosis is the major cause for effects of the brine on marine organisms. Exposed to higher salinity seawater they suffer osmotic stress due to more dissolved ions in the seawater compared to their body liquids. The impact of osmotic stress on the organisms depends on their individual sensitivity [42].

In several case studies the influence of brine discharge on the environment has been analysed. Even though brine dispersion is fast close to the discharge point, increased salinity is still present in some distance from the outlet. A study at
Dhekelia’s 40,000 m³/d SRWO desalination plant analysed the salinity increase in the vicinity of the discharge point. Brine was discharged with a salinity of 7.2% at a depth of 5 m and a distance of 250 m from the shore. Multi-point diffusers were used to increase dispersion. Salinity close to the discharge point was as high as 5.4% and salinities around 3.9% were measured up to a distance of 200 m [42].

A study undertaken in the province of Alicante in Spain by Fernández et al. [91] observed the dispersion of brine in the area around the discharge point and came to similar results. The study was conducted in the first year of operation starting in September 2003. For a plant capacity of 50,000 m³/d and a recovery rate of 40%, the discharged brine amounts to 75,000 m³/d with a salinity of 6.8%. During observations in February and April the maximum salinity was found at the bottom of the sea, due to the higher density of the brine compared to seawater. In August the maximum salinity was found in the middle of the water column due to lower temperature at the bottom.

Results showed that dispersion was lower than foreseen. While dispersion close to the discharge point was rather high, a layer of high salinity water expanded over several kilometres. Salinity values of 0.5% TDS higher than average were found up to 4 km from the discharge point [91].
Table 29
Chemical composition of brine and feedwater at the Dekhelia SWRO plant [42]

<table>
<thead>
<tr>
<th>Analysis</th>
<th>Feedwater, mg/L</th>
<th>Brine, mg/L</th>
<th>Ratio (feed/brine)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ca$^{2+}$</td>
<td>450</td>
<td>891.2</td>
<td>1.98</td>
</tr>
<tr>
<td>Mg$^{2+}$</td>
<td>1,452.3</td>
<td>2,877.7</td>
<td>1.98</td>
</tr>
<tr>
<td>Na$^{+}$</td>
<td>12,480</td>
<td>24,649.2</td>
<td>1.975</td>
</tr>
<tr>
<td>K$^{+}$</td>
<td>450</td>
<td>888</td>
<td>1.973</td>
</tr>
<tr>
<td>HCO$_3$</td>
<td>160</td>
<td>315.3</td>
<td>1.97</td>
</tr>
<tr>
<td>CO$_3^{2-}$</td>
<td>0.2</td>
<td>0.4</td>
<td>2</td>
</tr>
<tr>
<td>SO$_4^{2-}$</td>
<td>3,406</td>
<td>6,745.1</td>
<td>1.98</td>
</tr>
<tr>
<td>Cl$^{-}$</td>
<td>22,099</td>
<td>43,661.5</td>
<td>1.976</td>
</tr>
<tr>
<td>TDS</td>
<td>40,498.2</td>
<td>80,028.4</td>
<td>1.976</td>
</tr>
<tr>
<td>pH</td>
<td>8.1</td>
<td>7.8</td>
<td></td>
</tr>
</tbody>
</table>

The increased salinity resulted in a significant impact on marine organisms. After three years of operation it was observed that higher salinity resulted in significant degradation on some macro-algal populations, while some other species completely disappeared within a distance of 100 m from the discharge point. Changes in the observed marine ecosystem are shown in Table 30 [42]. Various organisms are of special importance for the marine environment because of their interaction with other organisms, e.g. the *Posidonia oceanica* seagrass. This seagrass population fixes sand banks, oxygenates the seawater and constitutes the breeding habitat of numerous species. It is reported that *Posidonia oceanica* is very sensitive to increased salinity of surrounding seawater [85]. Due to its significant positive influence on the marine environment *Posidonia oceanica* seagrass is classified as a high-priority habitat by the European Union [88].

In an environmental study at the desalination plant of Alicante in Spain, the effect of increased salinity in the area of brine discharged was observed. Great fields of *Posidonia* seagrass cover the seabed near the discharge point. Brine discharge increased mortality rates, increased leaf fall and reduced growth rates of the species. The mortality increased with salinity and became significant above 40,000 ppm. At a salinity close to 45,000 ppm, 50% of the plants died within 15 days. The loss of *Posidonia oceanica* results in higher turbidity, decreases water quality and favours sludge formation as well as a reduction of other biological life that inhabits the grassland [85].

The intensity of the discharge impact depends on hydro-geological factors such as waves, currents, depth of the water column, etc. The vulnerability of the marine environment depends on the discharge area and its marine habitat, which can range from coral reef, rocky beach or sandy surfaces. Therefore the discharge point should be wisely chosen to guarantee minimum effects [88]. Possible measures to mitigate environmental influence are [85,90]:

- dilution of the brine with seawater prior to the discharge to reduce salinity of the discharged brine,
- lower recovery rates to reduce brine salinity,
- higher permeate and thus reduced brine salinity where it is tolerable, e.g. for purely agricultural use,
- discharge devices such as multiple port diffusers placed 50–100 m along the end of the outfall, spreading the brine across a larger area and further increasing dispersion velocity,
- dilution of the brine with water from other processes, e.g. with cooling water from power plants.

Table 30
Benthic community at the Dhekelia plant [42]

<table>
<thead>
<tr>
<th>Benthic community</th>
<th>Before operation, %</th>
<th>After three years of operation, %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Polychaetes</td>
<td>27</td>
<td>80</td>
</tr>
<tr>
<td>Echinoderms</td>
<td>27</td>
<td>--</td>
</tr>
<tr>
<td>Scaphopods</td>
<td>26</td>
<td>--</td>
</tr>
<tr>
<td>Gastropods</td>
<td>20</td>
<td>--</td>
</tr>
<tr>
<td>Crustaceans</td>
<td>--</td>
<td>20</td>
</tr>
</tbody>
</table>
plants, reducing brine salinity at the discharge point,
- discharge in an area with strong currents,
- discharge at a larger depth (8–10 m),
- designing a brine discharge velocity of not less than 3.5 m/s,
- constant discharge volume per hour reducing discharge peaks.

Diffusers increase the volume of seawater in contact with the brine and therefore improve dispersion. The success of diffusers depends on their number and the space between them. To improve dilution, diffusers can be directed to the surface of the sea at an angle of 30–90° [87,90].

8.3.3. Water abstraction

The intake systems of desalination plants pump great amounts of water, which causes high water velocities close to the intake point. Screens are installed at the intake system to protect fish and other larger aquatic organisms. The mesh size of these screens is usually of the order of 5 mm. The abstraction system can cause two potential sources of impact to the marine environment.
- impingement of fish on the screen
- entrainment of biota into the feed water system

Collision of fish with the screen leads to physical damage, disorientation and stress, associated with high mortality rates of fish through disease and increased vulnerability to predation. Entrainment poses a significant threat to phytoplankton and zooplankton.

8.4. Waste disposal and land emissions

For desalination plants that are not located close to the shore such as inland brackish water desalination plants, rejected brine cannot be disposed in the open sea. Several disposal options are available for inland desalination plants, some of them are [92]:
- discharge into solar evaporation ponds,
- disposal to wastewater systems,
- land application (spray irrigation, percolation ponds),
- injection into deep saline aquifer (non-drinking water aquifer),
- disposal onto land surface
- disposal into the sea through long pipeline systems.

These disposal options are space and cost intensive and all types of land disposal procedures are associated with the environmental risk of groundwater contamination. Evaporation ponds are space consuming while the construction of long piping to the sea for the purpose of open sea disposal is cost intensive. The cost for brine disposal ranges from 5 to 33% of the total cost of desalination [92].

Groundwater contamination with rejected brine increases groundwater hardness and leads to a reduction of soil and plant productivity. The intrusion of sodium into the soil does not affect the intake of water by plants but affects the soil structure and the infiltration of water and therefore the growth of the plants. Additionally, heavy metals contained in the rejected brine may build up in soil and groundwater, deteriorating the surrounding environment [92].

Besides the environmental problems, brine intrusion into soil and groundwater may also affect the feed water quality, which is eventually depending on the salinity in the surrounding soil. A study on three brackish water desalination plants in the United Arabian Emirates concluded that rejected concentrate from the desalination process reached the groundwater. The disposal option applied in all three plants analysed was unlined pits [92].

The main approaches to prevent further deterioration of the groundwater can be [92]:
- The zero-discharge concept is the solution which least affects the environment. It includes recycling technology and requires the
Table 31
Overview of adverse environmental effects associated with desalination processes [86]

<table>
<thead>
<tr>
<th>Adverse impact</th>
<th>Impact level</th>
<th>Source of impact</th>
<th>Mitigation techniques</th>
</tr>
</thead>
<tbody>
<tr>
<td>Increased salinity, harmful effects to less salt tolerant species</td>
<td>M</td>
<td>Concentrated brine</td>
<td>• Dilution before discharge&lt;br&gt;• Salts recovery&lt;br&gt;• Selection of outfall location for maximum mixing and dispersion</td>
</tr>
<tr>
<td>Disinfectants</td>
<td>H</td>
<td>Chlorine and its compounds, reaction with organics</td>
<td>• Use of other disinfectants such as UV&lt;br&gt;• Protecting measures to the plant intake from pollutants</td>
</tr>
<tr>
<td>Heavy metals — toxicity</td>
<td>M</td>
<td>Corrosion of plant equipment</td>
<td>• Proper design and selection of plant equipment by using corrosion resistant materials</td>
</tr>
<tr>
<td>Chemicals: Eutrophication of receiving waters, Toxicity</td>
<td>H</td>
<td>Anticorrosion and antisalant additives</td>
<td>• Reduction of chemicals to minimum level&lt;br&gt;• Use of environmentally friendly additives</td>
</tr>
<tr>
<td>pH increase</td>
<td>L</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Air pollution:</td>
<td>L</td>
<td>Combustion of fuel</td>
<td>• Use of clean and renewable energy wherever possible&lt;br&gt;• Apply cogeneration and hybrid systems&lt;br&gt;• Scrubbing gases before release to the atmosphere</td>
</tr>
<tr>
<td>Acid rain</td>
<td>M</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Greenhouse effect</td>
<td>M</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Dust</td>
<td>M</td>
<td></td>
<td></td>
</tr>
<tr>
<td>• Sediments, turbidity and limitation of photosynthesis</td>
<td>M</td>
<td>Disturbance of sands by excavation and dredging activities</td>
<td>• Minimisation and control of cut and fill activities&lt;br&gt;• Proper management of runoff within the site area</td>
</tr>
<tr>
<td>• Reduced respiration of aquatic animals</td>
<td>M</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Noise</td>
<td>L</td>
<td>• Construction activities&lt;br&gt;• Pumps and other plant equipment during operation</td>
<td>• Limitation of construction activities to working hours&lt;br&gt;• Plant equipment with low noise level</td>
</tr>
</tbody>
</table>

extraction of chemicals from the brine. The production of salt from brine is possible, although not economically feasible.

• Solar evaporation ponds can be used for the production of electricity and may lead to a reduction of cost. Spraying of brine, creation of turbulences in the brine or creation of airflow above the brine increase evaporation rates and reduce space demand.

• Brine with high alkalinity and salinity combined with solar radiation and high temperatures can provide an ideal growth medium for e.g., *Spirulina*, which are of high commercial value.

8.5. Evaluation

The environmental effects associated with desalination processes can be classified according to Loizides [42] in three categories (Table 31). Classification is done by definition of impact levels ranging from L = low, M = medium to H = high. The strongest impact on the environment is therefore expected from disinfectants, namely
residual chlorine and eutrophication of receiving waters through anticorrosion and antiscaling additives. The impact from thermal pollution and high loads of heavy metals is relevant for thermal desalination plants, but is not an issue in operation of membrane-based desalination plants. Noise emissions, toxicity and acid rain through combustion of fuel to generate the necessary energy for the process are estimated to have a low level impact on the environment. Mitigation techniques to minimise the effects are also given in the overview [42].

9. Energy requirement and cost of RO desalination

Since the introduction of reverse osmosis membrane desalination in the 1970s, costs have steadily decreased. In the past, RO desalination has only been applied in special areas where water availability was limited. Mainly due to reductions in energy consumption, RO today can be seen as an alternative to natural water resources, offers advantages in reliability and is used independent of external factors [93].

9.1. Energy requirement

Desalination processes are very energy intensive compared to exploitation of natural water resources where energy is mostly needed to pump the water from the source to the consumer. In RO desalination, energy is needed among others for the intake system, the pumping system, pretreatment operation and most importantly to apply the necessary high pressure for the RO stage. Besides high cost arising from high energy requirement, energy consumption comes along with an environmental impact associated with the emission of greenhouse gases.

In the late 1970s early SWRO plants consumed as much as 20 kWh/m³. Through development of more efficient membranes, new membrane materials and the use of energy recovery devices, energy consumption was reduced to about 3.5 kWh/m³ by the end of the 1990s. Today, energy consumption below 2.0 kWh/m³ is technically feasible [86] (Fig. 47).

Due to reduced feed water salinity in BWRO, the energy requirements are lower than in SWRO. Plants for brackish water desalination run at energy consumption below 1 kWh/m³ [6].

In 2004 the Affordable Desalination Collaboration (ADC), a non-profit organisation of leading companies and agencies in the desalination industry introduced a project to show the technical and economical feasibility of water desalination to produce potable and irrigation water. In a test site in California (USA) with a capacity of 200–300 m³/d the ADC aims to produce potable water at an energy consumption of 1.7 kWh/m³ by combining and optimising the best available technologies in energy recovery, pump and membrane systems. The project operates with standard plant material, which is commercially available and thus, can be easily scaled up. Feed conditions have been 17.4 C, 31,600 mg TDS/l, boron concentration of 4.72 mg/l and an average turbidity of 1.7 NTU.

In 2006 the ADC produced potable water at an energy consumption of 1.58 kWh/m³ with a permeate flux of 15.3 L/m²h at an overall recovery of 43%. This low energy consumption was achieved using low energy membranes at the cost of per-
meate quality. At the most affordable operating point with 15.3 L/m²h and 50% recovery, permeate TDS has been 231 mg/L with a boron concentration of 1.11 mg/L [94]. The boron concentration is especially well above current standards for drinking water.

The ADC will try to achieve better permeate quality using different membranes in future tests. Further results have not been published so far. The ADC tests did not consider long-term operation issues such as membrane replacement rates, membrane fouling, cleaning cycles or overall system reliability but plant operation has been well within membrane manufacturers recommendations.

Energy is the largest cost component in the operation of a desalination plant and offers the greatest potential for further cost reduction. The share of energy on overall cost varies with the plant and its operation parameters and location. Wilf [6] presented a typical total water cost distribution of SWRO desalination plants.

Recent studies of Aqualyng, a partner within the Techneau project, on their RO plants confirm these results with a slightly lower share of energy cost. Aqualyng test data also show energy consumption in the range of 1.9–2.5 kWh/m³ for SWRO plants with capacities of 1,000 to 5,400 m³/d. Aqualyng plants use work exchanger isobaric chambers as ERD. A test site at Tenerife achieved a TDS reduction from 32,000 mg/L to 150 ppm at an energy consumption of 1.8 kWh/m³ [95]. This indicates that new low energy processes that take advantage of new isobaric chamber ERD are able to achieve good permeate water (Fig. 48).

9.2. Cost of reverse osmosis desalination

The cost of RO desalination has steadily decreased from the commercial introduction in 1970s until today, despite rising energy prices. However, desalination projects are generally more expensive than using natural water resources such as ground- or surface water. If these natural resources are over-exploited and do not cover municipal, agricultural and tourism fresh water demand, desalination offers an additional alternative water resource. Desalination has to compete with other alternative water resources such as wastewater recycling, water import from external sources etc.

Development in membrane materials, pumping and energy recovery systems and the use of so called build, own, operate, transfer (BOOT) contracts have led to desalination costs as low as 0.53 $/m³ for the Ashkelon desalination plant [96] and 0.48 $/m³ for the desalination plant at Tuas, Singapore [3] in 2003. However, recent water price bids for SWRO plant BOOT contracts exceeded these very low water prices (cf. Fig. 49) and it is likely that due to ever-increasing construction and energy costs further price increases will be seen, which will not be compensated by further technological development [97].

For brackish water desalination Wilf [6] estimates costs for BWRO to be in the range of 0.2 to 0.3 $/m³. Total life cycle cost as well as the different cost components vary with the plant location and associated site specific factors, available energy prices, labour or real estate costs, etc. Typical water cost contribution of an SWRO plant for an estimated life time of 25 years, as given in Fig. 50, shows high shares of energy cost and capital cost on total cost [6].
9.2.1. Fixed costs

Fixed costs include the acquisition of land and process equipment as well as the plant construction. Specific cost decreases with the size of the plant. Membrane modules, piping and pumping systems are standard plant equipment and can be easily scaled up. Plant capacity has steadily increased over the last decade taking advantage of the scale factor. Recently built large-scale SWRO desalination plants are tabulated in Table 32. However, membrane-based desalination is less affected by economy of scale than other processes due to the modular assembly. This has led to the trend to large diameter RO elements (cf. Section 4.2.1).

Overall investment cost of RO plants can be estimated by use of the total capacity $C$ [99].

- For seawater desalination plants according to:

$$J_{\text{INV}} = 7100 \cdot C^{0.85}$$

- For brackish water desalination plants:
Fig. 51. Investment costs for SWRO and BWRO plants [99].

\[ J_{\text{INV}} = 1850 \cdot C^{0.82} \]

Investment cost for BW desalination plants is approximately 3.5 times below that of an SWRO plant due to lower system pressure. A correlation of investment cost and plant capacity is shown in Fig. 51.

9.2.2. Membrane replacement cost

Membrane replacement cost depends on the rate membranes are damaged and irreversibly fouled. Fouling and membrane damage are minimised through efficient pre-treatment and cleaning. Membrane replacement cost generally accounts for about 5% of the overall life cycle cost of an SWRO desalination plant [6].

Avlonitis et al. [100] evaluated membrane replacement cost for small and medium size SWRO desalination plants in Greece with capacities ranging from 180 m³/d to 600 m³/d. An average replacement cost for spiral-wound elements of €1500/element was assumed. Results indicated that membrane replacement cost is related to plant capacity and operation time of the plant. With higher capacity and higher plant operation time membrane replacement cost decreases. Specific membrane replacement cost ranged from 0.11 €/m³ to 0.29 €/m³ for medium and small sized SWRO desalination plants [100].

9.2.3. Labour cost

Labour cost has a minor share on overall cost of SWRO desalination plants. Díaz-Caneja and Farinas [101] state that the number of operation staff is practically unrelated to the plant’s capacity for large scale desalination plants. The staff required to operate a 140,000 SWRO desalination plant is:

- manager of the plant (1)
- maintenance chief (1)
- clerk (1)
- analyst (1)
- mechanic (1)
- electrical and instrument worker (1)
- polyester technician (1)
- civil worker (1)
They estimate total labour cost to 1.12 €-cent/m³ [101].

In another cost estimation by Medina [102], labour cost is given depending on plant capacity as shown in Table 33. Labour cost is assumed to be in the range of 1 to 1.5% of investment cost for a desalination plant. Estimations by Côte et al. [72] on labour costs of an SWRO desalination plant using membrane pre-treatment with 4% of overall costs are in line with the presented data (cf. Fig. 50).

### Table 33
Labour cost depending on plant capacity [102]

<table>
<thead>
<tr>
<th>Plant capacity [m³/day]</th>
<th>No. of persons</th>
<th>€/m³</th>
<th>m³/year</th>
</tr>
</thead>
<tbody>
<tr>
<td>3,000</td>
<td>4</td>
<td>0.086</td>
<td>1,050,000</td>
</tr>
<tr>
<td>5,000</td>
<td>7</td>
<td>0.069</td>
<td>1,750,000</td>
</tr>
<tr>
<td>10,000</td>
<td>9</td>
<td>0.05</td>
<td>3,500,000</td>
</tr>
<tr>
<td>20,000</td>
<td>11</td>
<td>0.036</td>
<td>7,000,000</td>
</tr>
<tr>
<td>30,000</td>
<td>13</td>
<td>0.027</td>
<td>12,500,000</td>
</tr>
<tr>
<td>50,000</td>
<td>15</td>
<td>0.019</td>
<td>17,500,000</td>
</tr>
<tr>
<td>75,000</td>
<td>19</td>
<td>0.016</td>
<td>26,250,000</td>
</tr>
<tr>
<td>100,000</td>
<td>22</td>
<td>0.014</td>
<td>35,000,000</td>
</tr>
</tbody>
</table>

### 9.2.4. Chemical costs

Chemicals are used in pre-treatment as well as cleaning operation and vary from plant to plant with the applied pre-treatment technique. Membrane pre-treatment generally requires less chemical addition than conventional pre-treatment [63]. Therefore, chemicals cost is reduced with the application of membrane treatment. Furthermore, chemical addition depends on raw water characteristics, membranes in use, regulations and operation parameters.

Al-Malek [61] presented chemical consumption of a SWRO part of a hybrid-desalination plant at Al-Fujairah (UAE). The SWRO stage has a capacity of 170,000 m³/d and uses conventional pre-treatment with shock chlorination disinfection, flocculation, multimedia filtration and cartridge filtration. The amount of chemicals added in pre-treatment and associated cost are shown in Table 34.

Côte et al. [72] presented a comparison on cost of conventional and membrane based pre-treatment. For membrane pre-treatment, about 7% of total life cycle cost has been estimated to be caused by chemical addition. Total life cycle cost was assumed to be 0.615 $/m³ [72]. For a desalination plant with a capacity of 140,000 m³/d, Diaz-Caneja and Farinas [101] estimate total

### Table 34
Specific chemical consumption and costs for the Al-Fujairah desalination plant [61]

<table>
<thead>
<tr>
<th>Chemicals</th>
<th>Conc. ppm</th>
<th>$/kg</th>
<th>g/m³ (perm.)</th>
<th>$/m³ (perm.)</th>
<th>$/day</th>
</tr>
</thead>
<tbody>
<tr>
<td>Chlorine</td>
<td>3</td>
<td>0.55</td>
<td>0.08</td>
<td>0</td>
<td>0.35</td>
</tr>
<tr>
<td>Ferric chloride</td>
<td>3</td>
<td>0.27</td>
<td>20.58</td>
<td>0.00557</td>
<td>959.4</td>
</tr>
<tr>
<td>Cationic coagulant</td>
<td>0.85</td>
<td>1.94</td>
<td>2.33</td>
<td>0.00453</td>
<td>779.33</td>
</tr>
<tr>
<td>Sulfuric acid</td>
<td>25</td>
<td>0.18</td>
<td>68.63</td>
<td>0.0124</td>
<td>2132.52</td>
</tr>
<tr>
<td>Antiscalant</td>
<td>1.05</td>
<td>1.94</td>
<td>2.65</td>
<td>0.00515</td>
<td>886.67</td>
</tr>
<tr>
<td>Sodium bisulfite</td>
<td>6</td>
<td>0.5</td>
<td>0.63</td>
<td>0.0003</td>
<td>54</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td></td>
<td></td>
<td><strong>0.03</strong></td>
<td></td>
<td><strong>4812.27</strong></td>
</tr>
</tbody>
</table>
chemical costs for pre-treatment, conditioning, membrane cleaning and effluent treatment to 3.18 €-cent/m³

9.2.5. Maintenance costs

Maintenance of process equipment is necessary to guarantee stable and reliable operation throughout the lifetime of the desalination plant. Diaz-Caneja and Farinas [101] estimate maintenance costs for a 140,000 m³/d SWRO desalination plant to be 2.2 €-cent/m³ with highest costs for instrumentation, electricity, rotating mechanical equipment, centrifugal pumps, pipes, valves and accessories. In a study on desalination costing in Spain by Medina [102], maintenance costs for mechanical and electrical equipment are estimated to range from 2 to 2.5% of plant investment costs.

10. Current and future developments in RO desalination

Although reverse osmosis is a mature technology and widely used throughout the world, many developments in recent years have decreased energy consumption and cost of RO desalination. Development of high flux membranes and introduction of energy recovery devices have greatly reduced overall energy consumption resulting in a currently possible energy consumption even below 2 kWh/m³.

Recent innovations, which will also contribute to even more competitive desalination by reverse osmosis, have been mentioned in this report on the state-of-the-art of reverse osmosis. Some of these are

- development of large diameter spiral wound elements,
- interstaged design of different membrane elements inside one pressure vessel,
- new management in membrane replacement for longer membrane life,
- introduction of membranes with even higher rejection and higher flux.

Membrane developers also work on the development of membranes with high boron rejection to minimise the extent of post-treatment. Right now there is an intensive debate about proper pre-treatment to RO desalination. Membrane systems offer superior feed water quality at a cost comparable to pre-treatment with conventional dual-media sand filtration.

Recent combinations of membrane and thermal desalination plants, so-called hybrid systems, offer further improvements in desalination efficiency. Ion-selective nanofiltration membranes used in pre-treatment of RO feedwater offer a new method to reduce scaling potential. Aiming at higher system recovery and reduction in chemicals demand, application of NF membranes in pre-treatment could significantly increase RO flux, decrease energy consumption and overall cost. Hybrid desalination systems and nanofiltration as a new pre-treatment measure will be introduced in the following.

10.1. Hybrid desalination systems

Hybrid desalination concepts make use of different technologies to combine their different advantages and if possible eliminate their drawbacks. In a hybrid process reverse osmosis can be combined with other desalination concepts or power generation facilities.

Hybrid systems can be divided into three groups according to Awerbuch [103]:

- simple hybrid systems,
- integrated hybrids and
- power/water hybrids.

Power water/hybrid systems using RO desalination use the fact that water can be stored, which is not practically feasible with electricity. This way over-capacities of the network can be utilised. In addition, desalination facilities are a major, reliable and consistent customer of electricity, resulting in larger power plants with associated high efficiencies. An example of using
reverse osmosis in combination with power generation is aquifer storage for averaging desalination capacity, for strategic storage of fresh water or for general improvement of aquifer water quality [103].

Simple hybrid systems combine multi-stage flash (MSF) and reverse osmosis technologies. Typically an existing MSF plant is combined with a new RO plant to achieve the following advantages:

- A common seawater intake and outfall can be used for both plants, thus reducing capital investment [103].
- RO permeate and MSF product water can be blended to achieve contracted and/or required water standards.

Blending of the two products enables operation of the RO plant with relatively high TDS in the permeate, which will result in longer membrane lifetimes with reduced replacement costs [104] and lower energy consumption due to higher recovery rate [103]. Blending will also reduce strict requirements on boron concentration of the RO plant. This way, two or multiple pass membrane systems might become unnecessary [103,104].

The integrated hybrid MSF/RO plant in addition makes use of thermal wastes of the MSF plant in the RO facility and/or uses pressure energy otherwise wasted in the RO plant for the MSF process. The feed water temperature can be controlled to achieve high permeate flux using cooling water from the heat reject section of the MSF or from power plant [103]. Especially during winter times, seawater temperatures can be as low as 15°C and an increase in temperature will result in an increase in water production of 1.5 to 3% for each degree [103]. In addition, total feed flow can be significantly reduced.

In addition, low pressure steam can be used for de-aeration of RO feed water to minimise corrosion and to limit residual chlorine concentrations [103].

According to Awerbusch [103], de-aeration permits the use of more economical materials like SS 316 instead of SMO-254 and SS 317L for the RO plant. Further savings can be achieved by an integrated pre- and post-treatment section of the RO and MSF plant [103].

All installed hybrid desalination plants today such as the Jeddah and Fujairah hybrids, the latter being the biggest hybrid plant of reverse osmosis and thermal desalination technologies [103], make use of the simple hybrid approach. Cost savings only come from application of common plant intakes, respectively, outfall and a cogeneration power plant for the Fujairah plant.

10.2. Pre-treatment with nanofiltration

Membrane pre-treatment and the ongoing competition with conventional pre-treatment has been described in Section 6. A further application of membrane pre-treatment in seawater desalination is the use of nanofiltration (NF) membranes prior to the RO stage, whose application has been discussed by several authors during the last years.

The separation performance of NF membranes depends mainly on two effects. The sieving effect, which is important for neutral components, and the electrostatic effect as a result of charge interactions between the pore surface and the ions.

NF membranes have a very high rejection of divalent anions, while rejection of mono-valent ions is limited. The sulphate rejection of an NF membrane tested according to Eriksson et al. [105] was greater than 99%, independent of the solute concentration and the cation valence. Rejection for magnesium, calcium and bicarbonate was 98%, 92% and 44% respectively.

The NF membrane does not reject small, non-dissociated, dissolved inorganic species like silica or boric acid. The rejection of uncharged organic species depends on the molecular size. High rejection of divalent ions offers the possibility for
increased recovery in the RO stage, if scalants are removed in an NF pre-treatment step. The Saline Water Conversion Corporation (SWCC) has tested nanofiltration as pre-treatment of seawater feed at Umm Lujj, Saudi Arabia. Hassan et al. [106] found that the NF significantly reduced turbidity and microorganism concentration. Scale forming divalent ions could be significantly removed from the feed water resulting in an overall increase of 25% per RO element at 17% lower energy consumption.

The expected scaling problem on the nanofiltration membrane has been prevented with an extremely scaling resistant surface of the NF membrane and a very short residence time, using the induction time effect of scale crystallisation. However, according to Eriksson [107], so far the NF pre-treatment has not proven to be economically feasible. Water cost for a combined NF/RO plant has been higher than for double pass RO systems. Only under special circumstances which are not likely to occur in practical operation of a desalination plant, higher recoveries can make up for additional investment cost.

Application of NF in a hybrid NF/thermal might lead to significant improvements of distillation plants [103,108]. Top temperature in a thermal plant is determined by sulphate and calcium concentrations. Partial or complete removal of these ions from the feed to the thermal plant by nanofiltration would result in a temperature increase from currently 95–110°C to 120–150°C, resulting in an increase in water production by 25 to 45% [103].

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