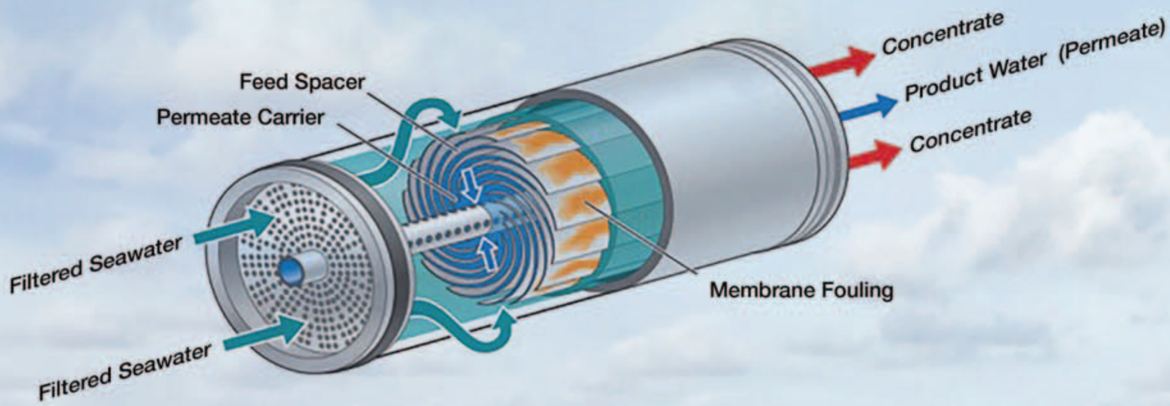

Pretreatment for Reverse Osmosis Desalination

Nikolay Voutchkov

PRETREATMENT FOR REVERSE OSMOSIS DESALINATION

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PRETREATMENT FOR REVERSE OSMOSIS DESALINATION

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Preface

At present, reverse osmosis (RO) membrane desalination is the predominant technology for production of fresh water from saline water sources (seawater, brackish water, wastewater, etc.). The engineered semipermeable membranes employed by this technology have the ability to allow transport of pure water molecules at an order-of-magnitude higher rate than they allow transport of salts.

However, the fine microstructure of the semipermeable membranes presently used for desalination by RO does not permit passage of particulates contained in the saline source water or formed during the desalination process. Therefore, if present in the feed water to the RO membranes in significant amount, these particulates may cause membrane fouling, which, in turn, may rapidly decrease membrane productivity and result in desalination plant-performance failure. Membrane foulants are typically organic and inorganic colloids and particulates, naturally occurring in the saline source water or generated on the surface of the membranes by aquatic microorganisms or physicalchemical processes that occur during RO-salt separation and concentration.

The purpose of the pretreatment system is to adequately and effectively remove foulants from the saline source water and to secure consistent and efficient performance of the RO membranes that process the pretreated water and produce desalinated water. The pretreatment system is typically located downstream of the desalination plant's intake facilities and upstream of the seawater RO-membrane system.

Depending on the source-water quality, desalination plant's pretreatment system may consist of one or more treatment processes including: screening, chemical conditioning, clarification by dissolved air flotation or gravity settling, granular media filtration, membrane microfiltration (MF) or ultrafiltration (UF), and cartridge filtration. Prior to processing through sedimentation and/or filtration, the saline source water used for membrane desalination is conditioned by the addition of a number of chemicals (biocide, coagulant, flocculant, antiscalant, etc.) in order to improve performance of downstream treatment processes.

This book provides a detailed overview of key processes, technologies, and equipment used for pretreatment of saline water used for membrane desalination, and discusses their areas of application, past track record, advantages, and disadvantages. The work describes typical causes and mechanisms of RO-membrane fouling and presents most recent developments in pretreatment technology and science. While the book makes reference, when applicable, to pretreatment of brackish waters, its content is mainly focused of pretreatment of seawater for RO desalination. It should be pointed out, however, that practically all technologies and source-water conditioning methods described in this book are equally applicable to pretreatment of seawater and brackish water.

The book's chapters address practically all aspects of pretreatment of saline source water used for production of fresh water by membrane desalination, such as the nature, origin, and characterization of key membrane foulants, the diagnostics of RO membrane fouling, the impact of the type of the desalination plant intake on the source water fouling potential and the configuration of the pretreatment system, the most commonly used source-water screening technologies and equipment, source-water conditioning for pretreatment, sand removal, clarification by gravity settling and dissolved air flotation, and saline source water granular media and membrane filtration. In addition, the book provides a comparative analysis of membrane pretreatment and includes guidelines for pretreatment system selection.

The book also includes guidance and examples for sizing and cost estimation of key desalination-plant pretreatment facilities. It is important to underline that the facility and equipment procedures presented in this book are not intended to serve as standard, all-inclusive design procedures, but their main purpose is to illustrate typical methodologies and approaches used by desalination professionals. References to particular technologies, equipment and membrane manufacturers should not be construed as an endorsement by the author or a recommendation for a preferential use or consideration.

Cost graphs incorporated in this book are recommended to be used only for the preparation of initial order-of-magnitude estimates of the construction costs of the respective pretreatment systems and are presented in 2017 US dollars. Site-specific project conditions, currency differences, and other factors may result in significant differences between the actual facility construction costs and the value of these costs determined using the cost graphs of this book. The reader is recommended to contact the suppliers of the specific pretreatment technologies, which are planned to be used for their desalination project in order to receive updated costs commensurate with local and international market conditions at the time of project implementation.

The book includes a total of 12 chapters, which follow the typical sequence of saline source-water pretreatment prior to membrane desalination and guide the reader toward the selection of the most suitable pretreatment technology or combination of technologies for the site-specific conditions of a given project. As indicated previously, the main emphasis of this book is the pretreatment of seawater for RO desalination.

Chapter 1 provides a brief overview of the key reasons of why seawater pretreatment is needed prior to membrane desalination. The chapter mainly emphasizes the role of pretreatment in the production of desalinated water and the interrelation of pretreatment system and other desalination plant components.

Chapter 2 describes the type of foulants typically contained in source seawater or brackish water and explains how these foulants impact RO-membrane performance. In addition, this chapter discusses the most common methods presently employed worldwide to characterize the RO-membrane fouling potential of saline waters and identifies threshold levels for key source-water quality parameters that trigger accelerated membrane fouling.

Chapter 3 features an overview of the typical RO-membrane fouling phenomena observed in full-scale desalination systems and the methods used for diagnostics of the type and severity of membrane fouling. The chapter also illustrates the use of RO-membrane autopsy for membrane fouling diagnostics.

Chapter 4 discusses the impact of the type and configuration of the desalination plant intake on the selection and design of the pretreatment system. The chapter encompasses practical pretreatment experience with desalination plants using both open intakes and subsurface intakes and contains guidance of how to select intake configuration that minimizes downstream pretreatment requirements for the site-specific conditions of a given desalination project. The chapter analyzes the key advantages and challenges associated with the use open and subsurface intakes and the trade-off between the use of costlier intakes and less complex pretreatment systems and vice versa.

Chapter 5 focuses on the type and configuration of the most commonly used screening equipment applied for removal of course particulate materials (large debris, algae, jellyfish, etc.) from the saline source water. The chapter addresses the criteria for selection of bar, band, and drum screens as well as microscreens, and cartridge filters.

Chapter 6 is dedicated to systems for addition of chemicals to the saline source water, which allow to condition this water prior to its further pretreatment or direct application for RO-system processing. Source-water conditioning is of critical importance for the efficient and cost-effective sedimentation, and filtration of the particulate and organic compounds contained in the water as well as for prevention of formation of scale on the RO-membrane surface. This chapter describes the purpose and dosing of commonly applied source-water conditioning chemicals such as coagulants, flocculants, scale inhibitors, biocides, acids, and bases.

Chapter 7 presents alternative pretreatment technologies that are commonly applied for removal of relatively large suspended solids contained in the source water. The chapter contains key design criteria used for sizing of lamella settlers and dissolved air flotation (DAF) clarifiers and provides construction cost curves for these facilities.

Chapter 8 is dedicated to the most commonly used type of technology for removal of find solids from the source water—granular media filtration. It discusses alternative types of filters used in desalination plants and their area of application and performance. The chapter incorporates construction cost curves for gravity and pressure-driven granular media filtration systems.

Chapter 9 discusses the use of MF and UF membranes for pretreatment of saline source water. The chapter presents key considerations associated with the selection, planning, and cost estimating of membrane pretreatment systems for seawater desalination plants. This chapter incorporates design examples for submerged and pressure-driven membrane pretreatment facilities.

Chapter 10 provides a comparative analysis of the key advantages and disadvantages of granular media and membrane pretreatment filters in terms of: the effect of source water quality and temperature on their performance; surface area requirements; quantity and quality of the generated residuals; chemical and power uses; and overall water production costs.

Chapter 11 features a methodology for identifying the type and configuration of the most suitable seawater pretreatment system for the site-specific conditions of a given desalination project. The selection methodology is based on the analysis of the source water quality collected by the desalination plant intake and is built upon the industry-wide practical experience with the implementation of alternative technologies and configurations over the past 20 years.

Chapter 12 delineates how to access the impact of the pretreatment system on the selection of the most-suitable configuration and key design criteria (flux, recovery, feed pressure, etc.) of the downstream seawater RO-desalination system.

This book is intended for water treatment professionals involved in the planning design and operation of desalination plants for production of fresh water from saline water sources (seawater, brackish water, high-salinity wastewater). The book is suitable for water utility managers and planners, consulting engineers, operators of desalination plants, and students and teachers in the desalination field.

Abbreviations and Units

AOC	Assimilable organic carbon
AOM	Algogenic organic matter
A_{ro}	Total membrane surface area of one RO element
ATD	Antitelescoping device
ATP	Adénosine-5′-triphosphate
AWWA	American water works association
B	Boron
Ba	Barium
$BaSO_4$	Barium sulfate
BFR	Biofilm formation rate
BOD	Biological oxygen demand
BOO	Build, Own, Operate
BOOT	Build, Own, Operate, Transfer
Br	Bromide
BW	Backwash water volume
BWRO	Brackish water reverse osmosis
°C	Degree Celsius
Ca	Calcium
$CaCO_3$	Calcium carbonate
$CaSO_4$	Calcium sulfate
CEB	Chemically enhanced backwash
CFU	Colony forming units
CIP	Clean-in-place
ClO_2	Chlorine dioxide
CO_2	Carbon dioxide
Cr	Chromium
Cu	Copper
d	Day
Da	Dalton, unit of weight
DAF	Dissolved air flotation
DB	Design build
DBNPA	2,2-dibromo-3-nitriloppropionamide
DBO	Design-build-operate
DO	Dissolved oxygen
DOC	Dissolved organic carbon
DP	Differential pressure
DWEER	Dual work exchanger energy recovery
EC	Electrical conductivity
EDaX	Energy dispersive X-ray
EDTA	Ethylenediaminetetraacetic acid
EPA	Environmental protection agency
EPC	Engineering, procurement, and construction
EPS	Extracellular polymeric substances
ERD	Energy recovery device
ERI	Energy recovery international
F	Fluor

°F	Degree Fahrenheit
$Fe(OH)_3$	Ferric hydroxide
$Fe_2(SO_4)_3$	Ferric sulfate
$FeCl_3$	Ferric chloride
FI	Fouling indicator
F_p	Feed pressure
FRP	Fiberglass reinforced plastic
ft	Foot
ft^2	Square foot
ft^3	Cubic foot
FTIR	Fourier transform infrared (spectroscopy)
gfd	Gallons per square foot per day
gpm	Gallons per minute
GRP	Glass-reinforced plastic
h	Hour
H_2S	Hydrogen sulfide
H_3PO_4	Phosphoric acid
HAA	Haloacetic acid
HCl	Hydrochloric acid
HDD	Horizontal directionally drilled
HDPE	High-density polyethylene
Hg	Mercury
hp	Horsepower
HTB	Hydraulic turbo booster
IDA	International desalination association
ISD	Interstage design
J	Membrane permeate flux
kg	Kilogram
km	Kilometer
L	Liter
lb	Pound
lb/in^2	Pound per square inch
lmh	Liters per square meter per hour
LOEC	Lowest observed effect concentration
LOET	Lowest observed effect time
LSI	Langelier saturation index
m	Meter
m^2	Square meter
m^3	Cubic meter
m^3/d	Cubic meter per day
MC	Membrane compaction factor
MF	Microfiltration
Mg	Magnesium
mg/L	Milligrams per liter
mgd	Million gallons per day
$MgSO_4$	Magnesium sulfate
mi	Mile
min	Minute
mL	Milliliter
mm	Millimeter
Mn	Manganese
MPN	Most probable number
mV	Millivolt

μg/L	Micrograms per liter
μm	Micrometer
μS/cm	Microsiemens per centimeter
N	Nitrogen
Na	Sodium
$Na_2S_2O_4$	Sodium hydrosulfite
$Na_2S_2O_5$	Sodium metabisulfite
$NaHSO_3$	Sodium bisulfite
NaOH	Sodium hydroxide
NDMA	*N*-nitrosodimethylamine
NDP	Net driving pressure
N_{epv}	Number of elements per RO vessel
NF	Nanofiltration
Ni	Nickel
NL	No leaks (for membrane integrity)
NOM	Natural organic matter
NPDES	National pollutant discharge elimination system
N_t	Number of RO trains
NTU	Nephelometric turbidity unit
N_{vpt}	Number of vessels per RO train
O&M	Operation and maintenance
O_p	Average osmotic pressure
ORP	Ooxidation-reduction potential
P	Phosphorus
PA	Polyamide
PACL	Polyaluminum chloride
P_d	Pressure drop
PE	Polyethylene
PES	Polyethersulfone
pH	Indication of acidity or basicity of solution
P_p	Permeate pressure
ppt	Part per thousand (1 ppt = 1000 mg/L)
P_r	Permeate recovery rate
psi	Pounds per square inch (unit of pressure)
PTFE	Polytetrafluoroethylene
PVC	Polyvinyl chloride
PVDF	Polyvinylidene difluoride
PVP	Polyvinilpyrrolidone
PX	Pressure exchanger
Q_{bw}	Daily volume of backwash water
Q_c	Daily volume of desalination plant concentrate
Q_f	Saline source water daily flow
Q_p	Volume of the plant fresh-water production (daily permeate flow)
Q_s	Volume of the saline source water (daily flow)
R	Plant recovery
R_{new}	Resistance of new membranes
RO	Reverse osmosis
R_t	Membrane resistance after given time of operation
s	Second
S	Membrane area of an element
SBS	Sodium bisulfite
SCADA	Supervisory control and data acquisition
SDI	Silt density index

SDSI	Stiff–Davis saturation index
SEM	Scanning electron microscopy
Si	Silicium
SMP	Specific membrane permeability
SO_4	Sulfates
S_P	Salt passage
S_r	Salt rejection
Sr	Strontium
SWRO	Seawater reverse osmosis
SUVA	Specific UV absorbance
TCF	Temperature correction factor
TDS	Total dissolved solids (salinity)
TDS_c	Concentrate salinity
TDS_f	Feed water salinity
TDS_p	Permeate salinity
THM	Trihalomethanes
Ti	Titanium
TN	Total nitrogen
TOC	Total organic carbon
TMP	Transmembrane pressure
TP	Total phosphorus
TSS	Total suspended solids
UAE	United Arab Emirates
UC	Uniformity coefficient
UF	Ultrafiltration
US	United States (of America)
USBR	US bureau of reclamation
US EPA	United states environmental protection agency
UV	Ultraviolet irradiation
UV_{254}	UV absorbance at 254 nm
VFD	Variable frequency drives
$V_{RO\ system}$	Volume of the RO system
WET	Whole effluent toxicity
WHO	World Health Organization
WWTP	Wastewater treatment plant
Y	Desalination plant recovery
yr	Year

Brief Book Summary Introduction

Pretreatment is an integral part of every desalination plant. The level and complexity of the needed pretreatment mainly depend on the concentration and type of particulate, colloidal, and dissolved organic foulants contained in the source seawater. At present, granular media filtration is the dominating technology for pretreatment. In the last 10 years, however, membrane pretreatment is emerging as an attractive alternative to granular media filtration.

This type of pretreatment is gaining a wider acceptance mainly due to its superior removal of particulate and colloidal foulants and its benefits in terms of consistent and reliable performance, and operational flexibility. Membrane pretreatment, however, is usually more costly than granular media filtration and has a very limited ability to remove easily biodegradable organics associated with algal blooms, which in most cases are the main culprit for reverse osmosis membrane fouling. Therefore, at its present state of development, MF- and UF technologies do not always offer the most viable and cost-effective solution for saline water pretreatment.

Taking under consideration the numerous factors affecting the overall pretreatment costs of a full-scale desalination plant, the selection of the most suitable pretreatment system for a given desalination project should be completed based on a thorough, site-specific, life-cycle cost analysis, which accounts for all expenditures and actual costs associated with the installation and operation of membrane and granular media pretreatment systems.

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CHAPTER

1

Introduction to Saline Water Pretreatment

OUTLINE

1.1 PURPOSE OF PRETREATMENT

As any other natural water source, saline water (e.g., seawater, brackish water) contains solids in two forms: suspended and dissolved. Suspended solids occur in a form of insoluble particles (particulates, debris, marine organisms, silt, colloids, etc.). Dissolved solids are present in soluble form (ions of minerals such as chloride, sodium, calcium, magnesium, etc.). At present, practically all desalination plants incorporate two key treatment steps designed to sequentially remove suspended and dissolved solids from the source water.

The purpose of the first step—pretreatment—is to remove the suspended solids from the saline source water and to prevent some of the naturally occurring soluble solids from turning into solid form, and precipitating on the reverse osmosis (RO) membranes during the salt separation process (see Fig. 1.1). The second step—the reverse RO system—separates the dissolved solids from the pretreated saline source water, thereby producing fresh low-salinity water suitable for human consumption, agricultural uses, and industrial applications.

Ideally, after pretreatment the only solids left in the source water would be the dissolved minerals, and as long as the desalination system is operated in a manner that prevents these

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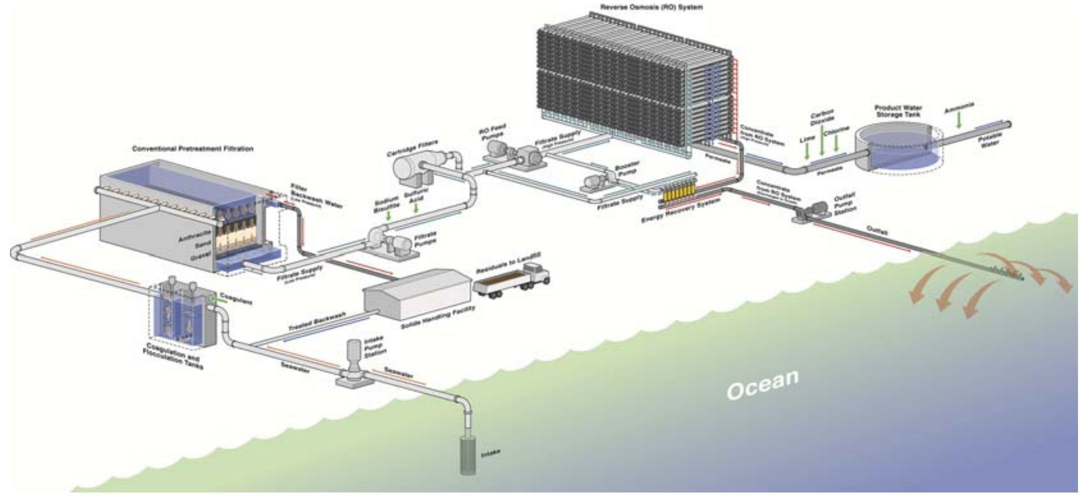

FIGURE 1.1 Schematic of typical seawater desalination plant.

minerals from precipitating on the membrane surface, the RO membranes could operate and produce freshwater of consistent quality at a high rate without the need to clean these membranes for a very long time. Practical experience shows that for desalination plants with high source-water quality and well-designed pretreatment system, the RO membranes may not need to be cleaned for one or more years and their useful life could extend beyond 10 years (Jamaly et al., 2014).

In actuality, however, pretreatment systems remove most but not all of the insoluble solids contained in the saline source water and may not always effectively protect some of the soluble solids from precipitating on the membrane surface. The suspended solids, silt, and natural organic matter (NOM) that remain in the saline water after pretreatment may accumulate on the surface of the RO membranes and cause loss of membrane productivity over time. In addition, because saline water naturally contains microorganisms as well as dissolved organics that could serve as food for these microorganisms, a biofilm could form and grow on the RO-membrane surface, causing loss of membrane productivity as well. The types of foulants contained in saline source water are described in detail in Chapter 2.

The process of reduction/loss of active membrane surface area and subsequently of productivity of RO membranes due to accumulation of suspended solids and NOM, precipitation of dissolved solids, and/or formation of biofilm on the RO membrane surface is defined as membrane fouling. Excessive membrane fouling is undesirable and besides having negative impact on RO-membrane productivity, it could also result in an increased use of energy for salt separation, and in deterioration of product water quality (Henthorne and Boysen, 2015; Villacorte et al., 2015).

Most RO systems are operated to produce a constant flow of freshwater (desalinated) at target content of total dissolved solids (TDS). Productivity of RO membranes is typically measured by the volume of desalinated water that the membranes can deliver through a unit of membrane surface (square meter/square foot) over a certain period of time (day/h). This freshwater membrane productivity is defined as membrane flux. For example, most seawater reverse osmosis (SWRO) desalination systems at present are designed to operate at a constant membrane flux in a range of 13.5–18.0 L/m^2 h (l/m^2 h or lmh) and brackish water RO (BWRO) systems for 25–35 lmh. For a given source seawater salinity, temperature, and target freshwater TDS level, producing a constant volume of desalinated water will require the source water to be fed to the desalinated system at a constant pressure (typically in a range of 55–70 bars for SWRO desalination systems and 15–25 bars for BWRO facilities).

As RO-membrane fouling occurs, to maintain membrane productivity (flux) and water quality constant, the desalination system would need to be operated at increasingly higher transmembrane pressure (TMP), which in turn means that the energy needed to produce the same volume and quality of freshwater would need to be increased. The increase in RO-system transmembrane pressure over time is an evidence of accumulation and/or adsorption of fouling materials on the surface of the RO membranes (i.e., membrane fouling).

It should be pointed out that membrane fouling is not only dependent upon the source water quality and the performance of the pretreatment system, but also upon the RO membrane properties such as charge, roughness, and hydrophilicity (Hoek et al., 2003; Hoek and Agarwal, 2006), as well as upon the flow regime on the membrane surface (Wilf et al., 2007). Membranes with higher surface roughness and hydrophobicity usually have higher fouling potential.

Typically, compounds causing RO-membrane fouling could be removed by periodic cleaning of the membranes using a combination of chemicals (biocides, commercial detergents, acids, and bases). In some cases, however, membrane fouling could be irreversible and cleaning may not recover membrane productivity, which in turn may require the replacement of some or all of the RO membranes of the desalination plant.

Criteria most commonly used in practice to initiate membrane cleaning are:

1. 10%–15% increase in normalized pressure drop between the feed and concentrate headers;
2. 10%–15% decrease in normalized permeate flow; and/or
3. 10%–15% increase in normalized permeate-TDS concentration.

All RO membranes foul over time. However, the rate and reversibility of fouling are the two key factors that have most profound effect on the performance and efficiency of the RO-separation process. These factors in turn are closely related to the saline source water quality and the performance of the desalination plant's pretreatment system.

1.2 MEMBRANE-FOULING MECHANISMS

1.2.1 External and Internal Fouling

Depending on the location of accumulation of insoluble rejected matter causing decline of membrane performance, fouling can be classified as:

1. external or "surface" fouling and
2. internal fouling.

External fouling involves accumulation of deposits on the surface of the membranes by three distinct mechanisms:

- formation of mineral deposits (scale);
- formation of cake of rejected solids, particulates, colloids and other organic and/or inorganic matter;
- biofilm formation—i.e., growth and accumulation of colonies of microorganisms on the surface of the membranes, which attach themselves by excretion of extracellular materials.

Although the three membrane-fouling mechanisms can occur in any combination at any given time, typically external membrane fouling of RO membranes is most frequently caused by biofilm formation (biofouling).

Internal fouling is a gradual decline of membrane performance caused by changes in the chemical structure of the membrane polymers triggered by physical compaction or by chemical degradation. Physical compaction of the membrane structure may result from long-term application of feed water at pressures above these the RO membranes are designed to handle (usually 83 bars for SWRO membranes) and/or by their prolonged operation at source water temperatures above the limit of safe membrane operation (typically 45°C).

Chemical degradation is a membrane-performance decline resulting from continuous exposure of membranes to chemicals that alter their structure such as strong oxidants (chlorine, bromamine, ozone, permanganate, peroxide, etc.) and very strong acids and alkali (typically pH below 3 or above 12).

While external fouling can be almost completely reversed by chemical cleaning of the membranes, internal fouling usually causes permanent damage of the microvoids and polymeric structure of the membranes, and therefore, is largely irreversible.

1.2.2 Concentration Polarization

A very important factor that may have significant impact on the extent and type of membrane fouling is concentration polarization. This phenomenon entails formation of a boundary layer along the membrane feed surface, which has salt concentration significantly higher than that of the feed water introduced to the feed/concentrate spacers of the RO membranes (Fig. 1.2).

The boundary layer is a layer of laminar feed water flow and elevated salinity that forms as a result of the tangential source water feed flow in the RO-membrane feed/concentrate spacers and of permeate flow in perpendicular direction through the membranes on the two sides of the spacer (Fig. 1.3).

In Fig. 1.3, C_b is the salt concentration at the boundary layer (e.g., feed saline water concentration); C_s is the salt concentration at the inner membrane surface, which typically is higher than that in the feed flow; and C_p is the salt concentration of the freshwater on the low salinity (permeate) side of the membrane.

Two different types of flow occur in the boundary layers of the feed/concentrate spacers (Fig. 1.3): a convective flow of freshwater from the bulk of the feed water though the membranes and diffusion flow of rejected solutes (salts) from the membrane surface back into the feed flow. Since the semipermeable RO membranes are designed such

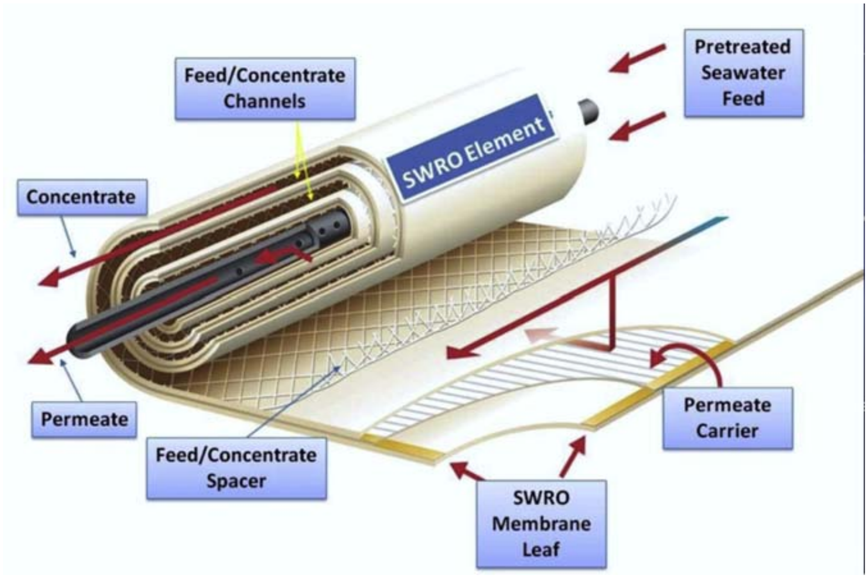

FIGURE 1.2 Reverse osmosis membrane. *SWRO*, seawater reverse osmosis.

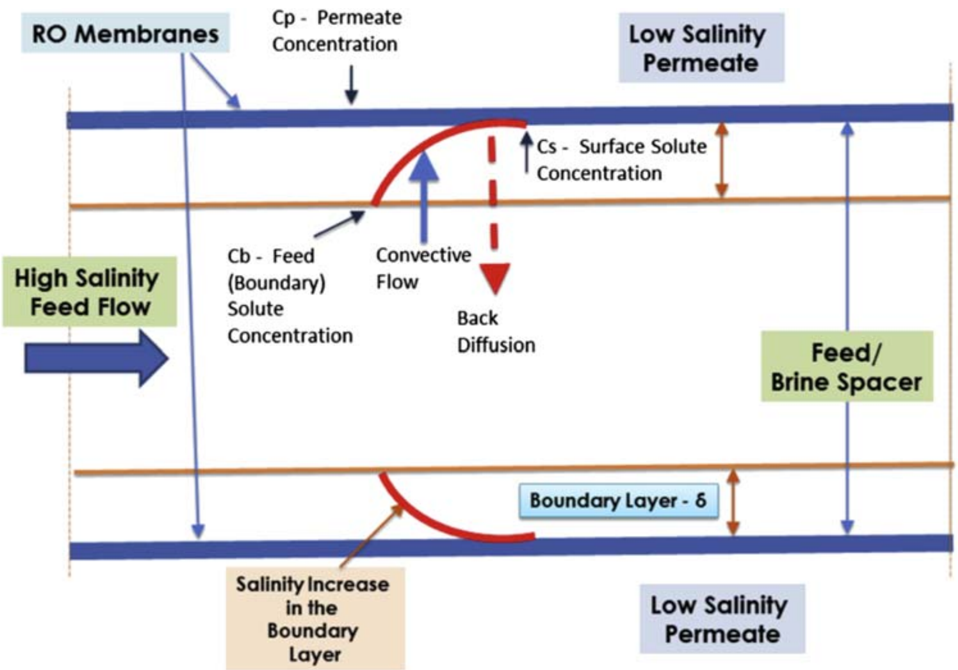

FIGURE 1.3 Boundary layers in a membrane feed spacer. *RO*, reverse osmosis.

that the rate of convective flow of water is typically higher than that of the diffusion of flow of salts, the salts rejected by the membrane tend to accumulate in the boundary layer with highest salt concentration occurring at the inner surface of the membrane (shown as C_s on Fig. 1.3). Besides salts, for the same reasons, the boundary layer also accumulates particulate solids.

This phenomenon of concentration of salts and solids in the boundary layer has four significant negative impacts on membrane performance:

1. It increases osmotic pressure at the membrane surface;
2. Increases salt passage through the membranes;
3. Creates hydraulic resistance of water flow through the membranes; and
4. Induces potential for accelerated scale formation and particulate fouling on the membrane surface because of the salt and solids concentration in the boundary layer.

The ratio between the solute (salt) content at the surface of the membrane (C_s) and in the bulk feed water (C_b) is referred to as concentration polarization factor and is denoted as Beta (β), that is:

$$\beta = C_s/C_b \tag{1.1}$$

The Beta value is always higher than 1, even when new RO elements are installed and the RO system is well designed and operated. The higher the Beta value, the greater the concentration difference, and the worst the negative impacts of this difference on membrane performance. Under best-case scenario the Beta could be in a range of 1.1–1.2, while in the worst case, the Beta value could exceed 2.0.

Eq. (1.2) indicates the impact concentration polarization has on RO-membrane flux (e.g., the freshwater production per unit membrane surface):

$$J = A \times [F_p - (\beta \times O_p + P_p + 0.5 \times P_d)] \tag{1.2}$$

where J is the membrane permeate flux; A is the membrane water permeability coefficient, which is unique for each type of commercial RO membrane; F_p is the feed pressure applied to the RO membranes; Q_p is the osmotic pressure of the saline water; P_p is the permeate pressure (also known as permeate backpressure); and P_d is the pressure drop (e.g., the difference in pressures) between the feed and concentrate sides of the RO membranes.

As seen from Eq. (1.2), the RO-membrane freshwater production (flux) decreases as concentration polarization increases. This formula indicates that in practical terms, the osmotic pressure that will need to be overcome by the RO-system feed pumps to produce the same volume of water will be increased proportionally to the concentration polarization factor.

In full-scale RO-desalination systems, A is constant specific to the RO membranes incorporated in the system; Q_p is also constant for a given source water salinity and target permeate salinity; P_p is constant determined by the target product water pressure; and P_d is of a preset value determined by the fouling propensity of the saline source water. Therefore, the only variable parameter the RO plant operator can adjust to maintain the same production flow as concentration polarization increases is the RO-system feed pressure (F_p). Such adjustment, however, is limited in practice by the maximum design delivery pressure of the RO feed pumps (which typically is within 10%–20% from their average capacity) and the maximum physical pressure the RO membrane elements are designed to handle (typically for most commercially available SWRO membrane elements, such pressure is 83–85 bars).

Because of these practical limitations, the typical impact of increase in concentration polarization on desalination plant operations is the reduction of plant's freshwater production (Herzberg and Elimelech, 2007).

The salt transport rate through the RO membranes could be presented as follows:

$$Q_s = B \times S \times (\beta \times C_b - C_p) \tag{1.3}$$

As it can be deducted from Eq. (1.3), the increase in concentration polarization will result in an increase in salt passage because the salt transport through the membranes is proportional to the difference between salinity on both sides on the membranes. Since the salinity at the feed side of the membranes is higher than the salinity in the feed solution, the salt transport is proportionally higher as well. This means that as the Beta value increases, the membrane salt rejection is reduced and the salinity of the produced freshwater is increased.

Due to the salinity gradient and particulate solids accumulation within the boundary layer the hydraulic resistance of this layer is higher than that of the feed water. As a result the head losses associated with the movement of freshwater through the membranes are increased, and assuming the same feed pressure is applied, the membrane flux is reduced. This flux reduction is the consequence of elevated hydraulic resistance of the membranes caused by concentrate polarization compounds with the flux reduction resulting from the elevated osmotic pressure that has to be overcome by the RO feed pumps.

If salt concentration in the boundary layer exceeds the solubility of sparingly soluble salts (such as calcium carbonate and sulfate) contained in the source water, these salts would begin precipitating on the membrane surface and would form mineral scale. Membrane scaling in turn will result in reduced membrane permeability and flux.

The magnitude of the concentration polarization factor, Beta, is driven by three key factors:

1. permeate flux;
2. feed flow; and
3. configuration and dimensions of feed channels and of feed spacer.

Increase in permeate flux (i.e., freshwater production) increases exponentially the quantity of salt ions and particulate solids conveyed to the boundary layer and, therefore, exacerbates concentrate polarization and particulate fouling of the membranes. Increase in feed flow, however, intensifies turbulence in the boundary layer and, as a result, decreases the thickness and salt and solids concentration of this layer. Depending on its configuration and geometry, RO-membrane feed/concentrate spacer and feed/concentrate channel may cause more or less turbulence in the concentrated boundary layer and, therefore, may reduce or enhance concentration polarization.

Since feed/concentrate spacer configuration and feed/concentrate channel size are constant for a given standard RO-membrane element, permeate flux and feed flow are the two key factors that determine the magnitude of concentrate polarization.

As indicated previously, the ratio between the permeate flow and the feed flow of a given RO-membrane element is defined as the permeate recovery rate of this element. Similarly, ratio between permeate and feed flows of the entire RO system is termed an RO system recovery rate.

As the recovery rate increases, the magnitude of concentrate polarization increases as well. For example, for SWRO systems using standard membrane elements, operation at recovery rate of 50% would typically result of approximately 1.2–1.5 times higher salinity concentration at the membrane surface than that in the source seawater. Beyond 75% recovery, the concentration polarization factor would exceed 2, which would have a significant impact on the efficiency of the membrane separation process.

In addition, at recovery rate above 75% and ambient salinity pH, many of the salts in seawater would begin precipitating on the membrane surface, which would require the addition of large amounts of antiscalant (scale-inhibitor) and would make SWRO desalination impractical. Since scaling is pH-dependent, an increase in pH to 8.8 or more, which often is practiced for enhanced boron removal, may result in scale formation at significantly lower SWRO system recovery (50%–55%).

To limit concentration polarization within reasonable limits, RO-membrane manufacturers recommend maintaining the maximum recovery rate per membrane element in a vessel within 10%–20%. As a result, with a typical configuration of six to eight elements per vessel, and taking under consideration the actual flux of the individual elements in the vessel, a single RO system is practically limited to a maximum of 50%–65% recovery. For brackish water desalination systems the typical maximum recovery rate is 85%–95%.

The concentration polarization phenomenon described above and its effect on membrane productivity (flux) decline is inherent not only to RO membranes but also occurs on the surface of ultrafiltration (UF) and microfiltration (MF) membranes used for saline water pretreatment. In this case, concentration polarization is accumulation of rejected particles (rather than salts) near the membrane surface causing particle concentration in the boundary layer that is greater than that in the raw seawater fed to the pretreatment system (which in turn results in UF/MF flux decline).

1.2.3 Membrane Fouling and Flux Redistribution

Membrane RO elements of a typical SWRO system are installed in vessels often referred to as membrane pressure vessels. Usually, six to eight SWRO-membrane elements are housed in a single membrane vessel (see Fig. 1.4).

Under typical RO-system membrane configuration, all of the feed water is introduced at the front of the membrane vessel and all permeate and concentrate is collected at the back end. As a result, the first (front) membrane element is exposed to the entire vessel feed flow and operates at flux significantly higher than that of the subsequent membrane elements. With the most commonly used configuration of seven elements per vessel and ideal uniform flow distribution to all RO elements, each membrane element would produce one-seventh (14.3%) of the total permeate flow of the vessel.

However, in actual SWRO systems, the flow distribution in a vessel is uneven and the first membrane element usually produces over 25% of the total vessel permeate flow, while the last element only yields 6%–8% of the total vessel permeate (see Fig. 1.4). The decline of permeate production along the length of the membrane vessel is mainly due to the increase in feed salinity and associated osmotic pressure as the permeate is removed from the vessel

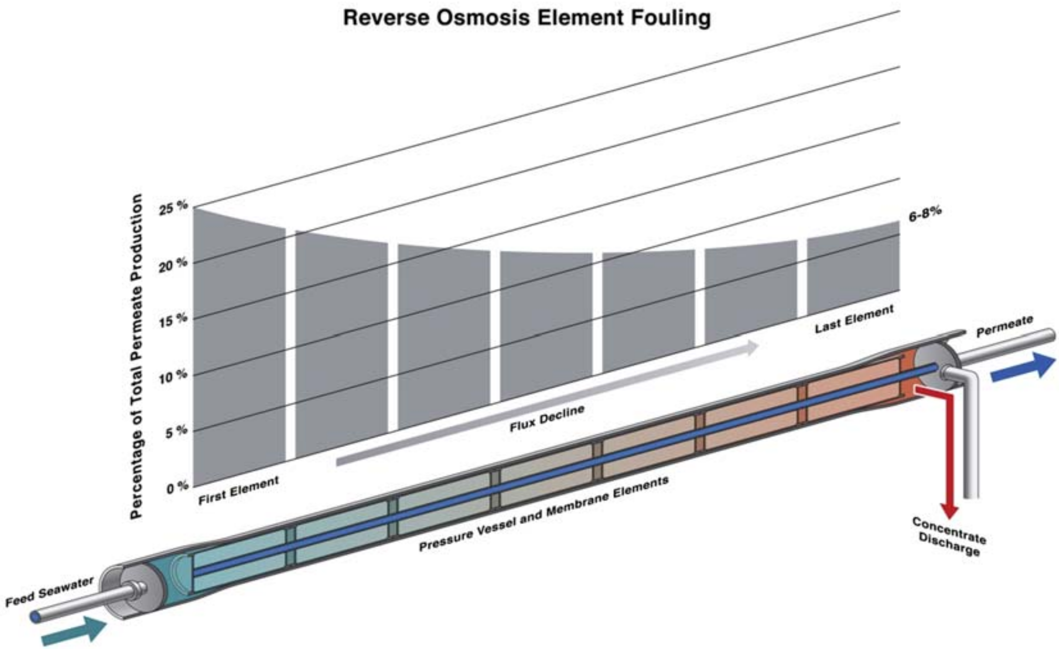

FIGURE 1.4 Membrane fouling and flux distribution in membrane vessel.

while the concentrate rejected from all elements remains in the vessel until it exits the last element.

Since the first element processes the largest portion of the feed flow it also receives and retains the largest quantity of the particulate and organic foulants contained in the source water and, consequently, is most impacted by biofouling. The reminder of the feed water that does not pass through the first RO element combines with the concentrate from this element and enters the feed channels of the second RO element of the vessel. This element is exposed to higher salinity feed water and lower feed pressure (energy) because some of the initially applied pressure (energy) has already been used by the first RO element of the vessel to produce the permeate. As a result, the permeate flow rate (flux) of the second element is lower and the concentrate polarization on the surface of this element is higher than that of the first RO element.

The subsequent membrane elements are exposed to increasingly higher feed salinity concentration and elevated concentration polarization, which results in progressive reduction of their productivity (flux). As flux through the subsequent elements is decreased, accumulation of particulate and organic foulants on these elements diminishes and biofilm formation is reduced. However, the potential for mineral scale formation increases because the concentration of salts in the boundary layer near the membrane surface increases.

Therefore, in RO-systems fouling caused by accumulation of particulates, organic matter and biofilm formation is usually most pronounced on the first and second membrane elements of the pressure vessels, while the last two RO elements are typically more prone to mineral scaling than the other types of fouling (Chesters et al., 2011).

The flux distribution pattern in RO-membrane vessels depicted on Fig. 1.4 could change significantly as a result of the membrane-fouling process. If the source seawater contains a large amount of foulants of persistent occurrence, as the first element if completely fouled, its permeability (flux) over time will be reduced below its typical level (±25% of total vessel production) and the flux of the second RO element will be increased instead. After the fouling of the second RO element reaches its maximum, a larger portion of the feed flow will be redistributed down to the third RO element, until all elements in the vessel begin to operate at a comparable and significantly lower flux.

Flux redistribution caused by particulate fouling, NOM deposition and/or biofouling can trigger scale formation on the membrane surface of the last RO element, which would not occur under normal flow distribution pattern (low-fouling conditions) shown on Fig. 1.4. The main reason for this phenomenon is that the concentrate polarization on the surface of the last RO element typically increases over two times as a result of this flux redistribution. As indicated previously, in a typical 7-element-per-vessel configuration and nonfouling conditions, the last element would operate at flux that is only 6%–8% of the average vessel flux. Under fouling-driven flux redistribution in the membrane vessel, the flux of the last element will increase to 12%–14% (i.e., it would be approximately two times higher than usual). Since membrane polarization is exponentially proportional to flux, if the RO system is operated at the same recovery, the likelihood for scale formation on the last one or two RO elements increases exponentially.

In addition to increasing the potential for mineral fouling (scaling) on the last one or two membrane elements, the long-term operation of fouled RO system is not advisable because of the higher feed pressure (energy) needed to overcome the decreased membrane permeability, if the system is operated to produce the same permeate flow. As the RO-system feed pressure reaches certain level (usually 75–85 bars for SWRO desalination systems), the external membrane fouling would be compounded by internal fouling due to the physical compaction of the membrane structure, which in turn would cause irreversible damage of the membranes.

Therefore, understanding of the causes and mechanisms of RO-membrane fouling are of critical importance for the successful design and operation of RO desalination plants.

References

Chesters, S., Pena, N., Gallego, S., Fazel, M., Armstrong, M., de Vigo, F., 2011. Results from 99 seawater RO membrane autopsies. In: Proceedings of IDA World Congress, Perth, Western Australia, September 4–9, 2011.

Henthorne, L., Boysen, B., 2015. State-of-the-art of reverse osmosis desalination pretreatment. Desalination 356, 129–139. Elsevier.

Herzberg, M., Elimelech, M., 2007. Biofouling of reverse osmosis membranes: role of biofilm-enhanced osmotic pressure. Journal of Membrane Science 295, 11–20. Elsevier.

Hoek, E.M.V., Bhattacharjee, S., Elimelech, M., 2003. Effect of surface roughness on colloid membrane DLVO interactions. Langmuir 19, 4836–4847.

Hoek, E.M.V., Agarwal, G., 2006. Extended DLVO interactions between spherical particles and rough surfaces. Journal of Colloid and Interface Science 298 (1), 50–58.

Jamaly, S., Darwish, N.N., Ahmed, I., Hasan, S.W., 2014. A short review of reverse osmosis pretreatment technologies. Desalination 354, 30–38.

Villacorte, L., Tabatabai, S.A.A., Anderson, D.M., Amy, G.I., Shippers, J.C., Kennedy, M.D., 2015. Seawater reverse osmosis desalination and (harmful) algal blooms. Desalination 360, 61–80.

Wilf, M., Awerbuch, L., Bartels, C., Mickley, M., Pearce, G., Voutchkov, N., 2007. The Guidebook to Membrane Desalination Technology. Balaban Desalination Publications, L'Aquila, Italy.

CHAPTER

2

Membrane Foulants and Saline Water Pretreatment

OUTLINE

Pretreatment for Reverse Osmosis Desalination
http://dx.doi.org/10.1016/B978-0-12-809953-7.00002-4

2.1 INTRODUCTION

Saline water is collected from a source using either subsurface intakes (wells, intake galleries, etc.) or open surface intakes. Subsurface intakes naturally prescreen and prefilter the collected saline water and thereby they remove coarse debris and most of the sand and particulates from this water. Open ocean intakes that collect ambient water directly from saline surface water source have equipment (bar racks, fine traveling screens, microscreens, and/or strainers) to prescreen large debris, floating materials, large aquatic organisms, coarse sand, and stringy materials from the source water.

As a result, after preliminary screening by the intake facilities, saline water typically contains the following five key groups of compounds that could cause RO-membrane fouling and therefore, would need to be removed by the pretreatment system:

1. Particulate foulants (mainly suspended solids and silt);
2. Colloidal foulants—compounds of relatively small size (0.2–1.0 μm) that are not in fully dissolved form, which when concentrated during the membrane separation process may coalesce and precipitate on the membrane surface (mainly clay-like substances);
3. Mineral-scaling foulants—inorganic compounds (i.e., Ca, Mg, Ba, and Sr salts), which during the salt separation process may precipitate and form scale on the membrane surface (such as calcium carbonate and sulfate, and magnesium hydroxide) or may block the membrane separation layer (such as iron and manganese);
4. Natural organic foulants—natural organic matter (NOM) that can attach to and foul the membranes;
5. Microbial foulants—marine organisms and organic compounds released by them, which can serve as food to the microorganisms that inhabit in the source water and can form fouling biofilm, reducing freshwater transport through the membranes.

2.2 PARTICULATE FOULANTS

2.2.1 Description

Particulate foulants are organic and inorganic particles contained in the source water such as fine debris, plankton, detritus, and silt. These solids cannot pass through the RO membranes. All suspended solids that naturally occur in insoluble form, if not removed by pretreatment, would be retained on the feed side of the RO-membrane leafs and depending on the hydrodynamic conditions on the membrane surface, as well as the size and charge of these particles, and they would either migrate along the membrane leafs and ultimately exit with the concentrate, or would be trapped on the membrane surface and would begin to accumulate there causing loss of membrane productivity over time. This type of foulants can be effectively removed by prefiltering of the source water prior to RO-membrane separation.

Particulate foulants in raw source saline water vary greatly in size. However, most of them, including picoplankton, are larger than 0.1 μm. A well-designed and operating pretreatment system will produce water that does not have particles larger than 20 μm in size. Typically, suspended solids of size larger than 100 μm contained in saline water are settleable and can be removed by clarification of the source water prior to filtration. However, many bacteria that inhabit seawater and cause RO-membrane biofouling are smaller in size than 0.1 μm and are practically impossible to remove completely from the source seawater even if the most advanced pretreatment technologies are applied. This makes pretreatment of seawater very challenging.

2.2.2 Measurement Parameters and Methods

2.2.2.1 Turbidity

Turbidity is a water-quality parameter, which measures the content of particulate suspended solids in water. The turbidity level in the source water is indicative of the content of clay, silt, suspended organic matter, and microscopic aquatic life, such as phyto- and zooplankton. It is expressed in *nephelometric turbidity units* (NTUs).

The turbidity of open ocean and surface brackish waters typically varies between 0.1 and several hundred NTU, although under normal dry weather conditions, it is usually in a range between 0.5 and 2.0 NTU. Rain events, algal blooms, storms, snowmelt, river discharges, and human activity (such as wastewater discharges, ship traffic, etc.) can cause significant turbidity increases and variations.

Usually, saline water with turbidity below 0.05 NTU causes very low particulate fouling of RO membranes. Most RO-membrane manufacturers require RO-system feed water turbidity of 1.0 NTU or less in order to maintain their performance warranty, although this level is relatively high in practical terms. Usually, it is desirable that the turbidity of the feed water to the RO-membrane elements is lower than 0.1 NTU in order to prevent accelerated particulate RO-membrane fouling.

Although turbidity is a good measure of the overall content of suspended solid particles in the source water, on its own, it is not an adequate quantitative parameter to characterize water's potential to cause particulate or other fouling of the downstream RO membranes. Turbidity measurement does not provide information regarding the type and size of particles in the source water and does not quantify the content of dissolved organic, inorganic and microbial foulants. The size of particles contained in the source water matters because RO-membrane feed and concentrate spacers, through which the saline source water is distributed inside the membranes, are of limited width (typically 0.7–0.9 mm).

2.2.2.2 *Silt Density Index (SDI)*

Silt density index (SDI) is a parameter that provides an indication of the particulate fouling potential of the saline source water. If RO system is operated at a constant transmembrane pressure, particulate membrane fouling will result in decline of system productivity (membrane flux) over time. SDI gives an indication of the rate of flux decline through a filter of standard size and diameter operated at a constant pressure for a given period of time.

A standard SDI test procedure is described in ASTM Standard D4189-07 (AWWA, 2007) and is based on the measurement of the time in seconds it takes to collect a 500-mL sample through a paper filter of size 0.45 μm and diameter of 45 mm both at the start of the test ($t_0 = 0$ min) and after the source water has passed through the filter under a driving filtration pressure of 2.1 bars (30 lb/in.^{2}) for a certain period, "n" measured in minutes (standard time for measurement of filtered water SDI, $t_n = 15$ min). The two sample durations (t_0 and t_n) are applied to a formula (Eq. 2.1), and the resulting SDI$_n$ value indicates the particulate fouling potential of the tested source water:

$$\mathrm{SDI}_n = (1 - (t_0/t_n))/n \times 100 \tag{2.1}$$

where n is the total test run time (which for the standard test of filtered water is 15 min) and t_0 and t_n are the respective times in seconds it takes to collect 500 mL of filtered water at the beginning of the test and after running water through the filter for the duration of the selected test run time.

Fig. 2.1 shows a typical SDI measurement system installed at a desalination facility.

It should be pointed out that while the standard time between the first and second measurements for SDI test of filtered (pretreated) source water is 15 min, this test is typically run for 5 or 10 min on untreated source water, depending on the solids' concentration. Based on this formula, the maximum value of SDI$_{15}$ is 6.7. This condition would occur if the time to collect 500 mL after 15 min of filtration were infinite.

Typically, source water with an SDI$_{15}$ lower than 4 is considered to have adequately low RO-membrane particulate fouling potential, and its use in membrane desalination is expected to result in a reasonably slow membrane flux decline over time (10%–15% RO system productivity reduction over a 3–4 month period). SDI$_{15}$ levels of the RO feed water in a range of 2–4 are indicative of well-performing pretreatment system, and if the pretreatment system produces water with SDI$_{15}$ less than 3 at all times, this pretreatment system would be considered of excellent particulate removal efficiency.

Source water with an SDI$_{15}$ lower than 2 typically has very low particulate fouling potential and is not likely to require RO-membrane cleaning more frequently than once every 6 months—industry standard is RO-membrane cleaning once every 3–4 months. Source water with SDI$_{15}$ lower than 1 usually results in very slow increase in fouling and associated differential pressure (DP) and would yield RO-membrane cleaning intervals of 12 months or longer.

FIGURE 2.1 Typical SDI measurement device used in full-scale desalination plants.

In order to maintain their performance warranties, membrane manufacturers usually require that the SDI_{15} of the source water fed to the RO membranes to be less than 5 at all times and less than 4 at least 95% of the time. If the source water's SDI_{15} is higher than 5, this typically means that this water has a very high content of particulate foulants and, therefore, it is not directly suitable for desalination, because it would cause accelerated fouling of the RO membranes. For source seawater with SDI_{15} value higher than 5, it is often useful to complete the SDI test at shorter (10, 5 or 2.5 min) intervals, which will usually provide more meaningful results (Mosset et al., 2008).

Color and appearance of the SDI pads is also indicative of the content of various foulants in the source water and the effectiveness of the plant pretreatment system. Fig. 2.2 shows filtration pads from SDI_{15} tests along with the SDI_{15} values of pretreated seawater. The SDI test pads referenced as "D2" are from iron-salt coagulated seawater pretreated via two-stage upflow sand media filters, while these referenced as "Zenon" are from noncoagulated UF filter effluent.

The top two SDI filter pads (5.3 SDI and 5.2 SDI) on Fig. 2.2 from left to right have a red discoloration caused by overdose of ferric coagulant. These SDI tests indicate that the pretreated influent is not suitable for SWRO treatment due to high particulate content in the source seawater. The other SDI pads are indicative of pretreated filter effluent of relatively low particulate-fouling potential.

It should be pointed out that although widely used in practice today, the SDI_{15} test has limited ability to measure the particulate fouling potential of saline water. The test is based

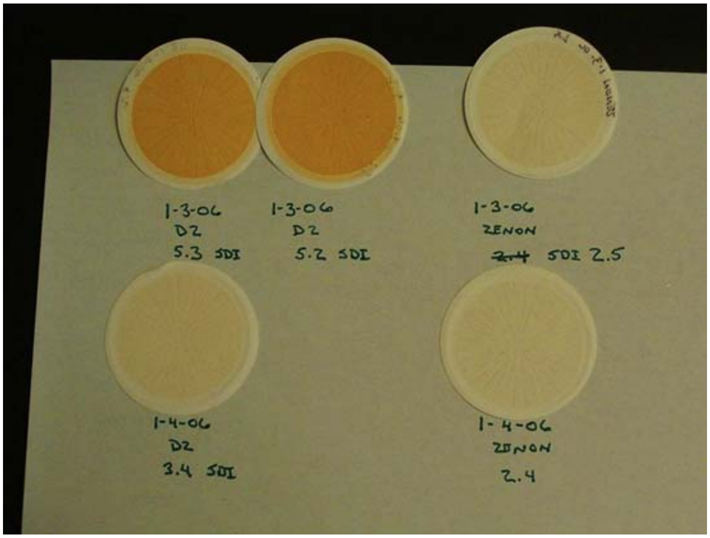

FIGURE 2.2 SDI pads from tests of pretreated seawater of various quality.

on filtration of saline water sample through microfiltration (MF) membrane pad with 0.45-μm pore openings. Therefore, the SDI measurement is mainly indicative of the content of particulates and colloids in the source water that can be retained by such size filters. The fouling mechanisms of permeable MF membranes and semipermeable RO membranes may differ significantly, depending on the types of foulants contained in the source water.

Fouling of MF and UF membranes usually occurs due to a combination of micropore plugging and cake formation on the membrane surface. In contrast, RO membranes are typically fouled by the formation of cake of deposits on the membrane surface (without pore plugging). Since smaller particles create significantly higher resistance in the filter cake than big particulates, their effect on RO-membrane fouling can be much more pronounced than that of the large-size particles captured by the SDI test.

In order to address the effect of smaller particles on SWRO-membrane fouling, a number of alternative filtration indices have been developed over the past 10 years. Advantages and disadvantages of these modified fouling indices are discussed in the following publications (Khirani et al., 2006; Boerlage, 2007; Mosset et al., 2008).

2.2.2.3 *Total Suspended Solids*

Total suspended solids (TSS) concentration is a measure of the total weight of solid residuals contained in the source water and is customarily presented in milligrams per liter (mg/L). TSS is measured by filtering a known volume of water (typically 1 L) through a preweighed glass-fiber filter, by drying the filter with the solids retained on it at 103°C, and then by weighing the filter again after drying. The difference between the weight of the dried filter and of the clean filter, divided by the volume of the filtered sample, reflects the total amount of particulate (suspended) solids in the source water.

It should be pointed out that because saline water contains dissolved solids that will crystallize and convert into particulate solids when the sample is heated at 103°C, often TSS analysis of saline water completed in accordance with the standard methods for water and wastewater analysis (APHA, 2012) yields an erroneously high TSS content in the water. The higher the source water salinity and the lower its particulate content, the more inaccurate this measurement is. In order to address the challenge associated with the standard method of TSS measurement, it is recommended to wash the solids retained on the filter by spraying the filter with deionized water before drying. Unless the source water solids sample is washed before drying, the results of this sample are meaningless.

If the laboratory TSS test is completed properly and the filtered sample is well washed, this parameter usually provides much more accurate measure of the actual content of particulate solids in the source water than does turbidity, because it accounts for the actual weight of the total particulate material present in the sample. For comparison, turbidity measurement is dependent on particle size, shape, and color, and typically is not reflective of particles of very small size (i.e., particles of 0.5 μm or less), such as fine silt and picoplankton.

In fact, a change in the ratio of TSS to turbidity is a good indicator of a shift in the size of particles contained in the source water, which may be triggered by algal blooms, storms, strong winds, and other similar events, which can result in resuspension of solids from the bottom sediments into the water column.

Typically, an increase in the TSS/turbidity ratio is indicative of a shift of particulate solids toward smaller-size particles. For example, during non-algal-bloom conditions, the TSS/turbidity ratio of an appropriately processed sample is typically in the range of 1.5–2.5, i.e., water with a turbidity of 2 NTU would have a TSS concentration of 3–5 mg/L. During heavy algal blooms dominated by small-size (pico- and micro-) algae, the TSS of the source water could increase over 10 times (e.g., to 40 mg/L), while the source water turbidity could be multiplied by 2 to 3 times only—for this example, it would be in a range of 4–6 NTU, with a corresponding increase in the TSS/NTU ratio from 2/1 to between 6/1 and 10/1.

2.2.2.4 *Particle Distribution Profile*

Particle distribution profile presents the number of solid particles in the source water for different particle size ranges. Usually, particles are classified in the following ranges: 1 μm or less; >1 and ≤2 μm; >2 and ≤5 μm; ≥5 and ≤10 μm; >10 and ≤20 μm; >20 and ≤50 μm; and higher than 50 μm.

Particles of sizes smaller than 1 micron (μm) are poorly removed by conventional granular media filtration and DAF clarification. Key pathogens such as *Giardia* and *Cryptosporidium* are in a range of 2–10 μm and typically 2 logs of these pathogens are removed by granular media filtration and 4–6 logs by MF or UF membrane filtration.

Particles that have sizes of higher than or equal to 20 μm are well removed by both membrane and granular media pretreatment filters. Typically, pretreatment system is considered to perform well if the pretreated water contains less than 100 particles per milliliter of size equal to or smaller than 2 μm and does not contain larger size particles.

2.2.3 Threshold Levels of Particulate Foulants

Table 2.1 presents a list of source water quality parameters used for characterizing particulate content that are recommended to be measured when deciding upon the type of pretreatment needed for a given source water.

TABLE 2.1 Water Quality Parameters for Characterization of Particulate Foulants

Source Water Quality Parameter	Pretreatment Issues and Considerations
Turbidity (NTU)	Levels above 0.1 NTU are indicative of high potential for fouling.
	Spikes above 50 NTU for more than 1 h would require sedimentation or DAF treatment prior to filtration.
Silt density index (SDI_{15})	Source seawater SDI_{15} levels consistently below 2 (year-around) indicate no pretreatment is needed.
	$SDI_{15} > 4$—pretreatment is necessary.
Total suspended solids (mg/L)	This parameter is needed to assess the amount of residuals generated during pretreatment.
	TSS does not correlate well with turbidity when it exceeds 5 NTU.

2.3 COLLOIDAL FOULANTS

2.3.1 Description

Colloidal foulants are inorganic and organic compounds that naturally exist in suspension and may precipitate on the RO-membrane surface, thereby causing membrane flux to decline over time. Colloidal solids have a particle size of 0.001−1 μm. For prevention of colloidal fouling, RO-membrane manufacturers usually recommend a feed turbidity of less than 0.1 NTU and an SDI_{15} lower than 4.

Metal oxide and hydroxide foulants most frequently encountered in brackish and seawater desalination are iron, manganese, copper, zinc, and aluminum. Typically, open ocean seawater contains very low levels of these metal foulants and, therefore, if such fouling is encountered on the membrane elements, the usual sources are overdosing of coagulant (iron salt) or corrosion of pipes, fittings, tanks, and other metal equipment located upstream of the RO system.

2.3.1.1 Iron and Manganese

Iron and manganese fouling may occur if source water is collected via subsurface intake from a brackish aquifer with high content of these metals or from a coastal aquifer, which is under the influence of fresh groundwater that contains high levels of these compounds in reduced form (i.e., iron of more than 2 mg/L as ferrous or manganese of more than 0.5 mg/L).

If iron and manganese are in reduced form and they are below 1.0 and 0.1 mg/L respectively, then they can be removed by RO membranes without causing accelerated fouling. However, if iron and manganese are in oxidized form, their levels should be reduced below 0.05 and 0.02 mg/L, respectively, to prevent mineral fouling of the membranes.

In addition to naturally occurring colloidal matter, iron colloidal fouling on the surface of RO membranes may be caused by corrosion of upstream piping and equipment or by overdosing or poor mixing of iron-based coagulant used for conditioning of the saline source water. If the source water contains chlorine, this colloidal iron tends to catalyze the oxidation process caused by chlorine, which in turn enhances the damage of the RO membranes even when residual chlorine in the saline source water is in very small doses (e.g., less than 0.05 mg/L).

2.3.1.2 Silica

Another colloidal fouling compound frequently encountered in brackish groundwater aquifers is silica. Total silica (silicon dioxide) in the source water consists of reactive silica, which is in soluble form, and unreactive silica, which is in colloidal form. While reactive silica is not a challenge for RO membranes, colloidal silica in the saline source water can cause significant membrane fouling. It should be pointed out, however, that elevated content of silica in colloidal form is mainly found in brackish water sources. Unreactive silica is present in very low levels in seawater and could pose a fouling challenge only when its level in the concentrate exceeds 100 mg/L.

The stability of colloids is reduced with an increase in source water salinity, and therefore, typical saline water with TDS concentration in a range of 30,000–45,000 mg/L would contain silica in dissolved and precipitated forms rather than in colloidal form. Open ocean seawater typically contains silica of less than 20 mg/L, and therefore this compound does not cause colloidal fouling of RO membranes.

However, if the source saline water is collected via a subsurface well intake that is under the influence of a brackish coastal aquifer with a high content of colloidal silica, or it is collected near an area where a silt-laden river enters into the ocean, then colloidal fouling may become a challenge. Colloidal foulants can be removed by coagulation, flocculation, sedimentation, and filtration, similar to particulate foulants.

2.3.1.3 Polymers

Another common source of colloidal fouling is overdosing or poor mixing of polymers used for conditioning (flocculation) of the saline source water prior to clarification or filtration. Such problem usually occurs when the source water has low turbidity (usually less than 0.5 NTU) and low electrical charge (zeta potential lower than −10 mV) and when polymer is added to enhance the flocculation of such particles. Practical experience shows that addition of polymers and coagulants to saline waters of low turbidity usually accelerates particulate and colloidal fouling of the RO membranes.

2.3.1.4 Hydrocarbons

The most common organic colloidal foulants are oil-product-based hydrocarbons. Such compounds are not contained naturally in most brackish waters or in open ocean seawater, and their occurrence indicates that the saline water intake area or aquifer is under influence of man-made sources of contamination—typically discharge from a wastewater treatment plant, from storm drains collecting surface runoff from urban areas (parking lots, industrial sites, etc.) or from waste discharges or oil leaks released by ships, boats, or near-shore oil storage tanks in port areas.

Even in very small quantities (0.02 mg/L or more), hydrocarbons can cause accelerated fouling of RO membranes. Therefore, it is desirable that the total hydrocarbon (THC) content in the source water fed to the RO be maintained below 0.02 mg/L at all times.

It should be pointed out that except for DAF clarification, which can reduce THC to less than 95% of their source water level, all other pretreatment systems provide very limited (5%−15%) hydrocarbon removal. Hydrocarbons cause practically irreversible fouling of the RO membranes, and if saline water with hydrocarbon levels over 0.1 mg/L is fed to the RO membranes, they will be permanently destroyed within 1 to 2 h after the hydrocarbons reach the membranes.

2.3.2 Measurement Parameters and Methods

Silica, colloidal iron, and manganese are measured by standard laboratory tests (APHA, 2012). Usually colloidal (also referred to as unreactive) silica is determined by measuring total and reactive silica in the source water.

Both oil & grease and THC are used as measurement parameters for hydrocarbon content in the saline source water. However, measurement of THC concentration is preferable because it could be completed continuously using online monitoring analyzer, which can detect THC content at levels significantly lower than the threshold level that can cause permanent fouling of the RO membranes. For comparison, oil and grease can be measured by laboratory analytical methods only and therefore is not suitable for continuous monitoring of hydrocarbon content in the source water.

Most desalination plants using surface water sources (e.g., saline river estuary or ocean water) are equipped with online THC analyzers because they allow for practically instantaneous detection of elevated THC content in the source water, which in turn permits the automatic shutdown of the desalination plant intake facilities as soon as THC concentration approaches the critical threshold level of 0.02 mg/L.

2.3.3 Threshold Levels of Colloidal Foulants

Table 2.2 presents the threshold levels of key colloidal foulants. If the saline source water contains colloidal foulants exceeding the threshold levels in this table, the RO membranes processing this water will be exposed to accelerated and potentially irreversible fouling.

TABLE 2.2 Water Quality Parameters for Characterization of Colloidal Foulants

Source Water Quality Parameter (mg/L)	Pretreatment Issues and Considerations
Iron	If iron is in reduced form, RO membranes can tolerate up to 2.0 mg/L.
	If iron is in oxidized form, concentration >0.05 mg/L would cause accelerated fouling.
Manganese	If manganese is in reduced form, RO membranes can tolerate up to 0.1 mg/L.
	If manganese is in oxidized form, concentration >0.02 mg/L would cause accelerated fouling.
Silica	Concentrations higher than 100 mg/L in the concentrate may cause accelerated fouling.
Total hydrocarbons	Concentrations higher than 0.02 mg/L would cause accelerated fouling.

2.4 MINERAL-SCALING FOULANTS

2.4.1 Description

All minerals contained in the saline source water are concentrated during the process of membrane salt separation. As their concentration increases during the desalination process, ions of calcium, magnesium, barium, strontium, sulfate, and carbonate can form salts, which could precipitate on the RO-membrane surface. The mineral scales that typically form during desalination are those of calcium carbonate, calcium and magnesium sulfate, and barium and strontium sulfate.

The potential for formation of these compounds (e.g., the scaling potential of source water) depends on their content in the concentrate flow stream, water temperature, pH, desalination plant recovery, and other factors. Table 2.3 indicates the impact of the key source water quality and operational factors on the scaling potential of the saline source water.

Formation of mineral scales on the membrane surface is balanced by the high salinity of the source water, which tends to increase the solubility of all salts. This means that the higher the salinity of the source water, the less likely mineral scale would form on the membrane surface at typical saline water pH of 7.6–8.3 and desalination system recovery of 45%–50%.

In brackish water desalination systems that typically operate at much higher recoveries (65%–85%) and use source waters that have relatively lower ionic strength, mineral scaling is a frequent problem.

In typical seawater desalination systems operating at recovery in a range of 40%–50%, mineral-scale fouling is usually not a challenge, unless the seawater pH has to be increased to 8.6 or more in order to enhance boron removal by the RO membranes.

Calcium carbonate is the most commonly encountered mineral foulant in brackish water desalination plants. Calcium sulfate and magnesium hydroxide are the most frequent causes of SWRO membrane scaling. Scale formation can be prevented by addition of antiscalant/dispersant to the source water.

TABLE 2.3 Factors Impacting Scaling Potential of Saline Source Water

Source Water Factor	Impact on Scaling Potential of Source Water	Note
Source water TDS concentration increase	Decrease in scaling potential	Seawater typically does not scale at ambient pH and temperature <35°C if plant recovery is <45%.
Ca, Mg, Ba, Sr, Si, F, SO_4, CO_3, HCO_3 concentration increase	Increase in scaling potential	$CaCO_3$ is common scale in BWRO plants while $CaSO_4$ and $MgSO_4$ are common scales in SWRO plants.
pH increase	Increase in scaling potential	Increase of seawater above 8.6 (pH > 8.6) accelerates scaling.
Temperature increase	Increase in scaling potential	Scaling accelerates significantly when temperature exceeds 35°C.
RO system recovery increase	Increase in scaling potential	Scaling potential is low if recovery <45%.

2.4.2 Measurement Parameters and Methods

Commonly used parameters that can be used to predict source water potential to form mineral scale of calcium carbonate are the Langelier Saturation Index (LSI) and the Stiff and Davis Saturation Index (SDSI). These indexes are function of the source seawater pH, calcium concentration, alkalinity, temperature, and TDS concentration/ionic strength.

Langelier Saturation Index can be calculated using the following equation:

$$LSI = pH - pH_s \tag{2.2}$$

where: pH is the actual pH of the saline source water;

$$pH_s = (9.30 + A + B) - (C + D) \tag{2.3}$$

$A = (Log_{10}\ (TDS - 1)/10)$, where TDS is in mg/L; $B = -13.2 \times Log_{10}(Temperature + 273) + 34.55$, where Temperature is in °C; $C = Log_{10}\ (Ca^{2+}) - 0.4$, where Ca^{2+} concentration is expressed in mg/L of $CaCO_3$; $D = Log_{10}$ (Alkalinity), where Alkalinity is expressed in mg/L of $CaCO_3$.

The value of LSI is indicative of the ability of the source water to form calcium carbonate scale only and is not reflective of formation of other scaling compounds. If LSI is higher than 0.2, the source water is likely to cause slight scaling and if it is above 1.0, the source water would cause severe scaling on the membranes. If LSI is negative, then the water has tendency to dissolve scale.

The LSI index predicts scaling potential of the source water only if its TDS is lower than 4000 mg/L. For saline source waters of higher TDS, the Stiff-Davis Saturation Index (SDSI) is applied. This index can be calculated as follows:

$$SDSI = pH - pCa - p_{alk} - K \tag{2.4}$$

where: $pCa = -Log\ (Ca^{2+})$, where Ca^{2+} is in mg/L as $CaCO_3$; p_{alk}: Log_{10} (Total Alkalinity), where Alkalinity is in mg/L as $CaCO_3$; K: constant that is function of total ionic strength and temperature.

Values of K, and nomographs and examples for calculation of LSI and SDSI indexes are presented elsewhere (AWWA, 2007).

2.4.3 Threshold Levels of Scaling Foulants

Table 2.4 presents the threshold levels of common mineral scalants in concentrate above which these compounds will begin to accumulate on the membrane surface and form mineral scales (Hydranautics, 2008). The mineral-scaling potential of the saline source water can be determined using proprietary software available from the manufacturers of antiscalants.

TABLE 2.4 Threshold Levels of Common Scaling Compounds

Scalant	Maximum Increase in the Concentrate as Compared to Source Water Above Which RO Membrane Scaling Is Likely
Calcium carbonate, expressed as LSI, in source water without scale inhibitor	+0.3
Calcium sulfate	230%
Strontium sulfate	800%
Barium sulfate	6,000%
Calcium fluoride	12,000%

2.5 NATURAL ORGANIC FOULANTS

2.5.1 Description

Depending on its origin, saline waters can contain naturally occurring or man-made organic compounds and aquatic microorganisms. Since all microorganisms and most organic molecules are relatively large in size, they are well rejected by the RO membranes. However, some of the organic compounds and aquatic species may accumulate on the membrane surface and form a cake layer that may significantly hinder membranes' main function—rejection of dissolved solids. Depending on the main source of the fouling cake layer, these RO membrane foulants are typically divided into two separate groups: natural organic matter (NOM) foulants and microbiological foulants (or biofoulants).

NOM is typically contained in surface saline waters (brackish water or open ocean seawater) and includes compounds that are produced by naturally decaying algal and other aquatic vegetation and fauna [i.e., proteins, carbohydrates, oils, pigments (i.e., tannins), and humic and fulvic substances (acids)]. High content of NOM in the source water used for production of drinking water is undesirable because it causes discoloration of the water, forms carcinogenic disinfection by-products (DBPs) when disinfected with chlorine, and results in complexation with heavy metals, which in turn causes an accelerated membrane fouling.

Under normal non-algal bloom conditions, typical seawater and brackish water collected using open intakes do not contain NOM concentration large enough to present significant challenge to desalination plant operations. High content of NOM is usually observed during algal blooms and/or when the desalination plant intake is located near the confluence with river or other freshwater source, or near wastewater-treatment plant discharge.

The easily biodegradable organic matter released by algae during their growth and respiration is referred to as extracellular organic matter or EOM (Edzwald and Haarhoff, 2011). When algae die off during the end phase of an algal bloom or their cells are broken by treatment or pumping processes, they also release intracellular organic matter (IOM) in the source water. The combination of EOM and IOM of algae is referred to algogenic organic matter (AOM). AOM is easily biodegradable NOM and provides food source for biogrowth of bacteria on the RO-membrane surface.

Humic acids are polymeric (poly-hydroxyl aromatic) substances, which have the ability to form chelates with metal ions in the saline water, such as iron. This feature of humic acids is very important for seawater or surface brackish water pretreatment systems using iron coagulants, because they can form a gel-like layer of chelates on the surface of the membranes, which would cause fouling. Typically, such fouling layer can be dissolved at pH of 9 or more, at which condition both the membranes and the humic substances carry a negative charge. This feature is used for membrane cleaning.

Humic substances are hydrophobic and therefore, hydrophilic membranes are less prone to fouling by humic acids.

Most NOM in saline water consists of compounds of relatively large molecular weights (500−3000 Da). One Dalton (Da) is equal to $1.666{,}054 \times 10^{-24}$ g. A typical saline water reverse osmosis desalination membrane would reject over 90% of the compounds that have molecular weight higher than 200 Da.

Humic and fulvic substances mainly differ by their ability to dissolve in strong acids. While humic substances are easily precipitated upon acidification of saline water, fulvic substances remain in solution at low pH.

Humic acids in their natural state are not a food source for most aquatic organisms. However, when oxidized with chlorine or other oxidants, humic acids could become easily biodegradable and serve as food source for aquatic bacteria growing on the RO-membrane surface. Therefore, continuous chlorination of source water containing large amount of humic acids often causes more membrane-biofouling problems than it solves.

Negatively charged NOM, which dominates in surface brackish water and seawater collected by open intakes, has the tendency to adhere to the surface of the RO membranes, which typically have slightly positive charge. Once adsorption occurs, the NOM begins to form cake/gel on the membranes and to affect their performance. It should be noted that depending on its properties and origin, NOM may also adhere to the surface of UF and MF pretreatment membranes and cause significant productivity loss by plugging membrane pores, adsorbing to the internal matrix of the membranes, and forming a cake of organic matter on the membrane surface.

Usually, saline water from surface saline water sources contains NOM that causes moderate fouling of UF and MF membranes, which could be removed by routine chemically enhanced backwash and periodic cleaning of the pretreatment membranes. This NOM can be removed with very little (typically less than 2 mg/L) or no coagulant addition to the saline source water. However, if the source water is influenced by surface runoff or large amount of alluvial organics, or if alluvial brackish aquifer is used for water supply, the NOM properties and ability to cause significant membrane productivity loss may increase dramatically. Under these circumstances, the efficient removal of NOM may require very high dosages of coagulant (usually over 20 mg/L).

A NOM fractionation study on surface waters completed by U.S. Bureau of Reclamation in 2002 demonstrates that this type of fouling depends on the type of membrane material and characteristics, the NOM polarity and molecular weight, and the chemistry of the source water.

The largest natural organic foulants are the polysaccharides, organic colloids and proteins, followed in size by humic substances, organic acids, and low molecular weight organics of neutral charge. These compounds have different potential to cause membrane fouling. Polysaccharides along with living bacteria that excrete them and attach on the membrane surface have the highest potential to cause RO-membrane biofouling and therefore, they are classified as a separate group of foulants—microbial foulants.

2.5.2 Measurement Parameters and Methods

The most frequently used parameters for quantification of natural organic fouling are total organic carbon (TOC) and UV_{254} absorbance.

2.5.2.1 Total Organic Carbon

Total organic carbon is one of the most widely used measures for organic content of saline source water. TOC concentration measures the content of both NOM and of easily biodegradable organics, such as polysaccharides, released during algal blooms.

This water quality parameter is widely used, because it is relatively easy to measure and it is indicative of the tendency of the source water to cause RO-membrane natural organic fouling and microbial fouling. TOC is measured by converting organic carbon to carbon dioxide in high-temperature furnace in the presence of a catalyst.

Typically, open ocean seawater—which is not influenced by surface freshwater influx (nearby river confluence); by man-made activities (i.e., wastewater or storm water discharges, or ship traffic); or by algal bloom event (i.e., red tide)—has a very low TOC content ($\leq$0.2 mg/L). When an algal bloom occurs, however, TOC concentration of the ocean water could increase by an order of magnitude (2–12 mg/L). Similar magnitude of TOC increase could be triggered by a storm water or river discharge during high-intensity rain event such as these occurring during rainy seasons in tropical and equatorial parts of the world.

Usually, an increase of TOC content in the source water above a certain threshold (2.0–2.5 mg/L) is observed to trigger accelerated biofouling of RO membranes.

Observations at the Carlsbad seawater desalination demonstration plant in California, USA (which is supplied by seawater collected using near-shore open ocean intake) indicate that when TOC concentration in the source water at that location exceeds 2.0 mg/L during algal bloom events, within one-to two-week period, the SWRO system experiences measurable biofouling and associated increase in operating pressure.

Similar TOC level observations at the Tampa seawater desalination plant, in Florida, USA, (where the typical background TOC level of the seawater is less than 4 mg/L) indicate that accelerated biofouling occurs when TOC concentration exceeds 6–8 mg/L. Usually, accelerated biofouling at the Tampa facility is triggered by one of two events—heavy rains, which increase the content of alluvial organics in the source seawater, or algal blooms, which cause elevated organic concentration due to massive die-off of algae. The increase in alluvial organics during rain events is caused by the elevated flow and alluvial content of Alafia River, which discharges into Tampa Bay several kilometers upstream of the desalination plant intake. During high-intensity rains in the summer months, TOC level in the river water discharging into the bay may exceed 20 mg/L.

TABLE 2.5 TOC Content and Fractions of Various Seawater Sources

Seawater Source	TOC (mg/L)	Polysaccharides (% of Total TOC)	Humic Substances & Building Blocks (% of Total TOC)	Low Molecular Weight Acids & Neutrals (% of Total TOC)	Other Low Molecular Weight Compounds (% of Total TOC)
Surface raw seawater—Perth, Australia	0.9	3	31	25	41
Surface raw seawater—Ashkelon, Israel (May 2005)	1.2	14	39	25	22
Surface raw seawater—Ashkelon, Israel (Nov. 2005)	1.0	7	52	22	19
Surface raw seawater—Carboneras, Spain	0.9	8	38	18	42
Well seawater—Gibraltar, Spain	0.6	1	26	22	51
Surface seawater—Gibraltar, Spain	0.8	5	28	25	42

Analysis of various sources of seawater (Leparc et al., 2007; Schrotter et al., 2006) summarized in Table 2.5 indicates that TOC concentration of seawater may contain various fractions of organics depending on the origin of this water and the type of seawater intake. These fractions may also change depending on the season as well.

The analysis of Table 2.5 leads to the conclusion that low molecular weight organic compounds are typically the greatest fraction of the TOC in seawater (40%–50%). Most of these compounds, however, have limited fouling potential and therefore, the higher percentage of low molecular weight compounds in the water, the lower fouling potential this source water has.

A combination of high percentage of compounds from the "other low molecular weight" category, low TOC, and low polysaccharide content of the TOC is a clear indication for source water of low fouling potential. Based on this "rule of thumb," the well seawater at the Gibraltar SWRO facility would have the lowest fouling potential from all sources listed in Table 2.5.

Comparison of the data from the Ashkelon seawater desalination plant in Israel indicates that the most easily biodegradable organics (polysaccharides) change seasonally and increase during the summer season along with the content of algal biomass in the Mediterranean Sea. This data also show that TOC concentration of seawater may not always correlate with the content of polysaccharides in this water. In this case, humic substances would be the main contributor of fouling to the SWRO membranes of the Ashkelon plant.

TABLE 2.6 Water Quality Parameters for Characterization of Organic Foulants

Source Water Quality Parameter	Pretreatment Issues and Considerations
Total organic carbon (mg/L)	If below 0.5 mg/L—biofouling is unlikely.
	Above 2 mg/L, then biofouling is very likely.
UV_{254} (cm^{-1})	If below 0.5 cm^{-1}, the saline source water has low organic and biofouling potential.
SUVA	If >4, then source water is dominated by aquatic humic matter and biofouling is unlikely.
	If <2, source water is experiencing algal bloom and biofouling is likely.

2.5.2.2 UV_{254} Absorbance

The ultraviolet (UV) absorbance of seawater sample at 254 nanometers (nm) is an indirect measure for content of NOM. The UV_{254} absorbance of saline water sample is determined by filtering of the sample through a 0.45-μm filter and measuring of the absorbance of UV light of the filtrate with spectrophotometer. This measurement is based on the fact that specific molecular structures (chromophores) within the NOM molecules adsorb UV light.

Because the NOM composition may vary from one water source to another, UV_{254} absorbance is not always easy to use for comparison of the fouling potential of different water sources. In addition, this parameter may not be reflective of the content of microbial foulants if the NOM contained in source water is not easily biodegradable.

Specific UV absorbance (SUVA) is defined as the UV absorbance divided by the concentration of dissolved organic carbon (DOC) in the source water. Edzwald and Haarhoff (2011) indicate that the SUVA can be used as an indirect indicator of occurrence of algal blooms in the source water. If SUVA is larger than 4, then NOM consists predominantly of aquatic humic matter and the source water is not exhibiting algal bloom conditions. If SUVA is between 2 and 4, then the source water NOM is a mix of AOM and aquatic humic matter, and the source water body from which the water originates is in an early stage of algal bloom. When SUVA is less than 2, then the NOM in the source water consists predominantly of AOM and the source water body is experiencing algal bloom.

2.5.3 Threshold Levels of Natural Organic Foulants

Table 2.6 presents threshold levels of key water quality parameters used for assessment of organic fouling potential of the saline source water.

2.6 MICROBIAL FOULANTS

2.6.1 Description

Microbial foulants are aquatic microorganisms and organic compounds excreted by them such as polysaccharides, proteins, and lipids, which are referenced as extracellular polymeric substances (EPS). The phenomenon of accumulation of aquatic organisms and their metabolic products (EPS) on the membrane surface is known as *biological fouling*, *biofouling*, *or microbial fouling*.

Although bacteria constitute the majority of the membrane biofilm, other microorganisms such as fungi, algae, and protozoa can also attach to the membrane surface and contribute to biofilm formation. Usually, the most predominant bacteria causing biofouling are *Pseudomonas*, *Bacillus*, *Arthrobacter*, *Corynebacterium*, *Flavobacterium*, and *Aeromonas*. Other microorganisms such as fungi (e.g., *Penicillium*, *Trichoderma*, *Mucor*, *Fusarium*, and *Aspergillus*) are typically present in the membrane biofilm in significantly lower levels than bacteria.

The biofilm formed on the membrane surface contributes additional resistance (pressure head losses) to the osmotic pressure that must be overcome in order to maintain steady production of freshwater by the membrane elements (Konishi et al., 2011).

In order to compensate for loss of productivity due to biofouling, the feed pressure of the RO-membrane system would need to be increased, which in turn would result in elevated energy use to produce the same volume of fresh desalinated water. Feed pressure increase beyond certain level (e.g., 83–85 bars for SWRO membranes) will cause irreversible damage of the membrane structure and ultimately will result in the need to replace all RO membrane elements. Similarly, the increase of DP above the threshold of 4.5 bars as a result of fouling would cause a permanent compaction and damage of the RO-membrane structure.

Biofouling is usually a significant operational challenge for saline waters of naturally elevated organic content and temperature, such as the seawater in the Persian Gulf and the Red Sea. Biofouling is also a challenge during intense algal blooms or periods when surface runoff from rain precipitation or nearby river water of high organic content enters the plant's open intake.

2.6.2 The Membrane Biofouling Process

Bacteria that are the main contributor to the biofilm formed on RO membranes typically exist in two states in the source water—metabolically active and inactive.

In their active state, bacteria have high metabolic rate, grow fast, consume relatively large quantity of organic materials contained in the source water, and store some of the assimilated organics on the surface of their cells in the form of EPS, mainly polysaccharides, lipids, and proteins. As a result, when in active state, the individual bacterial cells are encapsulated in a layer of EPS, which they store as food reserves. Because the EPS are very adhesive, bacterial capsules easily stick to each other to form microcolonies and adhere to the surface of the RO membranes to form biofilm (see Fig. 2.3). Bacteria are typically in active state when biodegradable organics are abundant in the source water (TOC >2 mg/L) such as conditions when algal blooms occur. When the algal bloom event is over and the organic content in the saline source water drops significantly (<0.5 mg/L), bacteria use the accumulated EPS as a food source.

It should be pointed out that bacteria create EPS not only during times of abundant content of biodegradable organics in the source water to store food, but also encapsulate themselves with EPS during conditions when ambient water contains compounds that can destroy the bacterial cells, such as chlorine and other oxidants, strong acids and bases, and other biocides. In this case, bacteria maintain their EPS protection capsules only as long as the biocides are present in the ambient water. The protective EPS capsules are very resilient to strong

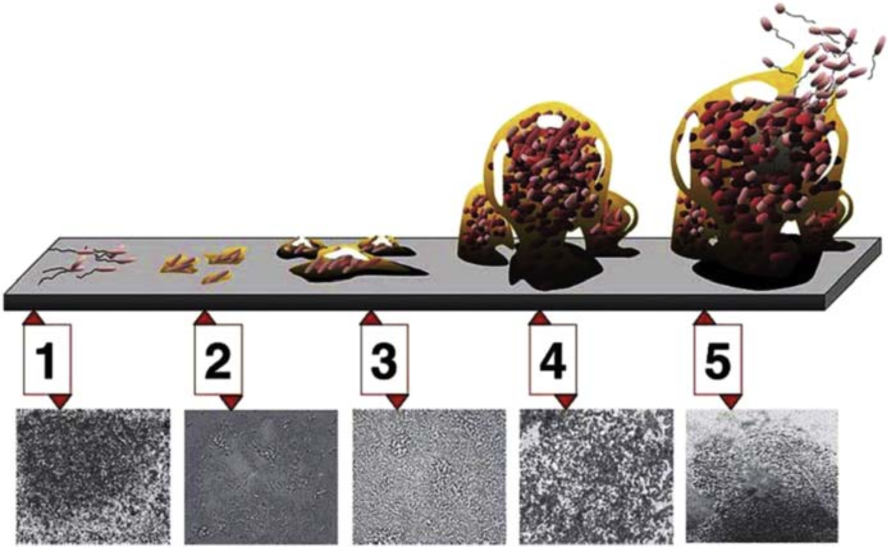

FIGURE 2.3 Key steps of the membrane biofouling process.

oxidants. For example, chlorine dosage needed to inactivate EPS-protected bacteria is 150 mg/L or more. Therefore, the typical practice of chlorine addition at the intake of desalination plants at dosages of 5–15 mg/L is not very effective in suppressing bacterial growth once bacterial cells are EPS protected.

Usually, within 48 h of the discontinuation of biocide addition, bacteria assimilate their EPS capsules using it as food. If bactericide is added again 48 h after its first application, a large portion of the bacteria without protective EPS cover will be destroyed. If bactericide is added continuously or more frequently than once every 48 h, most bacteria will continue to maintain their protective EPS capsules, will survive chlorination and will continue to cause RO membrane fouling. Therefore, intermittent chlorination with time between two bactericide applications of 48 h or more is significantly more efficient and cost-effective in suppressing bacterial growth than continuous chlorination.

It should be also pointed out that most bacteria that survive biocide addition will create protective EPS capsules within 4–6 h after the biocide feed to the source water is initiated. As a consequence, after these periods, all surviving bacteria will be protected against the damaging impact of the biocide. Therefore, continuation of biocide addition beyond 6 h at the dosages typically applied at desalination plant intakes is counterproductive and does not result in further control over the biofouling process. Quite the opposite, as indicated previously, continuous addition of biocide would cause bacteria to maintain their protective EPS capsule, reducing significantly the opportunity for sustainable long-term biofouling control.

When bacteria are in an inactive state of existence, which occurs in saline source waters with low content of biodegradable organic substances (typically TOC <0.5 mg/L), they have very low metabolic and growth rates and exist as single cells or small cell clusters that behave as microparticles. The cells of these particle-like bacteria do not have capsules of EPS and, therefore, do not tend to stick to the surface of the RO membranes. Since in inactive state, bacteria do not have ability to adhere to the membrane surface and they behave as most other small-size solid particles in the source water: they enter and travel through the space between the membrane leafs (feed/brine spacers) and exit the RO membranes with the concentrate. Because most bacteria in their inactive unicellular state are smaller than 1 μm in size and because RO membrane spacers are 700−800 μm wide, in this state, bacteria do no accumulate in large quantities within feed/brine spacers and on the membrane surface and do not cause accelerated membrane fouling.

At any given time, some of the aquatic bacteria naturally occurring in surface water bodies are in an active state and others are in an inactive state. The predominant state of aquatic bacteria (active or inactive) depends on how favorable the ambient environment is for bacterial survival and growth. Most aquatic bacteria will transform from an inactive to an active state under favorable environmental conditions, such as algal bloom events, when high concentrations of easily biodegradable organics released from the decaying algal biomass (which serve as food to these bacteria) are readily available in the source water. Since bacteria can attach to the membrane surface and grow colonies there at a very high rate, red tides or other intense algal bloom events are usually the most frequent cause of RO-membrane biofouling, especially in desalination plants with open intakes.

The membrane biofouling process (i.e., the formation of a microbial biofilm layer on the surface of an RO membrane) usually follows five key steps (Fig. 2.3):

1. Formation of a primary organic conditioning film;
2. Attachment of colonizing bacteria;
3. Formation of a biofilm matrix layer;
4. Establishment of a mature secondary biofilm; and
5. Biofilm equilibrium and die-off.

The primary organic conditioning film is a microthin layer on the surface of the membrane that is rich in nutrients and easily biodegradable organics, which creates suitable conditions for bacteria to convert from an inactive (particulate-like) state into an active state (Phase 1). In this state, bacteria are capable of producing EPS, which are adhesive substances that allow bacteria to attach to the membrane surface and to each other.

During the next step of the biofilm formation process, active bacteria adsorb to 10%−15% of the membrane surface and establish breeding ground from which they spread to the rest of the membrane surface (Phase 2). Bacteria multiply at an exponential rate, and within 5−15 days, they colonize the entire membrane surface and form a biofilm matrix layer that is several micrometers thick (Phase 3). The biopolymer matrix formed on the membrane surface entraps organic molecules, colloidal particles, suspended solids, and cells of other microorganisms (fungi, microalgae, etc.) over time to form a thicker layer with a higher resistance to permeate flow (Phase 4).

At the last phase (Phase 5) of membrane colonization, bacteria growth reaches equilibrium established by the availability and quantity of the food source, the rates of production and removal of site products from the bacterial growth process, turbulence of the cross-flow on the surface of the membranes, and space constraints. At this phase the membrane spacers and surface contain large quantity of microbial foulants that result in a significant increase (over 10% of RO feed pressure) and often over 50% of increase in DP.

In order for biological fouling to occur, aquatic bacteria need to have suitable low flow-through velocity conditions so they can attach to the RO membrane surface or to the surface of facilities upstream of the membrane system, such as cartridge filters. Besides on the desalination plant cartridge filters and RO membranes, bacteria would also attach to surfaces with low-velocity conditions such as cavities with stagnant water or high-retention time vessels along the feed water route to the RO membrane system (e.g., dead-end valves and pipes, pipe fittings, and oversized or hydraulically flawed cartridge filter housings).

The formation of a permanent biofilm layer occurs when membrane flux (flow through the membranes) exceeds a certain level (critical flux) at which cross-flow velocity through the membrane elements is low-enough to allow aquatic microorganisms to attach to the RO-membrane surface (Winters et al., 2007). The critical flux for aquatic bacteria is dependent upon the cross-flow velocity over the surface of the membranes and increases with the increase of this velocity. In practical terms, this means that the lower the permeate membrane flux (e.g., RO system recovery) for the same membrane feed flow, the higher the cross-flow velocity and the lower the rate of biofilm formation.

The critical flux is also a function of the concentration of active bacteria in the saline source water and it decreases as the concentration of bacteria rises. The concentration of active aquatic bacteria in turn mainly depends on the type of bacteria, the availability of easily biodegradable organic matter in the source saline water, and the water temperature. For a given RO system, decreasing the recovery from 50% to 35% would result in approximately two times lower fouling potential for the RO system operating in a typical flux range of 13.5–18.0 L/m^2 h (8.0–10.5 g/ft^2 day).

Although operation at low recovery may be attractive from the point of view of minimizing membrane biofouling, designing RO plants for low recovery is usually not cost-effective because of the associated increased size of the desalination plant intake, pretreatment, and RO systems, and the 30%–40% higher capital costs. Therefore, other approaches for biofouling reduction—such as control of the organic content in the source water and inactivation of aquatic bacteria by disinfection or UV irradiation—have found wider practical application.

2.6.3 Factors Impacting Biofouling Potential of Saline Source Waters

The biofouling potential of a given saline source water depends on many factors, including:

1. Concentration and speciation of microorganisms contained in the source water;
2. Content of easily biodegradable compounds in the source water;
3. Concentration of nutrients and the balance (ratio) between organic compounds and the biologically available nitrogen and phosphorus in the source water; and
4. Temperature of the saline source water.

2.6.3.1 Concentration and Speciation of Microorganisms

Most saline water sources contain microorganisms (bacteria, viruses, algae). The larger the concentration of microorganisms in the source water, the higher the potential for membrane biofouling, assuming that all other ambient conditions (e.g., food content, temperature, etc.) are conducive to biofouling. Shallow and warm surface waters may have several orders of magnitude higher content of bacteria and algae than source waters collected using deep intakes (i.e., intakes obtaining water at depth higher than 12 m, equivalent to 40 ft). Therefore, source water from shallow/onshore open intakes usually has higher biofouling potential than that collected using deep open water intakes or intake wells.

Not all aquatic microorganisms can generate EPS. Bacteria most commonly encountered in seawaters worldwide are capable of EPS generation. During algal blooms, aquatic organism speciation changes toward an incremental increase of bacteria, which can create EPS. In addition, especially in the initial first week to two-week phase of the bloom, the number of bacteria in the source water increases exponentially.

Brackish groundwater may contain bacteria, which are not found in other saline aquatic environments such as sulfur, iron, and manganese bacteria. Some sulphur-reducing bacteria can generate H_2S gas as a side product of their metabolism, which impacts source and product water quality in terms of odor.

Iron and manganese bacteria can thrive in both brackish water and saline water and can create brown or black discoloration of the water and leave residue of similar color on treatment plant structures and equipment. Some marine sulfur bacteria could be quite long in size and could form strings of biological material up to 5 cm long in source water wells and intake storage tanks/pump wet wells.

2.6.3.2 Content of Easily Biodegradable Compounds in the Source Water

Under normal, non-algal bloom conditions, saline source water and most brackish ground waters have relatively low content of biodegradable organics (TOC <0.5 mg/L). However, during algal bloom conditions, the level of biodegradable organics could increase significantly. If the TOC content of the saline source water reaches 2 mg/L or more, typically the biofouling potential of this water increases to a very high level and use of such water for desalination results in accelerated biogrowth on the surface of the RO membranes and in rapid increase of DP through the membrane vessels (as much as 1.0–2.0 bars/week vs. normal DP increase of 1–2.0 bars/3–4 months).

The source of easily biodegradable organics in the source water could not only be a natural event, such as an algal bloom, but also organic conditioning chemicals such as polymers and antiscalants (if they are overdozed) or it may be the type and operation of the pretreatment processes and systems used upstream of the RO facility.

In order to protect the aquatic environment of the area receiving the discharge from the desalination plant, all inorganic source water-conditioning chemicals (acids, bases, and coagulants) have to be environmentally safe and all organic conditioning chemicals (e.g., polymers and antiscalants) have to be biodegradable. If for example, an organic polymer or antiscalant are overdosed, the excess unreacted chemical will cause or accelerate biofouling because by design these chemicals are created to be easily biodegradable. Bioavailability of source water-conditioning chemicals also increases with their chlorination. Another potential source of biofouling are the biodegradable organic impurities contained in low-quality/low cost source water-conditioning chemicals, such as antiscalants, polymers, or acids (Vrouwenvelder and van der Krooij, 2008).

One reason for accelerated biofouling could be the continuous chlorination of the saline source water, which often is applied to inactivate aquatic microorganisms and reduce biofouling. Since chlorine is a strong oxidant, it can destroy the cells of active aquatic bacteria and algae, which naturally occur in the source water at any given time. The destroyed algal and bacterial cells release easily biodegradable organic compounds (such as polysaccharides) in the ambient water, which become food for the remaining aquatic bacteria that have survived chlorination by being in an inactive state. If the concentration of these organics reaches a certain threshold, it could trigger the conversion of these surviving bacteria from an inactive to an active state, followed by their attachment and excessive growth on the RO membrane surface, which in turn would manifest as membrane biofouling. Therefore, continuous chlorination or addition of other biocides often creates more membrane biofouling problems than it solves and usually does not provide sustainable long-term biofouling control solution. On the other hand, as indicated previously, intermittent chlorination has been found to provide effective control of microbial growth without generating a steady influx of easily biodegradable organics that can trigger a large-scale transfer of aquatic bacteria from their inactive to active state of existence.

Some pretreatment technologies such as granular media pressure filters and vacuum- and pressure-driven MF and UF membrane filters could potentially increase the biofouling potential of the saline source water by breaking the algal cells contained in the water as a result of the vacuum or pressure applied for MF or UF separation, and releasing their easily biodegradable cytoplasm into the feed water to the RO system. Although pressure filters provide effective removal of particulate and colloidal foulants, the filtration driving pressure applied by these systems could break some of the algal cells in the source water and cause the release of easily biodegradable organics, which in turn could result in accelerated RO-membrane biofouling. Most marine algae would break when the pressure applied to the UF/MF system feed water reaches 0.4–0.6 bars. Both pressure- and vacuum-driven filtration systems usually operate well above these thresholds. Similarly, pressure-driven granular media filters would have the same negative impact on algal breakage and ultimately would increase the biofouling potential of the saline source water.

From the point of view of minimizing biofouling associated with algal cell breakage, the most suitable pretreatment technologies are those that provide a gentle removal of the algal cells in the source water, such as down-flow gravity granular media filtration and clarification by dissolved air flotation and gravity sedimentation.

2.6.3.3 Concentration and Balance of Nutrients in Source Water

Results from research on desalination membrane fouling (Jiang and Voutchkov, 2013) indicate that the transformation of marine bacteria capable of producing EPS from passive to active state is triggered by imbalance of basic nutrients in the source water (i.e., organics, nitrogen, and phosphorus). Under normal non-algal bloom conditions, the ratios between the content of organics measured as TOC, nitrogen measured as total nitrogen (TN), and phosphorus, measured as total phosphorus (TP), are approximately equal (e.g., TOC:TN:TP = 1:1:1). Changes in ambient conditions which result in imbalance of the TOC:TN:TP ratio by introducing uneven amounts of nutrients to the saline source water (e.g., algal blooms, surface runoff during rain events) trigger the transformation of bacteria from passive to active state and turns on their genes responsible for production of EPS.

For example, based on testing completed at the Carlsbad RO desalination plant in California (USA) during non-algal bloom conditions, the saline water contains approximately 0.5 mg/L of TOC, TN, and TP and therefore, the ratio of these constituents is 1:1:1. During algal bloom events in 2010 and 2011, the TOC:TN:TP ratio has changed dramatically and

TOC has increased significantly. The TN and TP levels, while slightly elevated, increased at a lower magnitude than TOC (i.e., TOC = 4.6 mg/L; TN = 0.8 mg/L and TP = 0.7 mg/L, with TOC:TN:TP ratio of 6.6:1.1:1.0). Under this nutrient imbalanced environment, marine bacteria contained in the source water have switched from passive (non-EPS forming) state to an active state, where they have created large amount of EPS, which in turn has resulted in rapid biofouling of the membrane elements of the RO test facility in Carlsbad fed with seawater during the algal bloom period.

In the example case of nutrient balance presented above, the DP of the RO system has increased with approximately 1 bar (14.5 psi) over two-week period only. Under normal non-algal bloom conditions, such DP increase is observed over a period of 4–5 months. The accelerated fouling and associated drastic increase in DP during this test period required membrane cleaning after only 1 month of operation because the RO system productivity have decreased with over 15% from its steady-state average production level.

It is interesting to point out that not every condition of nutrient imbalance would trigger significant fouling effect on the RO membranes. While bacteria may switch from passive to active mode when the TOC:TN:TP ratio in source water is imbalanced, the amount of biomass created by their growth is limited by the TOC content in the source water. For example, during a short-term (three-week) period of mild algal bloom at the same test location in Carlsbad, the TOC of the source water has increased to 1.2 mg/L only. While the TOC:TN:TP was imbalanced, the elevated biofouling resulted in DP increase of only 0.28 bars (4.0 lb/in^2) over the 3-week algal bloom period and then the DP has stabilized back to normal (i.e., average DP of 0.3 bars or 4.3 lb/in.^{2} month). This mild algal bloom event did not trigger the need for more frequent RO-membrane system cleaning.

2.6.3.4 Source Water Temperature

Elevated source water temperature increases biological activity and, therefore, it accelerates biofouling. The effect of temperature on biofouling increases significantly when temperature exceeds a particular threshold, which for most marine bacteria is between 25°C (77°F) and 28°C (82.4°F). It is interesting to note that the growth rate of many anaerobic bacteria increases dramatically when the source water has temperature between 33 and 37°C (91 and 99°F). Since many side products of the metabolism of anaerobic bacteria are fine solids that are very difficult to remove by conventional filtration, and which tend to create accelerated fouling of UF and MF membrane pretreatment systems, it is advisable to control the source water temperature, if possible, in order to avoid this temperature to enter into the range for optimum growth of anaerobic bacteria.

Natural ambient saline water sources usually have temperatures outside of the range for optimal growth of anaerobic bacteria. Such temperature, however, may be reached when ambient source water is blended with cooling water from thermal desalination plants, which is practiced at a number of desalination facilities in the Middle East that have both RO and thermal desalination plants sharing common intake and outfall.

2.6.4 Biofouling Measurement Parameters and Methods

At present, there are no simple and inexpensive online methods to measure the biofouling potential of saline source waters. Therefore, biofouling potential and rate in practice are monitored by surrogate parameters such as total organic carbon, TOC:TN:TP ratio, RO train DP rate increase, chlorophyll a, and algal count and profile.

2.6.4.1 Total Organic Carbon

Most commonly, TOC concentration is used as an indicator of the biofouling potential of saline water, while the rate of DP increase is applied as an indicator of biofouling rate. Operational practice indicates that as the TOC concentration of the source saline water exceeds a certain threshold, typically 2 mg/L (Huang et al., 2013), the biofouling potential of this water increases dramatically. The same group of researchers have shown a strong statistically significant correlation between TOC and membrane fouling. In addition, their research study points out to chlorophyll *a* as a statistically significant indictor of source water biofouling potential during algal blooms.

2.6.4.2 TOC:TN:TP Ratio

Typically, TOC:TN:TP ratio is measured weekly to check whether the saline water potential is indicating appreciable increase of the biofoluing potential of the saline source water. If ratios are higher than 20% of 1:1:1, then typically the biofouling potential of the source water is elevated to the point that activates the generation of EPS by bacteria, which in turn increases significantly their ability to foul the RO membrane elements.

2.6.4.3 RO Train Differential Pressure Rate Increase

The most commonly used surrogate measure for the rate of biofouling of RO membrane elements is the RO system DP (i.e., the difference between RO system's feed and concentrate pressures, DP). Usually, in full-scale desalination plants the DP of all RO system trains is tracked continuously.

Biofouling of the feed channels and spacers of spiral-wound membranes (Fig. 2.4) typically results in significant increase in the membrane DP.

As biofilm forming bacteria colonize the RO membrane surface, often sections of the feed channels and spacers between the membrane leafs are blocked over time by patches of

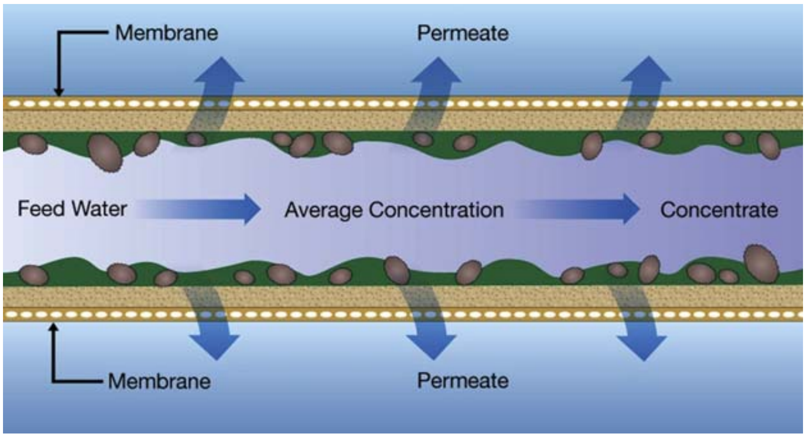

FIGURE 2.4 Biofilm accumulation in RO-membrane spacers.

biofilm-forming microorganisms such that the flow pattern within the membrane elements changes and the feed flow is completely blocked in some portions of the feed channels and increased in others. Flow channeling caused by random blockages of the feed channels and spacers results in sharp increase in salt concentration in the affected areas, which in turn triggers precipitation of sparingly soluble salts such as calcium carbonate and sulfate in the feed channels and further exacerbates and accelerates the membrane fouling problem. All of these impacts of membrane-fouling result in an increased head loss of source water traveling through the feed/brine spacers, which exhibits itself as an increase in difference between the feed pressure entering the membrane vessels and the pressure of the blend of feed water and concentrate exiting the pressure vessels (e.g., the DP).

Typically, DP of RO systems desalinating surface water collected by open intake increases with 0.1−0.4 bars (1.0−5.8 lbs/in.^{2})/month. However, if the RO system is exposed to heavy biofouling (e.g., during intensive algal bloom events), DP increase of the same magnitude could be observed within 1−2 weeks. The DP of RO systems using source waters collected by subsurface intake (e.g., saline water beach wells or brackish water production wells) could be significantly lower—0.04−0.20 bars (0.58−2.90 lbs/in^2). However, source waters collected by most of these systems have very low TOC concentration (typically <0.5 mg/L).

It should be pointed out that DP is not a parameter that measures biofouling exclusively. This parameter is reflective for all five types of fouling (particulate, colloidal, mineral, NOM, and microbial), which usually occur simultaneously, but at different rates. However, except for particulates and hydrocarbons, all other types of foulants have a lower rate of impact on the DP, and usually, if the pretreatment system performance and source water quality have not changed significantly in terms of turbidity and hydrocarbon content, the most likely type of fouling that can cause measureable DP increase within a short period of time (i.e., 1 week) is biofouling. Therefore, DP increase rate, especially during algal bloom events, is most often representative of the biofouling rate of the membranes.

2.6.4.4 *Chlorophyll* a

Chlorophyll *a* concentration of source water is an indicator of the content of algae with green pigmentation in the water. This parameter is measured using fluorometer or spectrophotometer. The content of chlorophyll *a* is proportional to the light transmission through the water sample at a given wavelength, which is detected by the instrument and converted into concentration units, typically either micrograms per liter (μg/L) or milligrams per liter (mg/L).

As a rule of thumb, source waters with chlorophyll *a* content below 0.5 μg/L have low biofouling potential, and algae levels are indicative of non-algal-bloom conditions. During severe algal blooms, the chlorophyll *a* level could exceed 10 μg/L.

It should be pointed out that content of algae and chlorophyll *a* naturally varies diurnally and seasonally, and also changes with depth. In general, algal content is proportional to the intensity of solar irradiation and typically increases significantly in the summer, as compared to the average annual algal content.

Tracking chlorophyll *a* on a daily basis and trending the collected data allow an operator to determine the occurrence of algal bloom events, because during algal blooms, the chlorophyll *a* content in the source water usually increases several times in a matter of only several days.

2.6.4.5 Algal Count and Profile

Algal count is a measure of the number of algal particles per unit volume of source water. The total algal count is the number of all algal cells contained in 1 mL or 1 L of source water. Typically, algal concentration (total algal count) in saline source waters below 10,000 cells/L does not present challenge for desalination plant performance (e.g., pretreatment system and RO-membrane fouling). Total algal count can be measured by online instrumentation or in laboratory setting. Higher content of algae in the source water may have three key performance impacts:

1. Reduction of pretreatment system filtration capacity and solids removal efficiency due to the filter overload;
2. Accelerated fouling of cartridge filters as a result of the increased content of fine solids (i.e., pico- and microalgae) in the source water and elevated biogrowth on the filter surface;
3. Loss of productivity and increase of DP of the RO membrane elements due to biofouling.

In addition to the total algal content, algal profile indicates the number, size, and type of algal species contained in the source water and is completed in laboratory setting by experienced biologist familiar with the aquatic species of the type of the analyzed source water. The algal count of individual species in the water is a very useful tool to verify the presence of algal bloom and identify the size and type of algae triggering the bloom.

Knowing algal size distribution is of critical importance for pretreatment system design and operation because it allows determining what pretreatment method or combination of methods will need to be applied and how the pretreatment facilities would have to be sized and configured. For example, if dissolved air flotation is used for pretreatment, than the DAF clarifiers have to be designed such that the size of the bubbles they create closely match the size of the majority of algae contained in the source water. If DAF design is not reflective of the actual size of algae in the water, this type of pretreatment facilities would have very limited algal removal efficiency and will be of limited or no benefit to plant operations.

2.6.5 Threshold Levels of Microbial Foulants

Table 2.7 presents a list of source water quality parameters used for characterizing particulate content that are recommended to be measured when deciding upon the type of pretreatment needed for a given source water.

Typically, algal bloom is defined as an event in which over 75% of the algae are from the same species. Knowing the type and size of the dominant algal species is critical to optimizing the source water chemical conditioning and pretreatment approach. Table 2.8 provides threshold levels of algal content, which characterize algal bloom severity and associated biofouling potential of the saline source water.

It is important to note that the levels of these parameters are thresholds and they do not necessarily correlate with each other for a given source water. For example, usually, there is no statistically significant correlation of the algal content measured as chlorophyll *a* and as number of algal cells per liter of source water. However, if the source water contains chlorophyll *a* at concentration higher than 0.5 μg/L, and/or if the total algal count of the source water is higher than 10,000 cells/L, then it is very likely the intake area of the desalination plant is experiencing algal

TABLE 2.7 Water Quality Parameters for Characterization of Microbial Foulants

Source Water Quality Parameter	Pretreatment Issues and Considerations
Total organic carbon (TOC), mg/L	If TOC < 0.5 mg/L the content of microbial foulants is low and the source seawater has minimal biofouling potential.
	In a range of 0.5–2.0 mg/L the water has moderate fouling potential.
	If TOC > 2.0 mg/L, the biofouling potential is high.
TOC:TN:TP ratio	If the ratio is within 20% of 1:1:1 the water has low biofouling potential.
	Any of the ratios exceeding 1.20:1 is an indication of source water with nutrient imbalance and elevated biofouling potential.
RO train differential pressure (DP) rate increase, bars/month	If DP increases with <0.1 bar/month, the source water has very low biofouling potential.
	If DP increase is 0.1–0.4 bars/month, the water has moderate fouling potential.
	If the DP rate increase is more than 1.0 bar/month, the source water has high fouling potential.
Chlorophyll *a*	This parameter is indicative of algal bloom occurrence.
	If water contains >0.5 μg/L source water may be in algal bloom condition.
	When heavy algal blooms occur, Chlorophyll *a* usually exceeds 4 μg/L.

TABLE 2.8 Algal Count Thresholds Typically Used to Characterize Algal Bloom Severity and Biofouling Potential

Total Algal Count, Cells/L	Algal Bloom Intensity	Biofouling Potential of Saline Source Water
<10,000	Non-algal bloom condition	Minimal
10,000–20,000	Low-intensity algal bloom	Low
20,000–40,000	Moderate algal bloom	Medium
40,000–60,000	High-intensity algal bloom	High
>60,000	Severe algal bloom	Very high

bloom. As indicated in the previous section, Huang et al. (2013) have found statistically significant correlation between chlorophyll *a* and the fouling indicator of RO membranes.

It should be pointed out that elevated content of TOC does not necessarily mean that the source water intake is experiencing algal bloom. TOC is measure of all types of organic compounds contained in the source water—originating from both living aquatic microorganisms and from NOM. For example, intake wells collecting source water from an alluvial aquifer yield saline source water with a high content of TOC, almost exclusively contributed by

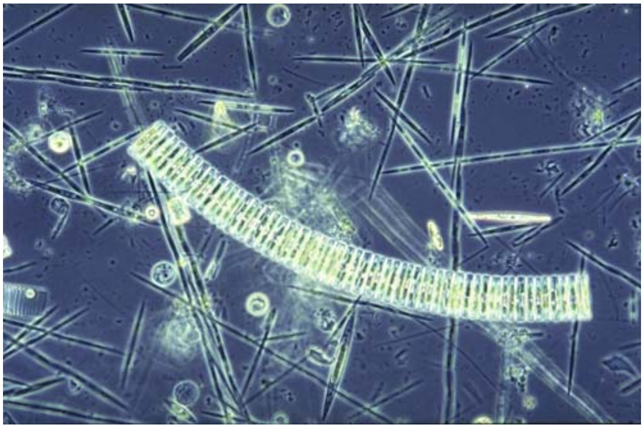

FIGURE 2.5 Algal bloom event dominated by *Pseudo-nitzchia*.

NOM and do not contain measurable amount of algae. Therefore, the TOC level has to be considered as supplemental rather than as a primary criterion for occurrence of algal bloom.

Not every algal bloom has significant impact on desalination plant operations. Most pretreatment processes, including granular media and membrane filters, can typically handle low intensity algal blooms with minimum to no impact on plant operation and RO-membrane fouling rate.

Single-stage gravity granular media filters and conservatively designed single-stage vacuum- and pressure-driven UF and MF filters can also provide adequate pretreatment during moderate algal blooms. If the intake area, however, experiences frequent high-intensity or severe algal blooms, two-stage pretreatment system would need to be employed to maintain plant productivity and reliable performance.

The occurrence of algal bloom can be verified by observation of the algal community in the source water by examining water sample under microscope. Under normal non-algal bloom conditions, the algae community is very diverse and contains variety of species in comparable numbers or algal cells per species. Under algal bloom condition, over 75% of the algal cells in the water sample are comprised by one dominating species of algae. For example, Fig. 2.5 illustrates algal bloom condition dominated by the algal species *Pseudo-nitzschia*. Excessive growth of this species is a frequent cause of red tides along the Pacific Ocean coast of California and Australia, as well as at a number of other locations worldwide. While examination of Fig. 2.5 shows the presence of a number of algal species in the source water sample, the majority of the algal cells visible on this figure have elongated needle-shape, which is characteristic for the *Pseudo-nitzshia* species.

2.7 COMBINED EFFECT OF VARIOUS FOULANTS ON MEMBRANE PERFORMANCE

Membrane fouling is a complex process, which often results from the additive impacts of several types of foulants. Particulate fouling usually causes a relatively quick and definitive deterioration of membrane permeate flux and RO system productivity, without significant impact on salt rejection. For comparison, colloidal fouling typically results in a marked deterioration in the RO system's salt rejection over time. In addition, colloidal fouling also causes rapid permeate flux decline over time. Such flux/RO system productivity decline is caused not only by the accumulation of a flow-resistant cake layer of colloidal particles on the surface of the membranes, but also by the back-fusion of salt ions within the colloidal cake, which in turn results in elevated salt concentration of the permeate. In contrast, fouling caused by NOM is accompanied by almost constant salt rejection over time.

Because of its complex nature, NOM often creates diverse interactions with other foulants and with the surface of the membrane elements. As indicated previously, hydrophobic humic substances are typically one of the most common foulants in source seawater, especially when the seawater is under the influence of river discharge. The high content of calcium in seawater reduces the solubility of humic acids and increases their aggregation, which in turn accelerates the accumulation of these NOM compounds on the surface of the RO membranes.

The calcium complexation of NOM in this case often forms a gel on the surface of the membranes, which is very difficult to remove. Therefore, source seawater originating from an area of river confluence into the ocean—especially if the river water has high content of NOM—would be very difficult to treat and use for membrane desalination.

As indicated previously, biofouling of the feed channels and spacers of spiral-wound membranes typically results in a significant increase in the membrane DP over a very short period of time (one to several weeks). As biofilm-forming bacteria colonize the RO membrane surface, they often block sections of the feed channels and spacers between the membrane leaves over time, so that the flow pattern within the membrane elements changes and the feed flow is completely blocked in some portions of the feed channels and increased in others (Fig. 2.6).

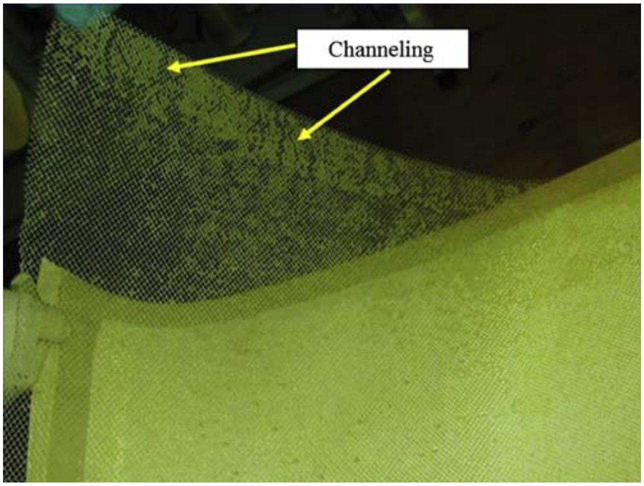

FIGURE 2.6 Flow channeling.

Flow channeling caused by random blockages of the feed channels and spacers results in a sharp increase in salt concentration in the affected areas, which in turn triggers precipitation of sparingly soluble salts such as calcium carbonate and sulfate in the feed channels and further exacerbates and accelerates the membrane fouling problem.

References

APHA, American Public Health Association, American Water Works Association, AWWA, Water Environment Federation, WEF, 2012. Standard Methods for Examination of Water and Wastewater. Washington, DC, USA, twenty-second ed., pp. 1368–1376

AWWA, American Water Works Association, 2007. Manual of Water Supply Practices, M46. Reverse Osmosis and Nanofiltration. Denver, CO, USA, Appendix C, second ed., pp. 205–206

Boerlage, S.F.E., 2007. Understanding SDI and modified fouling indices ($MFI_{0.45}$ and MFL_{UF}). In: Proceedings of IDA World Congress, Masspalomas, Gran Canaria, Spain, October 21–26. Ref: IDAWC/MP07-143.

Edzwald, J.K., Haarhoff, J., 2011. Seawater pretreatment for reverse osmosis: chemistry, contaminants and coagulation. Water Research 45 (17), 5428–5440. Elsevier.

Hydranautics, 2008. Chemical Pretreatment for RO and NF. Technical Application Bulletin No. 111, Revision C.

Huang, S., Voutchkov, N., Jiang, S., 2013. Investigation of environmental influences on membrane biofouling in Southern California desalination pilot plant. Desalination 319, 1–9. Elsevier.

Jiang, S., Voutchkov, N., 2013. Investigation of Desalination Membrane Fouling. WateReuse Research Foundation, Alexandria, Virginia, pp. 08–19. REF: WRF-.

Khirani, S., Aim, B.R., Manero, M.H., 2006. Improving the measurement of the modified fouling index using nanofiltration membranes (NF-MFI). Desalination 191, 1–7. Elsevier.

Konishi, T., Maruyama, K., Namba, M., Kobuke, M., Nozoe, M., Liu, E., Akiyama, T., Hirose, M., 2011. LD (Low-Differential-Pressure) technology for biofouling control – design concept supported by microbiological analysis. In: Proceedings of IDA World Congress, Perth, Western Australia, September 4–9, pp. 1–9. Ref: IDAWC/PER-081.

Leparc, J., Rapenne, S., Courties, C., Lebaron, P., Croute, J.P., Jacquemet, V., Tyrner, G., 2007. Water quality and performance evaluation at seawater reverse osmosis plants through the use of advanced analytical tools. Desalination 203, 243–255. Elsevier.

Mosset, A., Bonnelyle, V., Petry, M., Sanz, M.A., 2008. The sensitivity of SDI analysis: from RO feed water to raw water. Desalination 222, 17–23. Elsevier.

Schrotter, J.C., Gaid, K., Lafon, D., 2006. Reverse osmosis: improved operation through experience with SWRO plants. In: Proceedings of 2006 Desalination and Water Reuse International Forum & Exhibition, Tianjin, China, September 6–8, pp. 1–12.

USBR, 2002. Membrane Fouling: Influence of Natural Organic Matter Properties and Membrane Surface Treatment on Nanofiltration Treatment. Desalination and Research Development Report No. 83. US Bureau of Reclamation, Washington, DC, USA.

Vrouwenvelder, J.S., van der Kooij, D., 2008. Diagnosis, prediction, and prevention of biofouling on NF and RO membranes. Desalination 139, 65–71. Elsevier.

Winters, H.S., Fane, A., Nicolaisen, B., 2007. Microbial membrane fouling; microbial cake enhanced osmotic pressure (CEOP) or thermodynamic restriction?. In: Proceedings of IDA World Congress, Masspalomas, Gran Canaria, Spain, October 21–26, 2007 Ref: IDAWC/MP07-267.

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CHAPTER

3

Diagnostics of Membrane Fouling and Scaling

OUTLINE

3.1 PURPOSE OF MEMBRANE FOULING AND SCALING DIAGNOSTICS

Membrane fouling and scaling diagnostics aim to identify reasons for abnormal performance of RO membranes. The main performance goals of RO desalination systems are to

Pretreatment for Reverse Osmosis Desalination
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reliably produce permeate of target design flow rate and quality while minimizing operational expenditures for energy and for cleaning and replacement of membrane elements. Membrane fouling has direct impact on all plant-performance goals because it decreases RO system's freshwater production flow over time; increases energy needed to produce target freshwater flow; and accelerates the need for membrane cleaning and replacement. As indicated in Chapter 2, membrane fouling is a complex phenomenon often caused by a combination of foulants—particulate, colloidal, mineral scale, natural organics, and microbial biofilm.

The main purpose of RO-system membrane-fouling diagnostics is to identify the key types of foulants that exhibit the highest impact on membrane performance as well as to quantify the impact of these foulants on the condition of the RO-membrane elements. Often the loss of physical integrity of the RO membranes, including such extreme cases as membrane telescoping (Fig. 3.1) and cracked membrane fiberglass casing (Fig. 3.2) could be caused by numerous reasons: faulty RO system design; membrane production defects; mechanical damages due to pressure surges; operator errors during membrane installation or plant operation; or excessive membrane fouling.

Membrane-fouling diagnostics is a main component of the overall desalination plant performance diagnostics, which allows to discern the role and contribution of fouling to the plant operations' challenges from the impact of other factors described above and thereby to define the best course of action to address these challenges. Detailed guidance on how to troubleshoot RO-system performance based on the membrane diagnostics is presented elsewhere (Voutchkov, 2014).

3.2 TYPICAL MEMBRANE-FOULING PHENOMENA

Typical membrane-fouling phenomena include:

- Accelerated membrane train/vessel differential pressure increase due to blockage of the feed/brine spacer channels by colloids and particulates, by biogrowth, or by mineral scale deposits.

FIGURE 3.1 Telescoped membrane envelopes.

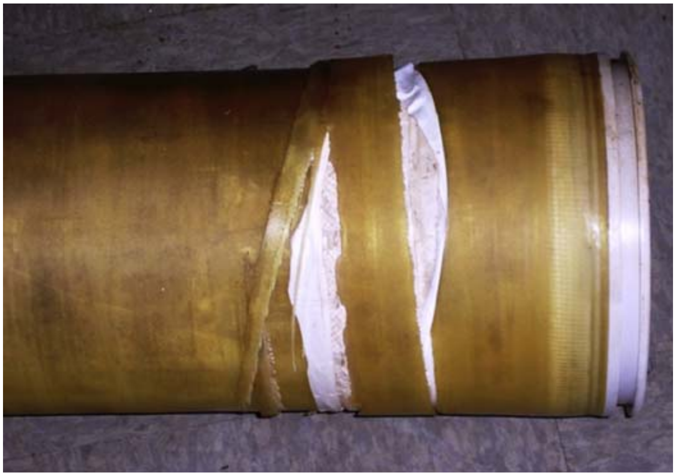


FIGURE 3.2 Cracked membrane fiberglass casting.

- Increase in feed pressure to maintain the same RO-system freshwater production.
- Permeate production flow (flux) decline due to:
  - Adsorption of organics on the membrane surface such as unconsumed polymer(-flocculant) added to the source water as well as accumulation of NOM on the membrane surface;
  - Accumulation of particulate solids and unused coagulant in the membrane spacers (Fig. 3.3);
  - Accumulation of biofilm in the membrane spacers (Fig. 3.4); and
  - Formation and accumulation of mineral scale on the concentrate side of the membrane elements (Fig. 3.5).

Table 3.1 summarizes the various types of membrane fouling/scaling and their impact of key desalination system performance parameters—for example, differential pressure (DP); RO system feed pressure; and RO membrane salt passage.

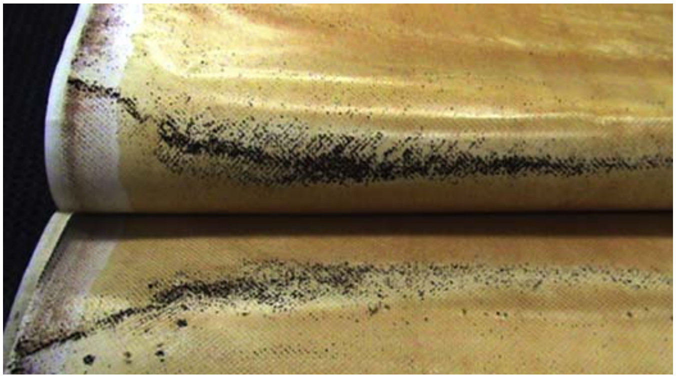


FIGURE 3.3 Accumulation of solids and organics on the membrane surface.

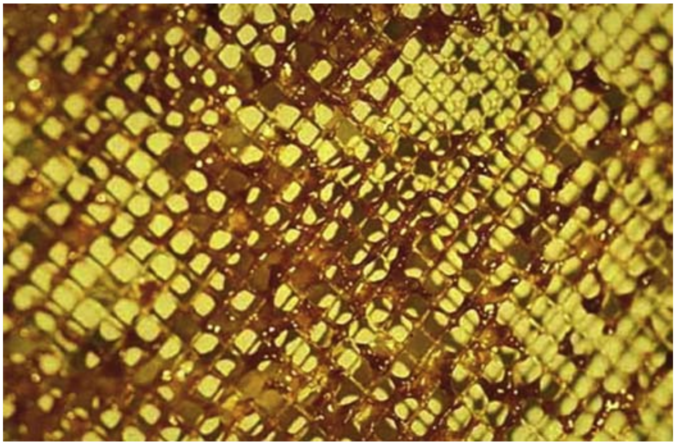


FIGURE 3.4 Accumulation of biofilm on membrane brine spacer.

FIGURE 3.5 Accumulation of scale on the membrane brine spacer.

3.3 TYPICAL DIAGNOSTICS PROCEDURES

As indicated previously, many of the same RO-system performance problems (e.g., reduced production flow, deteriorated permeate water quality, and unstable system performance) could be caused by both the loss of membrane integrity and/or biofouling. In addition, biofouling itself could be a cause for membrane integrity loss as well. Therefore, most often membrane fouling and membrane integrity are investigated at the same time.

RO-membrane integrity and fouling are diagnosed by first, completing sample collection from malfunctioning membrane vessels, and by inspecting the front and back (tail) elements of these vessels. Visual inspection is completed in parallel with performance data analysis of

TABLE 3.1 Types of Fouling/Scaling and Their Impact on RO-System Performance

Type of Fouling	Impact on Differential Pressure	Impact on Feed Pressure Needed to Maintain Production	Impact on Salt Passage
Particulate solids fouling	Rapid increase	Rapid increase	Rapid increase
Colloidal fouling (e.g., pretreatment polymer overdose)	Gradual increase	Minimal increase	Slight increase
Hydrocarbon/oil & grease fouling	Rapid increase	Rapid increase	Slight increase
Metal (Fe, Mn, Cu, Ni, Zn) oxide fouling	Rapid increase	Rapid increase	Rapid increase
Silica fouling	Slight increase	Increase	Slight increase
NOM fouling	Gradual increase	Increase	Decrease
Mineral scaling	Moderate increase	Slight increase	No significant increase
Antiscalant fouling	Slight increase	Increase	Slight increase
Microbial fouling (biofouling)	High-rate increase	High-rate increase	Slight increase

Rapid increase—within several hours to a day; high-rate increase—within several days to a week; gradual increase—within 1 month; slight increase—within 3–4 months.

key desalination plant components. Once malfunctioning vessels are identified by sample collection and visual inspection, a vessel probing and profiling is completed to identify the damaged membrane elements and/or the other components inside the vessels (O-rings, brine seals, etc.).

3.3.1 Visual Inspection, Performance Data Review, and Vessel Permeate Sampling

Visual inspection incorporates the following observations:

1. Color, foam, oxidation–reduction potential (ORP), salinity, and turbidity of the saline source water.
2. Water appearance after each treatment process (e.g., intake screening, pretreatment sedimentation/filtration, cartridge filtration, and RO permeate).
3. Settling and accumulation of solids in pretreatment facilities.
4. Color and odor of SDI pad and SDI values before and after the pretreatment filters and cartridge filters.
5. Conditions of interim storage tanks—septicity/odors, accumulation of solids.
6. Cartridge filter color, odor, and appearance along with track record of replacement.
7. Conductivity of the individual permeate vessels of the malfunctioning RO train.
8. Appearance of front and tail elements for the pressure vessels with poor conductivity.

RO-performance data review includes analysis of the plant normalized performance data (e.g., flow, salt passage) for the last month and for periods in the past of similar problems. The purpose of this analysis is to determine whether:

1. The problem is instantaneous (i.e., during startup/shutdown), it is instrumentation related, or it is mechanical/equipment related.
2. The problem is gradual—membrane fouling or degradation has occurred slowly over time.
3. The salt passage has increased—e.g., which RO trains/vessels have excessive salt passage.
4. The plant production has decreased and/or the feed pressure has increased since the last membrane cleaning.
5. The membrane maintenance history points out to significant changes in:
   a. Cleaning frequency and chemicals,
   b. Membrane replacement frequency.
6. Past problems with equipment and source water quality have occurred.
7. Instrumentation and calibration track record reveals frequent malfunction of monitoring equipment.

Data review and analysis above allows to determine whether the performance challenges of the RO system are:

- Membrane/vessel integrity loss-related problems—deteriorating the product quality.
- Membrane productivity loss-related problems—deteriorating normalized plant production flow.

Membrane integrity diagnostics includes two steps:

1. Completion of conductivity profile for all pressure vessels within a malfunctioning RO train to identify which vessels have highest conductivity/largest deviation from the average train conductivity.
2. Probing of the problematic vessels to identify which specific membrane elements or other internal components (e.g., O-rings, interconnectors, brine seals) inside the vessel are malfunctioning.

3.3.2 Vessel Profiling and Probing

Vessel profiling is completed by collecting permeate from all vessels though their sampling ports (Fig. 3.6) and analyzing the conductivity of the collected samples to identify vessels with conductivity which significantly differ from the rest.

For the vessels that are identified to have significantly higher than average RO train salinity, vessel probing test is completed. This test consists of insertion of plastic tubing into the vessel permeate tube to determine the conductivity level along the length of the vessel as shown in Fig. 3.7. Permeate plug is removed from the end cap and tubing is inserted through the length of the vessel. Probe tubing is pushed along the length of the vessel moving from the last to the first (front) element within the vessel. Samples are collected near the brine seal and permeate tube point of connection between the membranes. A more detailed procedure for vessel probing is presented elsewhere (AWWA, 2007).

FIGURE 3.6 Membrane pressure vessel sampling ports.

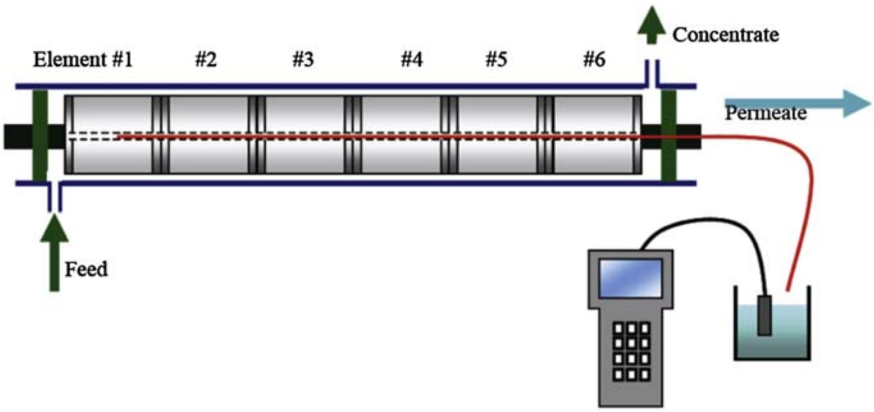


FIGURE 3.7 Probing of RO vessels.

The data from the permeate vessel probing should be interpreted taking into consideration the following rules of thumb derived from operational experience:

1. Conductivity of the first membrane element in the vessel is typically around two times lower than that of the last membrane element.
2. As probe is moved downstream, the conductivity is expected to increase uniformly.
3. Abrupt increase in conductivity indicates a leak or poorly performing element.
4. Feed water is separated from permeate via interconnector with O-rings installed on permeate tubes and brine seal (see Fig. 3.8); if they are broken, feed water will enter the permeate interconnector and will increase permeate salinity.
5. Salinity increases, if a given membrane element is defective, leaks, or it is chemically damaged.
6. High pressure drop causes glue-line leaks or telescoping.

Example result from vessel probing is presented in Fig. 3.9. As seen from this figure, the salinity of permeate has increased dramatically at the point of O-ring connector between the second and third elements. In this case, the most probable cause is that at this point the O-ring/O-rings are broken.

Membrane integrity testing could be used to uncover the following three types of problems:

1. Loss of rejection of individual RO elements.
2. Loss of integrity of membrane interconnectors:
   a. O-rings,
   b. Brine seals.
3. Membrane damage:
   a. Membrane breakage/collapse,
   b. Membrane telescoping.

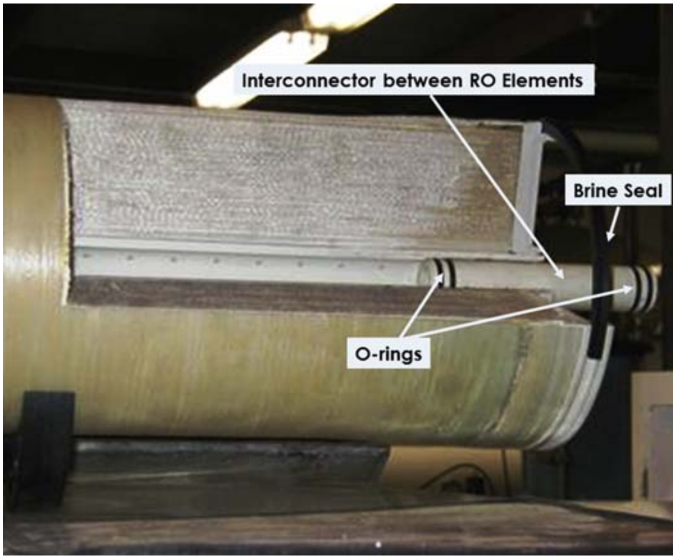


FIGURE 3.8 RO-membrane element with interconnector and O-rings.

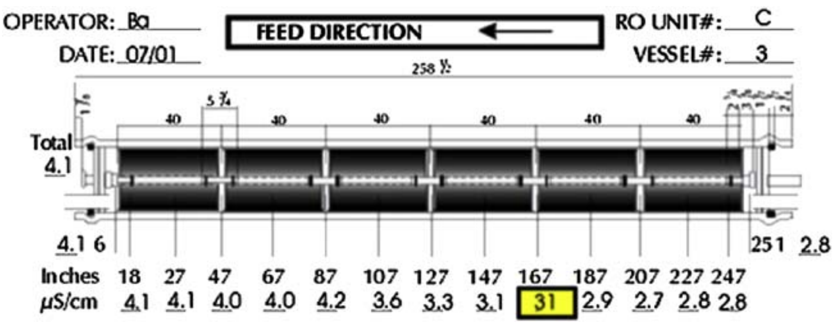


FIGURE 3.9 Example of vessel probing results.

3.4 MEMBRANE AUTOPSY

Membrane autopsy is one of the most common methods for identifying the nature and origin of RO-membrane fouling. This method incorporates sequential completion of series of standard tests in laboratory conditions on membrane elements harvested from RO trains with compromised performance (Chesters et al., 2011). The membrane autopsy procedure includes the following activities:

1. External visual inspection of the membranes,
2. Membrane weight measurement,
3. Bubble & vacuum tests of selected membranes,
4. Standard membrane performance test,
5. Membrane cleaning test,
6. Dye test,
7. Internal inspection and testing,
8. Cell test,
9. SEM,
10. EDaX,
11. Fourier Transform Infrared (FTIR) Spectroscopy,
12. Weight loss on ignition test, and
13. Fujiwara test.

The first five tests listed above are completed with all elements in the actual condition at which they were harvested and are referred to as nondestructive tests. After completion of tests 1 through 5, the membrane elements are dissected and samples of the membrane envelopes are collected to perform tests 6 through 13. Because the performance of these tests requires the membrane to be destroyed, they are also referred to as destructive tests. The process of cutting of the test elements to extract samples for the destructive tests is referred to as "membrane autopsy" (Fig. 3.10) and has widely be used in practice as a term for the entire membrane test.

A brief description of each of the 13 membrane autopsy activities along with their purpose and potential outcome are described herein. Usually, the tests are completed as a minimum

FIGURE 3.10 Membrane dissection.

on one front-membrane element, located in the first position near the entrance of the feed water to the RO vessel, and on one tail-RO element located at the last position in the RO vessel. The front RO-membrane element experiences the highest magnitude of particulate, organic, colloidal and microbial fouling while the last RO element is potentially exposed to the worst-case scaling conditions that may occur in the vessels.

Preferably, it is recommended to perform membrane autopsy on a complete set of all elements in at least one vessel within the RO train. The selected membrane set should be representative for the performance challenges of the entire RO train and of the desalination plant RO system, if possible. If the RO system is experiencing a number of different challenges, it is desirable to harvest several complete sets of RO elements operating at conditions reflective of respective RO system challenges.

Membrane autopsy is typically performed by commercial or university laboratories specialized in such services, which have all the necessary instrumentation, test equipment, and skilled staff to complete the tests. It is preferable that the RO membranes sent for autopsy are accompanied with the original factory documentation for these elements, if available.

3.4.1 External Visual Inspection

Once they arrive at the test laboratory, RO membrane elements are first inspected by visual observation to identify potential physical damages of the fiberglass casting, the core tube, brine seals, and antitelescoping devices (ATDs), as well for telescoping, extrusion of the brine spacer, and discoloration of the ATDs. In addition, the surface of the front end ATD is inspected for accumulation of solids, unusual odors, and biofilm. The surface of the tail end caps is typically inspected for accumulation of crystals of mineral scale.

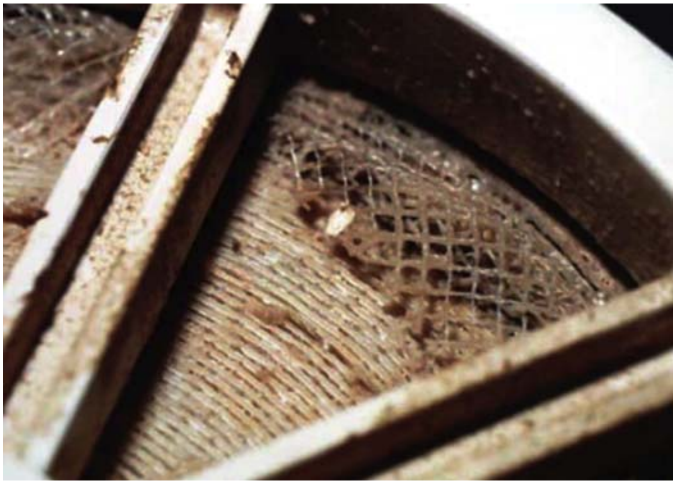


FIGURE 3.11 Membrane spacer protrusion.

Membrane telescoping (Fig. 3.1), physical damage of the fiberglass casting (see Fig. 3.2), and protrusion of membrane spacer (Fig. 3.11) are indicative of permanent membrane damage due to exposure of the membrane elements to excessive feed pressure surges, severe fouling, or mishandling. Reddish/brown discoloration of ATD (Fig. 3.12) is caused usually by overdosing of ferric coagulant.

3.4.2 Weight Measurement

All RO membranes are measured and their weight is compared against that of new RO elements of the same size. The magnitude of the weight gain is an indication of the extent

FIGURE 3.12 Reddish/brown discoloration of RO elements caused by overdosing of ferric coagulant.

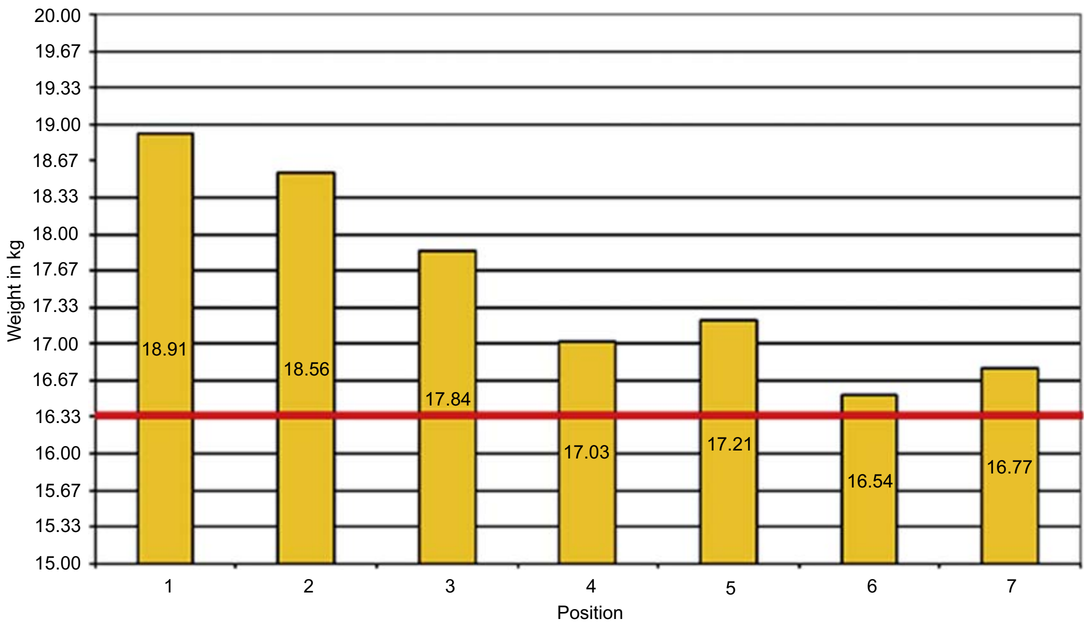


FIGURE 3.13 Example of the weight of SWRO elements in a vessel after 2 years of operation.

of membrane fouling. A standard new 8-in. RO element has weight of 16.4 kg (36 lbs). Under normal operating conditions, all membrane elements foul over time, and therefore their weight increases. Usually irreversible weight increase of the front RO-membrane elements over 18 kg (40 lbs) is an indication of heavy fouling.

Fig. 3.13 illustrates the weight of a set of SWRO elements, which were located in the same membrane vessel after 2 years of operation. The numbers indicated on the x-axis of the figure are the locations of the RO membranes within the vessel—"1" indicates the first (front) element; "2", the second element, etc. As seen from Fig. 3.13, the first two elements are heavily fouled and the fouling has propagated toward the fourth and fifth elements. For this specific desalination plant, such fouling has caused over 30% of permanent freshwater production (flux) reduction over the last 2 years the membranes were in service.

3.4.3 Mechanical Integrity Tests

Bubble test and vacuum decay test are used to identify whether the RO membranes may have been mechanically damaged by extreme fouling, backpressure, or may have membrane envelope glue line damage, and/or production defects. Such damages compromise the integrity of the RO membrane envelopes and usually result in increased salt passage, and ultimately, in unacceptably high salinity of the RO permeate.

3.4.3.1 Bubble Test

Bubble test involves closing one end of the RO membrane element permeate tube with an adapter plug, submerging the RO element into clean water and applying air at low pressure

FIGURE 3.14 Bubble test equipment.

(0.2 bars/3 psi) to the other side of the permeate tube for approximately 1 min. If continuous stream of air bubbles is released from the element into the water, this is an indication of loss of membrane integrity. If the element integrity is not compromised, it will stop releasing bubbles after the first minute and will hold the feed air pressure at a near-constant level. Failure of the membrane bubble test indicates that the element is damaged and is usually beyond repair. Typically, bubble test is used to identify membrane leaf glue damage and related salt rejection. Bubble test equipment is shown in Fig. 3.14.

3.4.3.2 Vacuum Test

This membrane integrity test involves applying vacuum of 20-in. mercury (Hg) level for 2 min to the open end of the RO membrane permeate tube. If the membrane element integrity is not compromised, the element will hold vacuum of over 19 bars at the end of the test. If more than 35% of the vacuum is lost at the end of the test, the RO element has a severe damage.

Table 3.2 illustrates an example of the results of bubble and vacuum tests of a set of seven SWRO elements. In this figure, “NL” and “CT” are abbreviations for No Leaks and Core-Tube leaks, respectively. As seen from this table, the tail element A608298 on position 7 has shown core-tube leaks during the bubble test and was not able to hold up vacuum (e.g., vacuum at the end of the test was 18.5 bars—less than the minimum threshold indicator level of 19 bars).

TABLE 3.2 Example Results From Bubble and Vacuum Tests of a Set of Seven SWRO Elements

	Bubble Test		Vacuum Test	
Serial Number	**Feed End**	**Brine End**	**Start**	**Finish**
A609590 position 1 (lead)	NL	NL	20.9	20.2
A609572 position 2	NL	NL	20.9	20.1
A609568 position 3	NL	NL	20.9	20.4
A609656 position 4	NL	NL	20.9	20.4
A611240 position 5	NL	NL	20.9	20.2
A619550 position 6	NL	NL	20.9	20.4
A608298 position 7 (tail)	NL	CT	20.9	18.5

3.4.4 Standard Membrane Performance Test

All RO-membrane elements are tested at a preset standard salinity, pressure, temperature, permeate recovery, and pH range after they are manufactured to confirm that they meet performance specifications in terms of permeate flow, salt rejection, and differential pressure (such test is referred to as "factory test"). The standard test conditions are different for SWRO, BWRO, and NF elements and they sometimes also vary among membrane manufacturers. When a given membrane element is purchased, its documentation contains the results of its original standard factory test.

As a part of the comprehensive performance analysis, the membrane elements provided for evaluation are retested at the same conditions as they were tested originally by the manufacturer and the values of membrane rejection, flow, and salt passage are compared. Significant reduction of membrane production and increase in differential pressure (over 8% per year of operation) are typically indicative of measurable membrane fouling.

Over 20% increase in salt passage could be attributed to damage of the membrane elements by oxidants, loss of mechanical integrity and/or membrane fouling. Comparison with the results of the other performance tests described herein is used to identify the most probable cause of the membrane performance problems.

If all membrane elements within the vessel have comparably high loss of production, this observation is indicative of propagation of the fouling throughout the length of the membrane vessel. If salt passage of all membranes in a given vessel has increased by approximately the same level, then it is very likely that these membranes have been exposed to oxidant. If only individual elements are exhibiting elevated salt passage, such phenomenon is likely caused by mechanical integrity loss of the elements.

3.4.5 Membrane Cleaning Test

Sample membrane elements are cleaned applying various combinations of cleaning chemicals and durations of soak and flush cycles to identify what is the optimal combination of cleaning solutions and their sequence of implementation to achieve maximum recovery of RO-membrane productivity and salt rejection and to minimize the membrane differential pressure. As a result of this test, the laboratory completing the membrane performance analysis prepares recommendations for efficient and cost-effective membrane cleaning.

3.4.6 Internal Inspection and Testing

Internal inspection and testing is completed as a part of the membrane autopsy. The anti-telescoping devices are first removed and the surface of the membranes is inspected for accumulation of solids, colloidal particles, biofouling, feed/brine spacer extrusion, and telescoping. Fig. 3.15 depicts the inner surface of the front ATD and the feed scroll. As seen from this figure this membrane is exposed to severe particulate fouling.

Once the membranes are opened along their length and unrolled into individual membrane leafs, the internal surface of the membrane leafs is visibly inspected for surface contamination, accumulation of debris, biofilm accumulation, and collection of scale on the RO membrane surface.

Fig. 3.16 presents membrane opened for internal membrane inspection. In this particular case, the membrane surface is covered with brown-colored biofilm, indicating heavy microbial membrane biofouling.

Besides the end caps and membrane surface, the membrane leafs are also inspected for the condition of their glue line integrity. As indicated in Chapter 1, membrane leafs are glued on the three sides and open to the permeate tube, forming an envelope. Glue lines of the membrane envelopes are inspected for failure (glue flaps, pouching, and delamination). Glue flaps are sections of unbounded (unglued) material, which could obstruct, and therefore hinder the spacer flow.

FIGURE 3.15 Accumulation of particulate debris on the front of the element and end cap.

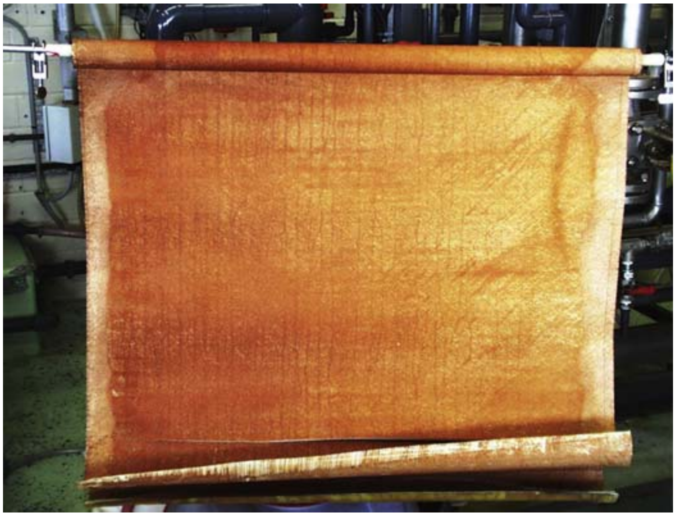


FIGURE 3.16 Inspection of membrane surface of RO element, showing heavy brown-colored microbial membrane fouling.

3.4.7 Dye Test

The methyl blue–violet dye test is applied to identify potential problems with membrane integrity. This dye at 10% solution is fed to the front (permeate) end of the tested RO element at nominal pressure and then the element is cut along its length and unrolled ("autopsied") to check whether this dye has been rejected or adsorbed on the membrane surface.

If the membrane surface is uniformly discolored with a very light blue/violet staining, this indicates that the membrane integrity is intact. If the membrane is physically damaged, then the applied dye will be absorbed on the damaged areas of the membrane surface and/or along the glue lines, if they are damaged, and will form highly visible purple stains as shown in Fig. 3.17. Fig. 3.18 presents the two sides [concentrate (A) and permeate (B)] of a damaged RO membrane.

3.4.8 Cell Test

Cell test is performed on a sample of the autopsied membrane element, which is first soaked in deionized (DI) water for 8–12 h to remove as much as possible the fouling from the surface of the membrane. This allows to assess the performance of the membrane element itself without the effect of fouling, so it can be compared against the results of the standard performance test, which will be reflective of the fouling impact of the membrane on performance (flux, salt rejection). The collected RO membrane sample, after cleaning in DI water, is installed in a special laboratory test cell (Fig. 3.19) in which the RO membrane is exposed to the same pressure, feed salinity, and temperature, as those used in the standard performance test.

Comparison of the salt rejection and flux of the same membrane, based on the results from the standard performance test and the cell test, allows to assess how much of the loss of

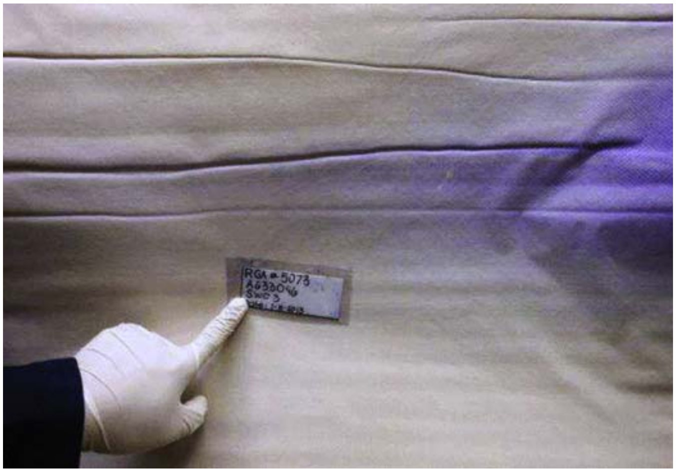


FIGURE 3.17 Damaged membrane area stained during dye testing.

(A) (B)

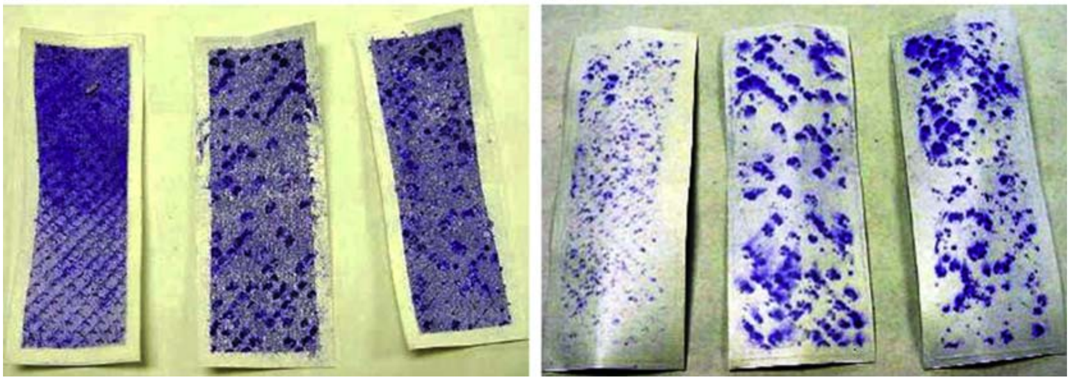


FIGURE 3.18 Concentrate (A) and permeate (B) sides of damaged membrane exposed to dye test.

membrane productivity and salt rejection is due to membrane fouling and how much is caused by damage of the membrane itself. It also allows determining, in general terms, how effective membrane cleaning would be in recovering the RO-membrane performance.

Example of cell test of front- and tail SWRO elements model SWC3 from Hydranautics is presented on Table 3.3. As indicated in this table, the "Test Data (Averages)" from the cell test for rejection and flux are significantly below the membrane factory test specifications.

Salt passage decrease of 26% combined with flux decline of 14% indicates that the first (lead/front element) is biologically fouled. Since the tested elements are 2-year old, the average flux decline per year is 7%, which is on the high-end of the acceptable maximum limit. Because the biological fouling has decreased salt passage, this element will be a good candidate for rotation, i.e., installation at the tail end of the membrane vessel.

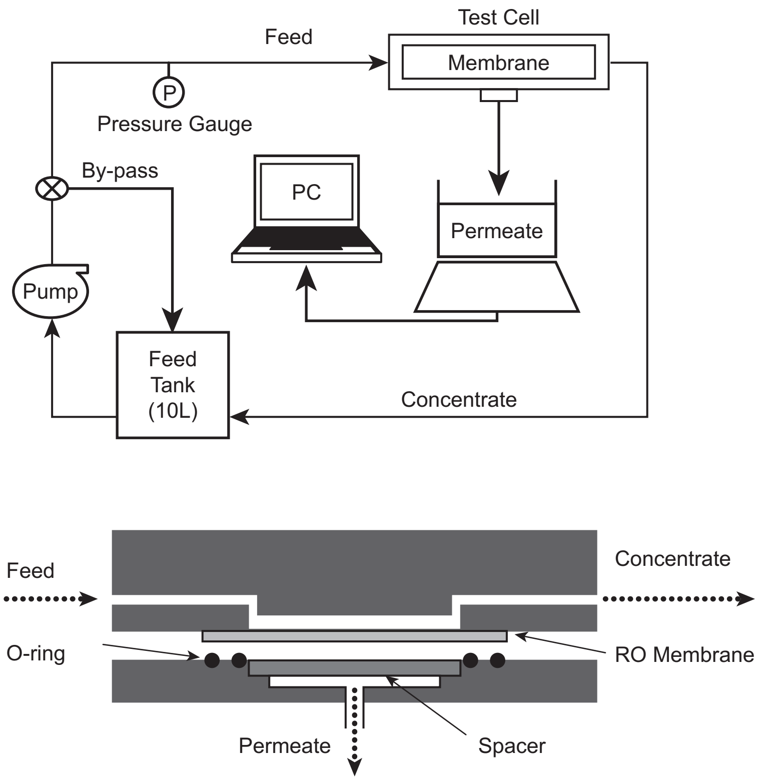


FIGURE 3.19 Schematic of standard membrane test cell.

TABLE 3.3 Example of Cell Test Results of Lead and Tail SWRO Elements

		Membrane Cell Test Performance Data					
		Membrane Specifications		Test Data (Averages)		Change in %	
Serial Number	**Membrane Type**	**Minimum Rejection**	**Nominal Flux Lmh (gfd)**	**Rejection**	**Flux, Lmh (gfd)**	**Salt Passage**	**Flux, Lmh (gfd)**
A609590 (lead)	SWC3	99.5%	32.3 (19)	99.63%	27.7 (16.3)	−26%	−14%
A608298 (tail)				98.78%	20.4 (12.0)	+144%	−37%

Test conditions:
Pressure: 15.2 bars (800 psi)
Feed concentration: 32,200 ppt NaCl
Flow: 0.34 m^3/h (1.5 gpm)
Feed temperature: 21.9 ± 0.3°C

However, the tested tail element A608298 has been exposed to both excessive flux loss (37%) and extremely high irreversible salt passage increase of 144%, and therefore is not suitable for rotation to lead position. As a consequence, this membrane element has to be replaced. As shown on Table 3.3, the results from the vacuum test indicate that this element cannot hold vacuum (e.g., it is irreversibly damaged), which confirms the observation from the cell test that the element is beyond repair and would need to be replaced.

3.4.9 SEM

Scanning electron microscope (SEM) photography is used to visibly assess the condition of the membrane surface and the topography of the membrane fouling. For example, Fig. 3.20 shows SEM photography of RO-membrane surface scaled with strontium sulfate.

3.4.10 EDaX

Energy Dispersive X-ray (EDaX) equipment generates electronic beams that strike the surface of the tested RO-membrane sample, which causes X-rays to be emitted from the material accumulated on the sample, and as a result, to generate graphs that show identified materials as peaks. Each peak is representative for a unique chemical element. The size of the peak is reflective of the amount of the chemical element contained in the membrane foulant. Besides the graph, the EDaX equipment also generates a table displaying the presence of the identified chemical elements by weight.

Fig. 3.21 presents an analysis of material collected from RO-membrane brine spacer. As it can be seen on this figure, the majority (71.3%) of the membrane material is carbon-based and the membrane fouling is mainly microbial (e.g., biofouling).

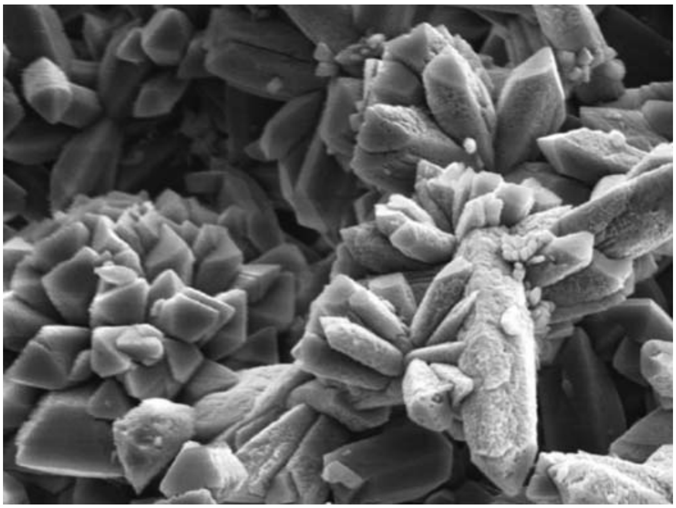
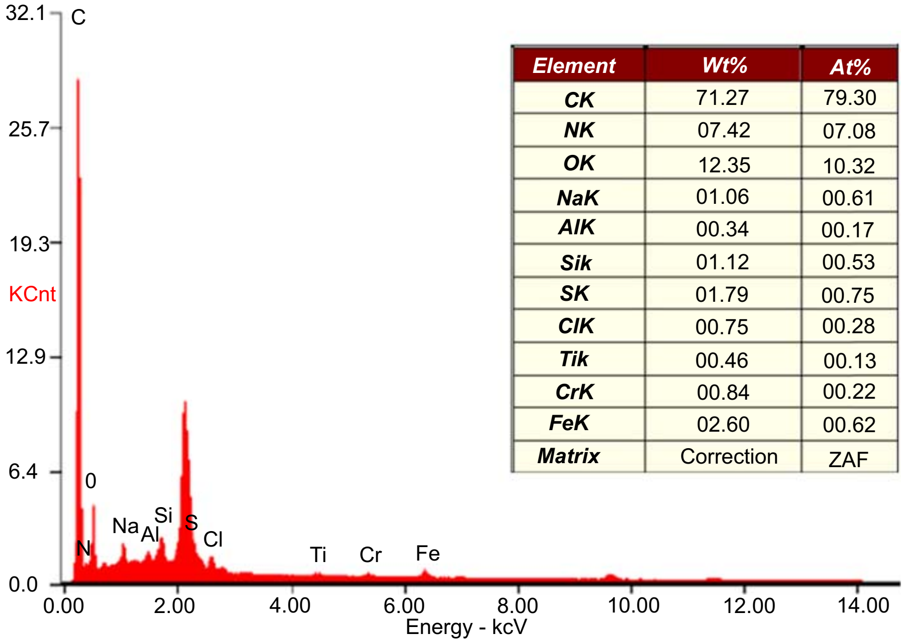


FIGURE 3.20 SEM photography of RO-membrane surface scaled with strontium sulfate.



Element	Wt%	At%
CK	71.27	79.30
NK	07.42	07.08
OK	12.35	10.32
NaK	01.06	00.61
AlK	00.34	00.17
Sik	01.12	00.53
SK	01.79	00.75
ClK	00.75	00.28
Tik	00.46	00.13
CrK	00.84	00.22
FeK	02.60	00.62
Matrix	Correction	ZAF

FIGURE 3.21 EDaX equipment-generated graph indicative of membrane fouling of organic origin.

3.4.11 Fourier Transform Infrared (FTIR) Spectroscopy

FTIR spectroscopy applies infrared (IR) radiation through a sample and by analyzing absorption and transmission of the IR it generates a molecular spectrum of the sample. The molecular spectrum (known also as "molecular signature" or "molecular fingerprint") of every compound is unique and the compound on the RO membrane could be identified by comparing its IR signature to a spectrum library of IR signatures of known compounds.

The FITR spectroscopy can be used for identifying of specific organic foulants such as polymers and antiscalants accumulated on the surface of the RO membranes. For example, in Fig. 3.22 the molecular spectrum of the RO membrane foulant matches with that of Nalco's polymer Cat-Floc 8103 Plus. As a result, the main source of organic fouling on the membrane surface in this case is polymer accumulation, caused by polymer overdosing.

3.4.12 Weight Loss on Ignition Test

The purpose of this test is to identify organic fouling on the surface of the membranes. For this test, sample of the foulant is collected from the surface of the RO membrane and its weight is measured. After the measurement, the sample is first dried at 110°C (230°F) and then heated to 550°C (1022°F). The weight of the residue from the sample ignition is measured and the difference (weight loss) as a result of the ignition process is calculated as percent of the initial weight. If the weight loss on ignition is over 35%, this result is indicative of organic foulant. Loss over 50% signifies heavy organic fouling.

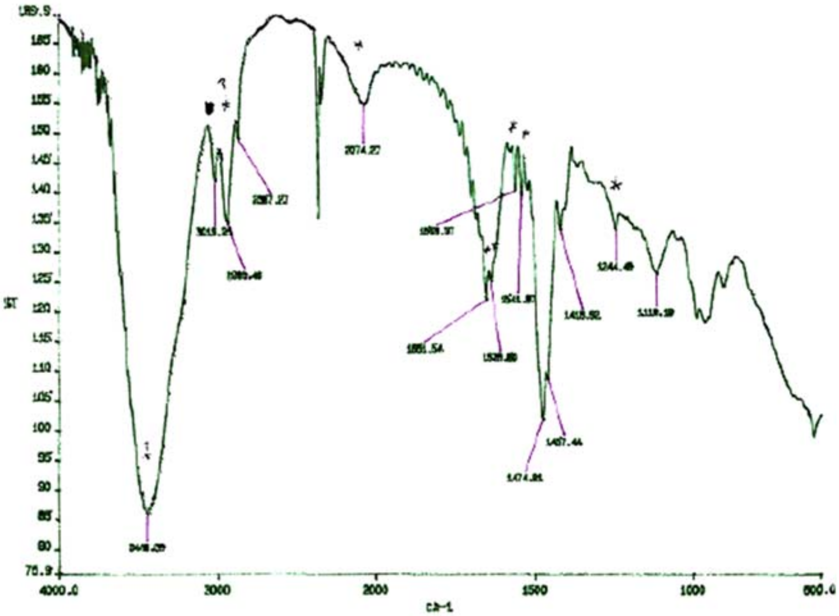


FIGURE 3.22 FTIR spectrum of RO-foulant matching the polymer Nalco Cat-Floc 8103 Plus.

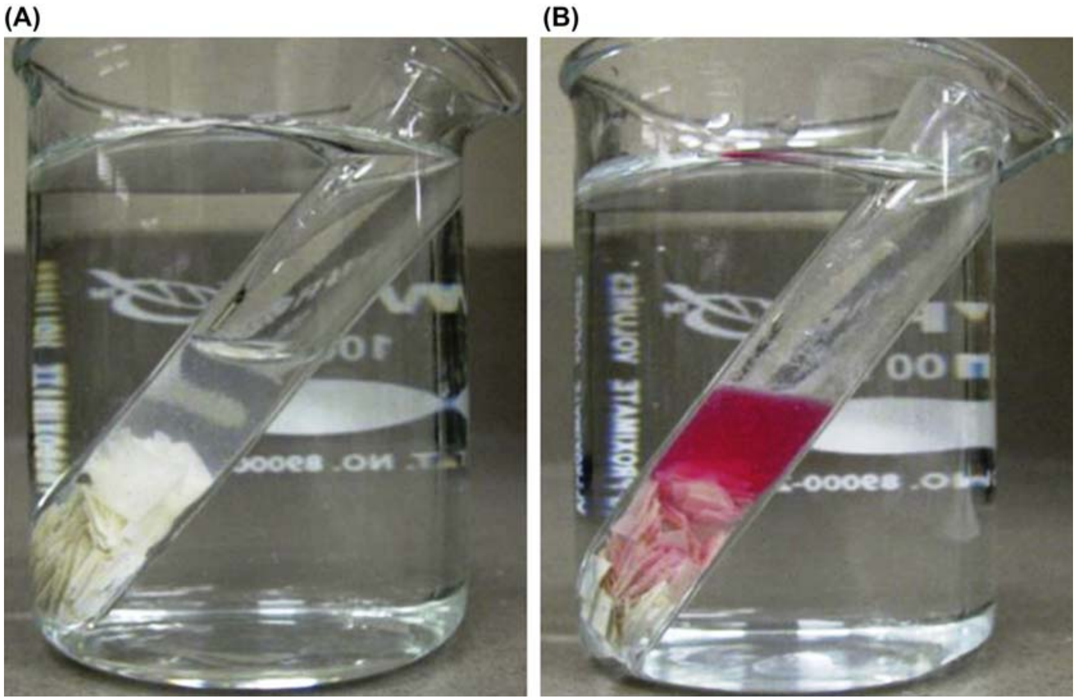


FIGURE 3.23 Color change from white (A) to red-violet (dark gray in print versions; B) during Fujiwara test indicating membrane exposure to halogens.

3.4.13 Fujiwara Test

This test is applied to determine whether the RO membrane has been exposed to oxidizing halogen (i.e., sodium hypochlorite, bromamine, iodine). The Fujiwara test is applicable for membranes made of polyamide (PA) material such as most spiral-wound RO elements used for desalination at present.

This test is qualitative only (e.g., provides only positive or negative answer regarding exposure to oxidants) and does not quantify the magnitude of exposure. Fig. 3.23 illustrates typical discoloration of halogen-exposed membrane sample during the Fujiwara test.

References

AWWA, 2007. Manual of Water Supply Practices, M46, Reverse Osmosis and Nanofiltration, second ed. American Water Works Association.

Chesters, S., Pena, N., Gallego, S., Fazel, M., Armstrong, M., de Vigo, F., 2011. Results from 99 seawater RO membrane autopsies. In: Proceedings of IDA World Congress, Perth, Western Australia, September 4–9, 2011.

Voutchkov, N., 2014. Desalination Engineering – Operation and Maintenance. McGraw Hill Education, New York.

CHAPTER

4

Saline Water Intakes and Pretreatment

OUTLINE

Pretreatment for Reverse Osmosis Desalination
http://dx.doi.org/10.1016/B978-0-12-809953-7.00004-8

4.1 INTRODUCTION

The saline source water intake is one of the key components of every desalination plant. The primary purpose of the intake is to ensure adequate and consistent flow and quality of source water over the entire useful life of the desalination plant. The type and configuration of intake selected for a given desalination project has a significant impact on the nature and quantity of foulants contained in the source water and, therefore on the complexity of the pretreatment system needed to control RO membrane fouling (Henthorne and Boysen, 2015).

Currently, there are two widely used types of desalination plant source water collection facilities—surface (open) and subsurface (groundwater, well) intakes. Open intakes collect source water directly from a surface water body (e.g., ocean, sea, brackish river, or lake) via onshore or offshore intake structure and pipeline connecting this structure to the desalination plant. The source water collected by this type of intakes usually contains debris, coarse and fine solids and silt, and aquatic organisms (e.g., fish, algae, bacteria, etc.), which make the pretreatment of this water more complex.

Subsurface intakes tap into brackish groundwater aquifer, or saline coastal aquifer to collect source water for the desalination plant. Source water obtained using subsurface intakes is conveyed through the soils of the aquifer, which naturally filter out debris, marine organisms and some organics, and thereby the RO-membrane fouling potential of the source water is reduced. As a result, the use of subsurface intakes for source water collection usually requires simpler pretreatment than source water collected by open intakes.

4.2 SUBSURFACE INTAKES

Subsurface intakes are the predominant type of source water collection facilities for brackish desalination plants of all sizes. Most BWRO plant intake wells collect water from deep confined aquifers of low to medium salinity (i.e., TDS of 600−3000 mg/L). Typically, deep brackish wells yield source water of low turbidity (<0.4 NTU) and silt content ($SDI_{15} < 2$), which can be processed through the RO system with minimal or no pretreatment (usually cartridge or bag filtration only).

Subsurface intakes for seawater desalination plants collect source water from either a saline near-shore (coastal) aquifer and/or from offshore aquifer under the ocean floor. The salinity of the coastal aquifers varies as a result of ocean water tidal movement and of changes in water level and salinity of the fresh and/or brackish groundwater aquifer/s hydraulically connected to the coastal aquifer. As a result, coastal aquifers are typically a source of saline water of TDS concentration lower than that of open-ocean water. For comparison, offshore aquifers yield water quality that is usually of the same TDS concentration as that of the ambient seawater. In addition, typically offshore aquifers collect seawater of lower RO-membrane fouling potential than coastal aquifers.

As compared to offshore aquifers, coastal aquifers have the disadvantage that their water quality is more variable and sensitive to changes in the water quality and volume of the fresh groundwater that drains into them. Therefore, for example, rain events could cause significant variations of the source water TDS, turbidity, iron and manganese and oxidationreduction potential, which typically have negative impact on the operation of the pretreatment and RO systems of the desalination plant.

In contract to brackish desalination plants where subsurface intakes are most common; at present less than 10% of seawater RO desalination plants worldwide use subsurface intakes. Such limited use of subsurface intakes for seawater desalination is mainly due to the fact that most coastal aquifers worldwide are limited in terms of sustainable volume of water that could be collected from them. Such limitation usually stems from the geology (e.g., limited natural water transmissivity) of coastal aquifers and the associated higher construction complexity and costs of subsurface intakes, especially for medium- and large-size desalination plants.

Several types of subsurface intakes have found application for collecting source water for SWRO desalination plants: (1) vertical wells; (2) horizontal directionally drilled (HDD) wells; (3) horizontal Ranney-type wells; and (4) infiltration galleries. The first three types of subsurface intakes are typically located on the seashore, in a close vicinity to the ocean. Vertical wells are used for both brackish and seawater desalination plants. The other three types of wells have found application mainly for seawater desalination plants.

The source water collected by subsurface intakes is pretreated via slow filtration through the bottom soil formations in the area of source water extraction. Therefore, this source water is expected to be of better quality in terms of solids, slit, oil & grease, and organic and microbial fouling potential, as compared to that collected by open intakes. However, practical experience shows that the notion that subsurface intakes collect better water quality than open intakes only holds true for site-specific conditions: usually when subsurface intakes are located on the shore or under the bottom of well naturally flushed surface water body (e.g., sea, ocean, brackish river); when they are not influenced of surface freshwater (e.g., not located in a confluence of river and ocean); and when they are collecting saline water from an aquifer of uniformly porous structure, such as limestone.

There are numerous small seawater desalination plants located in the Caribbean and several medium-size plants in Malta and Oman, which have such intakes and which require only minimal pretreatment (typically bag filters and/or sand strainers) ahead of RO pretreatment. However, the majority of the existing seawater desalination plants worldwide using subsurface intakes require an additional filtration pretreatment step prior to membrane salt separation. In addition, deep, well-configured and positioned open intakes often deliver saline source water of comparable or better quality than shallow coastal wells.

In summary, the assumption that subsurface intakes always deliver better water quality than open intakes does not hold true for all conditions and the selection of the most viable type of intake for a given desalination project has to be based on a detailed life-cycle cost analysis, which accounts for the site-specific project conditions and the impact of the quality of the source water collected by the intake on the configuration and costs of the downstream pretreatment system. The various types of subsurface intakes and their impact on desalination plant pretreatment needs are discussed below.

4.2.1 Vertical Wells

Vertical wells (Fig. 4.1) are the most commonly used type of subsurface intakes at present. Majority of existing BWRO desalination plants worldwide use vertical wells to collect groundwater from brackish aquifers. Similarly, over 30% of small SWRO desalination plants in operation at present apply vertical wells to obtain seawater from coastal aquifers for

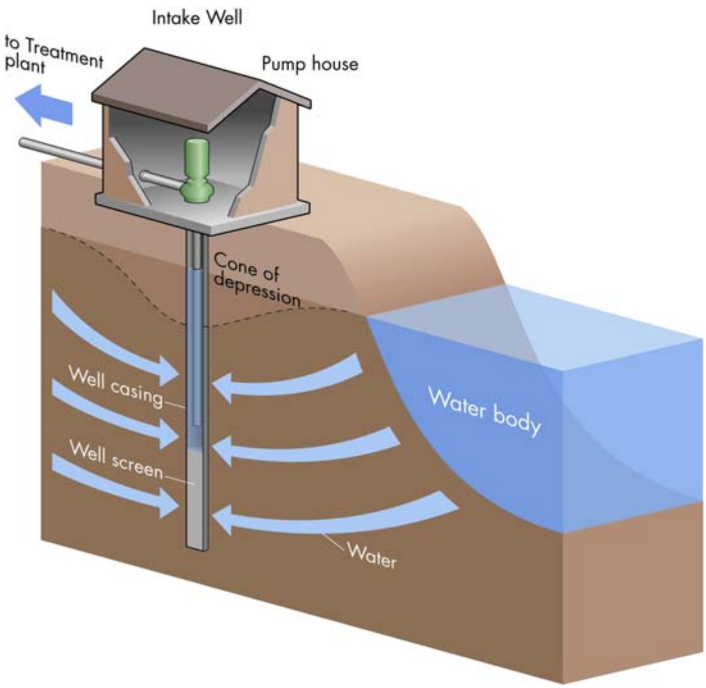


FIGURE 4.1 Vertical intake well.

freshwater production. Vertical wells that collect seawater from near-shore coastal aquifers are also referred to as beach wells.

Vertical wells consist of the following key components: casting, well screen, filter pack, well seal, and a surface seal. These wells are equipped with submersible or vertical turbine pump installed inside the well casting. Well casting is a steel or nonmetallic (typically, fiberglass) pipe, which lines the well borehole to protect wells from caving.

The diameter of the casting has to be of size adequate to house the well intake pump and to provide ample room for pump service. While the diameter of the well casting is determined mainly by the well screen size and yield, the well borehole diameter has to be at least 0.1 m (4 in.) larger than the well casting to accommodate the installation of the well seal. Usually, the well casting diameter is between 200 and 1200 mm (8 and 48 in.), and well depth is typically less than 75 m (250 ft).

The well screen is the intake portion of the well and is a sieve-like structure with slotted or perforated openings. The screen is located at a depth corresponding to the water-carrying zone of the aquifer. Screen depth, size of openings, diameter, and length are key well-performance design criteria. These well parameters are selected to maximize well's safe yield; control well entrance velocity; and to avoid excessive entrance of sand and other particulates, which have a negative impact on the well's useful life and water quality and on the downstream pretreatment system. Detailed guideline for selection of water-well castings and screens is provided elsewhere (Roscoe Moss Company, 2012; Voutchkov, 2013).

The performance of the well screen is enhanced by a gravel (filter) pack, which consists of clean, uniform, and well-rounded gravel and sand placed between the borehole wall and the well screen to prefilter the groundwater entering the well. Typically, the gravel pack depth extends at least 1 m (3 ft) above the well screen.

A well seal is installed above the filter pack to prevent soil and contaminants from entering the well-screen area. The well seal is a cylindrical layer of cement, bentonite, or clay placed in the annulus of the well between the well casting and the borehole. Typically, the well seal extends at least 0.6 m (2 ft) above the top of the gravel pack and usually through the elevation of the soil frost zone. The above ground portion of the well is finished with a concrete surface seal. The surface and well seals protect the well from surface runoff contamination and supports the casing.

Once constructed, the vertical well has to be monitored frequently to secure its long-term performance and identify early signs of potential malfunction and failure. The most common causes of well failure are borehole collapse, corrosion of the casing, improper or defective construction techniques, growth of organisms within the well borehole, and formation of mineral deposits or crusts in the open-hole or screened section of the well borehole.

The 80,200 m^3/day (21 MGD) Sur SWRO plant in Oman is the largest plant with vertical intake wells in operation at present. The intake area consists of limestone formations, which have average transmissivity of 7000 m^3/day m (David et al., 2009). The well field includes 33 (25 duty and 3 spare) beach wells capable of producing 70−100 Lps (1.6−2.3 MGD) each. Well depth is 80−100 m (260−330 ft) and the diameter of the wells is 14 in. Each well is equipped with 14-in. diameter PVC casting and screen with slot size of 3 mm. The wells are surrounded by gravel packs. Each well is equipped with a duplex stainless steel submersible pump and has an average drawdown on 12 m (39 ft).

4.2.2 Horizontal Ranney Wells

This type of wells consists of a concrete caisson that extends below the ground surface with water-well collector screens (laterals) projected out horizontally from inside the caisson into the surrounding aquifer (Fig. 4.2). Since the well screens in the collector wells are placed horizontally, higher rate of source water collection is possible than with most vertical wells. This allows the same intake water quantity to be collected with fewer wells. Individual Raney intake wells are typically designed to collect between 90 and 450 Lps (2 and 10 MGD) of source water.

The caisson of the horizontal collector well is constructed of reinforced concrete of 2.7−6.0 m (8.9−19.7 ft) inside diameter with a wall thickness from approximately 0.5−1.0 m (1.6−3.3 ft). The caisson depth varies according to site-specific geologic conditions, ranging from approximately 10 to over 45 m (33−148 ft).

The number, length, and location of the horizontal laterals are determined based on a detailed hydrogeological investigation. Typically, the diameter of the laterals ranges from 0.2 to 0.3 m (8−12 in.) and their length extends up to 60 m (197 ft). The size of the lateral screens is selected to accommodate the grain-size of the underground soil formation. If necessary, an artificial gravel-pack filter can be installed around the screens to suit finer-grained deposits. Usually, one well has 2−14 laterals oriented toward the source water body (ocean, river).

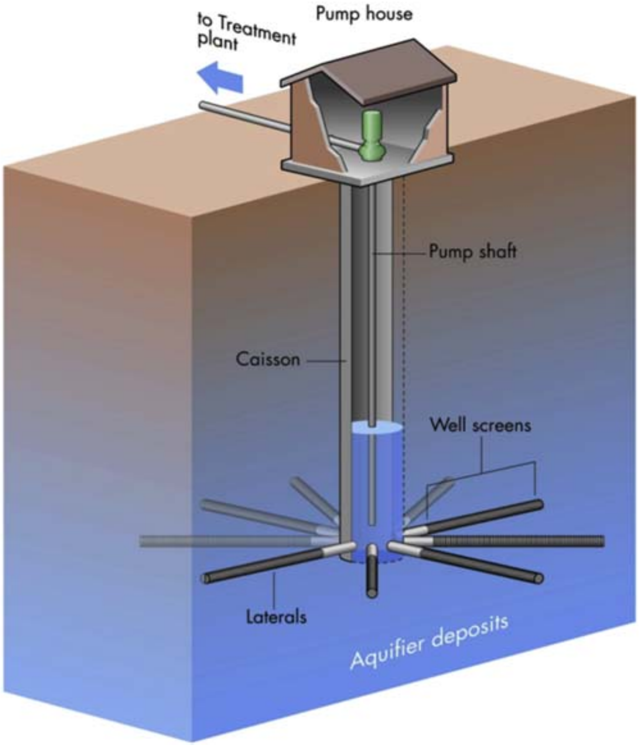


FIGURE 4.2 Horizontal Ranney-type intake well.

Ranney intake wells are typically coupled with a pump station installed above the well caisson. The pump station can be designed with submersible pumps to minimize noise levels. However, intakes that consist of mid- and large-size wells most frequently employ horizontal or vertical turbine pumps because these pumps usually have higher energy efficiency and require less power than submersible pumps.

Ranney wells are not as commonly used as vertical wells for seawater intakes. The largest horizontal-well installation is located in Salina Cruz, Mexico (Fig. 4.3) and consists of three wells, which are designed to deliver 168 Lps (3.8 MGD) of seawater, each. The wells are located on the beach and the water quality they collect is not directly suitable for seawater desalination. The source seawater from these wells contains high levels of iron and manganese and has to be pretreated in greensand filters prior to SWRO separation.

One potential challenge with Ranney wells is that if the source water contains hydrogen sulfide, this compound will likely be oxidized to elemental sulfur, which could subsequently cause RO membrane fouling (Missimer et al., 2010).

4.2.3 Horizontal Directionally Drilled Wells

The HDD intake wells consist of relatively shallow blank well casting with one or more horizontal perforated screens bored under an angle (typically inclined at 15–20 degrees)

FIGURE 4.3 Two Ranney wells for Salina Cruz SWRO plant, Mexico.

and extending from the surface entry point underground past the mean-tide line. This type of wells has found application mainly in seawater desalination installations.

One of the most widely used HDD well intakes today is the Neodren well intake system. A general schematic of Neodren HDD collectors is shown on Fig. 4.4. A detailed description of typical configuration of the HDD wells is presented elsewhere (Peters and Pinto, 2008).

HDD-well intakes collect source water via a number of perforated high-density polyethylene (HDPE) pipes with 120-μm pore openings at a relatively slow rate. The collected saline source water is naturally filtered through the ocean bottom sediments before it reaches the desalination plant. Typical collector pipe size is 450 mm (18-in.). However, collector pipes of diameters as large as 710 mm (28-in.) could be installed. Typically, the individual HDD collector pipes deliver the source water into a common wet well from where it is pumped to the desalination plant for further processing. The individual HDD well collectors of this type usually yield between 50 and 150 Lps (1.1–3.4 MGD).

The collector pipes are installed usually at a depth between 5 and 10 m (16–33 ft) below the surface of the ocean bottom in separate boreholes by drilling under the ocean seabed 200–600 m (660–2000 ft) into the coastal aquifer to a location that can yield seawater of

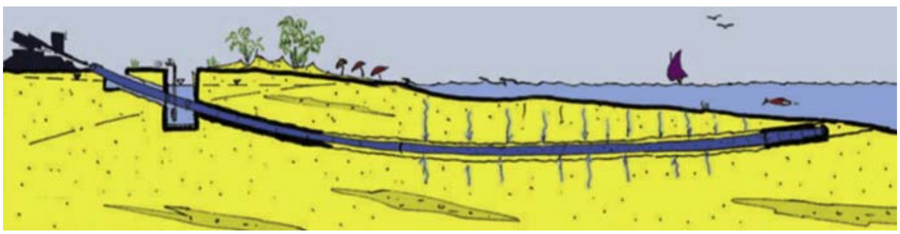


FIGURE 4.4 HDD intake.

ambient seawater salinity and avoid water collection from fresh near-shore aquifers in the vicinity of the plant site.

The Neodren HDD intake technology is patented by the Spanish company Catalana de Perforacions. This technology has been used for over 15 years in over a dozen small- and medium-sized seawater desalination plants in Spain, and currently it is under construction or consideration/pilot testing at a number of other plants worldwide.

The capacity of the collected source seawater depends on: the number and the diameter of the horizontally drilled perforated pipes; the length of the perforated portion of the pipes; and the transmissivity and depth of the seabed in which the collector pipes are drilled.

The natural seabed filtration process removes practically all coarse solids and particulates of a size of 50 μm or larger from the seawater and precludes marine organisms in all phases of development (adults, juveniles, and larvae) from entering the desalination plant (i.e., protects marine life against impingement and entrainment). This system is also an effective barrier against the heavy solid loads generated during algal blooms and oil spills.

Data available to date (Peters et al., 2007), however, indicates that while the HDD system can successfully reduce source seawater turbidity and TOC, this reduction typically is not adequate to directly apply the water collected by the HDD intakes to the SWRO membranes. The source water would need to be pretreated through filtration system prior to RO membrane separation.

One of the largest seawater desalination plants in operation at present that uses HDD wells is located in Spain—the 65,000 m^3/day (17 MGD) New Cartagena Canal (San Pedro del Pinatar) plant. The intake system consists of 20 HDD wells arranged in a fan shape—see Fig. 4.5.

The individual intake wells are between 500 and 600 m (1600 and 2000 ft) long and have 355 mm (14 in.) diameter. Each well produces between 100 and 140 Lps (2.3 and 3.1 MGD).

FIGURE 4.5 HDD Intake of San Pedro del Pinatar SWRO plant, Spain.

The plant operates at 45% recovery. The water is collected in a large wet well that is located underground and pumped to the plant using submersible pumps.

4.2.4 Infiltration Galleries

Infiltration galleries consist of a submerged slow sand media filtration bed incorporated into the bottom floor of the source surface water body (i.e., ocean, lake, or river). The filtration bed contains a number of equidistant horizontal perforated pipes that convey filtered source water to a wet well of an intake pump station located on shore (see Fig. 4.6).

Infiltration galleries are typically implemented when conventional horizontal or vertical intake wells cannot be used due to unfavorable hydrogeological conditions. For example, they are suitable for intakes where the permeability of the bottom soil formation is relatively low, or in the case of river or seashore filtration, where the thickness of the onshore and offshore sediments or the depth of the coastal aquifer is insufficient to install conventional intake wells.

The filtration bed of the infiltration gallery is sized and configured using the same design criteria as slow sand filters. The design surface-loading rate of the filter media is 0.12–0.25 m^3/m^2 h (0.05–0.10 gpm/ft^2).

Such subsurface intake systems are only feasible in locations where there is good periodic or continuous natural movement of water over the bed because they rely mainly on wave and current action to remove the solids retained and accumulated in the surface layer of the filtration bed. One potential challenge with infiltration galleries is the biofouling of the filtration media, which could reduce their production capacity over time.

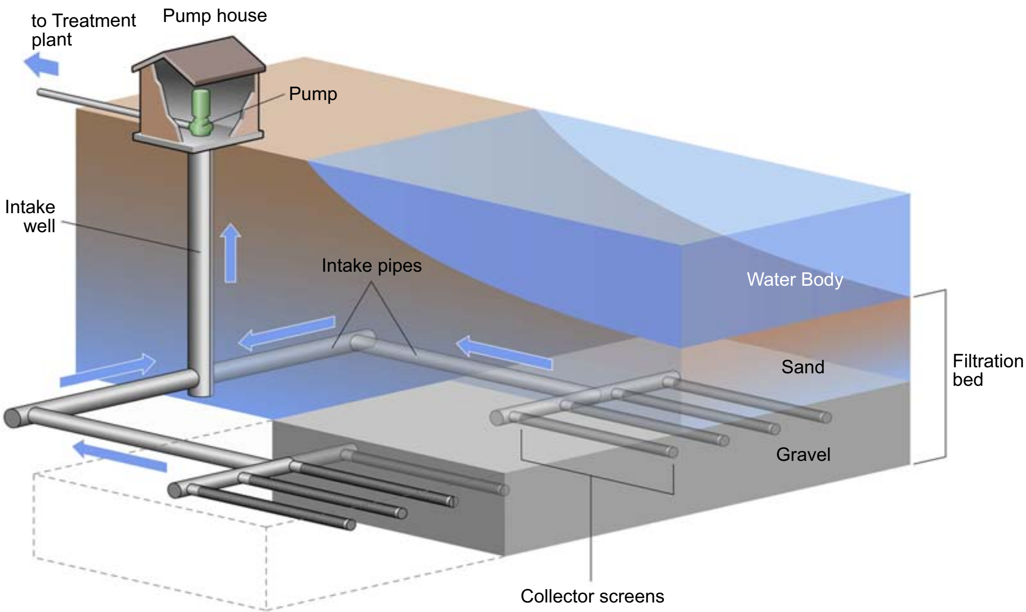


FIGURE 4.6 Infiltration gallery.

The largest seawater desalination plant with a seabed infiltration system in operation at present is the 50,000 m^3/day (13.2 MGD) Fukuoka District RO facility in Japan. This plant has been in operation since 2006. The infiltration bed is 313.6 m long and 64.2 m wide (2 ha/5 acres), and it is designed to collect source seawater flow of 130,000 m^3/day (34 MGD). The design infiltration velocity of this intake system is 0.25 m^3/m^2 h (0.10 gpm/ft^2).

The filtration media in the well is configured in three distinctive layers (see Fig. 4.7): a bottom—2.3-m (7.5 ft) layer of graded gravel pack with stone sizes between 20 and 40 mm, which surrounds the horizontal well collectors; in the middle—0.3-m (1.0 ft) deep interim layer of finer graded gravel of sizes between 2.5 and 13 mm; and on the top—1.5-m (5.0 ft) layer of natural sand excavated from the ocean bottom. The filtration media is submerged at 11.5 m (37.7 ft) below the ocean surface.

The filtered water collectors are 60-m (200 ft) long, 600-mm (24-in.) diameter polyethylene pipe screens (Fig. 4.8), which are installed at a distance of 5 m (16.4 ft) of each other. The collector pipes are designed for inflow velocity of 3 cm/sec (0.1 fps). The screens collect the source water flow into a central pipe with diameter of 1580 mm (62-in.) and length of 1178 m (3860 ft), which conveys it into a two-tank water collection well for pumping to the desalination plant. The collected water is pretreated with UF membrane filtration prior to desalination in SWRO membrane system.

4.2.5 Feasibility Considerations for Subsurface Intakes

4.2.5.1 Intake Site Location and Configuration

Site suitability for the construction of subsurface intake is determined by the following main factors: the transmissivity/productivity of the geological formation and aquifer from

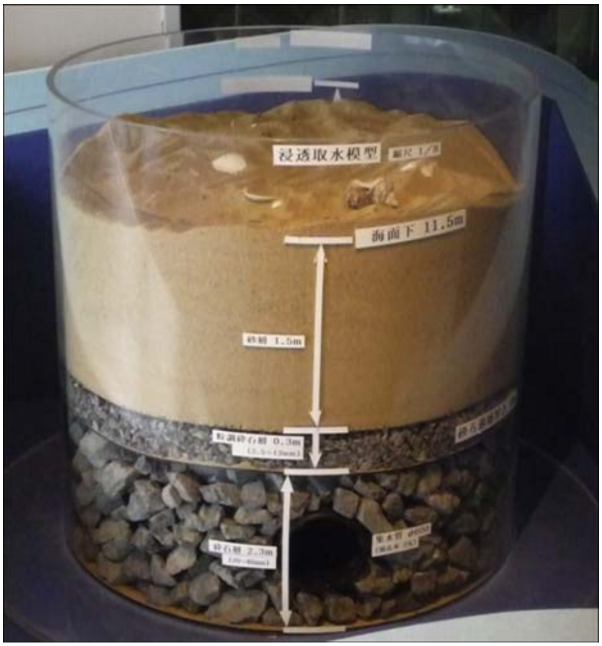


FIGURE 4.7 Seabed filtration media configuration of the Fukuoka SWRO plant.

FIGURE 4.8 Segment of 600-mm intake collector screen.

which saline water is collected; the depth and stability of the coastal zone through which the source water is naturally filtered; underground sources of contamination of the intake area (e.g., landfills, cemeteries, leaking underground oil tanks); and the existence of nearby freshwater source aquifers, which could be negatively impacted by the intake operations or have measurable effect on the quality of the collected water.

The geological conditions that favor the construction of subsurface intakes are permeable sand and limestone- or dolomite-type geological formations with high transmissivity and depth. Such conditions are not very common in coastal zones worldwide, except for some locations where subsurface intakes have been used successfully such as Malta, portions of the coastal zone of Oman, and some of the Caribbean islands.

Beaches of shallow bays that contain significant amount of mud/alluvial deposits and have limited natural flushing do not favor the use of wells for RO desalination plant intakes. High content of fine solids in the bay seawater in combination with low frequency of bay flushing and low transmissivity of the beach deposits may render shallow bay beaches less desirable or unsuitable for construction of well intakes. If the selected bay is shallow and poorly flushed, the water quality collected by the subsurface intake is likely to be variable and of high content of silt and other substances, which would require significant expenditures for source water pretreatment prior to reverse osmosis (RO).

It should be pointed out that both wells and near-shore open intakes use the same seawater as a source. In desalination plants with open intakes, the solids contained in the source seawater are removed in the desalination plant pretreatment filtration system in controlled and closely monitored conditions. In desalination plants with subsurface intakes, the same amount of solids is retained on the ocean floor in the area of source water collection while the filtered water is slowly conveyed through the ocean floor and the beach soils until it reaches the intake collectors.

The wave action near the ocean floor is the force that allows the solids separated from the source water by natural filtration to be dissipated in the ocean. If the bay area is not well flushed and the naturally occurring wave movement is inadequate to transport the solids away from the intake collection area at a rate higher than the rate of solids deposition and retention by natural filtration, then these solids would begin to accumulate on the ocean floor and would ultimately enter the filtration bed, reduce the well capacity, and deteriorate the source water quality over time. Since filtration process through the ocean bed is very slow, it usually takes 6–12 months until deterioration of well performance and capacity are observed. Therefore, if such conditions of poor flushing of the intake area occur, loss of well capacity over time is very likely and is difficult to predict.

Another important factor to consider when assessing the ability of subsurface intakes to pretreat the source water is the impact of heavy storms on the depth of the filtration layer through which the water passes before it enters the intake wells. For such natural filtration layer to perform comparably to dual media granular filters or membrane pretreatment filters, the depth of the natural filtration layer should be 10–30 m (67–98 ft). For example, if large storm or beach erosion due to tidal movement reduces this depth (distance) between the surface of the ocean bottom and the filter collectors to less than 10 m (33 ft), the source-water quality produced by the subsurface intake will deteriorate drastically. For comparison, man-made pretreatment systems for source water collected by open intakes are designed to handle the impact of storms and beach erosion on the source water quality, and under the same circumstances, their pretreatment performance would not be affected significantly.

Subsurface intakes whose area of influence (e.g., source water collection area) extends to nearby fresh/brackish groundwater aquifers may have a negative impact on the capacity and water quality of these aquifers, and in some cases their operation may result in enhanced seawater intrusion. In addition, if the coastal aquifer used for source water supply is hydraulically connected to nearby saltwater marches (coastal wetlands), the collection of large amounts of water may result in drainage or significant reduction in the water level in the marshes, which in turn may have a detrimental impact on the marsh ecosystem/drain the coastal wetlands. Such negative environmental impact of subsurface intakes is more pronounced in large desalination projects where the radius of influence of the subsurface intake on coastal aquifers could extend to 5 km (3 miles) or more inland.

If a nearby fresh/brackish groundwater aquifer hydraulically connected to the coastal intake aquifer is contaminated with pollutants, which are typically not present in the ambient open-ocean water in a measurable quantity (such as fuel oil contaminants, endocrine disruptors, heavy metals, arsenic, septic tank or landfill leachate, leachate of embalming solutions from cemeteries, etc.), then the RO plant may need to be provided with additional pretreatment and/or disposal facilities, which would erode the benefits of subsurface intake use.

The numerous factors that could impact subsurface intake-water quality described above point out to the fact that assessment of the potential sources of contamination of the source water in the zone of influence of the subsurface intake is a critical and integral part of the feasibility evaluation of using subsurface intakes and the complexity of source water pretreatment for the site-specific conditions of a given desalination project. The outcome of such evaluation would allow determining whether the natural source water pretreatment provided by the subsurface intake would be worse, comparable, or better than the man-made pretreatment system threating ambient seawater collected by an open intake.

Determining the feasibility, productivity and pretreatment performance of a subsurface intake system at a given site is typically based on a hydrogeological investigation, which entails the following key activities:

1. Completion of preliminary geological survey to identify if the selected site is generally suitable for the construction of subsurface water intake.
2. Implementation of a drill-test to collect samples of the aquifer formation deposits for visual classification and grain-size distribution analysis.
3. Installation of one or more test wells and observation wells, and conduction of pumping test to determine the site-specific hydraulic characteristics on the aquifer necessary for subsurface system design and to quantify the intake system's safe yield. The test should be minimum 72-h long. Usually, a number of incremental tests are completed in sequence to determine the optimum yield. This initial testing is followed with a longer-term test, which is used to determine the aquifer transmissivity.
4. Collection of adequate amount of samples of the source water and analysis of sample water quality, with special emphasis on the content of iron, manganese, barium, strontium, silica, radon, carbon dioxide, arsenic, and hydrogen sulfide in the source water. If the aquifer water quality is under the influence of a surface water source (i.e., river, lake) whose quality and quantity varies seasonally, then a year-around intake water quality sampling is completed to quantify seasonal fluctuations of the source water quality.
5. If the subsurface intake system would require the installation of multiple collection facilities (wells, infiltration galleries/river bank filtration facilities), then a computer model analysis is completed to establish the response of the production aquifer to pumping and potential impact of groundwater collection on adjacent fresh or saline water aquifers, which could be in interaction with the water supply aquifer.
6. Beach erosion, seismic, and 100-storm impact analyses of the intake area are completed to evaluate whether the intake system will be able to retain its integrity, productivity, and water quality over the useful life of the desalination plant.
7. Survey of potential sources of contamination of the source water aquifer within the area of influence of the subsurface intake (landfills, cemeteries, oil fields, gas stations, etc.) is completed to identify whether the desalination plant intake water quality will be impacted by such sources.
8. Survey of existing drinking water wells within the area influence of the desalination plant subsurface intake is implemented to determine whether the freshwater production capacity and quality of these wells will be impacted by the desalination plant operations.
9. Survey of the coastal wetlands, marshes, or other natural coastal habitats in the area of influence of the subsurface intake to determine whether they could be impacted (e.g., drained) by the operation of the subsurface intake.
10. Survey of seismic faults located in the area of influence of the subsurface intake and assessment whether future earthquakes could cause negative impacts on the aquifer from which saline source water is planned to be collected and on the structural integrity and productivity of the subsurface intake (e.g., wells, infiltration gallery, etc.)

The key information derived from the studies listed above is: the type of the source water aquifer (confined vs. unconfined); the aquifer permeability (hydraulic conductivity) that is a measure of the velocity of water movement through the ground (typically measured in m/s); the average specific yield (productivity) of the aquifer (in m^3/day per linear meter of riverbank or seashore along which the collector wells are located); the thickness of the production aquifer deposits and associated well depth; the source water quality of the subsurface intake to determine the desalination plant pretreatment; fatal flaws for use of subsurface intakes such as severe beach erosion and damaging seismic activity; sources of subsurface contamination in the intake area; and the existence of nearby fresh or brackish water aquifers, which could be negatively impacted by the intake well operations or may have measurable effect on intake well-water quality.

A confined aquifer (also referred to as an artesian aquifer) is a water-saturated geological formation between two layers of low permeability (i.e., bedrock), which restrict the vertical movement of the groundwater in or out of the aquifer. Confined aquifers are often pressurized by the surrounding geological formations and, therefore, collecting water from such aquifers may not require pumping. Unconfined aquifers are groundwater-saturated formations with fluctuating water level/table. Such fluctuation is driven by recharge from surface runoff (rain or snowmelt) or changes in the water table of surface water body (ocean, river, lake, etc.) hydraulically connected to the aquifer.

Coarse-grained porous and highly permeable geological formations (i.e., sandstones, beach sand and alluvial deposits, coarse-grained gravel and limestone) connected to a brackish riverbed (for brackish plant intakes) or to the ocean floor (for seawater intakes), whose specific yield (transmissivity) exceeds 1500 m^3/day m, and which have water-carrying zone of at least 6.0 m (20 ft) are most suitable for subsurface intakes. The higher the aquifer permeability, transmissivity, and thickness, the larger well yield the aquifer can support. Such soil conditions often exist along coastal dunes, reefs, and alluvial deltas. Fractured rock formations could deliver high volume of flow, but often the collected water is of inferior quality. The worst subsurface substrates for installation of subsurface intakes are those formed as a result of volcanic activity such as basalt and lava, as well as granite and clay formations.

One key criterion whether vertical wells can be used or horizontal wells would need to be installed is the presence of faults along the coast in parallel to the ocean shore. If such faults exist, typically, there is no hydraulic connection between the coastal aquifer and the sea and, therefore, vertical wells will not be suitable.

4.2.5.2 Need for Additional Pretreatment of Subsurface Intake Water

As mentioned previously, subsurface intakes typically yield better source-water quality than open intakes in terms of saline water turbidity and silt density index, which are two of the key parameters associated with the selection, sizing, and costs of the desalination plant pretreatment system. Therefore, often it is assumed that the use of subsurface intakes would eliminate the need for saline source water pretreatment prior to RO desalination.

However, the existing experience with the use of subsurface intakes for seawater desalination in California and at the largest beach-well seawater desalination plant on the West Coast in Salina Cruz, Mexico, indicate that some desalination plants using subsurface intakes may face a costly challenge—high concentrations of manganese and/or iron in the intake water. Unless removed ahead of the RO membrane system, iron and manganese may quickly foul the cartridge filters and RO membranes and render the desalination plant inoperable. The treatment of subsurface intake water that contains high concentrations of iron and/or manganese (see Chapter 2) requires chemical conditioning by oxidation and coagulation, and installation of conservatively designed "greensand" or membrane pretreatment filters ahead of the RO system. This costly pretreatment requirement may significantly reduce the benefits of using subsurface intake as compared to an open intake for a given project. Open seawater intakes typically do not have iron and manganese source-water quality-related pretreatment challenges because ambient ocean water does not contain these compounds in significant quantities to cause membrane-fouling problems.

Example of desalination plant with vertical beach wells that faced an elevated source water iron problem is the 4500 m^3/day (1.2 MGD) Morro Bay SWRO facility located in Northern California, USA (Kartinen and Martin, 2003). The plant source water is supplied by five beach wells with a production capacity of 1100 m^3/day (0.3 MGD) to 1900 m^3/day (0.5 MGD), each. The beach-well intake water has iron concentration of 5–17 mg/L. For comparison, open-intake seawater, typically, has several orders of magnitude lower iron concentration.

The Morro Bay facility was originally designed without pretreatment filters, which resulted in plugging of the SWRO cartridge filters within half-an-hour of starting operations during an attempt to run the plant in 1996. The high-iron concentration problem was resolved by the installation of pretreatment filter designed for a loading rate of 6.1 m^3/m^2 h. For comparison, a typical open-intake desalination plant is designed for pretreatment loading rates of 10–12 m^3/m^2 h—and, therefore would require less pretreatment filtration capacity if open-ocean intake was used for this project.

As indicated previously, the largest existing Pacific-coast seawater desalination plant in Salina Cruz, Mexico with subsurface intake has also faced an iron and manganese challenges, which were resolved by the installation of pretreatment filters and chemical conditioning of the Ranney well water. The existing experience shows that, the costs for pretreatment of saline water with high iron/manganese content collected by subsurface intake are typically higher than these for pretreatment of saline water collected using an open intake.

4.2.5.3 Source-Water Quality Variation

Open intakes provide relatively consistent saline source-water quality in terms of total dissolved solids concentration. For example, the intake source-water TDS concentration data collected for the development of the Huntington Beach and Carlsbad seawater desalination projects in Southern California, USA, indicate that the open intake salinity varied within 10% of its average value of 33.5 ppt.

Although in general, subsurface intakes produce source water of very consistent salinity as well, they could also yield water of unpredictably variable TDS concentration with swings exceeding over 30% of the average. For example, the TDS concentration of the two operational Ranney wells at the Salina Cruz desalination plant vary in a wide range—for well No. 2 between 16.8 and 21.8 ppt, and for well No. 3 between 17.8 and 19.8 ppt (Rovel, 2002). The wide range of source salinity concentration in this case is explained by fresh groundwater influence and intake location near river entrance to the ocean.

A similar trend was observed at the Morro Bay SWRO plant in California. During the plant's initial operation in 1992, the well-water TDS was approximately 26 ppt. In December 2001, the TDS of the intake water was 6.3 ppt. The December 2002 data for the same plant indicate intake salinity of 22 ppt (Kartinen and Martin, 2003). The wide range of source water salinity variation in systems using subsurface intakes over time would require the installation of variable frequency drives for efficient power use control, which would ultimately increase the construction cost of such system and complicate its operation.

One important issue to be taken under consideration when assessing the viability of using subsurface intakes for a given project is the fact that source water salinity could change unpredictably over time when influenced by freshwater inflow to the coastal aquifer. This uncertainty of intake water quality increases the risk of uncontrollable increase in unit cost of water production over time and has to be taken in consideration when comparing the overall life-cycle costs of the desalination plant operations. Therefore, subsurface intake water quality has to be thoroughly characterized by installing a set of test wells and collecting water quality samples under variety of operational conditions for a period of 12–18 months. A common practice of running such test for 72 h only and collecting one to three source-water quality samples have shown to be inadequate for predicting the potential source-water quality variations, which may occur over time and for selecting the most suitable desalination plant pretreatment system.

Temperature is a key source water quality parameter that has a measurable effect on the RO-system design feed pressure and membrane performance. Typically, the SWRO-system feed pressure needed for production of target freshwater production flow is reduced by 5%–8% on a linear scale for every 10°C source water temperature increment in a temperature range of 12–40°C.

Based on tests completed at the Carlsbad seawater desalination pilot plant on cold Pacific Ocean water in the winter, when the source water temperature drops below 12°C, the temperature effect is even more dramatic—the SWRO feed pressure increases with 5%–10% for every 2°C of temperature drop on an exponential scale until the source water temperature reaches 4°C, below which the source water would begin to freeze and seawater desalination is dramatically hindered. The accelerated exponential increase in the operational SWRO feed pressure for source water temperatures below 12°C is explained by similar curvilinear increase in source water density in the temperature range of 4–12°C combined with changes in membrane material behavior.

Source seawater of temperature above 40°C typically has two adverse effects on membrane performance that may negate the positive effect higher temperature on membrane pressure: (1) change in membrane material behavior (membrane compaction/internal fouling), which could result in shorter membrane useful life; and (2) accelerated membrane biofouling due to the effect of temperature on bacterial growth. An additional negative impact of temperature on membrane performance is the reduction in membrane salt rejection with the increase of source water temperature. Therefore, operation at high source water temperatures (typically, 30°C and higher) may compromise meeting product water quality goals in terms of TDS, chlorides, boron, sodium, and other product water quality requirements and may require the installation of additional treatment step—partial or full second RO pass—to address the negative effect of temperature on the product water quality.

The SWRO system construction-cost increase associated with the installation of partial or full second RO pass is typically in a range of 10%–25% of the cost of a single-pass SWRO system. The additional O&M costs associated with the operation of second pass-system vary between 3% and 10% of the costs for operation of the first pass.

Because of the higher depth of source water collection, subsurface intakes usually yield seawater of lower temperature and, therefore, their use may result in higher energy demand for production of the same volume of desalinated water. As discussed above, the difference in energy use may exceed over 20%, especially if compared to desalination plant collocated with power plant that collects water which is usually 5–15°C warmer than the ambient seawater. This benefit should, however, be assessed on a case-by-case basis, especially when the ambient seawater temperature exceeds 30°C, because of the negative impact of warm water on RO permeate quality and membrane biofouling.

Usually open-ocean intakes are considered less viable source of water for desalination plants in areas located in a close proximity to wastewater discharges or industrial and port activities. However, open-ocean intake seawater is typically free of endocrine disruptor or carcinogenic type of compounds such as: methyl tert-butyl ether (MTBE), *N*-Nitrosodimethylamine (NDMA) and 1,4-dioxane. Long-term water quality data collected for the development of the Huntington Beach and Carlsbad SWRO projects in Southern California and a number of other desalination plants worldwide confirm this observation.

Subsurface intake water, however, may contain difficult-to-treat compounds especially when they are under influence of contaminated groundwater. Example is the Morro Bay SWRO plant, where beach well-intake water was contaminated by MTBE caused by underground gasoline tank spill. MTBE is a gasoline additive. Similar problems were observed at the California's Santa Catalina Island 500 m^3/day (0.13 MGD) seawater desalination plant that uses beach-well intake.

The compounds of concern could be treated by a number of available technologies, including activated carbon filtration, UV irradiation, hydrogen peroxide oxidation, ozonation, etc. However, because these pretreatment systems will need to be constructed in addition to the RO system, this additional pretreatment may increase the overall desalinated water production cost measurably.

4.2.5.4 Oxygen Concentration of Plant Discharge

Subsurface intake water form many coastal alluvial aquifers, and brackish aquifers has very low dissolved oxygen (DO) concentration. This concentration is usually less than 2 mg/L, and often it varies between 0.2 and 1.5 mg/L. The RO treatment process does not add appreciable amount of DO to the intake water. Therefore, the RO-system product water and concentrate have approximately the same DO concentration as the saline source water. Often low DO concentration of the product water requires either product water reaeration or results in significant use of chlorine.

If the low DO concentrate from desalination plant with subsurface intake is to be discharged to an open water body such as an ocean, lake or a river, this discharge typically would not be in compliance with the United States Environmental Protection Agency's daily average and minimum DO concentration discharge requirements of 4 and 5 mg/L, respectively. Because large desalination plants using subsurface intakes that collect low-DO source water would discharge a significant volume of concentrate with low DO concentration, this discharge could cause oxygen depletion and stress to aquatic life. Therefore, concentrate with low-DO concentration has to be reaerated before surface water discharge.

For large RO plants, the amount of air and energy needed to increase the DO concentration of the discharge from 1 to 4 mg/L is significant and would have a measurable effect on the cost of desalinated water. Discharge of this low DO concentrate to a wastewater treatment plant outfall would also result in an additional power use to aerate this concentrate prior to discharge.

For comparison, concentrate from RO plants with open intakes would typically have DO concentration of 5–8 mg/L, which is adequate for disposal to a surface water body (ocean, brackish lake, or river) without reaeration. Therefore, the use of this type of intake is not likely to pose a challenge to the marine environment surrounding the outfall in terms of DO level in the discharge and pretreatment of the discharge prior to its release will not be needed.

4.2.6 Subsurface Intakes—Environmental Impacts and Mitigation Measures

Subsurface intakes are considered a low-impact technology in terms of impingement and entrainment. However, to date, there are no studies that document the actual level of entrainment reduction that can be achieved by this type of intakes.

Because subsurface intakes naturally filter the collected saline water through the granular formations of the aquifer in which they operate, using this type of intakes minimizes entrainment of aquatic organisms. The saline source water collected by this type of intake typically does not require mechanical screening and, therefore, subsurface intakes do not cause impingement impacts on the aquatic organisms in the area of the intake.

Subsurface intakes could have a number of environmental impacts, such as loss of coastal habitat during construction, and visual and aesthetic impacts, as well as it could affect nearby coastal wetlands depending upon the intake method of construction and design. The environmental impacts and mitigation measures of such intake wells are discussed in greater detail in the following publication— Voutchkov (2013).

4.2.7 Construction Costs of Subsurface Intakes

Table 4.1 provides a summary of the construction costs for the types of subsurface intakes presented in the previous sections. As seen from this table, vertical wells are usually less costly than any other type of subsurface intakes, which is one of the main reasons why vertical well intakes are used most commonly for small desalination plants. Infiltration galleries

TABLE 4.1 Construction Costs of Subsurface Intakes

Subsurface Intake Type	Typical Production Capacity (Yield) of Individual Well (ML/day)	Construction Cost of Individual Well (US$ million)
Vertical wells	0.1–3.5	0.3–2.5
Horizontal Ranney wells	0.5–20	0.8–6.4
HDD wells (i.e., Neodren)	0.1–5.0	0.4–1.8
Infiltration galleries	0.1–50	0.9–42.0

are the costliest type of intake but they allow collecting the largest volume of water per unit surface of coastal beach area and, therefore, they are more attractive choice for larger desalination plants. As indicated previously, infiltration gallery has been used for a 50,000 m^3/day (13.2 MGD) SWRO desalination plant in Japan but has not found wide application because the cost of the intake is over 50% of the cost of the desalination plant. Usually, the construction cost of open intakes and other types of subsurface intakes is 10%–30% of the total construction cost of the desalination plant.

4.3 OPEN INTAKES

Based on the location of their inlet structure, open intakes are typically classified as onshore and offshore. The inlet structure of onshore intakes is constructed on the banks of the source water body while the inlet structure of offshore intakes is typically located several hundred to several thousand meters away from the shore.

4.3.1 Onshore Open Intakes

To date, onshore intakes have found application mainly for very large thermal or hybrid seawater desalination plants. Such intakes typically consist of large and deep intake canal ending into a concrete forebay structure equipped with coarse bar screens followed by fine screens and intake pump station.

Depending on the coastal conditions, onshore intakes could be installed: on a sandy coast with low gradient; on a rocky coast; or in natural or artificial enclosure (i.e., ship turning basin, marina, industrial port or lagoon). Of the three coastal environments, rocky bottom conditions are the most favorable for construction of onshore intakes. The main factors associated with the feasibility and water quality of such intakes are: wind and swell regimes, water level variations, tidal regime, bathymetry, and coastal currents.

Onshore intakes on sandy coast with low gradient are usually constructed with a long entrance canal that is designed to protect the intake from littoral sediment transport by prevailing near-shore currents, winds, swells, and tides. The intake canal is constructed with jetties, which are oriented such that they create a protective shield of the intake against the prevailing current and prevent the littoral drift form carrying sediments into the intake area. Without the jetty the canal would fill up with sand and would need to be dredged frequently. The jetties are constructed of stone rock blocks and usually extend to elevation of 2–3 m (6.6–10 ft) above the mean water level. To avoid fish entrapment, the canal is designed for average velocity at mean water level of 0.3 m/s (1 fps) and at 0.15 m/s (0.5 fps) at high water level. The canal ends into an onshore inlet structure with trash racks.

Rocky-coast onshore intakes typically are concrete or metal structures, which are open directly to the water body. Depending on the site-specific conditions, such intakes may be designed with jetty protection. If possible, it is preferable to design the water entrance at a depth or at least 2.0 m (6.6 ft) below the low tide level and to protect the entrance with wave-breaking jetties. The main difference between rocky bottom and sandy bottom shoreline conditions is that wave action typically does not cause significant stir up of sediments and elevated source water turbidity.

Onshore intakes in natural or artificial enclosures are typically well protected against wave and wind action and therefore have more consistent water quality. However, one such intake location is that the embayment accumulates silt and sand, and unless the area in front of the intake is dredged periodically, the intake capacity decreases over time. Dredging operations have both cost and source water-quality implications.

The largest seawater RO-desalination plant with onshore open intake is the 155,000 m^3/day (40 MGD) Point Lisas desalination plant in Trinidad. In this case the intake consists of concrete forebay structure located on the shore of industrial port's ship turning basin (the Gulf of Paria) and is equipped with coarse and fine screens, and vertical turbine pumps.

Another desalination plant that uses source water from an onshore open intake is the 200,000 m^3/day (54 MGD) Carlsbad desalination plant in California, which is the largest SWRO desalination facility in the US and the Western Hemisphere. This plant collects most of its source water from the discharge of the Encina Power Station with which it is colocated (Fig. 4.9).

4.3.2 Offshore Open Intakes

Offshore open intakes are the most commonly constructed type of intakes for medium and large seawater desalination plants worldwide. These surface water delivery systems include the following key components: offshore intake structure (velocity-cap-type inlet structure);

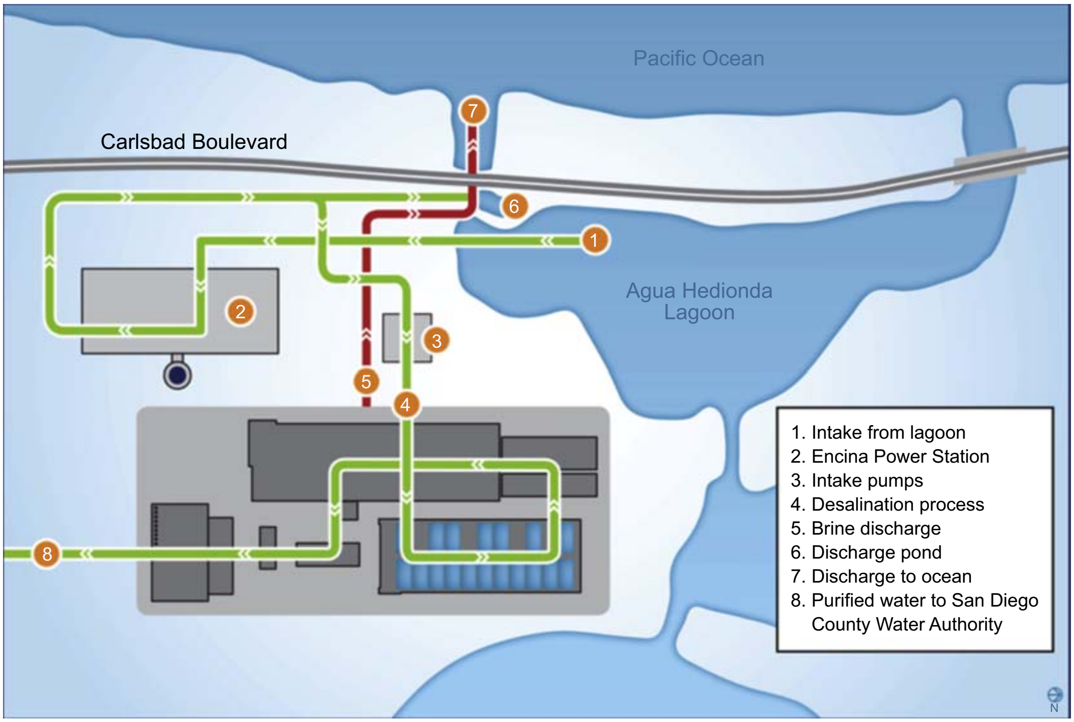


FIGURE 4.9 Segment of 600-mm intake collector screen.

FIGURE 4.10 Steel offshore intake tower.

one or more intake pipeline(s) or intake tunnels; an onshore intake chamber; trash racks; fine screens; source water intake pump station; electrical, instrumentation, and control equipment; and chemical feed equipment.

The inlet structure of offshore open intakes is usually a vertical concrete, copper–nickel or steel well (vault) (Fig. 4.10) or wedge wire screen (Fig. 4.11) located 4–10 m (13–33 ft) above the floor of the water body and submerged between 4 and 20 m (13 and 66 ft) below the water surface, which is designed to reliably collect adequate amount of seawater that has a minimum content of debris and aquatic organisms.

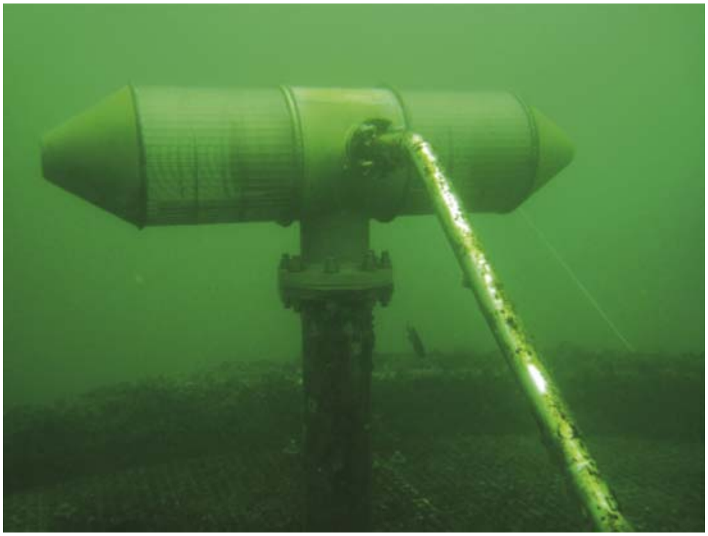


FIGURE 4.11 Copper–nickel intake wedge wire screen.

Typically, the offshore open-ocean intake structure is located several hundred to several thousand meters from the shore. The best location of the offshore intake structure in terms of source water quality is at ocean floor depths of 20 m (66 ft) or higher (deep water intake). Debris load in the source water and algal content during red tides at such depths are typically several times lower than that in the surface water or the shallow waters of the tidally influenced near-shore area.

Depending on the plant location and ocean floor formation, installing the intake structure at 20 m (66 ft) depth may require intake pipeline, which is between 200 and 5000 m (1320 and 16,500 ft) long. Because the construction cost for intake pipeline located on the ocean floor is usually very high (between 4 and 10 times higher than the cost of the same size pipe installed inland in the ground), the intake water quality benefits of locating the offshore intake structure in deep waters have to be compared against the costs for construction of the intake structure and pipeline and the cost of the source water pretreatment system.

The best location for an open offshore intake from a lifecycle cost point of view is typically such site where the ocean floor depth of 20 m (66 ft) or more can be reached within 500 m (1650 ft) from the shoreline. If such ocean floor location is not available within a reasonably close vicinity of the RO desalination plant, usually it is more cost-effective to collect source water of inferior water quality and build a more elaborate pretreatment system, than to install a costly offshore intake structure and a long intake pipeline.

Because of the high costs of deep intake structures and long pipelines, many existing SWRO desalination plants with open-ocean offshore intakes are located in shallow near-shore areas where the ocean floor depth is typically between 4 and 8 m (13–26 ft)—such intakes are also referred to as shallow water intakes. As a result, plants with shallow water intakes usually have source water with high content of debris, solids, and aquatic organisms, which requires elaborate pretreatment prior to RO-membrane separation.

Offshore intakes that usually extend several hundred meters away from the shoreline and 10–20 m (33–66 ft) below the water surface are typically not influenced by freshwater from surface runoff. Therefore, such intakes usually yield saline water of the same TDS content as that of ambient water.

One exception is offshore intakes located near the entrance of large rivers or other freshwater bodies into the saline water body (ocean, sea, brackish lake). Since old river beds and associated alluvial aquifers could often extend far beyond the tidal zone, the water quality of such offshore intakes may be influenced by the groundwater quality in the alluvial aquifers, which often is inferior than the open-ocean seawater in terms of content of iron, manganese, and other undesirable contaminants.

4.3.3 Colocated Intakes

The colocated desalination plant intakes are directly connected to the discharge outfall of an adjacently located coastal power plant using seawater for once-through cooling. This configuration allows using the power plant cooling water both as source water for the desalination plant and as a blending water to reduce the salinity of the desalination plant concentrate prior to the discharge to the ocean (see Fig. 4.12).

In seawater-cooled coastal power plants, the seawater enters the power plant intake facilities and after screening is pumped through the power plant condensers to cool them and thereby to remove the waste heat generated during the electricity generation process. The

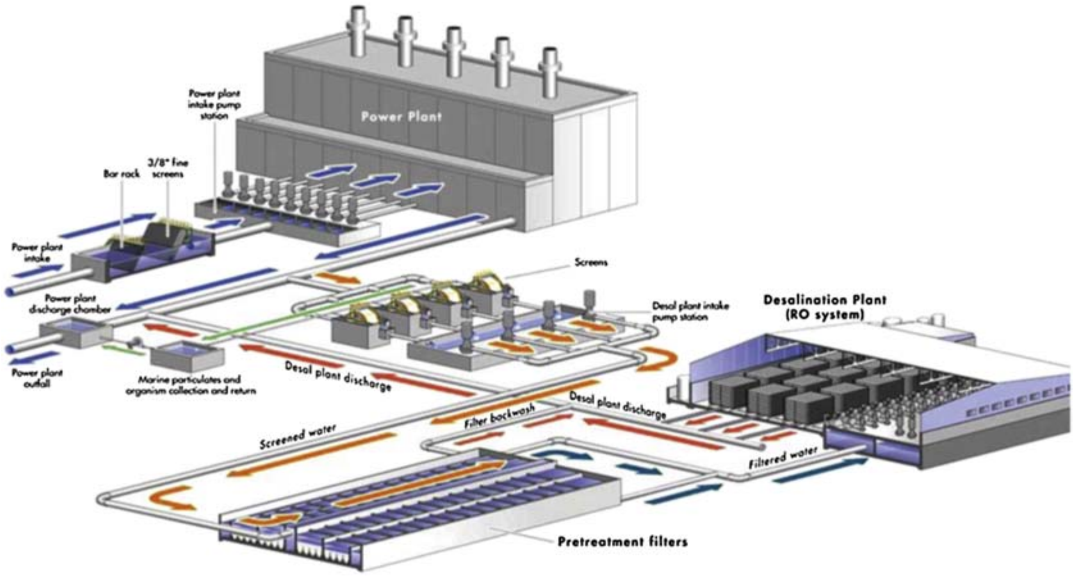


FIGURE 4.12 Colocation of desalination plant and power generation facility.

cooling water discharged from the condensers typically is 5–15°C warmer than the source ocean water and is usually conveyed to the ocean via a separate discharge outfall.

Under the collocation concept, the intake of the seawater desalination plant is connected to the discharge outfall of the power plant to collect a portion of the cooling water for desalination. After the desalination plant source seawater is pretreated, it is processed in an RO-membrane desalination system, which produces two key streams—low salinity permeate, which after conditioning is conveyed for potable water supply, and concentrate, whose salinity is typically two times higher than the source seawater. The desalination plant concentrate is conveyed to the power plant discharge-outfall downstream of the point of desalination plant intake connection. As a result the desalination plant does not have its own direct intake of ambient seawater and outfall to discharge concentrate; it uses the existing intake and outfall of the power plant with which it is collocated.

Usually, coastal power plants with once-trough cooling systems use large volumes of seawater. Because the power plant intake seawater has to pass through the small diameter tubes [typically 10-mm/(3/8-in.) or less] of the plant condensers to cool them, the plant discharge cooling water is already screened through bar racks and fine screens similar to these used at surface water intake desalination plants. Therefore, a desalination plant whose intake is connected to the discharge outfall of a power plant usually does not require the construction of a separate intake structure, intake pipeline, and screening facilities (bar-racks and fine screens). Since the construction cost of a new surface water intake structure and discharge outfall for a desalination plant is typically 20%–30% of the total plant construction expenditure, power plant colocation could yield significant construction-cost savings.

At present, this type of intake configuration has found application mainly for seawater desalination facilities co-sited with once-though cooling coastal power generation stations. However, such configuration could also be applied for inland brackish water desalination plants where the host power plant uses brackish intake and discharge wells for once-though cooling.

Sharing intake infrastructure also has environmental benefits because it avoids the need for new construction in the open water body area near the desalination plant. The construction of a separate new open-intake structure and pipeline for the desalination plant could cause significant disturbance of the benthic marine organisms inhabiting the water body.

Another clear environmental benefit of the colocation of power and desalination plants is the reduced overall impingement and entrapment of marine organisms as compared to construction of two separate open-intake structures—one for the power plant and one for the desalination plant. This benefit stems from the fact that total biomass of the impacted marine organisms is typically proportional to the volume of the intake saline source water. By using the same intake saline water twice (once for cooling and second time for desalination) the net intake inflow of saline water and aquatic organisms is minimized.

The colocation configuration has a number of discharge benefits: (1) the construction of a separate desalination plant outfall structure is avoided, thereby decreasing the overall costs for seawater desalination; (2) the salinity of the desalination plant discharge is reduced as a result of the mixing and dilution of the membrane concentrate with the power plant discharge, which has ambient seawater salinity; (3) because a portion of the discharge water is converted to potable water, the total amount of the power plant thermal discharge is reduced, which in turn lessens the negative effect of the power plant thermal discharge on the aquatic environment; (4) the blending of the desalination plant and the power plant discharges results in accelerated dissipation of both the salinity and the thermal discharges.

4.3.4 Selection of Open-Intake Type

Onshore intakes have one key advantage, that is, they are usually the lowest cost type of intake, especially for large desalination plants. However, such intakes typically produce the worst source water quality because in most cases they are designed to collect water from the entire depth of the water column and because they are located in the surf zone where breaking waves continuously lift particles from the bottom into suspension, thereby significantly increasing water turbidity as compared to deeper waters.

The first 8–10 m (26–33 ft) of the water column in the surf zone typically have several times higher levels of turbidity, algae, hydrocarbon contamination, silt, and organics than deeper waters. The quality of water collected from this upper zone of the water body could vary in a wide range because of the influence from wind, tides, currents, ship traffic, storms, and surface freshwater runoff. Onshore intakes could also be exposed to beach erosion and direct wave action with irreversibly damaging consequences.

Because onshore open intakes usually yield source water of worse quality than offshore intakes, desalination plants with such onshore intakes have complex multistage pretreatment system. Practical experience shows that in most full-scale desalination projects, the savings of lower cost onshore open intakes are negated by the higher expenditures needed for more elaborated source water pretreatment and more conservative RO-system design.

Due to the lower quality source water they collect, onshore open intakes have found very limited application for RO desalination plants, and unless the specific site location or costs dictate the need to use this type of intake, they are not used for such plants. However, onshore intakes are often the prime choice for thermal desalination plants, where source water quality in terms of suspended solids, organic, and algal content is of secondary importance and has minimal influence on the desalination process.

4.3.5 Feasibility Considerations for Open Intakes

4.3.5.1 Intake Site Location and Configuration

Key factors that determine the selection of the location of open intakes are: potential for beach erosion in the intake area; location and direction of underwater currents; presence and location of active seismic faults; topography and geology of the floor of the water body; location of environmentally sensitive habitats along the intake pipe route and near the intake inlet; location and size of municipal and industrial wastewater discharges within 1 km (0.6 miles) radius from the intake; size of waves and depth of wave impacts; ship and boat traffic; tide and wind characteristics in the intake area.

Open intakes should be located away from coastal areas of active beach erosion and seismic faults; high waves; strong underwater currents carrying debris, silt, plankton, sea grass, weeds, and other stringy materials; and locations with heavy ship and boat traffic. If sensitive marine habitats are encountered, the intake route should either be modified or the intake conduit/s should be installed via directional drilling or tunneling under the sensitive area instead of using open trench construction or laying the conduit on the bottom of the water body.

The proper design of open-ocean intakes requires the collection of detailed source water quality data from the proposed site of the intake, characterization of aquatic life in the vicinity of the intake and completion of detailed sanitary survey assessing the potential sources of SWRO plant source water quality contamination in the vicinity of the intake location (such as waste discharges of industrial and municipal wastewater plants, stormwater discharges, or large industrial port or municipal marina activities, which may result in oil and gasoline spills and in other ocean water contamination).

The exact location and depth of the offshore intake structure must be determined based on a hydrological study to ensure that the intake will be adequately submerged at low tide; protected from the damaging orbital storm wave motion; and will be far enough offshore to avoid the near-shore sediment transport zone where storms can cause suspension or deposition of large quantities of silt and sediment, and can ultimately damage the intake structure and the interconnecting piping.

Diurnal and seasonal source water quality fluctuations should also be considered when determining the location of the intake structure. At minimum low tide conditions, inlet mouth should be submerged at least 3 m (10 ft) below the water surface. In addition, the distance between the inlet mouth and the ocean floor should be no less than 3 m (10 ft) to prevent excessive sand carryover into the downstream intake facilities. The intake water supply can be protected against large aquatic organisms and large floating debris by installation of wire net across the intake mouth.

The proper design of open-ocean intakes requires the collection of detailed source water quality data from the proposed site of the intake, characterization of aquatic life in the vicinity of the intake and completion of detailed sanitary survey assessing the potential sources of RO-plant source water-quality contamination in the vicinity of the intake location (such as waste discharges of industrial and municipal wastewater plants, stormwater discharges or large industrial port or municipal marina activities, which may result in oil and gasoline spills and in other ocean water contamination).

4.3.5.2 Inlet Screens

Inlet screens are typically coarse bar screens with distance between the bars of 50–300 mm (2–12 in.). The bar length is usually between 1.0 and 3.0 m (3.3–9.9 ft) and is determined based on the selected maximum distance between the screen velocity cap and the water surface and the bottom.

Design maximum through screen velocity varies depending on the content of jellyfish in the source water and is usually in a range of 0.10–0.15 m/s (0.3–0.5 fps). It is important to note that the design through-screen velocity should be calculated for 50% of the total area between the screen bars to allow for growth of shellfish and accumulation of debris over time. It is anticipated that the inlet screening area will be reduced by 30%–50% every 18–24 months and, therefore, the screens would need to be cleaned periodically by divers.

A maximum through-screen velocity of 0.10 m/s or less is typically selected if the source water contains jellyfish load of 0.5 kg/m^3 of source water or more. This lower flow through-screen velocity allows to minimize the intake of jellyfish into the plant and thereby to minimize their negative impact on downstream screening facilities. Jellyfish outbreaks are typical for warmer, shallower, and polluted waters, especially if the intake is located in the middle of an underwater current though which jellyfish travel most frequently.

Usually, small jellyfish specimen that can enter the intake weigh between 200 and 400 g (0.9 lbs) per individual and for large plants with few fine screens jellyfish could add significant load to the screens, which in extreme conditions could cause equipment damage. In addition, jellyfish are very difficult to clean from the screens and often standard debris jet sprays with which fine screens are equipped cannot remove them. From this perspective, drum screens are usually more difficult to clean once the jellyfish attach on the screens and therefore, reducing their intake by reducing the inlet through-screen velocity is of critical importance, especially for plants with fine drum screens. If large quantities of jellyfish enter the intake, they could render the fine screens inoperable or reduce significantly plant source water flow rate.

4.3.5.3 Inlet Materials

Inlet screens/towers are typically built from corrosion-resistant materials such as copper–nickel alloy, concrete, or stainless steel. The screen bars are either made of 90/10 Cu/Ni or super duplex stainless steel for seawater applications and duplex stainless steel for brackish water intakes.

4.3.6 Feasibility Considerations for Colocated Intakes

For the collocation concept to be cost-effective and possible to implement, the power plant cooling-water discharge flow has to be larger than the capacity of the desalination plant with which it is collocated, and the power plant outfall configuration has to be adequate to avoid entrainment and recirculation of concentrate into the desalination plant intake. It is preferable that the length of the power plant-outfall downstream of the point of connection of the desalination plant discharge is adequate to achieve complete mixing prior to the point of entrance into the ocean.

A special consideration has to be given to the effect of the power plant operations on the cooling water quality, since this discharge is used as source water for the desalination plant. For example, if the power plant discharge contains levels of copper, nickel, or iron significantly higher than these of the ambient seawater, this power plant discharge may not be suitable for collocation because these metals may cause irreversible fouling of the membrane elements. In such cases, the pretreatment costs for colocated desalination plant are likely to be higher than those for plant with separate open-ocean intake.

Another potential problem could be the location of the disposal of the power plant intake screenings. In most power plants the screening debris are removed from the intake cooling water and disposed off-site. However, this disposal practice may change during the course of the power plant and desalination plant operations.

For example, in the case of the Tampa Bay seawater desalination plant, during the final phase of the desalination plant construction, the power plant decided to change their screenings' disposal practices and to discharge their intake screenings just upstream of the desalination plant intake rather than to continue disposing them off-site. This change in power plant operations had a dramatic effect on the Tampa Bay Water desalination plant start-up and operations, and especially on the pretreatment system performance.

Since the desalination plant was pilot tested and designed around the original method of power plant operations under which all screenings were removed from the cooling water, the desalination plant was not built with its own separate intake screening facilities. The presence of power plant-waste screenings in the desalination plant intake water had a detrimental effect of the pretreatment filter operations because the screening debris frequently clogged the filter distribution piping, airlifts, and sand media.

Although this problem has a significant effect on the desalination plant operations, it also has relatively straightforward solutions—either installing separate fine screening facilities for the desalination plant or moving the point of the power plant screening debris-discharge downstream of the desalination plant intake. A summary of key advantages and disadvantages of the collocation approach is presented in Table 4.2.

4.3.7 Open Intakes—Environmental Impacts and Mitigation Measures

Use of open-ocean intake for collecting saline source water for the desalination plant would result in some entrainment of aquatic organisms (WRA, 2011) as compared to subsurface intakes because they take saline source water directly from the water column of the saline water body (e.g., ocean, sea), rather than source water which is prefiltered through the soils of the coastal aquifer. In addition, some aquatic organisms may be impinged on the screening facilities.

Impingement and entrainment of aquatic organisms is not an exception for seawater intakes. The same phenomena and extend of environmental impact may be observed at the intakes of conventional water treatment plants collecting fresh surface water from rivers, lakes, and

TABLE 4.2 Key Advantages and Disadvantages of Desalination Plant Collocation

Advantages	Disadvantages
• Capital cost savings by avoiding the construction of separate intake pipeline and structure, and new discharge outfall. • Decrease of the required RO system feed pressure and power cost savings as a result of using warmer water. • Reduction of unit power cost by connecting directly to power plant generation facilities and avoiding power transmission charges. • Accelerated permitting process as a result of avoidance of construction of new intake and discharge outfalls in the ocean. • Reduction of marine organism impingement and entrainment because the desalination plant does not take additional seawater from the ocean. • Reduction of the impact on marine environment as a result of faster dissipation of thermal plume and concentrate. • Reduction of the power plant thermal discharge to the ocean because a portion of this discharge is converted to potable water. • Use of already disturbed land at the power plant minimizes environmental impact.	• Use of warmer seawater may accelerate membrane biofouling, especially if the source water is rich in organics. • RO membranes may be exposed to iron, copper, or nickel fouling if the power plant condensers and piping are built of low-quality materials. • Source seawater has to be cooled if its temperature increases above 40°C to protect RO-membrane integrity. • Permeate water quality diminishes slightly with the increase of source water temperature. • Use of warmer water may result in lower boron rejection and require feed water pH adjustment to meet stringent boron-water quality targets. • RO-plant source water screening may be required if the power plant disposes off its screenings through their outfall and the point of disposal is upstream of the desalination plant intake. • Desalination plant operations may need to be discontinued during periods of heat treatment of the power plant facilities.

estuaries as well. The impingement and entrainment effects of open intakes may be mitigated significantly if the source water velocity of the intakes is reduced below 0.15 m/s (0.5 fps).

4.3.8 Construction Costs of Open Intakes

4.3.8.1 Costs of Onshore Intakes

Fig. 4.13 depicts the construction costs of onshore open intakes as a function of the desalination plant intake flow. Because of the significant impact of the site-specific conditions of the actual intake costs, such costs may vary within ±30% of the values indicated on this figure.

4.3.8.2 Costs of Offshore Intakes

The construction costs for intake systems with offshore inlet structures and HDPE pipelines, and with concrete tunnels are depicted on Fig. 4.14. These costs are presented as a function of the plant intake flow and are expressed in US $ per meter of intake conduit. Similar to the onshore intake construction costs, the costs for these types of intakes will vary from one location to another and will be influenced by site-specific conditions such as water depth, geology, and currents, and they could be within a 30% envelope of the values indicated on Fig. 4.14. Analysis of this figure indicates that the construction of desalination plant intakes with deep tunnels is typically several times more costly than the installation of HDPE pipeline on the bottom of the source water body.

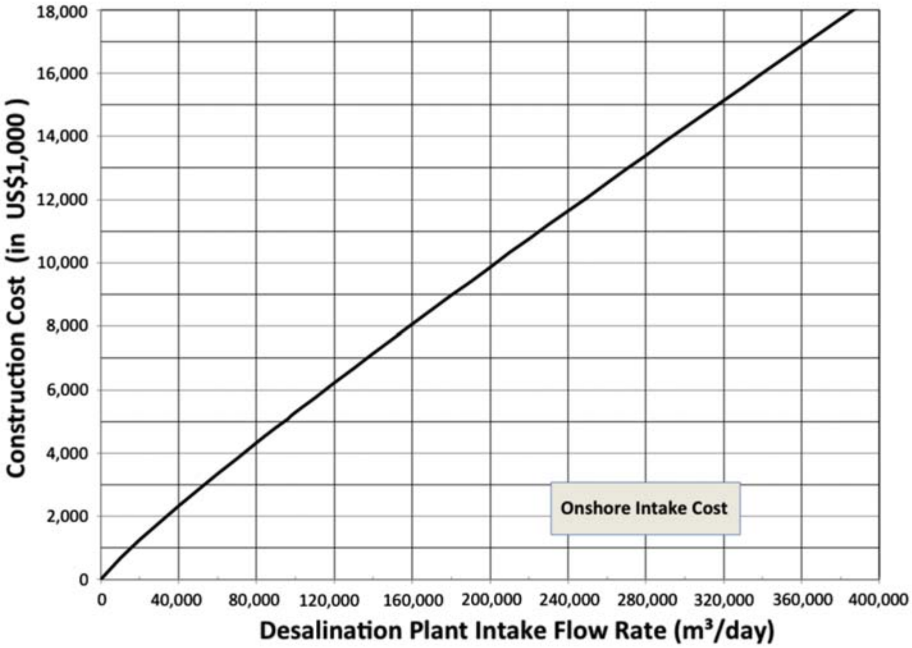


FIGURE 4.13 Onshore intakes—construction costs.

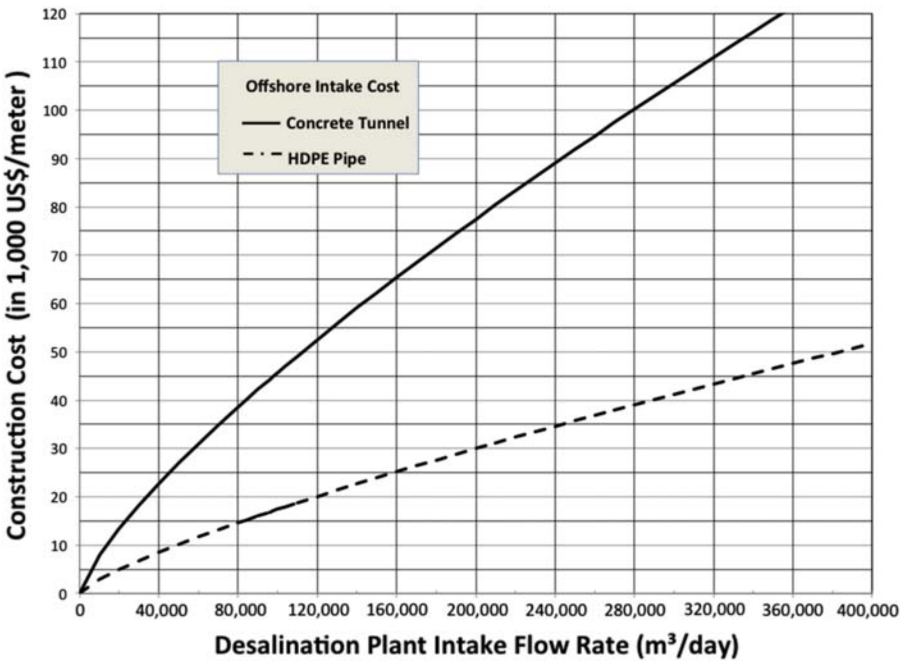


FIGURE 4.14 Offshore intakes—construction costs.

References

David, B., Pinot, J.P., Morrillon, M., 2009. Beach wells for large-scale reverse osmosis plants: the Sur case study. In: Proceedings of IDA World Congress, Atlantis, the Palm-Dubai, UAE, November 7−12, 2009. REF: IDAWC/DB09-106.

Henthorne, L., Boysen, B., 2015. State-of-the-art of reverse osmosis desalination pretreatment. Desalination 356, 129−139. Elsevier.

Kartinen Jr., E., Martin, C., 2003. Solving Morro Bay's seawater reverse osmosis plant's iron problem. In: Proceedings of AWWA Membrane Technology Conference, Atlanta, Georgia, 2003.

Missimer, T.M., Maliva, R.G., Thompson, M., Manahan, W.S., Goodboy, K.P., 2010. Reduction of seawater reverse osmosis treatment costs by improvement of raw water quality: innovative intake designs. Desalination and Water Reuse 20 (3), 12−22.

Peters, T., Pinto, D., 2008. Seawater intake and pretreatment/brine discharge − environmental issues. Desalination 221, 576−584. Elsevier.

Peters, T., Pinto, D., Pinto, E., 2007. Improved seawater intake and pre-treatment system based on Neodren technology. Desalination 203, 134−140. Elsevier.

Roscoe Moss Company, 2012. A Guide to Water Well Casting and Screen Selection. Los Angeles, California.

Rovel, J.M., 2002. Current and future trends in SWRO (pretreatment). In: Proceedings of IDA World Congress, Bahrain, November, 2002.

Voutchkov, N., 2013. Desalination Engineering Planning and Design. McGraw Hill.

WRA, 2011. Desalination Plant Intakes − Impingement and Entrainment Impacts & Solutions, White Paper. WateReuse Association (WRA), Alexandria, Virginia.

CHAPTER

5

Pretreatment by Screening

OUTLINE

5.1 INTRODUCTION

Screens are the first pretreatment step of every desalination plant. Depending on their type and use, they are incorporated into the intake system and/or are located downstream of the intake and upstream of the reverse osmosis (RO) system. The main purpose of screens is to remove large solid particles contained in the saline source water to protect the other pretreatment facilities or the RO system from equipment and structural damage, accelerated filter media clogging and fouling, and accelerated reduction of product water treatment capacity.

Depending on the type of intake and the quality of the saline source water it collects, screening equipment may be as simple as cartridge or bag filters, or as sophisticated as series of mechanical screens of various sizes designed to sequentially remove large debris and

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aquatic organisms, as well as microscreens that aim to retain finer solids such as silt, plankton, sand, shell particles, and other debris in the saline source water.

Open ocean intakes are typically equipped with coarse bar screens followed by smaller size ("fine") screens with openings of 1–10 mm (0.04–0.4 in.), which preclude the majority of the adult and juvenile aquatic organisms (fish, crabs, etc.) from entering the desalination plant. While coarse screens are always stationary, fine screens could be two types—stationary (passive) and periodically moving (i.e., rotating) screens.

The saline source water collected by open intakes always contains solid particles and aquatic organisms, which naturally exist in the ambient aquatic environment. Most of the particles and organisms larger than 20 μm are removed by screening and downstream filtration before this water enters the RO desalination membranes for salt separation. After screening, the source water is typically processed by finer pretreatment filters, which usually have sizes of the filtration media openings (pores) between 0.01 and 0.04 μm for membrane ultra- and microfilters and 0.25–1.2 mm for granular media filters.

If the saline source water is collected using subsurface intake, the screening of most of the coarse particles and aquatic organisms is completed by the natural soil layer of the intake system and, therefore, the desalination plant screening facilities are simplified and often only incorporate cartridge or bag filters located upstream of the RO membrane system. It should be pointed out, however, that subsurface intakes are not an absolute screening barrier, and their ability to provide efficient screening is dependent on the type and thickness of the soil formations through which the saline source water travels before it reaches the intake collection system (e.g., well screens).

5.2 BAR, BAND, AND DRUM SCREENS

A typical surface water intake system for medium and large membrane desalination plants with open intakes includes a set of manually cleaned bar racks followed by automated traveling fine bar screens or fine-mesh band or drum screens.

5.2.1 Coarse Bar Screens (Bar Racks)

Bar racks usually have a distance between bars of 50–150 mm (2–6-in.) and their purpose is to retain large-size debris and aquatic life from entering the plant intake. For offshore intakes, the screens are installed on the intake's vertical tower (see Fig. 5.1).

The design flow-through velocity for clean screens is usually 0.10 m/s (0.33 fps). Such low design velocity is not only selected to minimize impingement of aquatic life on the screens but to also account for the loss of flow-through surface as a result of growth on shellfish and accumulation of debris on the surface of the coarse bars.

Typically, the screen bars are manufactured of super duplex steel or copper-nickel alloys (the latter are preferred) to suppress marine growth. Shellfish growth on the screens could narrow the open space between the bars with over 50% and as a result either the bars have to be cleaned manually every several years, or the through-screen velocity is increased from 0.10 cm/s (0.33 fps) to between 0.12 and 0.15 m/s (0.40–0.50 fps) within 1 to 2 years from installation (cleaning).

FIGURE 5.1 Vertical intake tower with coarse screen.

The on-shore screens (Fig. 5.2) are usually equipped with automated raking mechanism, which periodically removes the accumulated debris and allows maintaining the open space between the bars at approximately the same flow-through velocity over time. Therefore, the design flow-thorough velocity of these screens is higher than that of the coarse screens of open intake towers, that is, 0.12–0.15 m/s (0.40–0.50 fps). It is important to maintain the intake through-screen velocity below 0.15 m/s (0.5 fps) at all times to minimize impingement of aquatic organisms on the screens.

5.2.2 Fine Screens

5.2.2.1 Rotating Screens

Fine self-cleaning mechanical bar screens typically have 3–10 mm openings. They are installed vertically in water intake channels downstream of the coarse screens and are equipped with rotating cleaning equipment often combined with water-spraying nozzles to remove the debris from the screen surface. These nozzles periodically spray wash water supplied by pumps sized for 45–68 m^3/h (200–300 gpm) with pressures of 4–7 bars (60–100 psi). Because one of the main functions of the fine screens is to protect the intake pumps from damage, the actual screen openings should be selected to be smaller than the distance between the intake pump impellers.

FIGURE 5.2 Coarse screens of onshore intake.

Two types of rotating fine screens that have found the widest application in desalination plants are band and drum screens. Typically, band screens are installed at smaller and medium-size plants with intake capacity, while drum screens have found wider implementation for some of the largest size desalination facilities worldwide. Fine bar screens are sometimes also used in open intakes but they are not as common and band and drum screens.

5.2.2.2 Band Screens

These vertical traveling screens consist of individual screening panels with fine mesh openings, which are attached on support roller chains, and which in turn are installed on metal-framed guide tracks (see Fig. 5.3).

As the intake source water travels through the screens, the debris contained in the water are removed and accumulated on the screen panels. The screen panel mesh is typically made

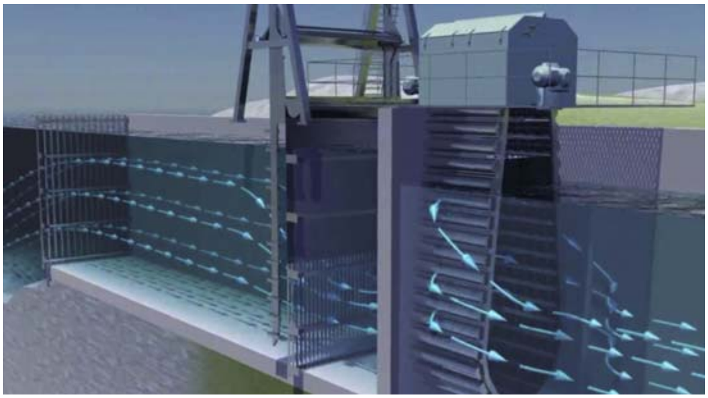

FIGURE 5.3 Intake structure with bar rack followed by band screen.

of polyamide, polyester, or super duplex stainless steel for seawater applications and duplex stainless steel for brackish water intakes. The debris accumulation causes gradual increase in screen headlosses. Once the headlosses reach a certain preset level (or based on a preset time), the screening panels are rotated upwards the debris collected on the panels are moved to the deck level where they are removed from the panels into collection troughs by low-pressure water sprays. The debris is either conveyed to collection bins and disposed as solid waste or is recycled back to the source water body.

Typically, the band screens are designed to enter cleaning cycle at 0.10–0.20 m (0.33–0.66 ft) water elevation differential for a 30% reduction of the screening area. Most commercially available band screens travel at velocities of 2–10 m/min (6.6–33 ft/min). These screens are typically designed for two-speed or variable speed operation. The design screen area efficiency factor for band screens is usually 0.5–0.6 (i.e., 50%–60% of the screening area is active filtration area used to determine through-screen velocity). These fine screens are designed to maintain through-screen velocity below 0.15 m/s (0.5 fps) at all times and normally operate at velocities of 0.06–0.10 m/s (0.18–0.33 fps). The band screens could be configured in three flow patterns—through-flow, center-flow, and dual-flow (see Fig. 5.4).

While the through-flow pattern is commonly used, its main disadvantage is that if the backwash spray does not effectively remove the screenings from the surface of the panels, these screenings will be conveyed to the back (clean) side of the screen and released into the screened water.

In the center-flow configuration the screen panels are oriented in parallel with the flow—the water enters in the space between the two screens and passes out through both sides of the screen simultaneously. This configuration allows retaining all debris on the inner side of the screens, thereby addressing the main challenge associated with the through-flow pattern. The key disadvantage of this configuration is that it produces divergent turbulent flow, which is not favorable for the intake pump hydraulics. In the dual-flow pattern the feed water enters from both outer sides of the screen and is collected inside the screen, which improves channel-flow hydraulics while retaining solids only on the outer side of the screen.

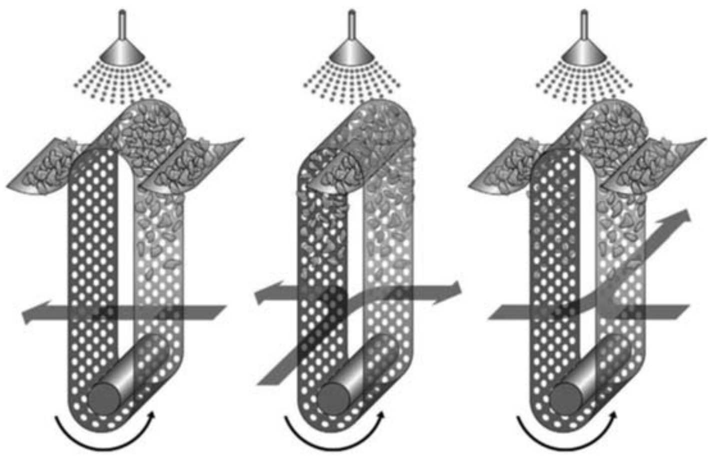

FIGURE 5.4 Through-flow, center-flow, and dual-flow screen patterns (left to right).

Despite the benefits of the center-flow and dual-flow screen patterns, the most widely used configuration is that of flow-through pattern, mainly because of the higher construction and installation complexity of the two other configurations.

Newer installations are more commonly designed in dual-flow in-to-out configuration. An example of the application of such screens is the 300,000 m^3/day Adelaide desalination plant, where three units (each capable to treat 100% of the intake flow of 624,600 m^3/day−165 MGD) are installed in individual channels. The screens are of effective width of 2.8 m (9.2 ft) and have mesh openings of 3.0 mm (0.1 in.) and 50 mesh panels. The screen material is super duplex stainless steel.

Fine-mesh screens are modified band screens, which use finer screening panels—with screen openings of 0.5−1.0 mm (0.02−0.04 in.) and which sometimes are equipped with buckets, that allow capturing fish and other aquatic organisms. Such screens have been found to reduce significantly entrainment of aquatic organisms and sometimes are installed downstream of the conventional size band screens. Such is the case of the Tampa Electric Power Company power-plant intake, which also serves as a cold-water intake for the Tampa Bay seawater desalination plant. The fish and other organisms captured in the screen buckets are conveyed to the source water through a low pressure/low speed pump system.

Besides the fine-mesh screens, there are other modified traveling band screens that have been specifically designed to reduce impingement and entrainment of marine species. Such screening technologies are discussed in greater detail in the following publication (Mackey et al., 2011).

5.2.2.3 *Drum Screens*

Drum screens have found wide application for intakes of large seawater desalination plants in Australia, the Middle East, and Europe. These screens consist of rotating cylindrical frame covered with wire-mesh fabric. This frame is located in a screen structure and supported by a horizontal center-shaft, which rotates slowly on roller bearings. The screens are rotated by a drive located at shaft level.

The most commonly used water pattern for such screens is "in-to-out" (also referred to as "double-entry")—the source water enters the inner side of the cylinder and moves radially out, creating hydraulically beneficial converging flow pattern. Debris deposited on the inner surface of the screens are removed by a jet water spray located on the top of the screen and collected in a water trough—see Fig. 5.5.

Drum screens have unit capacities of up to 3,000,000 m^3/day (270 MGD). Similar to band screens, they are also available in single entry, double exit out-to-in and in-to-out configurations, as well as double entry-single exit (out-to-in) flow pattern (Rogers, 2009).

Drum screens are more advantageous for applications where the source water debris and materials may fluctuate significantly because they are less susceptible to overtorque due to the large influx of solids to the screen over short period of time, which could occur in on-shore open intakes or shallow off-shore intakes. In addition, drum screens typically create lower flow-through head-losses at the same flow.

Besides hydraulic loading, drum- and band screens are also frequently designed based on solids' load—especially for jellyfish outbreaks—when the amount of these marine organisms in the water could exceed 300 tons/h. When jellyfish outbreaks occur, they can completely blind the screens and the removal of the jellyfish from the screen mesh is very cumbersome. From that prospective, the manual removal (scrubbing) of jellyfish from the screens is usually

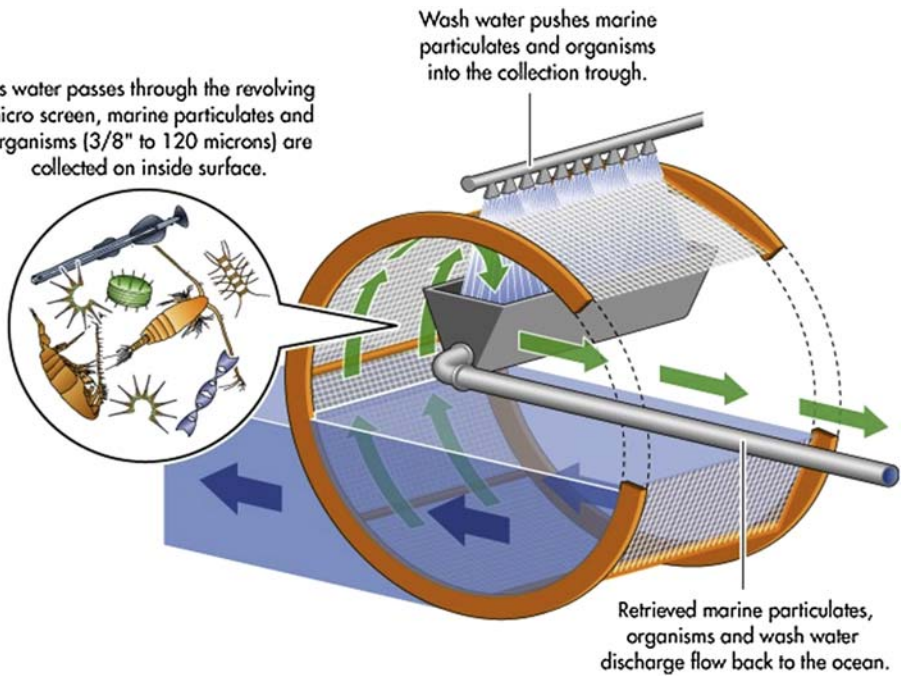

FIGURE 5.5 Drum screens.

easier to do for band screens than for drum screens. If drum screens, however, are designed to handle large debris/jellyfish loads, they can perform with minimum maintenance during jellyfish outbreak episodes.

One important difference between drum screens and band screens is that the latter have 2.5–3.0 times smaller overall footprint. Therefore, if space is at premium, band screens may be a preferred option. In addition, band screens are usually 30%–40% less costly. However, they typically have higher maintenance (mainly equipment) costs.

5.2.2.4 *Wedge-Wire Screens*

Wedge-wire screens are passive screening equipment located offshore, which is directly connected to the suction end of the intake pump station, thereby eliminating the need for additional coarse- or fine-screening facilities. These screens do not have any mechanical moving parts and, therefore, are also referred to as "passive" screening technology. The desalination plant with the largest size wedge-wire screen intake in operation at present is the 150,000 m^3/day (40 MGD) Beckton plant in UK (see Fig. 5.6). The wedge-wire screens are located at 3 m (10 ft) above the bottom. They are made of copper-nickel alloy and have 3-mm openings. The screen flow-through velocity is 0.15 m/s (0.5 fps)—(Moore et al., 2009).

Wedge-wire screens are cylindrical metal screens with trapezoidal-shaped "wedge-wire" slots with openings of 0.5–10 mm. The screen size most commonly used for desalination plants at present is 3 mm.

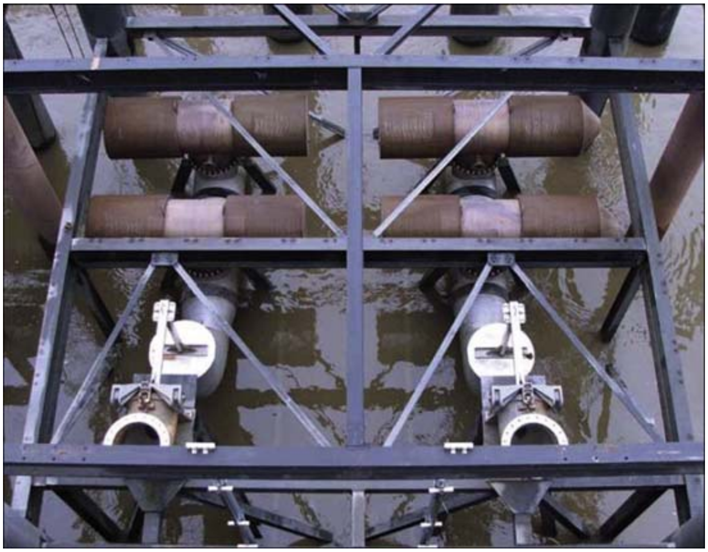

FIGURE 5.6 Wedge-wire screens of Beckton desalination plant, London, UK.

Wedg-wire screens combine very low flow-through velocities (10–15 cm/s/0.3–0.5 ft/s), small slot size, and naturally occurring high screen-surface sweeping velocities to minimize impingement and entrainment. These screens are designed to be placed in a water body where significant prevailing ambient cross-flow current velocities [$\geq$0.3 m/s (1.0 fps)] occur over 90% of the time. This high cross-flow velocity allows organisms that would otherwise be impinged on the wedge-wire intake, to be carried away with the flow. Therefore, wedge-wire screens are considered by the USEPA best technology available for impingement and entrainment reduction (WRA, 2011).

An integral part of a typical wedge-wire screen system is an air burst back-flush system, which directs a charge of compressed air to each screen unit to blow off debris back into the water body, where they are carried away from the screen unit by the ambient cross-flow currents.

The screens will need to be installed at a minimum of 1 m (3.3 ft) from the bottom to avoid entrance of sand and silt into the screens. If the intake is located in relatively shallow, tidally influenced area the depth from the bottom is recommended to be increased to a minimum of 2.0 m (6.6 ft) to prevent the entrance of bottom sediments in the intake water.

Typically, the material used for such screens is copper-nickel alloy (Cu/Ni 90/10), supper duplex stainless steel, or titanium. Copper-nicker alloys usually offer optimum combination between reasonable costs and corrosion and erosion resistance.

5.2.3 Construction Costs of Drum- and Band Screens

The graphs presented in Fig. 5.7 provide budgetary-level construction cost estimates for band- and drum screens as a function of the desalination plant feed-water flow. As shown

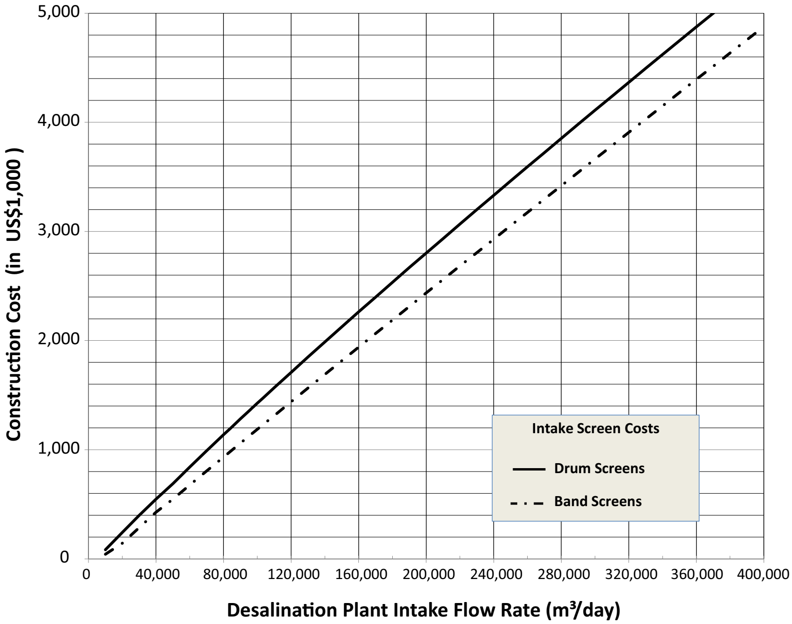

FIGURE 5.7 Construction costs for drum- and band screens.

on this graph in general, band screens are less costly than drum screens for the same size project. Fig. 5.8 presents graph indicating order-of-magnitude costs of wedge-wire screens as a function of the plant source water flow.

5.3 MICROSCREENS

5.3.1 Types and Configurations

If the filtration pretreatment system selected for the RO desalination plant is of membrane type, the types of fine screens described in the previous sections do not provide sufficiently effective removal of source water particles to protect the integrity of the membrane pretreatment system. Typically, microscreens—microstrainers (Fig. 5.9) or disk filters (Fig. 5.10)—could be used for this application.

Most self-cleaning microstrainers (Fig. 5.9) consist of screen with small openings located in filtration chamber. The source water enters on the inner side of the strainers, moves radially out through the screens, and exits through the outlet. The gradual buildup of solids on the inner surface of the screens creates a filter cake, which increases the differential pressure between intake and filtered water over time. When the differential pressure reaches certain preset value, the deposited solids are removed by jet of backwash water. The self-cleaning process typically takes 30–40 s.

The disk filters (Fig. 5.10) are equipped with polypropylene discs, which are diagonally grooved on both sides to a specific micron size. A series of these discs are stacked and compressed on a specially designed spine. The groove on the top of the disks runs opposite to the

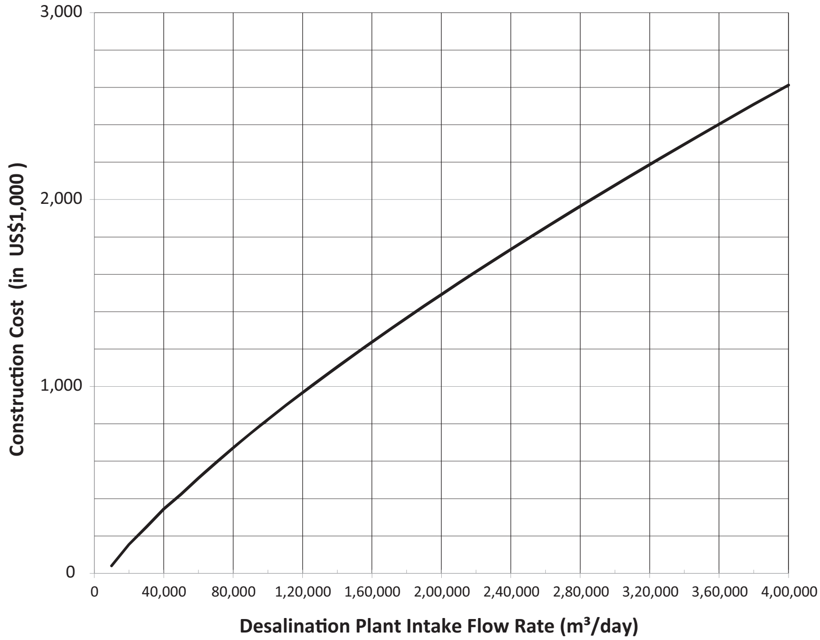

FIGURE 5.8 Construction costs for wedge-wire screens.

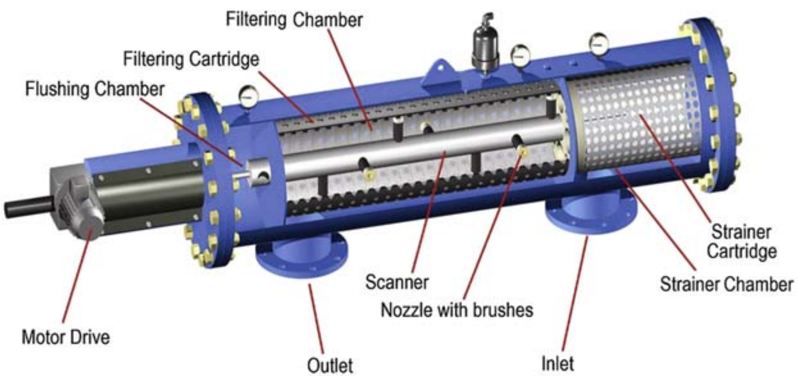

FIGURE 5.9 Self-cleaning microstrainer.

groove below, creating a filtration element with series of valleys and traps for source water debris. The stack is enclosed in corrosion- and pressure-resistant housing.

Desalination plants with disk filter microscreens are usually also equipped with conventional coarse screens or a combination of coarse and fine screens, which retain debris larger than 10 mm (0.4 in.). During the filtration process, the filtration discs are tightly compressed together by the spring's power and the differential pressure, thus providing high filtration efficiency. Filtration occurs while water is percolating from the peripheral end to the core

FIGURE 5.10 Disk filters.

of the filtration units. Source water debris and aquatic organisms (mainly plankton) of sizes smaller than the size of the microscreens are retained and accumulated in the cavity between the filter disks and the outer shell of the filters, thereby increasing the head loss through the filters. Once the filter head-loss reaches a preset maximum level (typically 0.35 bars/5 psi or less), the filters enter backwash mode. All debris retained on the outer side of the filters are then flushed by tangential water jets of filtered seawater flow under 0.15–0.2 bars (2.0–3.0 psi) of pressure and the flush water is directed to a pipe, which returns the debris and marine organisms retained on the filters back to the ocean.

Because of the relatively low differential pressure the filters operate at, these filters are likely to minimize impingement of the marine organisms in the source water. Since the disk filtration system is equipped with an organism return pipe, the entrained marine organisms are returned back to the source water body, thereby reducing their entrainment.

One of the key issues associated with using membrane pretreatment is the puncture of membrane fibers by sharp objects contained in the source water (e.g., broken shells or sharp sand particles). In addition, seawater contains barnacles, which in their embryonic phase of development are 130–180 μm in size and can pass through the screen openings unless these openings are 120-μm or smaller.

If barnacle plankton passes the screens, it could attach to the walls of downstream pretreatment facilities, grow on these walls, and ultimately could interfere with pretreatment system operations. Once barnacles establish colonies in the pretreatment facilities and equipment, they are very difficult to remove and can withstand chlorination, which is otherwise very effective biocide treatment for most other marine organisms. Therefore, the use of fine microscreens or disk filters (80–120-μm size) is essential for reliable operation of the entire desalination plant using membrane pretreatment. Microscreens or disk filters are not needed for pretreatment systems using granular media filtration because these systems effectively remove fine particulates and barnacles in all phases of their development.

5.3.2 Design Example

Disk filters have found wide application as microscreens for source seawater prior to membrane pretreatment. The following example illustrates disk filter-type microscreen application for a 50,000 m^3/day (13.2 MGD) seawater desalination plant designed for 43% SWRO system recovery and a total plant seawater intake flow of 127,910 m^3/day (5329 m^3/h or 33.8 MGD):

Manufacturer	Amiad/Arkal or equal
Model	SpinKlin Galaxy6″
Unit disk filter (module) capacity	16,000 m^3/day (667 m^3/h)
Number of arrays	2
Number of disk filter modules per array	4
Number of disks (spines) per disk filter	8
Number of jets per spine	48
Filter size	100 μm
Array inlet- & outlet-piping diameters	300 mm (12 in.)
Inlet and outlet diameters of disk filters	150 mm (6 in.)
Pressure loss (clean filter)	0.15 bars (2.1 psi)
Pressure loss triggering backwash	0.30 bar (4.2 psi)
Average pressure loss during operation	0.22 bar
Number of filters washed at one time	4 (one array)
Backwash flow	16 m^3/min per array
Backwash cycle length	6 min
Backwash frequency	14 washes/day
Total backwash volume	0.3%–0.5% of intake flow
Backwash pumps (horizontal centrifugal)	1 duty + 1 standby
Backwash pump head	3.0 bars
Disk filter material	Polypropylene

This example is developed for a particular type of popular disk filter (SpinKlin, manufactured by Arkal, Israel—see Fig. 5.10). These filters have found applications on a number of desalination plants with membrane pretreatment such as the 300,000 m^3/day (80 MGD) Adelaide and Southern (Perth II) Desalination plants in Australia, the 100,000 m^3/day (26 MGD) Chennai SWRO Plant in India, and other plants in the Middle East and Europe.

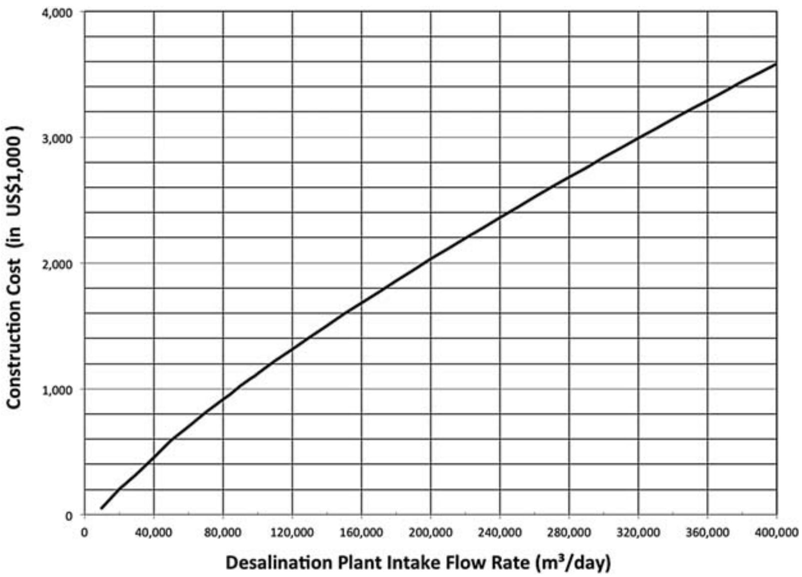

FIGURE 5.11 Construction costs of microscreens.

A number of other manufacturers provide similar equipment, and the specific unit sizes and design criteria vary (Gille, 2003). Equipment manufacturer should be consulted to identify the microscreen system design criteria for a specific desalination project.

5.3.3 Costs of Microscreens

Fig. 5.11 depicts a budgetary cost graph for microscreen systems as a function of the intake flow they are designed to process. These costs are presented in year 2017 US$.

5.4 CARTRIDGE FILTERS

5.4.1 Types and Configurations

Cartridge filters are fine microfilters of nominal size of 1–25 μm made of thin plastic fibers (typically polypropylene), which are wound around a central tube to form standard size cartridges (see Fig. 5.12). Often, they are the only screening devices between the intake wells and the RO system in brackish and seawater desalination plants with well intakes producing high-quality source water. Cartridge filters are RO membrane protection rather than screening devices and the main purpose they serve is to capture particulates in the pretreated source water that may have passed through the upstream pretreatment systems.

FIGURE 5.12 Cartridge filter loading.

Although wound (spun) polypropylene cartridges are most commonly used for seawater and brackish water applications, other types such as melt-blown or pleated cartridges of other materials have also found application.

Standard cartridge filters for RO desalination plants are typically 101.6−1524 cm (40−60 in.) long and are installed in horizontal or vertical pressure vessels (filter housings). Cartridges are rated for removal of particle sizes of 1, 2, 5, 10, and 25 μm, with the most frequently used size being 5 μm.

5.4.2 Planning and Design Considerations

Cartridge filters are typically installed downstream of the granular media pretreatment system (if such system is used for pretreatment) to capture fine sand, particles, and silt that could be contained in the pretreated seawater following granular media filtration. When the source seawater is of very high quality (SDI < 2) and does not need particulate removal by filtration prior to desalination, cartridge filters are used as the only pretreatment device, which in this case serves as a barrier to capture fine silt and particulates that could occasionally enter the source water during the start up of the intake well pumps or due to equipment/piping failure.

The main function of cartridge filters is to protect the downstream RO membranes from damage, not to provide removal of large amount of particulate foulants from the source seawater. A typical indication whether the pretreatment system of a given desalination plant operates properly is the seawater SDI reduction through the cartridge filters. If the pretreatment system performs well, then the SDI of the source water upstream and downstream of the cartridge filters is approximately the same.

If the cartridge filters consistently reduce the SDI of the filtered source water with over 0.5 units, this means that the upstream pretreatment system is not functioning properly. Sometimes, SDI of the source water increases when it passes through the cartridge filters. This condition almost always occurs when the cartridge filters are not designed properly or are malfunctioning and provide conditions for growth of biofouling microorganisms on and within the filters.

A frequently debated question is whether cartridge filters are needed downstream of MF or UF membrane pretreatment systems, taking under consideration that the membrane filters have one to two orders of magnitude smaller pores than cartridge filters. The answer to this question is highly dependent on the quality of the pretreatment membrane fiber material and the type of flow pattern through the pretreatment system.

For UF- or MF filtration systems that have a direct flow-through pattern where the desalination plant feed pumps convey water directly through the membrane pretreatment system without an interim pumping, the pretreatment membranes are more likely to be exposed to pressure surges. If pretreatment membrane fiber material is weak and it easily breaks under pressure surge conditions, such pretreatment systems experience fiber breaks more frequently. Broken membrane fibers would release small amount of particles into the RO feed water, which could cause accelerated membrane fouling.

In addition, if the broken membrane fibers release sharp particles of shellfish origin, these particles could also damage the RO membranes. Shellfish particles my find their way into the UF- or MF-pretreated water if shellfish plankton that is contained in the source water passes through the microscreens, then grows to adult shellfish organisms (i.e., barnacles) on the walls of the feed pump station, and the shells of the organisms released from the walls are broken into small sharp particles by the feed pumps of the membrane pretreatment system. The broken sharp-shell particles would be pressurized onto the filter fibers causing punctures and ultimately entering the filtered flow. In such cases, the use of cartridge filters downstream of the membrane pretreatment system is a prudent engineering practice.

Cartridge filters are operated under pressure and the differential pressure across these filters is monitored to aid in determining when filter cartridges should be replaced. In addition, valved sample ports should be installed immediately upstream and downstream of the cartridge filter vessel(s) for water quality sampling and testing (including SDI field testing).

Cartridge filtration systems are typically designed for hydraulic loading rates of 0.2—0.3 Lps/250 mm (3—5 gpm/10-in.) of length. Additional filtration capacity is normally provided to allow cartridges to be replaced without interruption of water production. Pressure vessels are typically constructed of duplex stainless steel for seawater RO installations.

The clean cartridge filter pressure drop is usually specified as less than 0.2 bars (2.8 psi). Commonly, cartridges are replaced when the filter differential pressure reaches 0.7 or 1 bar (10—14 psi). The operational time before replacement depends on source water quality and the degree of pretreatment. Typically, a cartridge filter replacement is needed once every 6 to 8 weeks. However, if the source seawater is of very good quality (SDI_{15} consistently less than 2), cartridge filters may not need replacement for 6 months or more.

For RO systems where sand in the feed water might be anticipated, rigid melt-blown cartridges or cartridge filters with single open ends and dual O-rings on the insertion nipple (rather than conventional dual open-end cartridges) are commonly used. The single open-end insertion filters have positive seating and an insertion plate, which do not allow deformation of the filter cartridge under pressure caused by sand packing. Double open-end cartridge filters are held in place by a spring-loaded pressure plate.

5.4.3 Design Example

This example presents the sizing and configuration of the cartridge filtration system for 50,000 m^3/day (13.2 MGD) seawater desalination plant with total plant seawater intake flow of 127,910 m^3/day (33.8 MGD):

Design feed flow (Qin)	98,440 m^3/day = 1480 Lps
Type of cartridge filter material	Pleated polypropylene
Cartridge filter size	5 μm
Cartridge filter length, Lcf	1016 mm (40 in.)
Selected design loading rate, DLR	0.25 Lps/250 mm
Number of cartridge filters needed	Qin/[DLR × (Lcf/250 mm)] = 1480 Lps/[0.25 Lps × (1016 mm/250 mm)] = 1457 cartridge filters (5.1)
Number of cartridge vessels	6 (selected to match RO trains)
Material of cartridge vessels	GRP
Number of cartridges per vessel	1457/6 = 243 (selected 240)
Actual cartridge filter loading rate	1480 Lps/[240 × 6 × (1016 mm/250 mm)] = 0.253 Lps per 250 mm (4.01 gpm/10 in.)

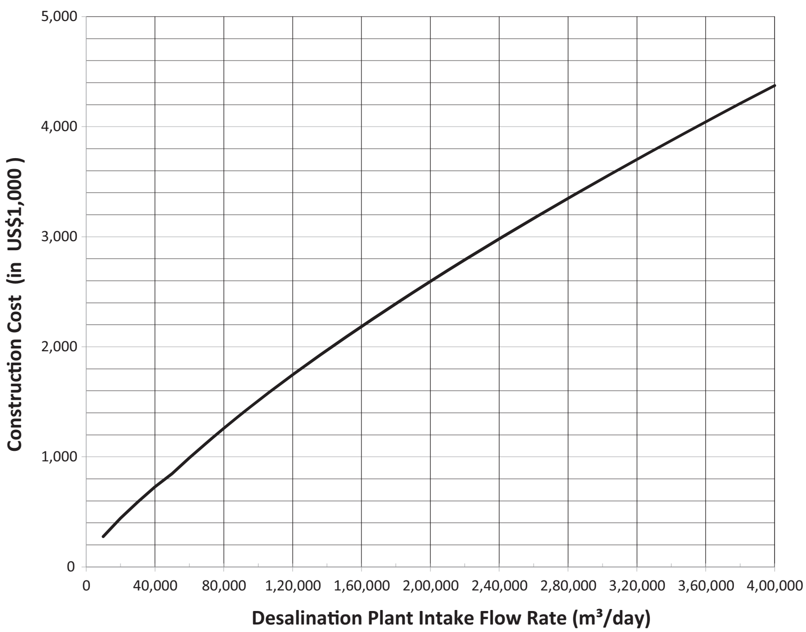

FIGURE 5.13 Construction costs of cartridge filter systems.

In summary, the cartridge filtration system for the 50,000 m^3/day (33.8 MGD) desalination plant will consists of 6 cartridge vessels each of which contains 240 cartridge filters of 5-μm size and length of 40 in.

5.4.4 Construction Costs of Cartridge Filter System

A graph of cartridge system construction costs as a function of intake flowrate is presented in Fig. 5.13. This graph is determined based on actual desalination projects and may vary in a range of ±30% of the values presented on this figure.

References

Gille, D., 2003. Seawater intakes for desalination plants. Desalination 156, 249−256.

Mackey, E.D., Pozos, N., James, W., Seacord, T., Hunt, H., Mayer, D.L., 2011. Assessing Seawater Intake Systems for Desalination Plants. Joint Report of the Water Research Foundation, WateReuse Research Foundation, the US Bureau of Reclamation and the California Department of Water Resources.

Moore, B.J., Malfeito, J.J., Zorrilla, J., 2009. An overview of the 150,000 m3/day Beckton desalination plant in London. In: Proceedings of IDA World Congress, Dubai, UAE, November 7−12. Ref. IDAWC/DB09-089, 2009.

Rogers, N., 2009. A review of onshore screening technologies for fine screening of large intakes. In: Proceedings of IDA World Congress, Dubai, UAE, November 7−12. Ref. IDAWC/DB09-070, 2009.

WRA, 2011. Desalination Plant Intakes − Impingement and Entrainment Impacts & Solutions, White Paper. WateReuse Association (WRA), Alexandria, Virginia.

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CHAPTER

6

Conditioning of Saline Water

OUTLINE

Pretreatment for Reverse Osmosis Desalination
http://dx.doi.org/10.1016/B978-0-12-809953-7.00006-1

6.1 INTRODUCTION

To reduce the fouling potential of the saline source water, this water is conditioned before reverse osmosis (RO) separation using various chemicals: coagulants, flocculants, scale inhibitors, oxidants (i.e., chlorine, chlorine dioxide), oxidant reduction compounds (i.e., sodium bisulfite), and pH adjustment chemicals (e.g., acids and bases). Coagulants and flocculants are added to enhance the removal of particulate and colloidal foulants in the downstream pretreatment facilities. Scale inhibitors are introduced in the saline source water after pretreatment filtration to suppress the crystallization of mineral scaling foulants on the surface of the RO membranes.

Oxidants (typically sodium hypochlorite or chlorine dioxide) are fed to the saline source water to control biofouling of the pretreatment and RO membrane systems and excessive growth of aquatic organisms (i.e., shellfish) on the inner surface of the intake piping, equipment, and structures. Sodium bisulfite or other reducing chemicals are added to the pretreated source water to remove the residual chlorine and/or other oxidants before the introduction of the source water into the RO membranes.

6.2 COAGULATION

Coagulant addition is accomplished ahead of the pretreatment sedimentation tanks, dissolved air flotation units, or filters. Coagulants most frequently used for membrane plant source water conditioning before sedimentation or filtration are ferric salts (ferric sulfate and ferric chloride). Aluminum salts (such as alum or poly-aluminum chloride) are not typically used because it is difficult to maintain low levels of aluminium in dissolved form and small amounts of aluminum may cause irreversible mineral fouling of the downstream RO membrane elements (Edzward and Haarhoff, 2011).

The optimum coagulant dosage is pH dependent and should be established based on an onsite jar or pilot testing for the site-specific conditions of a given application. Practical experience indicates that the optimum pH for coagulation of particles in saline waters is highly temperature dependent. As the temperature decreases, the optimum pH for coagulation increases and vice versa. For example, the optimum pH for seawater with temperature of 10°C (50°F) is 8.2, whereas at temperature of 35°C (95°F) the optimum pH decreases to 7.4 (Edzwald and Haarhoff, 2012).

Use of coagulant is critical for the effective and consistent performance of granular media pretreatment filtration systems. Coagulation reduces the surface charge of the source water particles and facilitates their agglomeration into larger size particles, which are easier to settle and/or filter by granular media filtration. The amount of coagulant needed for source water conditioning is dependent on the size and charge of the particles dominating in the source water. Coagulation allows granular filtration process to also remove fine particulate debris (e.g., silt) and microplankton from the source water.

Well-operating filters can remove solid particles, which are as small as 0.5 μm. However, if the source water has low turbidity [(usually < 0.5 nephelometric turbidity unit (NTU)] and the prevailing size of particles is less than 5 μm (which, e.g., is common for deep intakes), coagulant addition does not yield a significant improvement of the granular media filtration process. In this case, adding minimal amount of coagulant (i.e., 0.5 mg/L or less) or even not adding any coagulant is viable and much more efficient than adding excess coagulant. In such conditions, however, it is critical to have an extended contact period for coagulation and flocculation (i.e., coagulation and flocculation times of 10 min or more) because for source waters containing small amount of particles, the main mechanism for flock formation is physical collision and contact of the source water particles rather than their agglomeration by charge-based attraction.

Coagulation is critical for source waters of high turbidity and/or natural organic matter (NOM) content, especially if turbidity/NOM spikes are caused by surface runoff (e.g., rain events), river water, or wastewater discharges as well as in the case of bottom sediment resuspension triggered by frequent boat traffic, source water area dredging, periodic occurrence of strong currents near the intake, or high-intensity winds in shallow intake areas. In such cases, a rule-of-thumb based practice is to add coagulant of dosage approximately two times higher than the level of the actual source water turbidity.

Membrane pretreatment can remove particles as fine as 0.04 μm (microfiltration membranes) or 0.01 μm (ultrafiltration membranes) without coagulation. Therefore, for these systems coagulation is typically applied only when the saline source water contains NOM particles with a high negative charge that could be coagulated easily and removed via filtration or during heavy algal blooms or oil spill events.

6.2.1 Types of Coagulation Chemicals and Feed Systems

This chemical conditioning of source water for accelerated settling and filtration of particulate foulants includes three key components: chemical feed systems, coagulation, and flocculation tanks. The purpose of coagulation tanks is to achieve an accelerated mixing of the coagulant and the source water and to neutralize the electric charge of the source water particles and colloids.

The subsequent agglomeration of the coagulated particles into larger and easy to remove flocks is completed in flocculation tanks. It should be noted that flocculation tanks are always added downstream of the coagulation tanks, independent of whether additional flocculant chemical is fed to the source water. Although coagulation is a relatively rapid chemical reaction, flocculation is a much slower process and typically requires longer contact time and enhanced mixing conditions. Therefore, coagulation and flocculation system design requirements differ.

The main purpose of the coagulant feed system is to achieve uniform mixing of the added coagulant with the source water, which promotes accelerated attraction of the coagulant particles to the source water solid particles (e.g., to facilitate efficient coagulation). The two types of coagulant mixing systems most widely used in desalination plants are in-line static mixers with chemical injection ports (Fig. 6.1) and mechanical (flash) mixers installed in coagulation tanks (Fig. 6.2).

Under normal operational conditions, coagulant is dosed proportionally to the flow rate of the saline source water and to the content of particulate solids in the water. Optimum coagulant dosage is determined based on jar testing and is dependent on source water quality (content of particulate solids and organics, water temperature, and pH). Jar testing is recommended to be completed if source water turbidity and organic content change over 50% from their last daily average level or if source water quality changes significantly due to the

FIGURE 6.1 In-line static mixer.

FIGURE 6.2 Flash mixers in coagulation tank.

occurrence of algal bloom, source water foaming, discoloration, increase in silt content, significant decrease in temperature, or other unusual ambient conditions, which result in significant changes in source water quality.

As a rule of thumb, under normal nonalgal bloom conditions, 1.0 mg/L of ferric chloride or 1.2 mg/L of ferric sulfate is recommended to be added for every 1 NTU of turbidity in the saline source water. During algal bloom or intense rain events, coagulant dosage is recommended to be increased approximately 1.5–2 times compared to that used under normal conditions to achieve enhanced removal of organics and solids. It should be pointed out, however, that overdosing of coagulant is undesirable because it would shorten the useful life of the downstream cartridge filters and cause accelerated fouling of the RO membrane elements. Therefore, under unusual or extreme source water conditions, the optimum dosage of coagulant should be determined by jar testing.

At present, ferric chloride is more commonly used than ferric sulfate for saline water pretreatment because of the lower application dosage and lower unit chemical costs. However, the use of ferric chloride could have a significant disadvantage if it is not of high purity and contains significant amount of manganese—i.e., accelerated fouling of the plant cartridge filters and potential oxidation of the RO membranes by spontaneous generation of permanganate oxidant in the concentrate as it travels through the membrane vessels. In general, ferric chloride is cheaper coagulant because it is generated as a waste product of iron ore pickling. Because all iron ore naturally contains manganese, typically commercially available ferric chloride also has this metal impurity.

If a manganese content in ferric coagulant exceeds certain threshold, this manganese tends to precipitate on the cartridge filters downstream of the plant pretreatment facilities and ultimately on the RO membranes. As a result, the useful life of a cartridge filters is often shortened from a normal duration of 6—8 weeks down to 1—2 weeks and RO membranes are exposed to slow oxidation, which results in the permanent loss of membrane rejection within 1—2 years of their installation.

The most common approach to identify whether the ferric chloride used for coagulation is of inadequate quality (i.e., contains unacceptably high concentration of manganese) is to observe the color of the cartridge filters downstream of the desalination plant pretreatment facilities—excessive content of manganese creates typical dark brown/black discoloration of the cartridge filters (Fig. 6.3).

RO membrane oxidation is caused by the spontaneous generation of permanganate in the concentrate from the magnesium in the ferric chloride as concentrate salinity increases along the length of the RO membranes. The process of permanganate generation is not well known and studied at this time. Nevertheless, operational experience indicates that permanganate oxidation has a slow but irreversible degrading impact on RO membrane salt rejection in seawater reverse osmosis (SWRO) desalination plants using low-cost/low-quality ferric chloride which contains more than 50 micrograms of manganese per gram of ferric chloride.

FIGURE 6.3 Cartridge filters with dark brown/black discoloration as a result of the use of poor quality ferric chloride.

A relatively simple method to diagnose the spontaneous generation of oxidant in the membrane vessels is to measure the oxidation–reduction potential (ORP) of the saline source water fed to the individual RO trains and that of the concentrate from the same RO trains. If the spontaneous generation of a significant amount of oxidant occurs in the membrane vessels of a given RO train, the ORP of the RO train concentrate is measurably (typically over 10%) higher than the ORP of the feed water to this train.

If such a condition occurs and sodium bisulfite is being fed to the RO system at the same time, it is recommended to discontinue the use of this chemical and check the ORP of the RO train feed water and concentrate within 20 min of the discontinuation of the bisulfite feed to verify whether the problem is solved. If this problem persists, it is desirable to reduce or discontinue the addition of ferric chloride and consider the supply of new batch of this product from another supplier or to switch to the use of ferric sulfate.

Due to the uncertainty of the quality of ferric chloride available to the desalination plant operators and the potential for changes of content of manganese and other impurities from one batch of chemical to another, the most suitable solution to the challenges described earlier is to use ferric sulfate instead of ferric chloride for coagulation. Ferric sulfate is generated from the chemical reaction of pure iron and sulfuric acid and therefore, it usually does not contain impurities and does not cause the problems associated with the use of low-quality ferric chloride. Therefore, although of higher unit chemical cost and higher application dosage for the same amount of solids, the use of ferric sulfate is preferable and strongly recommended to troubleshoot the negative impact of the use of low-purity ferric chloride on plant cartridge filters and RO membranes.

Source water temperature is an important factor affecting the coagulation process, especially for source waters with solid particles that have low electric charge. Because in this case, the driving force for the coagulation process is mainly random particle movement and collision, lower temperatures tend to hinder the available kinetic energy for particle movement and therefore cause lower coagulation efficiency. Such negative impacts are typically observed for source water temperatures below 20°C (68°F).

It is important to note that in many cases, plant operators try to compensate for such temperature impact by increasing coagulant dosage. This approach, however, is not recommended because the overdosing of chemical will result in an elevated fouling rate of the downstream cartridge filters and RO membranes.

Instead, it is preferable to increase the flocculation time to 15 min or more by reducing the volume of the pretreated saline source water.

Two other approaches suitable to troubleshoot this operation challenge are pH adjustment and increase of the motor speed of the flash mixers in the coagulation tank. These corrective actions are applicable if the coagulation system is equipped with coagulation contact tank and the flash mixer motors have two-speed or variable frequency drives. Usually, lower temperature results in higher optimum pH—for source water temperature in a range of 10–20°C (50–68°F) the optimum pH is usually between 8.2 and 7.8. Therefore, in this case, if the acid is added for optimizing of the coagulation process, it is recommended that the acid feed is discontinued when the source water temperature drops below 20°C (68°F). The pH adjustment-based troubleshooting approach usually works if the source water particles are negatively charged because the pH adjustment chemical increases the charge of the coagulant and its electrostatic attraction force. If source water particles do not have significant electric charge (e.g., their zeta potential is <-20 mV), pH adjustment is not likely to have a measureable effect on the coagulation process.

Ferric chloride and sulfate are corrosive liquids and moderately toxic. Inhalation of their fumes could cause throat and respiratory tract irritation. It is important to keep in mind that when dry chemicals are diluted with water, the chemical reaction releases a large amount of heat.

In-line static mixers have lower energy and maintenance requirements and are relatively easy to install. They typically operate at a velocity range of 0.3–2.4 m/s (1–8 fps) and are designed to operate in a plug-flow hydraulics to provide uniform mixing within the entire pipe cross-section. The velocity gradient × contact time for such mixers can be determined by the formula:

$$G \times T = 1212 \times d \Big/ (D_p \times L/Q)^{0.5} \tag{6.1}$$

where G, velocity gradient (s^{-1}); T, time (s); d, diameter of the static mixer (in.); D_p, differential pressure through the mixer (psi); L, length of the static mixer (in.); Q, flow (gpm).

Although in-line static mixers are simple to install and significantly less costly, they have two disadvantages: (1) their mixing efficiency is a function of the flow rate because the mixing energy originates from the flow turbulence; (2) static mixers are proprietary equipment and the project designer would need to rely on the equipment manufacturer for performance projections. Static mixers also create additional head losses of 0.3–1.0 m (1.0–3.3 ft), which need to be accounted for in the design of the intake pump station. Another important issue is to provide adequate length of pipeline (at least 20 times the pipe diameter) between the static mixer and the entrance to the pretreatment filters to achieve adequate flocculation.

Mechanical flash mixing systems consist of coagulation tank with one of more mechanical mixers and chambers. The coagulation tank is designed for velocity gradient $G \times T =$ 4000–6000. The power requirement for the mechanical mixers is 2.2–2.5 horsepower/ 10,000 m^3/day of processed flow. This type of mixing usually provides reliable and consistent coagulation, especially for desalination plants designed for significant differences in minimum and maximum plant production (e.g., more than 1:10).

6.2.2 Planning and Design Considerations

Overdosing of coagulants and their inadequate mixing with the source water are some of the most frequent causes for RO membrane particulate and colloidal fouling. When overdosed, coagulant accumulates in the downstream pretreatment facilities and can cause fast-rate fouling of the cartridge filters (Fig. 6.4) following the pretreatment step and colloidal iron fouling of the RO membranes (Fig. 6.5).

The effect of overdosing of a coagulant (iron salt) on the silt density index (SDI) level can be recognized by visually inspecting the SDI test filter paper. In Fig. 6.6, the first two SDI test pads are discolored as a result of coagulant overdosing. The numbers below the pads are the SDI readings.

In such a situation, a significant improvement of source water SDI can be attained by reducing coagulant feed dosage or, in the case of poor mixing, modifying the coagulant mixing system to eliminate the content of unreacted chemical in the filtered seawater fed to the RO membrane system.

FIGURE 6.4 Coagulant accumulation on cartridge filters due to iron coagulant overdosing.

FIGURE 6.5 Coagulant residue on the reverse osmosis membrane feed due to iron coagulant overdosing.

6.2.3 Design Example

6.2.3.1 In-line Static Mixer

Eq. (6.1) is applied for seawater desalination plant with a fresh water production capacity of 50,000 m^3/day [13.2 million gallons per day (MGD)] and intake flow of 127,910 m^3/day (33.8 MGD = 28,190 gpm), a 30-in. diameter mixer with a length of 5.0 ft, and D_p of 0.5 psi, and the $G \times T =$ (1212 × 30-in.)/[(0.5 psi × 5 ft × 12 in.)/23,472 gpm]$^{0.5}$ = 1300. The optimum $G \times T$ mixing range is usually between 500 and 1600.

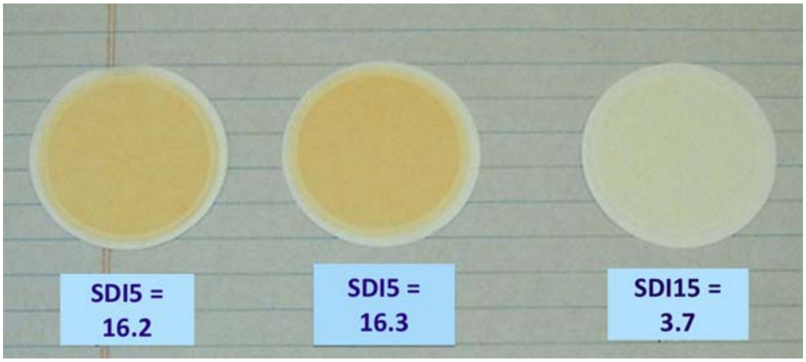

FIGURE 6.6 Iron accumulation on SDI test pad due to coagulant overdosing.

6.2.3.2 Coagulation Chamber With Flash Mixers

The key design criteria for such flocculation systems are listed below:

Number of coagulation tanks	= 4
Tank width × length × depth	= 1.2 m × 1.2 m × 1.4 m (3.9 ft × 3.9 ft × 4.6 ft)
Tank volume (V)	= 2.0 m^3 (21.5 ft^3)
Contact time (T)	= (2.0 m^3 × 4 tanks × 24 h × 60 min)/127,910 m^3/day = 0.09 min (5.4 s)
Mixing energy per tank at 2.5 hp/10,000 m^3/day (W)	= [(2.5 hp × 127,910 m^3/day)/4]/10,000 = 8 hp = 5966 W
Absolute viscosity of water (μa)	= 0.0014 N s/m^3
Velocity gradient multiplied by time G × T	$= W/(\mu a \times V)^{0.5} \times T$ $= [5966\ W/(0.00114\ N\ s/m^3 \times 2.0\ m^3 \times 4)]^{0.5} \times 5.4\ s$ = 4368
Type of mixer	Vertical-shaft with hydrofoil blades
Blade area/tank area	0.15%
Shaft speed	40–80 rpm

It is important to point out that the velocity gradient G × T provided by the coagulation chambers (4368) is significantly higher than that provided by the static mixer (1300), which would correspond to a more robust mixing.

6.3 FLOCCULATION

Flocculants (polymers) are sometimes applied in addition to coagulants to improve pretreatment. However, polymer addition, even if slightly overdosed, may also cause organic fouling of the RO membranes. Often, the potential for RO membrane fouling due to polymer overdosing is of a greater concern than the benefit of polymer use. Therefore, many desalination plants do not condition the coagulated saline source water with polymer. If used, only nonionic or anionic polymers are usually applied because most RO membrane elements carry a negative surface charge. The use of cationic polymer is likely to form a polymer film on the membrane surface, which will foul the RO membrane elements.

6.3.1 Types of Flocculation Chemicals and Feed Systems

The type and dosage of the most suitable polymer (nonionic or anionic) for a given project is typically determined by jar and/or pilot testing. Such a testing is usually completed along with the coagulant jar testing (Jamaly et al., 2014). Once the optimum coagulant dosage is established, flocculant of different dosages is added to determine its optimum application rate. Usually several different flocculants are tested in series. The main criteria for selecting optimum flocculant type and dosage are the size of the formed flocs, and the percent of turbidity reduction in the supernatant which are determined by direct visual observation and the settling rate over 30 min to 1-h period and by measuring the turbidity of the supernatant at time zero and at the end of the settling period.

Similar to coagulants, the solids removal efficiency of flocculants depends on temperature. Usually, the flocculation process is hindered in cold waters, i.e., waters with temperatures lower than 12°C (54°F) and could be enhanced by applying higher intensity mixing in the flocculation chambers.

Many polymers are inactivated by temperatures higher than 35°C (95°F). Therefore, if the source water temperature is likely to exceed this threshold limit for prolonged periods of time, it is recommended to select flocculants that are specifically designed to perform in high-temperature waters.

One of the most common challenges with flocculant addition is overdosing. The application of flocculants at dosage higher than 1.0 mg/L often results in a high content of unused polymer in the pretreated water. Such polymers usually accumulate in the filtration media (if direct filtration is used for pretreatment), on the surface of the cartridge filters, and on the RO membrane elements causing their premature fouling. To confirm that the flocculant is not overdosed, it is recommended to inspect the surface of the cartridge filters periodically and look for the formation of transparent to yellow-colored film on the surface of the cartridge filters, which is slimy on touch and appearance. Normally, the surface of the cartridge filters should be naturally rough and should not have a surface accumulation of polymer. Similar observation holds true for the front-end cap of the RO membrane elements located downstream of the cartridge filters.

Another common problem with flocculants is the content of biodegradable organics and metal (i.e., copper, nickel) impurities that can cause fouling or damage on the RO membranes. To determine the relative contribution of flocculant to the source water total organic carbon (TOC) content, it is recommended to measure the TOC concentration of the saline source water before and after the addition of flocculant to it at a target dosage. Flocculants suitable for RO desalination systems should not cause more than 0.2 mg/L of TOC increase. Use of flocculants contributing over 0.5 mg/L of TOC to the source water is likely to result in accelerated RO membrane biofouling and cartridge filter replacement.

6.3.2 Planning and Design Considerations

The formation of large flocks that can be removed easily by the downstream sedimentation, dissolved air flotation (DAF), or filtration processes is a slower process, and therefore, it requires a longer retention time than coagulation. The flocculation systems most widely used for pretreatment of saline source waters are mechanical flocculators with vertical mixers (Fig. 6.2).

6.3.2.1 Key Design Criteria

The key design criteria for such flocculation systems are listed below:

Minimum number of tanks	4
Velocity gradient	30–120 s^{-1}
Contact time	10–40 min
Flocculation chambers in series	2–4
Water depth	3.5–4.5 m
Blade area/tank area	0.1%–0.2%
Shaft speed	2–6 rpm

6.3.3 Design Example

An example of a flocculation tank for 50,000 m^3/day seawater desalination plant, which is part of the DAF clarifier system, is presented in Chapter 7.

6.4 ADDITION OF SCALE INHIBITORS

The formation of mineral deposits (scaling) on the surface of the RO membranes is caused by the precipitation of low-solubility salts such as calcium carbonate, magnesium carbonate, barium sulfate, strontium sulfate, and silica. The stability of these compounds depends on their concentration in the concentrate flow stream, water temperature, pH, desalination plant recovery, and other factors.

As plant recovery increases, more salts in the source water are likely to reach to the point at which their solubility is exceeded and they begin forming crystals (referenced as solubility limit) and ultimately will begin accumulating on the membrane surface and cause mineral membrane fouling (scaling). Therefore, for a given saline water mineral composition, RO desalination systems face a threshold of maximum recovery when the scaling process destabilizes membrane performance due to excessive accumulation of salt crystals or amorphous scale on the membrane surface and the impact of membrane productivity.

The determination of the scaling potential of saline source waters is rather complex and, therefore, it is usually performed using the computer software. This type of software is available from key manufacturers of antiscalants such as Nalco, Avista, Genesys, etc. In their software, chemical manufacturers relate antiscalant requirements to the scaling potential of each salt that can be formed in the concentrate in large enough quantities to create a measurable scale. The antiscalant dosages are then determined for each scale forming mineral and recommendations are made based on the salt that would require the largest amount of antiscalant.

It is important to point out that the results of this type of software are very sensitive to the accuracy and completeness of the source water quality analysis. Sometimes parameters such as barium, strontium, and fluoride are not measured in the source water and assumed concentrations are used instead. Such practices may often cause the projection results to be inaccurate. General steps to determine the limiting salts and to manually calculate allowable permeate recovery for saline waters of given mineral content are illustrated elsewhere (AWWA, 2007).

El-Manhaeawy and Hafez (2001) have studied the use of molar ratio of sulfates and bicarbonates as a tool to predict scaling potential of saline source waters. Table 6.1 summarizes study results.

The table is indicative of the fact that in typical open-ocean seawater the main cause of scaling is sulfate, whereas in low-salinity brackish water, the predominant type of scaling is caused by carbonate salts. This does not, however, mean that seawater cannot cause carbonate scale—it is indicative of the fact that carbonate scale formation is of relatively low rate, typically not more than 10% (by weight) of the scale observed on SWRO membranes, and is contributed by calcium carbonate.

One of the key factors associated with the scaling potential of saline source waters is their ionic strength. Typically, the higher the ionic strength [i.e., total dissolved solids (TDS) concentration] of the source water, the higher the RO system recovery threshold at which scaling would occur at the same temperature and mineral composition.

Another important factor associated with the scaling potential of the source water is temperature—usually the scaling potential of calcium carbonate increases with the increase in source water temperature. In seawater, the scaling is accelerated significantly when the water temperature exceeds 35°C.

Silica, especially if it is in a colloidal state, sometimes creates scaling challenges in brackish waters. Usually, silica is considered to be of concern when its concentration in the RO concentrate exceeds 140 mg/L (Wilf et al., 2007). In typical open-ocean waters, silica is usually below 20 mg/L and is not considered a compound of high scaling potential because SWRO systems typically operate at relatively low recoveries (40%–50%).

Barium and strontium concentrations in saline source waters correlate closely with the concentrations of these compounds observed in membrane scales. Usually, these compounds are of very low levels in seawater to be significant sources of scale. However, in some brackish waters they can be at an order of magnitude higher levels than in seawater

TABLE 6.1 Sulfate/Bicarbonate Molar Ratios and Scaling Potential

Molar Ratio of SO_4/HCO_3 (in mMols)	Chloride Concentration (mg/L)	Sulfate Scaling Potential	Carbonate Scaling Potential
>15	≥20,000	High	Low
10–15	≥10,000	Medium to high	Medium
1–10	≥3,000	Medium	Medium to high
<1	<3,000	Low	High

and may result in scaling, especially in brackish plants operating at high recoveries (85% and above).

Scales vary in texture and appearance—typically calcium sulfate scales that form on RO membranes, which treat seawater, have orderly prismatic crystalline structure. Sulfate crystals could reach lengths of 20 mm and widths of 5 mm (El-Manharawy and Hafez, 2001). However, sulfate scale formed from high-salinity brackish waters can vary measurably from one water source to another. On the other hand, carbonate scales formed on brackish water membranes are typically fine amorphous white deposits.

Scaling control depends on the particular mineral salts that precipitate on the membrane surface. For example, calcium carbonate scales can be prevented from forming by source water acidification. Acids convert carbonate ions $\left(CO_3^{2-}\right)$ into soluble bicarbonate ions and carbonic acid and ultimately into carbon dioxide.

To prevent calcium carbonate and other scaling, often commercially available scale inhibitor chemicals (antiscalants) are added to the source water or, alternatively, scaling foulants are removed by softening or nanofiltration pretreatment facilities located upstream of the RO system.

Some of the compounds naturally contained in seawater (such as humic acids) serve as natural chelating agents and scale inhibitors. Therefore, acidification of seawater prior to membrane salt separation is not usually needed and commonly practiced.

When a high level of boron removal is targeted, seawater acidification will have a negative impact on boron removal. In addition, an overdose of acid can cause piping and equipment corrosion and create iron-based colloidal fouling on the RO membranes. Therefore, the benefits associated with acid addition will need to be weighed against the potential problems that acidification could cause.

Often, SWRO systems have to be designed to remove boron to levels below 1 mg/L. In this case, a most common practice for enhanced boron removal is to increase the source seawater pH from ambient levels (7.8–8.3) to a range between 8.8 and 11. At this high pH range, RO membrane scaling is very likely to occur and, therefore, scale inhibitors are typically added to prevent it.

Some scale inhibitors prevent the formation of seed crystals, whereas others deform the seed crystals so that they cannot grow and cause problems in the membrane system. In some cases, dispersants are added to the scale inhibitor formulations to aid in preventing colloidal material deposition.

It is important to note that antiscalants are designed so that they do not pass through the membranes and are, therefore, contained in the concentrate. This is an important issue in terms of their potential environmental impacts and toxicity.

Routine operation of antiscalant systems involves feeding of the selected scale inhibitor product at a rate proportional to the plant flow. As indicated earlier, antiscalant dosage and type are determined based on the mineral makeup of the saline source water and can be determined using specialized software packages available from the antiscalant manufacturers or with the direct assistance from these suppliers.

Most antiscalants are delivered to the desalination plant site in a liquid form at 99% concentration (as neat solution) and are stored in plastic containers or storage tanks. Usually, antiscalant is added to the pretreated saline source water before it is fed to the RO system. Antiscalant is first diluted from its neat (99% concentration) to a working concentration of

20% as weight and then injected into the pretreated water. Antiscalant mixing with the source water is achieved by in-line static mixer. Sometimes antiscalant is fed upstream of the cartridge filters to use them as a mixing system.

Conditions accelerating membrane scaling are high content of scaling minerals (mainly calcium and magnesium salts) combined with low overall mineral content of the saline water (i.e., low TDS), operation at very high RO system recovery (over 50% for seawater desalination and over 65%−70% for brackish water desalination), increase in source water pH (usually pH increase above 8.8), and increased source water temperature (usually scaling rate accelerates dramatically for temperatures above 35°C (95°F)). The scaling thresholds listed earlier may change from one source water to another depending on the presence and concentration of compounds prone to precipitation at normal operational conditions.

In most cases, scaling could be avoided by lowering the RO system recovery. However, such operational strategy would result in reduced desalination plant productivity and may not be easy to use as a sustainable long-term solution. In the case of carbonate scale, lowering pH to 7.6 or less could inhibit its formation. However, pH decrease is not a remedy for most of the other types of scales.

It should be noted that antiscalants are intentionally designed to be biodegradable because they are removed from the RO system with the concentrate and are disposed to the receiving water body along with it. Therefore, antiscalant overdosing typically causes accelerated RO membrane biofouling. To avoid overdosing, it is recommended to apply the minimum viable dosage of this chemical. Such a dosage could be determined by pilot testing by starting at a minimum dosage of 0.25 mg/L and applying incrementally higher dosages of antiscalant until they reach levels of 3−4 mg/L (or higher if recommended by potential supplier of antiscalant based on their specialized software). For most antiscalants, the optimum feed dosage varies between 0.25 and 1.0 mg/L.

It should be pointed out that most antiscalants can be oxidized by chlorine and other strong oxidants such as bromamines. Therefore, these source water conditioning chemicals should always be added after source water dechlorination. Exposure to chlorine will not only inactivate the applied antiscalant but will also break down its complex organic molecules into easily biodegradable compounds that would accelerate RO membrane biofouling.

6.4.1 Acids

Sulfuric acid feed upstream of RO membrane systems is commonly used for calcium carbonate control. Calcium carbonate is the most common scaling compound in brackish water (Wilf et al., 2007) and, therefore, brackish water desalination plants are usually equipped with acid addition systems.

The addition of acid lowers the carbonate concentration by converting bicarbonate to carbon dioxide. The carbon dioxide passes through the RO membranes and is removed or used in the posttreatment system. Sulfuric acid is only effective against carbonate scale and because this chemical adds sulfates to the source water, it actually increases its calcium sulfate scaling potential.

Sometimes hydrochloric acid is used as a scale inhibitor instead of sulfuric acid. Sulfuric acid is usually preferred over hydrochloric acid because of cost and safety reasons. However, hydrochloric acid may be used if the sulfate introduced to the source water with the feed of sulfuric acid significantly affects the system design and cost.

Because source water acidification could be needed for both enhanced coagulation and prevention of mineral scaling, the actual applied dosage should be driven by the larger of the two dosages. Usually, this is mineral scaling for brackish water reverse osmosis desalination applications and enhanced coagulation for SWRO applications. The sulfuric acid solution is dosed proportional to the source water flow.

Sulfuric acid is typically delivered on site in liquid form and in a concentration of 98% by weight. The density of this chemical at this concentration is 1.83 kg/L. The commercial product is usually diluted to 20% or less before its application to the saline source water to avoid/minimize corrosion of metallic piping and equipment installed on the conveyance piping and mixing chamber structures.

Applying sulfuric and hydrochloric acids at concentrations over 25% by weight can cause accelerated corrosion of steel equipment and structures and has to be avoided by frequent checking of the application concentration of these acids, which is preferably to be set at 20% or less.

6.4.2 Other Scale Inhibitors

Sodium hexa-metaphosphate (SHMP) has been one of the most commonly used scale inhibitors in the past but in recent years SHMP is frequently replaced by proprietary chemical formulations because of their improved effectiveness, long storage life without losing strength, microbial growth resistance while in the feed tank, ease of handling, or other reasons. SHMP can serve as a bacterial nutrient and because it contains phosphates, its use/overdose could result in discharge of concentrate with high phosphorus content, which in turn can trigger algal blooms in the discharge area. Therefore, the use of this otherwise popular and effective sale inhibitor is limited.

Phosphonates such as amino-tris-methylenephosphonic acid, 1-hydroxyethyllidene-1, 1,1-diphosphonic acid, and 2-phosphonobutane-1,2,4-tricarboxylic acid have found wide application for high-temperature waters and they are usually very suitable for calcium and barium sulfate scale prevention and calcium carbonate scale inhibition.

Phosphonates decrease the precipitation rate of salts that have exceeded their solubility thresholds. These antiscalants are particularly efficient for SWRO systems with high pH operation for enhanced boron removal. Besides calcium scale, other important scaling compounds that impact system operations are magnesium carbonate and magnesium hydroxide. Phosphonates can also react with and remove low levels of iron in the source water and inhibit silica fouling.

Polymeric dispersants based on polyacrylic acid and maleic acid (polyacrylates) are commonly applied as calcium carbonate scale inhibitors. They act by distorting the crystalline growth of the scales on the surface on the membranes. However, they are incompatible with coagulants used for seawater pretreatment and are, therefore, not recommended for such applications.

It is important to select the correct scale inhibitor for the specific application. For example, the presence of iron in the source water can cause precipitation and membrane fouling with some types of antiscalants. Scale inhibitor feed systems typically include positive-displacement metering pumps (or centrifugal pumps for large systems) drawing from a day tank or other storage devices (for small plants) such as 55-gallon drums or larger-capacity totes.

6.5 ADDITION OF SODIUM HYDROXIDE

Sodium hydroxide systems are typically used to adjust the pH of the feed seawater to the first or second pass of RO systems designed for enhanced boron removal. Addition of sodium hydroxide allows the conversion of boron, which naturally occurs in seawater as boric acid, into borate, which has larger molecule of stronger charge, and is, therefore, easier to reject by the RO membranes than boric acid.

In SWRO plants with two-pass membrane systems, sodium hydroxide (NaOH) is usually introduced into the first-pass permeate because this water does not contain a significant amount of scaling compounds and the antiscalant does not need to be added for its treatment through the second RO pass. Sodium hydroxide dosage usually varies between 2 and 20 mg/L, and based on practical experience, most SWRO plants apply 12–15 mg/L of sodium hydroxide for enhanced boron removal.

In single-pass SWRO systems, sodium hydroxide is added to the feed of the RO system—typically downstream of the point of addition of antiscalant. In such systems, feed of antiscalant is needed because source water pH increase above 6.8 even at plant recovery of 40%–45% would trigger scale formation on the last two SWRO elements within several days to a week. SWRO operation at recoveries higher than 45% may result in scaling at even lower pH. Practical experience shows that if sodium hydroxide is added to feed of the first pass of the RO system, achieving the same boron rejection effect requires approximately 15%–20% higher NaOH application dosage.

Ultimately, NaOH could be added to both the first- and the second-pass RO system feed waters. Practical experience shows that in this case, if the pH of the first-pass feed is increased to only 8–8.6, addition of antiscalant could be avoided for waters of high-salinity (i.e., TDS concentration >40 ppt). The NaOH dosage needed to achieve this pH is usually only 5–8 mg/L. Under this two-pass NaOH feed arrangement, the total boron removal could be increased with 10%–15% compared to the conventional practice of NaOH addition to the second-pass feed only.

In most cases, overdosing of sodium hydroxide would not have a negative impact on plant RO system performance except for increased scaling potential if this chemical is added to the feed of the first-pass RO system. As indicated previously, such scaling could be controlled successfully by the addition of antiscalant.

If sodium hydroxide is underdosed, the RO permeate will have elevated boron level. To determine the optimum dosage of sodium hydroxide, it is recommended to incrementally increase its concentration in the range of 10–20 mg/L in 2 mg/L increments every 2 h. For each test, NaOH dosage and boron levels in the RO feed and permeate should be measured and boron removal rate determined. The optimum dosage of sodium hydroxide is determined as the minimum dosage at which boron levels in the RO permeate reach steady state and further increase in NaOH does not yield significant additional boron removal.

6.6 ADDITION OF BIOCIDES

Oxidants such as sodium hypochlorite and chlorine dioxide are often used to suppress the growth of aquatic organisms (i.e., shellfish, barnacles) on the inner surface of intake pipes, equipment, tanks, distribution channels, and other structures in contact with the source seawater, as well as to minimize biofouling of RO membranes.

6.6.1 Sodium Hypochlorite

Sodium hypochlorite (NaOCl) is the most commonly used oxidant at present. When added to water, sodium hypochlorite generates hypochlorous acid (HOCl) and sodium hydroxide (NaOH):

$$NaOCl + H_2O \rightarrow HOCl \tag{6.i}$$

Hypochlorous acid in turn dissociates to hydrogen (H^+) and hypochlorite (OCl^-) ions:

$$HOCl \leftrightarrow H^+ + OCl^- \tag{6.ii}$$

The sum of sodium hypochlorite, hypochlorous acid, and hypochlorite ions is termed and measured as free residual chlorine. Chlorine in all of its forms is a toxicant that attacks all aquatic organisms and typically destroys them by the oxidation of their tissue and cells. It should be pointed out, however, that the use of chlorine has several drawbacks. Chlorination cannot destroy all forms of biofouling organisms and, therefore, it is not an absolute barrier to RO membrane biofouling. Chlorine or other oxidants added to the source water will need to be removed before they reach the RO membranes because these oxidants will cause a permanent damage of the RO membrane polymeric structure and ultimately the salt rejection of the membrane elements.

In addition, chorine and other oxidants break down otherwise nonbiodegradable NOM into biodegradable organic compounds and destroy the outer walls of bacterial cells and thereby cause the release of intracellular material into the source water. Because the intracellular material released from algal and bacterial cells as a result of oxidation is rich in easily biodegradable organics, it serves as a food to bacteria that have already colonized the RO membranes or survived the chlorination process.

Long-term exposure to chlorine triggers the production of extracellular polysaccharides or deoxyribonucleic acid by some of the microorganisms in the source water as a defense mechanism, which in turn protects the biofilm-forming bacteria. As a result, although continuous use of chlorine may have a short-term benefit in controlling RO membrane biofouling, in the long term it usually does not solve this problem and, therefore, it is not recommended. Once microorganisms build a protective layer around their cells, the chlorine dosages that are needed to break this layer are very high—usually 200–500 mg/L or more. Typically, it takes 4–6 h for most aquatic organisms to build a protective layer that shields them from the harmful impact of chlorine at dosage of 3–15 mg/L usually added at desalination plant intake facilities. Therefore, after 6 h, further chlorination becomes ineffective and continuous addition of chlorine becomes counterproductive.

It is important to note that within 48 h of discontinuation of chlorination or addition of any other bactericide, most aquatic organisms retain their protective layer. However, after this period, the aquatic organisms begin to absorb the protective layer over their cells and again become vulnerable to destruction by chlorination. Therefore, if chlorination is reapplied again after a 48-h (or longer) period, the bactericidal effect of the disinfectant will be reinstated for at least 4–6 h, after which the bacteria will build another protective layer for their cells.

In the light of the practical observations presented earlier, intermittent chlorination for a period of 4 to 6 h with at least of 48 h between chlorination events has been found to be a more efficient method for RO membrane biofouling control than continuous chlorination. Typical chlorine dosage for periods of intermittent disinfection is between 2 and 5 mg/L and is determined based on a target amount of residual chlorine left in the pretreated water downstream of the cartridge filters (or membrane filters if cartridge filters are not used) of 0.1 mg/L or less.

Such intermittent addition of chlorine to the source water for biofouling control is sometimes referred to as "shock" chlorination. Taking under consideration that for the shock chlorination to be efficient, the time between chlorination events has to be at least 48 h (e.g., 2 days), in most plants, such biofouling control approach is practiced two to three times per week.

Because aquatic organisms (shellfish and bacteria) are very adoptive to ambient conditions, usually a random schedule of intermittent chlorination works better than a preestablished chlorination schedule. The random nature of chlorination is of critical importance for controlling shellfish growth on the internal surface of the plant intake pipes because if chlorine is applied at a preset time schedule, within 1 week, shellfish will begin closing their shells a few minutes before the scheduled chlorination events occur, making the use of chlorine for shellfish growth control highly ineffective.

6.6.2 Chlorine Dioxide

This oxidant is weaker than chlorine but is fairly effective for most aquatic microorganisms and at the same time it is not as aggressive in terms of RO membrane oxidation. Therefore, if used intermittently and in low dosages (0.2–0.5 mg/L) and low pH, it could be applied without the need for dechlorination—that is, it is weak enough not to cause permanent damage to the RO membrane polymeric structure.

The feasibility of using chlorine dioxide for biofouling control is pH dependent and may not cause RO membrane degradation if the pH of the source water is below 8 (Erikson and Dimotsis, 2012). Ambient seawater has a pH in the range of 7.8–8.3 and the use of chlorine dioxide without dechlorination may not always be suitable, unless the source water is pH adjusted.

RO membrane manufacturers differ in their views regarding the use of chlorine dioxide without subsequent dechlorination and the RO membrane supplier for the specific desalination plant has to be consulted regarding application dosages and the need for dechlorination before RO separation if chlorine dioxide is chosen as a biocide.

Because of its short useful life, chlorine dioxide has to be generated at the desalination plant site. However, most of the chlorine dioxide generators available on the market do also produce small amounts of chlorine in the form of HClO and ClO^-, which could oxidize the membrane elements over time. Therefore, the chlorine dioxide system selected for a given RO desalination project would have to be equipped with provisions to remove this residual amount of chlorine or with dechlorination system.

An undesirable site product from the generation of chlorine dioxide is the creation of chlorites (ClO_2). They are carcinogenic and are not removed by the pretreatment process. Chlorites, however, are usually well removed by SWRO desalination membranes and partially removed by brackish and nanofiltration membrane elements. Therefore, the impact of chlorites generated by the chlorine dioxide system on the quality of desalinated water has to be investigated on a project-by-project basis.

6.6.3 Chloramines

Another type of oxidants, which have found wide applications for water reclamation, are chloramines. Chloramines are created by the sequential addition of chorine and ammonia to the source water and have been found to be very efficient because they are weak enough not to cause the oxidation of the RO membrane film.

Although chloramination is a very common and an efficient practice for the biofouling of RO membranes used for treating wastewater or brackish water of low bromide content (i.e., bromide concentration below 0.05 mg/L), it is not recommended for desalination applications where the saline source water has a high content of bromides, such as seawater, for example, where bromide levels are usually in a range of 68–95 mg/L. When blended with ammonia, bromides in seawater create bromamines, which are several times stronger oxidants than chlorine and cause a permanent loss of RO membrane salt rejection within a very short period (usually less than 1 week). Therefore, chloramination is not commonly practiced for seawater desalination applications.

6.6.4 Nonoxidizing Biocides

Most nonoxidizing biocides are proprietary formulas, which have toxic effects on the aquatic species and are usually small-molecular-weight compounds that can penetrate bacterial cell walls and inhibit their metabolism and enzymatic system. Experience with nonoxidizing biocides indicates that they are efficient only when applied at large dosage over a short period of time (Voutchkov, 2014). Long-term applications at small dosages usually have limited benefit because most biofilm-forming bacteria can adapt to nonoxidizing biocides over time.

6.7 ADDITION OF REDUCING COMPOUNDS

Because RO membranes are damaged by exposure to oxidants, such as chlorine, when the saline source water is conditioned with chlorine or other strong oxidants, these oxidants will need to be removed before membrane separation. Typically, RO membranes would degrade irreversibly after an exposure of 200–1000 h at a free chlorine dosage of 1 mg/L. Higher chlorine dosages would shorten this time to only several days. Usually, membrane degradation is expedited if the pH of the water is alkaline.

To protect membrane integrity, residual chlorine or other oxidants, which are not consumed by the source water impurities, are typically removed by the addition of reducing compounds (oxidant scavengers), which react with the oxidants in the seawater and create nonoxidizing side products.

The most commonly applied reduction chemical is sodium metabisulfite ($Na_2S_2O_5$). When introduced to the source water, it creates sodium bisulfite ($NaHSO_3$), which reduces hypochlorous acid to sulfuric acid (H_2SO_4), hydrochloric acid (HCl), and sodium bisulfate (Na_2SO_4), all of which are not oxidizing compounds:

$$Na_2S_2O_5 + H_2O \rightarrow 2NaHSO_3 \tag{6.iii}$$

$$2NaHSO_3 + 2HOCl \rightarrow H_2SO_4 + 2HCl + Na_2SO_4 \tag{6.iv}$$

Approximately, 3.0 mg/L of sodium metabisulfite is needed to remove 1.0 mg/L of free chlorine. Typically, the sodium metabisulfite dosage is optimized based on the reading of the ORP of the saline source water at the entrance to the RO system trains, which should be maintained to less than 200 mV to protect the RO membrane integrity.

Sodium metabisulfite is usually introduced to the feed water immediately after cartridge filtration on the suction side of the booster pumps, which feed pretreated source water into the RO system high-pressure feed pumps. Because the reaction between residual chlorine and sodium bisulfite is practically instantaneous, no separate mixing device is needed. Mixing provided by the booster pumps and high-pressure RO pumps is typically adequate to achieve practically compete removal of chlorine or other strong oxidants before membrane separation.

Overdosing of sodium metabisulfite is not recommended for two reasons: (1) after consuming chlorine and other strong oxidants in the source water, this reducing compound will react with oxygen naturally occurring in the source water and will reduce the content of oxygen in the desalination plant concentrate, which in turn may have a negative impact on the marine environment receiving this concentrate; (2) sodium bisulfite could serve as food to some of the biofouling bacteria growing on the RO membranes and, therefore, could exacerbate membrane biofouling.

Another compound, besides sodium metabisulfite, which can be used as reducing agent, is activated carbon. Activated carbon, however, is more costly and the reaction is slower. Therefore, this compound has not found a wide use for dechlorination.

6.8 PLANNING AND DESIGN CONSIDERATIONS FOR SOURCE WATER CONDITIONING

All chemical feed systems for source water conditioning chemicals have two key components—storage and feed solution preparation tanks and chemical feed pumps. These chemicals are stored on site in areas, which allow their safe loading, containment, and handling—storage buildings or contained storage areas.

In conditions, where the ambient air temperatures may be reduced below freezing for portions of the year or frequently exceed 35°C (95°F), the chemical storage and feed facilities are installed in buildings. Otherwise, they are located under a shed providing protection against direct sunlight exposure.

Most chemicals are delivered to the site as liquid solutions because they are easier to handle and store. Some chemicals, however (i.e., dry ferric chloride, dry ferric sulfate and polymer), are sometimes delivered in the powder form. Chemicals are stored in tanks made of materials suitable for their safe containment. Usually chemical storage tanks are designed for 15—30 days of supply.

Before their use, chemicals are diluted from the concentration at which they are delivered to an application concentration to simplify their pumping and mixing with the source water. Dilution could either be completed in line or in a batch mode in day tanks. In the second case, the diluted chemicals are stored in separate tanks and often referred to as day tanks because they typical store 1 day of the needed volume of chemical at its application concentration. Both the chemical storage tanks and the day tanks are typically equipped with ultrasonic level transducers to monitor storage level. The chemicals are delivered at application solution to the point of their injection using diaphragm-type metering pumps. These pumps usually have adjustable diaphragm positioner/stroke rate that allows controlling the dosage of the delivered chemical.

6.8.1 Overview of Key Water Conditioning Chemicals

Table 6.2 provides a summary of the key characteristics of most commonly used saline source water conditioning chemicals. The product concentration referenced in this table is the concentration of most commonly available commercial products. The user should consult a chemical supplier for the properties of the specific product they are purchasing.

TABLE 6.2 Properties of Commonly Used Conditioning Chemicals

Chemical	Typical Application	Typical Product Concentration (%)	Bulk Density (kg/L)	Application Concentration (%)
Liquid ferric chloride	Coagulation	40	1.42	5
Liquid ferric sulfate	Coagulation	40	1.55	5
Sulfuric acid	pH adjustment	98	1.83	20
Sodium hypochlorite	Biogrowth control	13	1.23	5
Sodium bisulfite	Dechlorination	99	1.48	20
Antiscalant	Scale control	99	1.0	20
Sodium hydroxide	pH adjustment	50	1.525	20

6.8.2 Example Calculations

This section presents the calculations associated with the design of the ferric chloride storage and feed system for 50,000 m^3/day (13.2 MGD) seawater desalination plant designed for an intake flow of 127,910 m^3/day (33.8 MGD) and feed average and maximum feed dosage of 15 and 50 mg/L, respectively.

6.8.2.1 *Chemical Use*

The daily amount of the needed chemical in kg/day is calculated using Eq. (6.2):

$$Q_{dc} = \left[\text{Concentration } (\text{mg/L}) \times \text{Flow } (\text{m}^3/\text{day})\right]/1000 \tag{6.2}$$

For this specific example, the daily average and maximum chemical use is

$$\text{Qavg}_{dc} = \left[15\ \text{mg/L} \times 127,910\ \text{m}^3/\text{day}\right]/1000 = 1919\ \text{kg/day};$$
$$\text{Qmax}_{dc} = \left[50\ \text{mg/L} \times 127,910\ \text{m}^3/\text{day}\right]/1000 = 6395\ \text{kg/day};$$

6.8.2.2 *Chemical Storage Tanks*

The daily chemical use of 1919 kg/day is at 100% chemical concentration. Because the actual commercial product of liquid ferric chloride is delivered to the plant at 40% solution, the amount of this chemical that will need to be stored on site for 30 days is

$$\begin{aligned}\text{Ast} &= [\text{Qavg}_{dc}/(\text{storage concentration }\%)] \times \text{Storage time}\\ &= [1919\ \text{kg/day}/(0.4)] \times 30\ \text{days} = 143,925\ \text{kg of 40\% liquid FeCl}_3\text{ for 30 days}\end{aligned} \tag{6.3}$$

Taking under consideration that the bulk density of this chemical D_d is 1.42 kg/L = 1420 kg/m^3, the actual storage volume is

$$\text{Vst} = \text{Ast}/(D_d) = 143,925\ \text{kg}/1420\ \text{kg/m}^3 = 101\ \text{m}^3$$

Typically, the actual storage tank volume is selected to be 10%–15% larger volume to allow the venting space/free board on the top and sediment accumulation at the bottom, which are inactive storage areas. As a result, the actual tank storage would be $101 \times 1.15 = 116\ m^3$. Assuming four individual storage tanks at 3 m (10 ft) diameter each, the depth of each tank will be $(116\ m^3/4 \text{ tanks}) \times (3.14 \times 3\ m \times 3\ m/4) = 4.1$ m (13.5 ft).

6.8.2.3 *Water Dilution Flow*

The average dilution flow (in L/h) needed to reduce the chemical concentration from its delivery concentration (C_d) to its application concentration (C_a) can be calculated by the following formula:

$$\text{Qd}_{avg} = \text{Qavg}_{dc} \times [(C_d/C_a) - (1/D_d)]/24 \tag{6.4}$$

$$\text{As a result: } Qd_{avg} = 1919\ \text{kg/day} \times [(40\%/5\%) - (1/1.42)]/24 = 583\ \text{L/h}\ (2.6\ \text{gpm})$$

The maximum dilution flow will be calculated for the maximum concentration of chemicals that will have to be delivered at a design flow.

$$Qd_{max} = 6395\ \text{kg/day} \times [(40\%/5\%) - (1/1.42)]/24 = 1945\ \text{L/h}\ (513\ \text{gph}/8.5\ \text{gpm})$$

6.8.2.4 Chemical Metering Pumps

The chemical metering pumps have to designed for the maximum capacity of chemicals they have to deliver:

$$\begin{aligned} Qc_{max} &= Qmax_{dc}/(D_d \times 24\ \text{h}) \\ &= 6395\ \text{kg/day}/(1.42 \times 24) = 188\ \text{L/h}\ (50\ \text{gph}) \end{aligned} \tag{6.5}$$

Typically, for plants of this size and chemical fluctuation, at least 2 + 1 pumps will be provided for chemical feed—i.e., the individual pumps will have a capacity of 94 L/h (25 gph).

References

AWWA, 2007. Manual of Water Supply Practices M46. Reverse Osmosis and Nanofiltration, second ed. American Water Works Association, Denver, Colorado.

Edzwald, K.J., Haarhoff, J.J., 2012. Dissolved Air Flotation for Water Clarification. American Water Works Association and McGraw-Hill.

Edzward, K.J., Haarhoff, J.J., 2011. Seawater pretreatment for reverse osmosis: chemistry, contaminants and coagulation. Water Research 45, 5428−5440. Elsevier.

El-Manhaeawy, S., Hafez, A., 2001. Molar ratios as a useful tool for prediction of scaling potential inside RO systems. Desalination 136, 243−254. Elsevier.

Erikson, P., Dimotsis, G., 2012. Field experience: can chlorine dioxide be used as a biocide in RO plants? IDA Journal, International Desalination Association 4 (1), 14−20 (IDA), First Quarter.

Jamaly, S., Darwish, N.N., Ahmed, I., Hasan, S.W., 2014. A short review of reverse osmosis pretreatment technologies. Desalination 354, 30−38. Elsevier.

Voutchkov, N., April 2014. Desalination Engineering − Operation and Maintenance. McGraw Hill.

Wilf, M., Awerbuch, L., Bartels, C., Mickley, M., Pearce, G., Voutchkov, N., 2007. The Guidebook to Membrane Desalination Technology: Reverse Osmosis, Nanofiltration and Hybrid Systems Process Design, Applications and Economics. Balaban Publishers, Rehovot.

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CHAPTER

7

Sand Removal, Sedimentation, and Dissolved Air Flotation

OUTLINE

7.1 INTRODUCTION

The purpose of sand removal, sedimentation, and dissolved flotation pretreatment systems is to minimize the content of coarse materials such as grit, debris, and suspended solids collected by the plant intake and protect downstream filtration facilities from solids' overload. The source water collected by onshore intakes and shallow open offshore intakes usually does not contain large quantities of sand but could have elevated content of floating and suspended solids. Well intakes typically have a very low content of suspended

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solids but depending on their design and the subsurface soil conditions they could produce source water of elevated sand content, especially when they are brought into service after a long shutdown.

7.2 SAND REMOVAL SYSTEMS

Well-designed desalination plant intakes usually produce source water of low sand and silt content. Therefore, most desalination plants do not have separate sand removal facilities. Small quantities or sand and coarse silt contained in the source water are typically retained by the plant sedimentation or filtration facilities. However, in locations where desalination plant's open intake is located adjacent to an area of significant ship traffic, of turbulent underwater currents, or in area where frequent dredging activities occur, a large amount of sand and silt may enter the desalination plant continuously and would need to be removed in separate facilities.

Sand removal facilities may or may not be followed by sedimentation basins. If the saline source water contains low level of turbidity but large amount of sand, than construction of grit removal facilities instead of clarifiers is more appropriate and cost-effective.

7.2.1 Settling Canals and Retention Basins

Some large onshore intakes are designed with long canal that delivers the source water into retention basin where the water is presettled and sand and large debris are accumulated. The source water from the retention basin overflows into the forebay of the screening facilities/intake pump station of the desalination plant from where it is conveyed into the main pretreatment system. Such canals and retention basins are dredged periodically or are equipped with sediment removal/flushing systems to minimize solids accumulation over time.

While such retention reservoir configurations are suitable for dampening the effect of heavy rain events, winds, currents, ship traffic, and other sources of elevated content of solids in the source water, they may present problems such as excessive algae accumulation, especially if the flow velocity is relatively low and the water remains in the reservoirs for a long time. Usually, it is more prudent and cost-effective to build an offshore intake with depth of at least 8 m (26 ft) below the water surface rather than to build an onshore open intake and settling canal/retention basin system to manage high sand and silt content in the source water.

7.2.2 Strainers

Depending on the size of the desalination plant, grit removal facilities most widely used in practice are 200–500-μm strainers (Fig. 7.1). Strainers of this size can remove sand and silt particles of 0.10 mm or larger.

Strainers are typically applied for small- and medium-size desalination plants [i.e., plants of capacity of 20,000 m^3/day (5.3 MGD) or less], which face large content of sand originating from shallow onshore open intake or intake wells with frequent failures.

FIGURE 7.1 Sand strainers.

7.2.3 Cyclone Separators

Cyclone separators have found application for removal of sand from groundwater, especially for small desalination plants. In such systems the inlet pressure from the intake well pumps drives the source water into the top of the separator chamber at a tangent, causing rotation and formation of vortex in the center of the separator. The vortex action forces the separation of heavy particles from the water. These particles accumulate at the bottom in a collection chamber from where they are periodically removed. In most recent desalination plant designs, cyclone separators and strainers are replaced by microscreens described in Chapter 5.

7.3 SEDIMENTATION TANKS

7.3.1 Introduction

Sedimentation is typically used upstream of granular media and membrane filters when the membrane-plant source water has daily average turbidity higher than 30 NTU or experiences turbidity spikes of 50 NTU or more which continue for a period of over several hours. If sedimentation basins are not provided, large turbidity spikes may cause the pretreatment filters to exceed their solids' holding capacity (especially if granular media filters are used), which in turn may impact filter pretreatment capacity. If the high solids load continues, the pretreatment filters would enter a condition of continuous backwash, which in turn would render them out of service and effectively will shut down the desalination plant operations.

Sedimentation basins for saline source water pretreatment should be designed to produce settled water with turbidity of less than 2.0 NTU and SDI_{15} below 6. To achieve this level of turbidity and silt removal, sedimentation basins are typically equipped with both coagulant (most frequently iron salt) and flocculant (polymer) feed systems. The needed coagulant and flocculant dosages should be established based on jar and/or pilot testing.

If the source water turbidity exceeds 50 NTU, then conventional sedimentation basins are often inadequate to produce settled water of the desired turbidity target level of less than 2 NTU. Under these conditions, sedimentation basins should be designed for enhanced solids removal by installing lamella plates (lamella settlers) or using sedimentation technologies that combine lamella and fine granular media for enhanced solids removal.

It is important to note that sedimentation tanks do not remove oil, grease, and other hydrocarbons to levels that protect downstream RO membranes from colloidal fouling. In addition, clarifiers do not settle well algae contained in the seawater because in most waters such algae are very small in size and are difficult to coagulate.

Typically, the use of enhanced sedimentation technologies is needed for treating source water from open shallow intakes that are under the strong influence of high-velocity currents, river water, or wastewater discharges of elevated turbidity. This condition could occur when the desalination plant intake is located in a river delta area, ship channel, industrial port, or is influenced by a seasonal surface water runoff.

For example, during the rainy season, the intake of the Point Lisas source water desalination plant in Trinidad is under the influence of the Orinoco River currents, which carry a large amount of alluvial solids. As a result, the desalination plant intake turbidity could exceed 200 NTU (Irwin and Thompson, 2003). To handle this high-solids load, the plant source water is settled in lamella clarifiers prior to conventional single-stage dual-media filtration. While this plant has lamella clarifiers, it does not incorporate separate sand removal facilities or strainers upstream of it (Fig 7.2).

FIGURE 7.2 Trinidad desalination plant.

7.3.2 Conventional Sedimentation Tanks

Conventional sedimentation tanks (clarifiers or settlers) are used for removal of suspended solids prior to filtration when the source water turbidity exceeds 20 NTU but is lower than 50 NTU. These clarifiers cannot produce water adequate for direct feed to the RO membranes and the clarified effluent will have to be filtered by granular media or membrane filtration prior to desalination.

Since conservatively designed membrane filtration systems can handle up to the same level of source water turbidity (e.g., 50 NTU) without presedimentation or other upstream treatment of the source water, in this case, it is preferable to use a conservatively designed single-stage membrane pretreatment system (e.g., MF or UF system with design flux of 40 lmh or less) instead of constructing a two-stage pretreatment which consists of clarification followed by higher rate granular media filtration or membrane filtration (e.g., MF or UF system with design flux of 65 lmh or less).

Conventional sedimentation tanks could be configured as rectangular or circular structures. To date, rectangular sedimentation tanks have found most common application for pretreatment of saline source waters because of their lower costs and slightly superior performance.

Key design criteria for this type of tanks are presented below:

Minimum number of tanks	Four
Water depth	3.0–4.5 m (10–15 ft)
Mean flow velocity	0.3–1.1 m/min (1.0–3.6 m/min)
Detention time	2–4 h
Surface loading rate (clarifier area)	1.0–2.0 m^3/m^2 h (0.4–0.8 gpm/ft^2)
Length-to-width ratio	Minimum of 4:1
Water depth-to-length ratio	Minimum of 1:15
Sludge collector speed	0.4–0.8 m/min (for collection path)

7.3.3 Lamella Sedimentation Tanks

Lamella sedimentation tanks (clarifiers or settlers) usually have superior performance and three to four times smaller footprint as compared to conventional clarifiers and they can handle up to four times higher source water turbidity (e.g., up to 200 NTU). Therefore, they have found wider application for saline water pretreatment than conventional sedimentation basins. These clarifiers contain plastic lamella plate modules installed in the upper portion of the clarifier tanks (see Fig. 7.3), which enhance the sedimentation process by shortening the path of solid particles to the bottom of the clarifiers.

Lamella clarifiers can be configured both as rectangular or circular structures. However, rectangular lamella clarifiers have found the widest application for pretreatment of saline

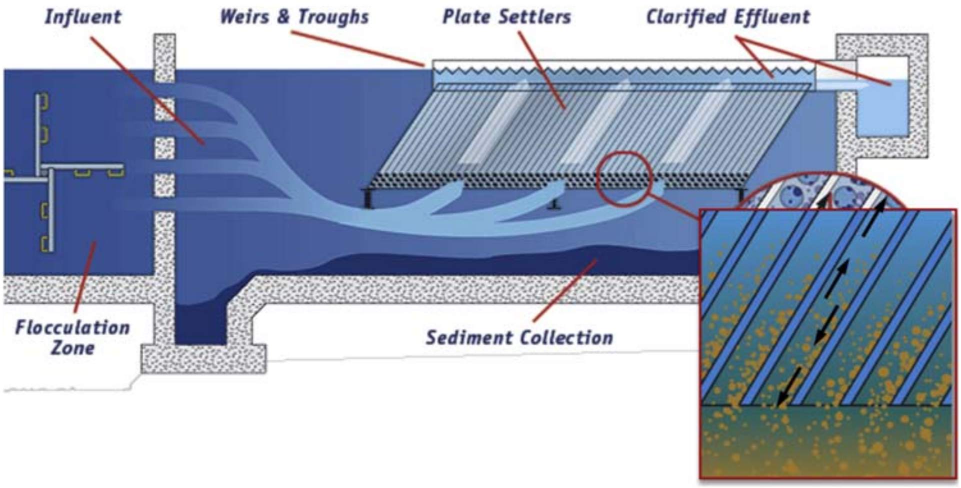

FIGURE 7.3 Schematic of lamella sedimentation tank.

source waters, especially for removal of high turbidity spikes from seawater. Key design criteria for this type of sedimentation tanks are presented below:

Minimum number of tanks	Two
Water depth	3.5–5.0 m (11.5–16.4 ft)
Mean flow velocity	0.3–1.1 m/min (1.0–3.6 m/min)
Detention time (in the lamella module)	0.2–0.4 h
Surface loading rate (lamella module)	1.0–2.0 m^3/m^2 h (0.4–0.8 gpm/ft^2)
Surface loading rate (clarifier area)	4.0–8.0 m^3/m^2 h (1.6–3.2 gpm/ft^2)
Sludge collector speed	0.4–0.8 m/min (for collection path)

Lamella modules used in the high-rate settlers are proprietary products and design engineer should consult equipment manufacturers regarding lamella module configuration, number and size of modules as well as design surface loading rate and depth of the sedimentation tank.

7.3.4 Lamella Settler—Design Example

The example below illustrates general design criteria for a lamella settler for pretreatment of the source water for a 50,000 m^3/day (13.2 MGD) seawater desalination plant designed for a 43%-SWRO system recovery. The source water turbidity reaches levels of 80 NTU during storm events, which may last several days and therefore, the plant is equipped with a combination of lamella settlers followed by a single-stage-dual granular media filters. The source water quality is relatively low in terms of hydrocarbon content—with maximum concentration of 0.04 mg/L or less. The source water is not frequently exposed to algal blooms, and when such events occur periodically they are of low intensity—with algal content of the source water of 20,000 cells/L, or less.

The plant filter backwash flow is 8.5% of the intake flow and lamella clarifier waste stream (sludge) flow is 0.6% of the intake plant flow. The pretreatment system is designed to operate with addition of coagulant, flocculant, and pH adjustment of the source water flow.

The lamella settler system is designed to treat a total of 127,910 m^3/day (33.8 MGD)–[50,000 m^3/day/43%)/(100%–(8.5% + 0.6%) = 127,910 m^3/day]. Key design parameters of this system are shown in Table 7.1.

In Table 7.1 the sizes (width, length, and depth) and the net surface area per lamella module are provided by the lamella supplier. The surface loading rate is calculated by dividing the total feed flow to the lamella settlers by the total surface area of all lamella modules. This loading rate should be comparable to the loading rate used for design of conventional settlers [i.e., 1.0–2.0 m^3/m^2 h/(0.4–0.8 gpm/ft^2)].

However, if the surface loading rate is calculated by dividing the feed flow by the physical total surface area of the lamella settlers, this loading rate will be approximately five times higher in this example (7.7 m^3/m^2 h/1.42 m^3/m^2 h = 5). This comparison illustrates the fact that lamella settlers are significantly more space-efficient and economical than conventional settling tanks. Therefore, they have found wider implementation for desalination plant pretreatment than conventional clarifiers.

TABLE 7.1 Example of Lamella Settler Pretreatment System for 50,000 m^3/day (13.2 MGD) Desalination Plant

Component/Parameter	Specifications/Design Criteria
FEED WATER	
• Design flow rate, m^3/day (MGD)	127,910 (33.8 MGD)
• Turbidity, NTU	0.5–80
• $SDI_{2.5}$	6–16
• Algal content, cells/L	<20,000
DESIGN CHEMICAL DOSAGES	
• Ferric chloride, mg/L	15 (0.5–50 mg/L)
• Cationic polymer, mg/L	0.5 (0.0–1.0 mg/L)
• Sulfuric acid, mg/L—target pH—6.7	8 (0–30 mg/L)
• Number of settler tanks	4
• Number of lamella modules per tank	4
• Width of lamella modules, m (ft)	1.24(4.1 ft)
• Length of lamella modules, m (ft)	8.67 (28.4 ft)
• Depth of lamella modules, m (ft)	2.588 (9.8 ft)
• Net surface area per lamella module	235 m^2 (2528 ft^2)
• Surface loading rate/module area	1.42 m^3/m^2 h/(0.6 gpm/ft^2)
• Setter tank surface area	43 m^2 (463 ft^2)
• Settler tank surface loading rate	7.7 m^3/m^2 h (3.1 gpm/ft^2)
• Water depth	5.5 m (20.8 ft)

7.4 DISSOLVED AIR FLOTATION CLARIFIERS

7.4.1 Introduction

Dissolved air flotation (DAF) technology is very suitable for removal of floating particulate foulants such as algae, oil, grease, or other contaminants that cannot be effectively removed by sedimentation or filtration. DAF systems can typically produce effluent turbidity of <0.5 NTU and can be combined in one structure with dual-media gravity filters for sequential pretreatment of seawater.

DAF process uses very small size air bubbles to float light particles and organic substances (oil, grease) contained in the source water (Fig. 7.4). The floated solids are collected at the top of the DAF tank and skimmed off for disposal, while the low-turbidity source water is collected near the bottom of the tank.

A typical DAF system consists of the following key components: coagulation and flocculation chambers; air saturation zone, flotation chamber, air saturation system, and clarified water recycling system (see Fig. 7.4).

As indicated in Chapter 6, coagulation and flocculation chambers are designed to enlarge the size of the particulate solids naturally contained in the saline source water in order to enhance their removal in the flotation chamber.

After coagulation and flocculation, the saline source water is mixed with clarified water, which is saturated with air to expose the particles in the saline source water in contact with the air bubbles that will carry them to the surface of the clarifier. The clarified water is recycled from the effluent end of the DAF units and is pumped through an air saturator at a rate of 10%–15% of the flow rate of the source water entering the DAF clarifier.

Typically, 8–12 g/m^3 of air has to be introduced for effective DAF process. As a rule of thumb, the air dosage is determined from the weight ratio of air to suspended solids of 0.12:1.0. The air is dissolved in the recycle water under pressure of 6–8 bars in pressure vessels (air saturators) equipped with an educator on the inlet side for adding air or with a packed column. In packed column saturators the depth of the packing is usually

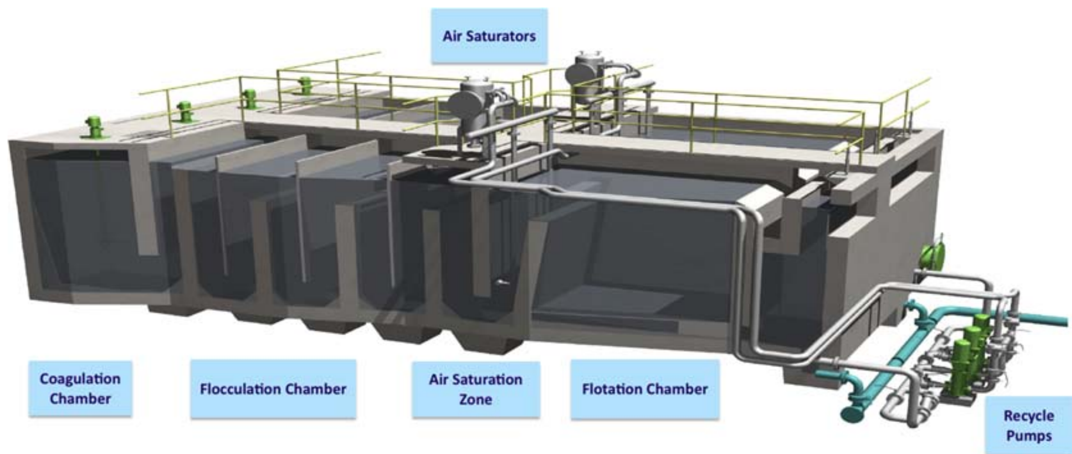

FIGURE 7.4 General schematic of DAF clarifier.

0.8−1.2 m of polypropylene rings. The design hydraulic loading rate of the air saturators is 60−80 m^3/m^2 h (25−33 gpm/ft^2). Saturator efficiency for educator systems is 65%−75% while for packed column units is 90%−95%. Air saturated water is recycled to the entrance of the flotation chamber through a series of nozzles to release the air bubbles unto the coagulated saline water in a white water curtain and maximize the contact of the air bubbles and the solid particles. Most DAF systems available on the market have air nozzles that deliver air bubbles in sizes of 10−100 μm and average between 30 and 50 μm. The diffusers are spaced at 0.1−0.3 m (0.3−1.0 ft).

The flotation chamber is a rectangular tank designed for a surface loading rate of 20−35 m^3/m^2 h (8−14 gpm/ft^2). Typical tank depth is 2−3 m (7−10 ft) and the preferred length to width ratio is 1.5−2.5:1. In the flotation chamber the air bubbles carry the solids attached to them to the surface of the chamber where they accumulate and form a sludge layer (see Fig. 7.5).

The sludge layer (blanket) collected on the surface of the DAF tanks is removed by hydraulic means or by mechanical scrapers and directed for subsequent solids-handling system. Based on existing practices, the coagulation process is often enhanced by an acid feed that allows to adjust the pH of the source water to optimum level for formation of large, strong, and light flocs.

Some of the solids contained in the source water may settle rather than float during the DAF clarification process. Such solids accumulate at the bottom of the DAF tanks over time and are periodically removed from there via bottom sediment collection and evacuation system. Under normal operational condition, the sludge blanket level has solids concentration of 1%−3% and is 0.5−1.0 m (1.7−3.3 ft) thick.

Full-scale experience at SWRO desalination plants, to date, shows that DAF systems do not perform well if the source water turbidity is less than 5 NTU. In such low-turbidity conditions, the DAFs are shut down and bypassed. Typically below this turbidity level, the source water solids are very difficult to float and to form thick solids blanket. If solids blanket is not formed within 30−45 min, many of the air bubbles that have attached particles will burst and

FIGURE 7.5 Sludge layer on the surface of DAF clarifier.

the particles will settle down. Because clarified water is collected at the bottom of the DAF tanks, settling of solids is undesirable and often results in turbidity spikes in the clarified water higher than the turbidity of the source water entering the DAF clarifiers.

Practical experience, to date, shows that the most common problems associated with DAF clarifier performance are caused by:

1. Inadequate coagulation and flocculation of the suspended solid particles in the source water due to the low charge of these particles;
2. Mismatch between the size of the source water particles and that of the DAF air bubbles;
3. Low air bubble release pressure that does not provide adequate energy for the bubbles to overcome the drag forces of the dense saline water and results in bubble burst in mid-water column rather than carrying of the solids to the top of the DAF clarifiers;
4. Short-circuiting caused by lack of baffling devices inside most proprietary DAF systems available on the market at present;
5. Ineffective sludge removal—especially for DAF clarifiers where the sludge blanket is removed by hydraulic overflow rather than by mechanical sludge collection mechanism.

Overdosing of coagulant and flocculant are often the two most common operational challenges of DAF systems when the source water turbidity is low (<5 NTU). In such cases, DAF shutdown and bypass is a more desirable operational strategy than adding coagulant, because this coagulant does not have particles to react with and ultimately is conveyed to and captured by the downstream plant filtration facilities (filters, cartridges, and RO membranes) and causes their premature fouling.

The surface loading rate for removal of light particulates and floatable substances by DAF is several times higher than that needed for conventional sedimentation. Another benefit of DAF as compared to conventional sedimentation is the higher density of the formed residuals (sludge). While residuals collected at the bottom of sedimentation basins typically have concentration of only 0.3%–0.5% solids, DAF residuals (which are skimmed off the surface of the DAF tank) contain solids' concentration of 1%–3%.

In some full-scale applications, the DAF process is combined with granular media filters to provide a compact and robust pretreatment of source water with high algal and/or oil and grease content. Although this combined DAF/filter configuration is very compact and cost-competitive, it has three key disadvantages:

1. complicates the design and operation of the pretreatment filters;
2. DAF loading is controlled by the filter loading rate and, therefore, DAF tanks are typically oversized; and
3. flocculation tanks must be coupled with individual filter cells.

7.4.2 Planning and Design Considerations

The feasibility of DAF application for pretreatment of saline surface waters is determined by the source water quality and is governed by turbidity concentration and overall lifecycle pretreatment costs. The DAF process can handle source water with turbidity of up to 50 NTU. Therefore, if source water is impacted by high turbidity spikes or heavy solids (usually related to seasonal river discharges or surface runoff), then DAF may not be a suitable pretreatment option. In most algal bloom events, however, source water turbidity almost never exceeds 50 NTU, so the DAF technology can handle practically any algal bloom event.

Although DAF systems have much smaller footprint than conventional flocculation and sedimentation facilities, they include a number of additional equipment associated with air saturation and diffusion, and with recirculation of a portion of the treated flow, and therefore, their construction costs are typically comparable to these of conventional sedimentation basins and higher than lamella settlers of the same capacity.

Usually, the O&M costs of DAF systems are higher than these of sedimentation tanks due to the higher power use for the flocculation chamber mixers, air saturators, recycling pumps, and sludge skimmers. The total power use for DAF systems is usually 0.05–0.075 kWh/10,000 m^3/day of treated source water, which is significantly higher than that for sedimentation systems—0.01–0.03 kWh/10,000 m^3/day of treated water.

DAF clarifiers for seawater applications have several key differences as compared to these for fresh surface waters: (1) they have to remove smaller size algal cells and, therefore, have to have diffusers that create smaller size bubbles; (2) seawater has significantly higher density than freshwater and therefore requires operation at higher air pressures to provide adequate solids removal; (3) seawater particles and algae have lower charge than freshwater solids, which makes them more difficult to coagulate and flocculate and requires larger contact chambers than there of freshwater DAF systems. The differences between seawater and freshwater applications of DAF are discussed in greater detail in the following publication (Edzwald and Haarhoff, 2012).

Practical experience shows that DAF system design that is not adopted to the specific water quality challenges of seawater pretreatment often does not meet performance expectations of high algal content removal, especially during normal (non-algal bloom) source water conditions when the content of algae in the water is low (<500 cells/L) and source water turbidity is <5 NTU.

Smaller ocean water algal particles require smaller size air bubbles for effective removal. The optimum range of the size of the air bubbles is directly related to the predominant size of algal cells in the source water, which can be determined by the completion of algal profiles of this water.

Most existing commercially available DAF technologies have been created for wastewater and freshwater applications and, therefore, the majority of the bubbles generated by their diffuser systems are in a range of 30 and 100 μm. Often, the type of algae dominating during red tide events in the Persian Gulf for example have an order of magnitude smaller size than freshwater algae—i.e., they are pico-plankton (0.2–2 μm) and nano-plankton (2–20 μm). If such small size plankton is the main cause of algal blooms, conventional DAF systems designed to remove larger-size (40–100 μm) freshwater algal cells are likely to have limited removal efficiency. In addition, as indicated in a recent study (Zhu and Bates, 2012) commonly applied source water chlorination practice may result in algal cell destruction and further diminish the benefits associated with DAF pretreatment.

Because of its higher density and viscosity than fresh water, seawater requires 20%–30% higher air saturation and introduction of the air at higher pressures. As a result, while the required pressure of the feed water recycled to the DAF for freshwater is 4–6 bars, the actual pressure needed for seawater DAF operations to form large percentage of smaller size bubbles is typically 6–8 bars.

Low-charge particles in seawater as compared to high charge particles in fresh water require longer contact time and better mixing in the coagulation and flocculation chambers to form large-enough flocks for effective removal in the flotation zone of the DAF clarifiers. With freshwater particles and algae that carry strong negative charge, addition of coagulant (ferric chloride or sulfate) that carries positive charge will result in creation of large flocks in a very short time, based on strong opposite-charge attraction.

With fine uncharged seawater particles, the main mechanism of flock formation is direct physical contact with the coagulant particles, which requires more time, especially if the solids concentration is very low (i.e., at low feed water turbidity). As a result, a typical 5–7-min contact time used to design the flocculation chambers of DAFs for freshwater applications will be insufficient for adequate size flock formation of seawater particles—a contact time of at least 10–15 min is needed for DAF systems processing seawater. One proprietary DAF system designed for seawater pretreatment applications addresses this challenge by installing a device referenced as "the Turbomix," which increases particle collision and flocculation (Gaid, 2012).

The key design criteria for the coagulation and flocculation chambers, flotation chamber, and recycle system of typical DAF clarifier are presented below:

IN-LINE STATIC MIXER (OR COAGULATION CHAMBER)	
Velocity gradient (G × T)	500–1600 s^{-1}
FLOCCULATION CHAMBER	
Contact time	10–20 min
Flocculation chambers in series	2–4
Water depth	3.5–4.5 m (11.5–15.0 ft)
Type of mixer	Vertical shaft with hydrofoil blades
Blade area/Tank area	0.1%–0.2%
Shaft speed	40–60 rpm
FLOTATION CHAMBER	
Minimum number of tanks	2 (same as filter cells if combined with filters)
Tank width	3–10 m (10–33 ft)
Tank length	8–12 m (26–39 ft)
Tank depth	2.5–5 m (8–16 ft)
Length-to-width ratio	1.5–2.5 to 1
Surface-loading rate	20–35 m^3/m^2 h (8–14 gpm/ft^2)
Hydraulic detention time	10–20 min
TREATED WATER RECYCLE SYSTEM	
Recycling rate	10%–15% of intake flow
Air loading	8–12 g/m^3
Saturator loading rate	60–80 m^3/m^2 h (25–33 gpm/ft^2)
Operating pressure	6.0–8.0 bars (87–116 psi)

Since most existing proprietary DAF systems were developed for removal of freshwater algae that usually are an order of magnitude larger in size than seawater algae, the air bubble systems of existing DAFs are designed to generate bubbles of size that are significantly larger than optimum. This flaw could be addressed by modification of the air-bubble nozzle system to produce smaller size bubbles and fit the size of the smallest size of algae, which occur in the ambient saline source water during the algal bloom season. The most appropriate bubble size could be determined based on source water particle size and algal speciation analyses.

Pressurizing the air-saturated clarified DAF stream to higher levels (8–10 bars vs. standard 6–8 bars) would improve DAF operation but typically would require the redesign of the DAF's air-saturation system.

Short-circuiting that occurs in some of the exiting proprietary DAF systems could be addressed by the installation of baffles within the DAF tanks, which break the flow pattern and increase the contact time between the air bubbles and source water particles.

If the DAF system has an ineffective sludge removal system, which does not allow easy evacuation of particles collected on the tanks' bottom, such tanks would need to be taken out of service and cleaned periodically. Otherwise, the solids accumulated at the bottom of the DAF tanks will begin to digest anaerobically and disintegrate into finer much more difficult to filter particles, which in turn, will deteriorate the performance of the downstream filtration facilities.

It is important to note that DAF clarifiers usually do not remove significant amount of alluvial organics and biopolymers, i.e., UV_{254} and DOC are not likely to be reduced by DAF. This flotation process removes some of the particulate organics, mainly contained in the source water algae and bacteria attached to them. Such removal rate would be highly dependent on the size of algae in the source water and could vary between 5% and 20%.

DAF process with built-in filtration (DAFF) is used at the 136,000 m^3/day (36 MGD) Tuas seawater desalination plant in Singapore (Kiang et al., 2007). This pretreatment technology has been selected for this project to address the source-water quality challenges associated with the location of the desalination plant's open intake in a large industrial port (i.e., oil spills) and the frequent occurrence of red tides in the area of the intake.

The source seawater has total suspended solids concentration that can reach up to 60 mg/L at times and oil and grease levels in the seawater that could be up to 10 mg/L. The facility uses 20 built-in filter DAF units, two of which are operated as standby. Plastic covers shield the surface of the tanks to prevent impact of rain and wind on DAF operation as well as to control algal growth. Each DAF unit is equipped with two mechanical flocculation tanks located within the same DAF vessel. Up to 12% of the filtered water is saturated with air and recirculated to the feed of the DAF units.

A combination of DAF followed by two-stage dual-media pressure filtration has been successfully used at the 45,400 m^3/day (12 MGD) El Coloso SWRO plant is Chile, which at present is one of the largest SWRO desalination plants in operation in South America. The plant is located in the City of Antogofasta, where seawater is exposed to year-round red-tide events, which have the capacity to create frequent particulate fouling and biofouling of the SWRO membranes (Petry et al., 2007).

The DAF system at this plant is combined in one facility with a coagulation and flocculation chamber. The average and maximum flow rising velocities of the DAF system are 22 and

33 m^3/m^2 h (9–14 gpm/ft^2), respectively. This DAF system can be bypassed during normal operations and is typically used only during algal bloom events.

The downstream pressure filters are designed for surface loading rate of 25 m^3/m^2 h (10.2 gpm/ft^2). Ferric chloride at a dosage of 10 mg/L is added ahead of the DAF system for source water coagulation. The DAF system reduces source seawater turbidity to between 0.5 and 1.5 NTU and removes approximately 30%–40% of the source seawater organics.

Another example of large seawater desalination plant incorporating DAF system for pretreatment is the 200,000 m^3/day (53 MGD) Barcelona facility in Spain (Sanz and Miguel, 2013). The pretreatment system of this plant incorporates 10 high-rate SeaDAF units equipped with flocculation chambers, followed by 20 first-stage gravity dual-media filters and 24 second-stage pressurized dual-media filters. The purpose of the DAF system is to mainly remove algae and to reduce source-water organic content. Because the plant intake is located near a large port area, the DAF unit is also designed to handle potential oil contamination in the source water.

The intake of the desalination plant is located 2200 m from the coast and 3 km away from the entrance of a large river (Llobregat River) to the ocean, which carries significant amount of alluvial/NOM reach organics. After coagulation with ferric chloride and flocculation in flash-mixing chambers, over 30% of these organics are removed by the DAF system.

7.4.3 DAF—Design Example

This example DAF clarifier is designed for seawater desalination plant with production capacity of 50,000 m^3/day (13.2 MGD) with SWRO system with 43% recovery—the same conditions used for sizing of the lamella settlers discussed in Section 7.3.2. The plant source water turbidity reaches levels of 80 NTU during storm events and up to 40 NTU during algal blooms. This source water is planned to be treated by a combination of DAF clarifier and granular dual media filter.

The plant filter backwash flow is 5% of the intake flow, and lamella clarifier waste stream (sludge) flow is 0.5% of the intake plant flow. Maximum algal count in the source water is 60,000 cells/L, and the hydrocarbon levels can reach levels of 0.5–1.0 mg/L. The pretreatment system is designed to operate with addition of coagulant, flocculant, and pH adjustment of the source water flow.

The pretreatment system will need to be designed to treat a total of 127,910 m^3/day. Source water coagulation will be completed by in-line static mixers. Design parameters of the DAF clarifier are summarized in Table 7.2.

7.5 CONSTRUCTION COSTS OF LAMELLA SETTLERS AND DAF CLARIFIERS

The graph of Fig. 7.6 depicts the construction costs of lamella settlers and DAF clarifiers. As can be seen on Fig. 7.6, lamella settlers are less costly than DAF clarifiers for the same volume of pretreated source water. However, lamella settlers do not remove algae and hydrocarbons well and, therefore, often DAF clarifiers are the preferred primary treatment step of choice for desalination plants using open intakes.

TABLE 7.2 Example of DAF Clarification System for 50,000 m^3/day (13.2 MGD) SWRO Desalination Plant

Component/Parameter	Specifications/Design Criteria
FEED WATER	
• Design flow rate, m^3/day (MGD)	127,910 (33.8 MGD)
• Turbidity, NTU	0.5–80
• SDI	6–16
DESIGN CHEMICAL DOSAGES	
• Ferric chloride, mg/L	15 (0.5–50 mg/L)
• Cationic polymer, mg/L	0.5 (0.0–1.0 mg/L)
• Sulfuric acid, mg/L – Target pH – 6.7	8 (0–30 mg/L)
FLOCCULATION TANK	
• Number per DAF tank	1
• Total number	6
• Width, m (ft)	4.85 (15.9 ft)
• Length, m (ft)	8.0 (26.4 ft)
• Depth, m (ft)	4.9 (16.1 ft)
• Number of mixers per tank	2
• Total retention time	13 min
DAF TANKS	
• Number	6
• Width, m (ft)	4.85 (15.9 ft)
• Length, m (ft)	10.0 (32.8 ft)
• Depth, m (ft)	4.9 (16.1 ft)
• Total surface area	291 m^2 (3132 ft^2)
• Surface contact zone area	57 m^2 (620 ft^2)
• Surface flotation area	234 m^2 (2548 ft^2)
• Surface loading rate @ 15% recycle	26.2 m^3/m^2 h (10.7 gpm/ft^2)
CIRCULATION PUMPS	
• Number	6 + 1
• Capacity	133 m^3/h (680 gpm)
• Delivery head	7 bars (100 psi)
AIR COMPRESSOR	
• Number	6 + 1
• Capacity	15 m^3/h (66 gpm)
• Delivery pressure	10 bars (142 psi)
DAF SATURATOR TANKS	
• Number	6
• Capacity per tank	100 m^3/h (440 gpm)
• Net volume per tank	4 m^3 (1060 gallons)

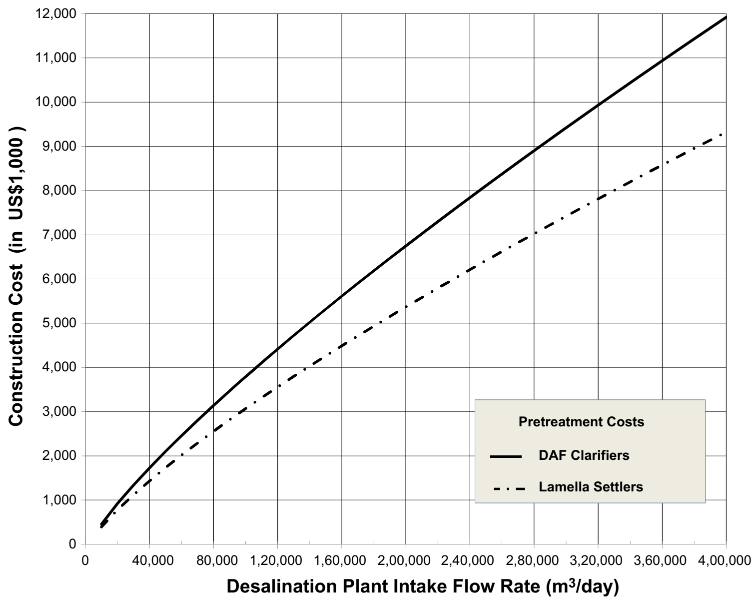

FIGURE 7.6 Construction costs of lamella settlers and DAF clarifiers.

Based on readings from Fig. 7.6, the estimated costs of the example lamella settlers and DAF clarifiers for 50,000 m^3/day SWRO desalination plant (127,910 m^3/day of intake water), described in Sections 7.3.2 and 7.4.3, are US$3.7 and 4.7 million, respectively. These costs are in US$2017 dollars.

References

Edzwald, K.J., Haarhoff, J.J., 2012. Dissolved Air Flotation for Water Clarification. American Water Works Association and McGraw-Hill.

Gaid, A.K., May 2012. Overcoming Red Tide. H2O Magazine 26–34.

Irwin, K.J., Thompson, J.D., 2003. Trinidad SWRO – Orinoco Fluctuations Fail to Make Filters Falter, Desalination and Water Reuse, International Desalination Association. November/December, 13/3, pp. 12–16.

Kiang, F.H., Young, W.W.L., Ratnayaka, D.D., 2007. The Singapore solutions. Journal of Civil Engineering 77 (1), 62–69. American Society of Civil Engineers, January.

Petry, M., Sanz, M.A., Langlais, C., Bonnelye, V., Durand, J.P., Guevara, D., Nardes, W.M., Saemi, C.H., 2007. The El Coloso (Chil1e) reverse osmosis plant. Desalination 203, 141–152. Elsevier.

Sanz, M.A., Miguel, C., 2013. The role of SWRO Barcelona-Llobregat plant in the water supply system of Barcelona area. Desalination and Water Treatment 51, 111–123. Taylor & Francis.

Zhu, I.X., Bates, B.J., 2012. IDA Journal, International Desalination Association 4 (1), 34–37. First Quarter.

CHAPTER

8

Granular Media Filtration

OUTLINE

Pretreatment for Reverse Osmosis Desalination
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8.1 INTRODUCTION

Granular media filtration, also often referred to as conventional filtration, is the most commonly used source water pretreatment process for reverse osmosis (RO) desalination plants today (other than cartridge filtration). This process includes filtration of the source water through one or more layers of granular media (e.g., anthracite coal, silica sand, garnet). Conventional filters used for saline water pretreatment are typically rapid single-stage dual-media (anthracite and sand) units (Fig. 8.1).

However, in some cases where the source water contains high levels of organics (TOC concentration higher than 6 mg/L) and suspended solids (monthly average turbidity exceeds 20 NTU), two-stage filtration systems are applied. Under this configuration, the first filtration stage is mainly designed to remove coarse solids and organics in suspended form. The second-stage filters are configured to retain fine solids and silt, and to remove a portion (20%–40%) of the soluble organics contained in the saline water by biofiltration. All pretreatment systems are designed to meet the filtered-water quality specifications listed in Table 8.1. These specifications

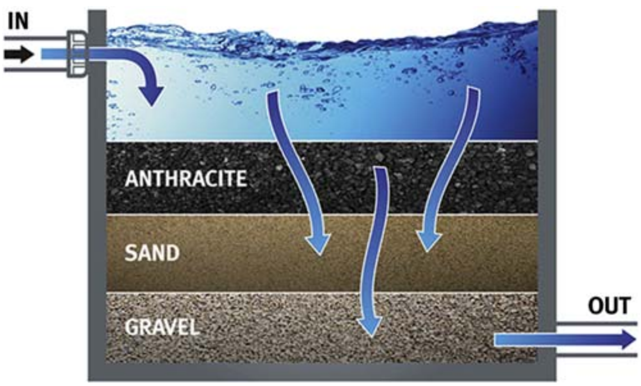

FIGURE 8.1 Typical configuration of single-stage dual-media filter.

TABLE 8.1 Target Pretreated Water Quality

Parameters	Concentrations/Levels	Units
Turbidity (daily avg./max)	<0.1/0.5	NTU
SDI_{15}	<3 (at least 95% of the time)	No unit
	<5 (at all times)	
Total organic carbon	<1.0	mg/L
pH (min)/(max)	4.0/9.0	pH units
ORP	Less than 200	mV
Chlorine residual	≤0.02	mg/L
Total hydrocarbons	≤0.04	mg/L

aim to protect the downstream RO membranes from premature damage, loss of permeability and productivity and salt rejection, and excessive frequency of cleaning and replacement.

8.2 THE FILTER OPERATION CYCLE

Granular media filtration is a cyclical process that incorporates two sequential modes of operation: (1) source water processing (filtration) mode; and (2) filter media backwash mode.

8.2.1 Source Water Filtration

During the filtration cycle the water moves in the direction of size gradation of the media and suspended solids and organics in the water are retained on and around the media grains. Depending on the direction of the flow through the media, filters are classified as upflow (from bottom to top) or downflow (from the top to the bottom). As the filtration cycle is continuous, the solids removed from the source water are accumulated in the filtration media and slowly fill up the media cavities around the filter grains.

As the feed water is filtered through the media, the content of solids and silt in this water decreases. Usually, well-operating filters remove 90%–99% of the suspended solids, particulate organics, and silt contained in the saline source water. Some of the aquatic microorganisms (e.g., algae, bacteria, and some viruses) are also retained on the filter media. These microorganisms consume a portion of the dissolved organics in the source water as it passes through the filtration bed. The efficiency of the filters to remove dissolved organics is a function of three main factors: media depth, surface loading rate, and temperature. Removal of dissolved organics by the filters increases with depth and temperature, and with the decrease of the filter loading rate. Elevated temperature increases the dissolved organic content removal because it accelerates the growth rate of the microorganisms retained in the media. The higher depth and lower filter loading rate increase the contact time between the microorganisms in the media bed and the dissolved organics in the water, creating opportunity for their enhanced assimilation.

The solids and microorganisms retained in the pore volume between the filter media grains reduce this volume over time and create hydraulic losses through the filter media (filter bed resistance). Most filters used for saline water pretreatment operate at constant filtration rate, which means that the feed pressure of these filters increases over the duration of the filtration cycle to compensate for the headlosses in the filter bed caused by accumulation of solids. Once the filter media headlosses reach a certain preset maximum level, the filtration cycle is completed, the filter is taken out of service and media backwash is activated to remove the solids accumulated in the media.

8.2.2 Filter Media Backwash

Granular media filters are configured in individual cells (units), which operate independently. Once a given filter cell reaches a preset target headloss through the filtration bed (for constant flow filters), this cell is backwashed using filtered saline source water or concentrate from the RO-membrane system. Backwashing is always completed in a direction opposite to the direction of the source water flow during the filtration cycle to remove the solids out of the media.

Filter-cell backwash frequency is usually once every 24–48 h. Deeper filters of larger surface area have higher capacity to retain solids and, therefore, usually have longer filtration cycles. The actual length of the filtration cycles will depend not only on the depth and loading rate (e.g., the retention time) of the filters but also on the content and size of solids in the saline source water.

Spent (waste) backwash volume is typically 2%–6% of the intake source water. Use of RO concentrate instead of filtered effluent to backwash filter cells allows reducing backwash volume and energy needed to pump source water to the desalination plant. However, because of the corrosive nature of the concentrate, backwashing with RO concentrate is not commonly practiced.

During backwash of downflow filters, the backwash water flows upward through the filter media, scours the filter grains, removes the solids accumulated in the filter bed, expands the bed, and transports the removed solids toward the backwash troughs for their evacuation out of the filter cell.

Practical experience shows that backwashing of filter media grains smaller than 0.8 mm with water only is difficult. Therefore, at present, a typical backwash cycle includes a combination/sequence of air and water washing of the filtration media.

Typically, the backwash sequence includes three steps: (1) water wash; (2) air wash; and (3) second water wash of the media. The first water wash aims to mainly dislodge the coarser solids accumulated in the filter media pores. The following air wash creates greater turbulence and enhances particle scrubbing allowing to also dislodge the finer solids and some of the bacterial film accumulated on the grains of the filtration media. The air wash is followed by another water wash, which aims to remove the solids and air accumulated in the media and to prepare the filtration bed for another filtration cycle. The length of each of the water and air backwashing cycles is a function of the content of solids in the source water and the depth of the filtration media and typically it is between 5 and 8 min.

Filter backwash is initiated either at preset interval of time, or at predetermined maximum filter bed headloss. Sometimes, filter backwash is also initiated if the filtered water quality begins to deteriorate steadily and exceeds the acceptable threshold for RO-membrane processing in terms of filtered water turbidity (usually set to 0.5 NTU) and/or SDI_{15} (usually established at 4 or 5).

Once a given filter cell/vessel is taken out of service for backwash, the rate and length of the water–air–water backwash sequences is determined based on the length of the filtration cycle and the source water turbidity, which occurred for the duration of the filtration cycle. The shorter the filtration cycle and the higher the feed water turbidity, the longer the filter backwash cycle.

The optimum length of the entire filter backwash cycle is determined by collecting grab sample of spent backwash water from the surface of the filter (for gravity filters) or from the spent-filter backwash pipe (for pressure filters) every 2 min after the initiation of the filter backwash and measuring the turbidity of this waste stream with handheld turbidimeter. Based on a rule of thumb derived from practical experience, the filter backwash cycle should be discontinued when the turbidity of the spent filter backwash water is reduced down to less than 15 NTU. Further continuation of the backwash cycle will result in ovescrubbing of the filter media and increased maturation period which are undesirable.

After the backwash of a given filtration cell is completed and the cell is returned to normal filtration cycle, the filtered water produced in the first 15–45 min usually exceeds the levels of turbidity and/or SDI acceptable for processing of this water through the downstream RO system. During this initial period after backwash, the filter media rebuilds its ability to filter out fine solids via the internal accumulation of solids around the media grains. This time for recovery of the filter cell performance after backwash is often referred to as maturation period. Because the quality of the filtered water produced during the maturation period is not adequate for desalination, typically, this water is wasted (e.g., released via the desalination plant discharge). To optimize the filtration and backwash cycles, it is imperative that the maturation period is reasonably short—preferably not longer than 15–20 min. The length of the maturation period typically depends on the length and mode of filter backwash.

If the filters are over-backwashed (e.g., the backwash is too long and/or the backwash rate is too high), the filter cell will take longer to mature, and vice versa. On the other hand, if the backwash cycle is too short, some of the solids accumulated in the media during the filtration cycle will not be removed and the filtration cycles will become shorter and shorter over time. Therefore, the backwash time has to be adjusted to fit the amount of solids in the source water and these retained in the filter media—the lower the source water turbidity the shorter the backwash cycle should be. Optimizing the backwash schedule usually results in long (24–48 h) filtration cycles and short (15–20 min) maturation times, and ultimately in improved overall pretreatment filter performance.

For the filters to operate well, the suspended solids in the feed water have to be well coagulated and aggregated into large particles (typically $>20\ \mu m$), which can be retained by the filtration media. As indicated in the previous section of this chapter, coagulation and flocculation process is dependent on many factors and requires careful monitoring and optimization to achieve satisfactory pretreatment. Overdosing of coagulant hinders filter performance and results in accelerated fouling of the downstream cartridge filters and RO-membrane elements by unreacted coagulant and flocculant. On the other hand, under-dosing of coagulant often results in poor retention of fine particles and silt, and causes particulate fouling of the RO elements.

The applied bed expansion depends on the size of the filter media—the smaller the size, the larger media expansion is needed. For example, media of diameter of 1.2 mm requires expansion of only 10%–15%. Filter media with size of 0.8 mm needs an expansion of 20%–25%. For sand filter media of 0.4–0.6 mm the backwash rate should provide 30%–50% of media bed expansion for optimal filtration performance.

The number of filter cells and the individual production capacity of each cell are typically selected to allow full flow operation with one filter cell out of service in backwash and one out of service for maintenance. Additional information on the design of granular media filtration systems is provided elsewhere (Wilf et al., 2007; AWWA, 2007).

8.3 KEY FILTRATION SYSTEM COMPONENTS

8.3.1 Filter Cells

As indicated previously, typical granular media filtration system consists of number of individual units (cells or vessels), which operate in parallel. The number of filter cells is mainly dependent on the total flow these filters are designed to process. The construction cost of the filtration system is usually reduced when fewer individual cells are used. However, the minimum number of filters is limited by the following key factors: (1) the practical maximum size of individual filter bed (100–150 m^2/1080–1610 ft^2)—larger area beds are likely to result in nonuniform backwash of the filter bed; (2) the increase of the filtration rate of the filters remaining in operation, when one or two filters are in a backwash mode; (3) the configuration of the RO system, i.e., the number of individual trains and the planned mode of operation of the desalination plant.

To maintain a consistent, high-quality filter performance, the number of filter cells should be selected in such that when one cell is out of service, the hydraulic loading rate of the filters remaining in operation does not exceed 20% of the average loading rate with all units in service, and when two units are out of service, this rate is less than 30% of the average loading rate.

In general, even for very small desalination plants, the minimum number of individual pretreatment filters is recommended to be at least four. For plants of capacity higher than 5000 m^3/day (1.3 MGD) 6–8 filter units are preferable (Kawamura, 2000).

For desalination plants larger than 10,000 m^3/day (2.6 MGD), filter cells are usually divided into two groups that can be operated independently and paired with one-half of the desalination plant RO trains. In plants larger than 200,000 m^3/day (53 MGD), the desalination plant is typically divided into at least two sets of two filter groups, each with 8–32 individual filter cells.

8.3.2 Filter Media

Filter media type, uniformity, size, and depth are of key importance for the performance of the pretreatment filters. Dual-media filters have two layers of filtration media—typical design includes 0.6–1.0 m (2.0–3.3 ft) of anthracite or pumice over 0.4–1.6 m (1.3–5.2 ft) to of sand. Usually, filters with a total depth of the media of 1.8 m (6 ft) or more are referred to as deep media filters. The bottom 25%–30% of the media of these filters develops and sustains biofilm of microorganisms, which are capable of biodegrading a portion of the dissolved organics contained in the source water. In this case, the depth of the anthracite level is enhanced to between 1.2 and 1.8 m (4.9–5.9 ft).

If the source water is relatively cold (i.e., the average annual temperature is below 15°C/59°F), and at the same time is of high organic content, a layer of granular activated carbon (GAC) of the same depth is used instead of deeper layer of anthracite, because the biofiltration removal efficiency will be hindered by the low temperature. While during biofiltration a portion of the soluble organics in the source water is metabolized by the microorganisms that grow on a thin biofilm formed on the granular filter media, the GAC media removes a portion of the source water organics mainly by adsorption.

Trimedia filters have 0.45–0.6 m (1.5–2.0 ft) of anthracite as the top layer, 0.2–0.4 m (0.7–1.3 ft) of sand as a middle layer, and 0.10–0.15 m (0.33–0.50 ft) of garnet or limonite as the bottom layer. These filters are used if the source water contains a large amount of fine silt or the source water intake experiences algal blooms dominated of microalgae (0.5–20 μm). Filter media density varies as shown on Table 8.2.

The effective size of the media is the size of the opening of the sieve for which 10% of the grains by weight are smaller in diameter. Uniformity coefficient is the ratio between the opening sizes of a sieve for which 60% of the grains by weight are smaller, divided by the effective size of the media:

$$UC = d60/d10 \tag{8.1}$$

Uniformity coefficient is an important parameter because it indicates how similar the media particles are in size. In general, for the same size media, increasing uniformity coefficient of the media allows to increase the length of the filter cycles.

The d60 value of the filter media can also be used to determine the filter backwash rate at 200°C/680°F (Qasim et al., 2000) using the following formulas:

$$U_b = d60 \quad \text{for sand} \tag{8.2}$$

$$U_b = 0.47 \times d60 \quad \text{for anthracite} \tag{8.3}$$

For temperatures different from 20°C, the backwash time can be calculated applying an adjustment coefficient for water viscosity:

$$U_{bt} = U_b \times K^{0.333} \tag{8.4}$$

where K, is absolute viscosity of water at a given temperature (kg/m s).

TABLE 8.2 Typical Filter Media Characteristics

Media Type	Typical Effective Grain Size (mm)	Specific Density tons/m^3 (lb/ft^3)	Uniformity Coefficient
Pumice	0.8–2.0	1.2 (75)	1.3–1.8
Anthracite	0.8–2.0	1.4–1.7 (87–104)	1.3–1.8
Silica sand	0.4–0.8	2.60–2.65 (162–165)	1.2–1.6
Garnet	0.2–0.6	3.50–4.30 (218–268)	1.5–1.8

The size of the media and uniformity coefficient should always be configured to decrease while the specific density should increase in the direction of the flow. This configuration allows to prevent the intermixing of the different types of media materials during backwash. Intermixing of the media results in shorter filter cycles and the need for more frequent backwashing.

The depth of the filter bed is typically a function of the media size and follows the general rule thumb that the ratio between the depth of the filter bed (L—in mm) and the effective size of the filter media (d_e—in mm), L/d_e, should be in a range of 1000−1500. For example, if the effective size of the anthracite media is selected to be 0.65 mm, the depth of the anthracite bed should preferably be approximately 1.0 m/3.3 ft (0.65 × 1500 mm = 0.975 mm).

The depth of the GAC media is estimated based on the average contact time in this media, which is recommended to be 10−15 min. For example, if a filter is designed for a surface loading rate of 9 m^3/m^2 h (4 gpm/ft^2), the depth of the GAC media should be at least 1.5 m/4.9 ft (9 m^3/m^2 h × 10 min/60 min per hour = 1.5 m).

When each of the filter media layers is first placed in the filter cells, and additional 3−5 cm (1.2−2.0 in.) of media should added to the design depth of the layer to account for the removal/loss of fine particles from the newly installed bed after backwashing.

It should also be pointed out that if the filters are designed to achieve TOC removal by biofiltration, it would take at least 4−6 weeks for the filters to create sustainable biofilm on the surface of the filter media grains that can yield steady and consistent filter performance and TOC removal. If the source water temperature is relatively low (i.e., below 20°C), than biofilm formation process may take several weeks longer.

8.3.3 Media Support Layer and Undertrain System

The filtration media is typically supported by a layer of gravel bed that has depth of 0.3 m (12 in.) and is preferably graded in six layers − 1/6−$^3/_4$ in., with depth of 0.05 m (2 in.) each (Kawamura, 2000). The gravel bed in located on the top of a filter underdrain system. There are two types of filter underdrain systems widely used at present: underdrain blocks or false bottom with nozzles.

The nozzle type of underdrain has found a wider application for desalination applications. In this system, nozzles penetrate the bottom of the underdrain and allow to collect filtered water uniformly, and to distribute the backwash water more evenly during the backwash cycle. Typically, the flow velocity in the channel, pipe and false bottom below the underdrain system is designed to be relatively low [0.6 m/s (2 fps)] to provide uniform flow pattern distribution. The ratio between the nozzle area and the bed area varies from 0.2% to 1.5% depending on the type of underdrain system, backwash rate applied and type of the filter.

8.3.4 Service Facilities and Equipment

At present, most filters used for saline water pretreatment have air and water backwash systems. As a result each filtration system is equipped with air blowers (usually one or two duty and one standby) and water backwash pumps (usually two to three duty and one standby). As a rule of thumb, the backwash pumps should be designed to deliver water allowing to wash the filter at a surface loading rate, which is three to five times higher than the design filtration rate.

8.4 FILTER TYPES AND CONFIGURATIONS

8.4.1 Single-, Dual- and Trimedia Filters

Single-media filters are not very commonly used for saline water pretreatment because of their limited ability to perform under varying source water conditions. Typically, such filters could be applied for desalination plants with subsurface intakes, producing turbidity of <2 NTU, TSS of <3 mg/L, and $SDI_{15} < 6$. The majority of the desalination plants worldwide use dual-media filters with the top layer of pumice or anthracite and the bottom layer of sand.

Trimedia filters are used for pretreatment of saline waters only when the source water quality varies significantly and the intake area experiences frequent algal blooms dominated by small-size algae (average algal cell size < 20 μm) and/or very fine silt. Such filters are suitable for capturing small and diverse size particles, which cannot be well retained by the top two layers—anthracite and sand. Since the cost of filter cells increases with depth, often instead of a deep single trimedia gravity filter, a combination of first-stage coarser media (anthracite-sand) gravity filter followed by a second-stage pressure filter containing finer (sand and garnet) media is used.

8.4.2 Single- and Two-Stage Filters

Two-stage filtration is typically applied when the source water contains high levels of turbidity (usually above 20 NTU) and organics (TOC > 6 mg/L) for long periods of time (i.e., weeks/month). Such conditions occur in: desalination plant intake areas exposed to prolonged red-tide events (which sometimes could last for several months); river estuaries that are exposed to elevated turbidity levels occurring during the wet season of the year; or intakes influenced by seasonal winds or currents, which for prolonged periods of time collect source water of elevated content of particulate or organic foulants.

Two-stage filtration systems typically consist of coarse (roughing) filters and fine (polishing) filters operated in series. Usually, the first-stage filter is a mono-media type (i.e., coarse sand or anthracite) or dual media (e.g., anthracite and sand) while the second-stage filter is configured as a dual-media filter with design criteria described in the previous section. The first (coarse-media) filter typically removes 60%–80% of the total amount of solids contained in the source water and is designed to retain all large debris and floating algal biomass. The second-stage filter removes over 99% of the remaining solids and fine silt as well as microalgae contained in the source seawater, typically producing effluent turbidity of less than 0.05 NTU.

Two-stage filters have several advantages. The filtration process through the coarse media filters not only removes large particulate foulants but also enhances coagulation of the fine particulates contained in the source water, which makes their removal in the second-stage filters less difficult and allows the second-stage filters to be designed as shallow-bed- rather than deep-bed filters and to operate at higher surface loading rates. This benefit results in reduced size of the dual-media filters and in lower total amount of coagulant (ferric salt) needed to achieve the same final filter effluent water quality, as compared to single-stage dual-media filters.

Two other benefits of the two-stage filters are that: (1) they can handle larger fluctuations of intake source water turbidity because of the larger total filter media volume/solids retention capacity; (2) when the second stage filters are designed as deep-bed (rather than shallow depth) filters, they can achieve enhanced TOC removal by biofiltration. While deep single-stage dual-media filters can typically reduce 20%–30% of the TOC contained in the source seawater, the two-stage systems with deep second-stage filters can achieve 40%–60% TOC removal, mainly due to enhanced fine particle coagulation and biofiltration.

8.4.3 Downflow and Upflow Filters

Most filters used in pretreatment of seawater and brackish water are downflow filters. This flow direction allows large algal particles to be retained at the top of the filter media and removed with the backwash water with minimum brakeage and release of organics. If upflow filtration is used, algae contained in the source water are pressed against the filter media and unwanted dissolved organics are released from the broken algal cells into the filtered water, which is undesirable because it enhances biofouling of the downstream RO membranes.

8.4.4 Filters Combined With DAF Clarifiers

In cases where the saline source water contains a large amount of algal particulates, and/or oil and grease, and space is at premium, DAF and granular media filtration processes can be combined in one structure, where the DAF clarifier is located is located above the filter cell (see Fig. 8.2).

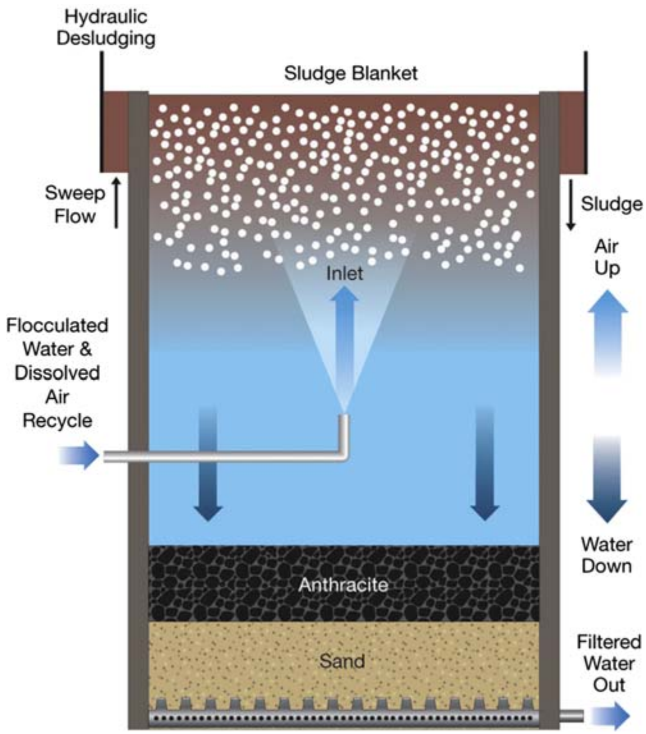

FIGURE 8.2 Combined DAF clarifier and granular media filter.

Under this configuration, granular media filters are typically designed as dual-media (anthracite and sand) downflow filters. The design surface loading rate of these filters is usually two to three times higher than that of single-stage dual-media filters (i.e., 15–35 m^3/m^2 h/6 to 14 gpm/ft^2).

Since the O&M expenditures for he DAF clarifier are relatively high, the filtration portion of the pretreatment system is recommended to be designed for the lower end of that range 15–20 m^3/m^2 h (6–8 gpm/ft^2), which would allow to operate only the filtration portion of the DAF-filter system, when the source water quality is good and the levels of turbidity and organics in the water are low.

8.4.5 Gravity and Pressure Filters

Depending on the driving force for water filtration, granular media filters are classified as gravity and pressure filters. The main differences between the two types of filters are the head required to convey the water through the media bed, the filtration rate, and the type of vessel used to contain the filter media.

Because of the high cost of constructing large pressure vessels with proper wetted surfaces for corrosion resistance, pressure filters are typically used for small- and medium-size capacity RO plants. Gravity pretreatment filters are used for both small and large RO-desalination plants.

8.4.5.1 Gravity Filters—Description

Typically, gravity filters are reinforced concrete structures that operate at water pressure drop through the media of between 1.8 and 3.0 m (6–10 ft). The hydrostatic pressure over the filter bed provides the force needed to overcome the headloss in the media. Single-stage dual-media downflow gravity filters are predominant type of filtration pretreatment technology used in desalination plants of capacity higher than 40,000 m^3/day (10.6 MGD). Some of the largest SWRO desalination plants in the world in operation today—such as the 200,000 m^3/day Carlsbad desalination plant in California (Fig. 8.3), the largest SWRO desalination plant in the western hemisphere, and the 624,000 m^3/day Sorek desalination plant (Fig. 8.4), the largest SWRO plant in the world—are equipped with single-stage, dual-media gravity filters.

8.4.5.2 Gravity Filters—Key Advantages

8.4.5.2.1 BETTER REMOVAL OF ALGAL MATERIAL FROM THE SOURCE WATER

Surface saline water (i.e., seawater) always contains a measurable amount of algae, whose concentration usually increases several times during the summer period and may increase up to 10 times during periods of algal blooms.

There is a large variety of algal species in the seawater. Some algal species that occur during red-tide events have cells that are relatively easy to break under pressure as low as 0.4–0.6 bars. When the algal cells break, they release cytoplasm in the source water that has a very high content of easily biodegradable polysaccharides. When the amount of polysaccharides released by the broken algal cells exceeds certain level in the filtered seawater, they would typically trigger accelerated biofouling of the RO membranes.

FIGURE 8.3 Single-stage gravity filters of the Carlsbad desalination plant, California.

FIGURE 8.4 Single-stage gravity filters of the Sorek desalination plant, Israel.

Two practical approaches to address this type of problems are: (1) use of dissolved air flotation facility ahead of the pretreatment filters to gently remove algal cells and prevent their breakage (which is preferable); (2) installation of GAC media layer ("activated carbon cap") on the surface of the filters to remove some of the polysaccharides and other organics in the source water.

Pressure filters usually operate at several times higher filtration pressure than gravity filters. Because the operating pressure of these filters is often higher than the algal cell break-pressure threshold, pressure filters would have the disadvantage to cause an accelerated biofouling when filtering source water of very high algal content. This effect is likely to manifest itself mainly in the summer and during algal blooms when the level of TOC in the source water exceeds 2 mg/L.

Pressure filters are used in medium- and large-size desalination plants in Spain, Algeria, and Australia. However, in most successful applications the source water quality is very good (TOC < 1 mg/L, SDI < 4 and turbidity < 4 NTU). Most of the Spanish desalination plant intakes are relatively deep and the algal content in the source water is fairly low. At depth of 10–20 m (33–66 ft), the concentration of algae is significantly lower than that at the water surface and, therefore, as long as desalination plant intake is deep (e.g., 10 m or more), biofouling caused by breakage and decay of algal biomass may not be as significant problem as it would be for shallow intakes or intakes located at the surface of the water body (i.e., near-shore open intakes).

8.4.5.2.2 LONGER USEFUL LIVE OF THE FILTER STRUCTURE

Typically, gravity filters are concrete structures that have useful life of 50–100 years. Pressure filters are steel structures with a lifespan of 25 years or less. The internal surface of the pressure filters used for seawater desalination is typically lined up with rubber or polyurethane coating that needs to be replaced every 5–10 years and inspected occasionally. In recent years, several manufacturers began to offer plastic pressure filters, which are expected to address the issues associated with the limited longevity of steel filter vessels. However, at present these filters have limited track record to establish their actual useful life.

8.4.5.2.3 LOWER POWER USE

Gravity filters operate without the need to pressurize the feed water through the filtration media. Therefore, they usually consume several times less energy than pressure-driven filters of the same volume of pretreated water.

8.4.5.2.4 HIGHER SOLIDS RETENTION CAPACITY AND BETTER HANDLING OF TURBIDITY SPIKES

Gravity media filters have approximately two to three times larger volume of filtration media and retention time than pressure filters for the same water production capacity. Therefore, this type of filters can retain proportionally more solids and as a result, pretreatment filter performance is less sensitive to occasional spikes in source water turbidity. If the source water consistently has low content of particulate solids (e.g., <2 NTU), the gravity and pressure filters perform comparably.

Pressure filters usually do not handle solids/turbidity spikes as well because of their smaller solids retention capacity (i.e., smaller volume of media pores that can store solids before the filter needs to be backwashed). If the source water is likely to experience occasional spikes of high turbidity (20 NTU or higher) due to rain events, algal blooms, ship traffic, ocean bottom dredging operations, strong seasonal winds, seasonal change in underwater current direction, or spring upwelling of water from the bottom to the surface, than pressure filters will produce effluent with inferior effluent quality (SDI and turbidity) during such events and therefore, their use would likely result in a more frequent RO cleaning.

8.4.5.2.5 SIMPLER INSPECTION AND MAINTENANCE

Gravity filters are typically covered with light plastic covers that protect the filter cells from direct sunlight or are installed in buildings to prevent excessive algal growth on the filter troughs and media. If covers are used, they can be easily removed and the filter cells can be inspected visually for irregularities—malfunctioning of filter backwash nozzles, weir corrosion, poorly backwashed areas of filter media, formation of "mud-balls," etc.

Pressure filters are completely enclosed and very difficult to inspect for the condition of the media, and integrity of the internal equipment and underdrain. As a result, these filters have to be designed with higher contingency factor (reserve capacity). A 15%–20% reserve capacity is recommended if pressure filters are used to accommodate for potential flow-distribution problems and uneven backwash air and water distribution.

8.4.5.2.6 EASIER TO ACCOMMODATE MEMBRANE PRETREATMENT IN THE FUTURE

As membrane pretreatment technology evolves and new membrane systems available on the market are designed to better-handle algal bloom challenges, for some existing plants, it may be advantageous to modify the exiting conventional granular media filters into submersible membrane pretreatment filters. This upgrade would be possible as long as gravity filter cells are designed with adequate depth and configuration to accommodate submersible UF/MF membranes. Pressure filter vessels would not be possible to modify into membrane holding tanks.

8.4.5.3 Pressure Filters—Description

Pressure filters have filter bed configuration similar to that of gravity filters, except that the filter media is contained in steel or plastic pressure vessel. They have found application mainly for small- and medium-size desalination plants—usually with production capacity of less than 20,000 m^3/day (5 MGD). However, there are a number of installations worldwide where pressure filters are used for pretreatment of significantly larger volumes of water. An example is one of the largest SWRO desalination plants in the Middle East—the 190,000 m^3/day Al-Ghubrah desalination plant in Oman. The pressure filters for this plant are shown on Fig. 8.5.

In most cases for good source water quality (SDI < 5 and turbidity less than 5 NTU) the pressure filters are designed as single-stage, dual-media (anthracite and sand) units. Some

FIGURE 8.5 Horizontal pressure filters—Al-Ghubrah desalination plant, Oman.

FIGURE 8.6 Vertical pressure filters.

plants with relatively poor water-quality use two-stage pressure filtration systems. Pressure filters are available in two vessel configurations—vertical and horizontal.

Vertical pressure filters (see Fig. 8.6) are customarily used in smaller plants and individual vessels have maximum diameter of 3 m. Horizontal pressure filters (Fig. 8.5) are used more frequently in desalination plants and are more popular for medium- and large-size facilities. Horizontal filters allow larger filtration area per filter vessel as compared to vertical units. However, usually vertical vessels can be designed with deeper filter media, if deep filters are needed to handle spikes of source seawater turbidity.

Compared to gravity media filters that operate under a maximum water level over the filter bed of up to 2.5 m (8.2 ft), pressure filters typically run at feed pressure equivalent to 15–30 m of water column (49–98 ft). The magnitude of the feed pressure is often driven by the suction pressure requirements of the high-pressure feed pumps of the downstream RO system. One key advantage of the use of pressure filters is that they could allow avoiding intermittent pumping of the pretreated source water.

A typical RO system with gravity pretreatment filters requires installation of filter-effluent transfer pumps to convey the filtrate from the filter effluent well to the high-pressure RO-feed pumps. Use of pressure filters could eliminate the need for such interim filter-effluent transfer pumps because the filtrate is already pressure driven by the intake pumps and the pretreatment filters do not break the hydraulic grade line.

8.4.5.4 Pressure Filters—Key Advantages

8.4.5.4.1 LOWER CONSTRUCTION COSTS

Pressure filters are prefabricated steel structures and their production costs per unit filtration capacity are lower than these of concrete gravity filters. Since pressure filters are designed at approximately two to three times higher surface loading rates than gravity filters [25–45 vs. 8–15 m^3/m^2 h (10–18 gpm/ft^2 vs. 3–6 gpm/ft^2), their volume and size are smaller and, therefore, they usually are less costly to build and install.

8.4.5.4.2 SMALLER FOOTPRINT

Because of their smaller volume and filtration area, pressure filters occupy smaller footprint. If the available site is of limited footprint, this is an important factor to consider, when selecting granular media filtration technology.

8.4.5.4.3 SIMPLER INSTALLATION

Since pressure-filter vessels are prefabricated, the installation time of this pretreatment system is approximately 20%–30% shorter than that of gravity filters with concrete structures.

8.4.5.4.4 NO EFFECT OF SUNLIGHT ON ALGAL GROWTH ON FILTER WEIRS

Pressure filters are completely enclosed and sunlight cannot reach the filter weirs, distribution system, and media, and to induce an algal growth that would have negative impact on filter performance. For comparison, gravity filters (if they are not located in a building or covered with nontranslucent panels) would grow algae on all components of the filters exposed to direct sunlight.

8.5 FILTER PERFORMANCE

8.5.1 Removal of Solids

Typically, saline source water collected by open intakes has high content of both fine silt and suspended solids. The purpose of the pretreatment filters for RO plants is to remove both of these particulate foulants. Because removing fine silt from saline-source water to levels below SDI_{15} of 4 by granular media filtration is usually much more challenging than the reduction of source-water turbidity by 99%, the design of these pretreatment facilities is typically governed by the filter effluent SDI target level rather than by the target turbidity or pathogen removal rates.

Filter solids-removal efficiency (reduction of turbidity/total suspended solids) is not directly related to its silt and fine-colloid removal efficiency (SDI reduction capability). Dissolved organics and coagulant (iron salts) can absorb on/in the SDI-filter test pad and result in increased SDI values. Full-scale experience at many granular media pretreatment filter installations indicates that filters can consistently reduce source-water turbidity to less than 0.1 NTU, while at the same time filter effluent could have SDI_{15} frequently exceeding 4. In many cases, granular media filters at RO-desalination plants need to be designed more conservatively than similar filters in conventional surface water treatment plants to capture fine solids, silt, and colloidal organics contained in the saline source water.

8.5.2 Removal of Organics

Typical gravity and pressure dual-media filters with filter-media bed depth of 1.0–1.4 m (3.3–5.3 ft) have relatively low organics removal rate, i.e.,15%–20%. This removal rate, however, increases significantly with depth and could reach 25%–35% for filters of total filter depth of 1.8 m (6.6 ft) or more. If a carbon cap is installed on the top of the filter media (above the layer of anthracite), filter TOC removal rate could be increased to 40%–50%.

8.5.3 Removal of Microorganisms

8.5.3.1 Algae

The rate of algae removal by the filters will mainly depend on the size of the algae in the source water and the size of the filter media. Most algae larger than 100 μm are typically retained on the surface of the top media (anthracite/pumice). Practical observations indicate that the closer the desalination plant is located to the equator, the larger the percentage of micro- and picoalgae in the source seawater. Such algae are not well removed by conventional sand media of sizes 0.4—0.6 mm (400—600 μm) and require the installation of a third layer of finer filter media or use of two-stage filtration system. Depending on the size of media and size of the algae dominating in the source water, the algal removal could typically vary between 20% and 90%.

8.5.3.2 Bacteria and Viruses

Desalination pretreatment filters would typically remove 99% (2 logs) of pathogens, but sometimes may have lower removal rates in terms of marine bacteria because these bacteria are very small in size and would pass through the filters.

8.5.4 Monitoring of Filter Performance

Besides filter effluent quality flow, length of filtration cycle, and maturation time, granular media filters are also monitored for the depth and condition of their media, the filter underdrain, and the volume and quality of the backwash they generate.

Loss of filter media is usually assessed annually—2—4 cm of media loss from the filter cells per year is considered normal. Media is topped off when its depth is reduced by approximately 10 cm below design level.

Filter media condition is usually inspected every 6 months to 1 year. Over time, pretreatment filters accumulate debris, coagulant residuals, and dead organic matter, which tend to deteriorate their performance. Therefore, many plants worldwide practice periodic 1-day soaking of their granular media pretreatment filters in citric or sulfuric acid in the winter months and in caustic soda (sodium hydroxide) in the summer/algal bloom season.

Low-pH sulfuric, hydrochloric, or citric acid filter media soak is equivalent to the performance of clean-in-place (CIP) of the RO membranes and aims to loosen calcite deposits and aggregates that may be blocking the filter media and to maximize the performance of the filters in terms of removal of solids, organics, and silt, and to extend the filter cycle length. High-pH (caustic soda) filter-media soaking aims to dissolve patches of organic matter, sand, silt, and anthracite cemented with organic residuals from the source water. The caustic-soda soak mainly targets filter-media solid patches of organic origin, while the acidic soak aims to dissolve filter blockages formed from calcium and magnesium precipitates and overdozing of coagulant.

Depending on the flow distribution configuration of the pretreatment filtration system, filter soak should be completed on individual filter cells rather than on all cells at the same time. The recommended low-pH/high-pH filter-soak procedure is as follows:

1. Drain down the filter cell that is planned to be cleaned.
2. Refill the cell with filtered water that contains sulfuric (hydrochloric or citric) acid of concentration of 80—90 mg/L for acid soak, and with sodium hydroxide of dosage of

120–140 mg/L for alkaline soak. The target pH for the acid soak is 4 while that for the alkaline soak is 11.

3. Let the filter cell soak in acid (or caustic soda respectively) for 16–24 h. Measure the pH of the water every 6 h to and once the pH reaches 7.5–7.8, drain the cell and backwash it.
4. Send the filter effluent to waste for 30 min after the low- or high-pH soak.

The most common operational problems of granular media pretreatment filters and proven solutions for maintaining a steady and consistent filter performance are summarized in Table 8.3.

TABLE 8.3 Granular Media Filters—Typical Problems and Solutions

Indicator/Problem	Potential Causes	Solution
• Reddish brown discoloration of SDI test pads used for filter effluent testing	• Overdosing of iron-based coagulant	• Reduce the coagulant dosing rate, by at least 50% • Adjust coagulant dose by jar testing with addition of flocculant and pH adjustment
• Black/dark brown discoloration of SDI test pads	• Use of low-purity ferric chloride with high content of manganese	• Change the coagulant and use ferric sulfate • Check the ORP of water fed and permeate of all RO trains to avoid membrane oxidation
• Solid particles observed on SDI test pads	• Broken nozzles or cracks on the bottom of filter cells	• Check underdrain system and repair or replace the faulty devices
• High frequency of backwashing (short cycles up to 2–6 h)	• Intermixing of sand and anthracite media on the filter surface	• Replace both sand and anthracite with media of adequate specifications (avoid the use of overly heavy anthracite)
	• Improper upstream coagulation • Polymer overdosing	• Check/adjust the coagulant/polymer dosage
	• Low efficiency of the backwash procedure	• Check/adjust backwash rates of water and air according to recommended good practices—backwash rate at least 2.5 times higher than filtration rate
	• Nozzles clogging	• Empty the filter and start a recovery cleaning procedure
	• High content of fine silt in source water	• Check the filter surface after filtration cycles for silt accumulation and apply preventive measures (see below)
	• Algae growth • Surface clogging	• Remove algae/debris from the filter surface, add/increase the dose of coagulant/flocculant

TABLE 8.3 Granular Media Filters—Typical Problems and Solutions—cont'd

Indicator/Problem	Potential Causes	Solution
• Mud ball formation and solidification of some portions of the filter media	• Inadequate backwash rate • Nozzle clogging	• Check/adjust backwash rates of water and air, and apply recovery cleaning • If uniform backwashing is not possible, empty the filter media, clean the nozzles, and replace filter media with material and layers with adequate specifications
• Filter media "cratering" or geyser formation during backwashing	• Broken filter nozzles	• Empty the filter media and replace broken nozzles
• Filter media mounding	• Plugged filter nozzles • Inadequate thickness of the underdrain gravel pack on the nozzles	• Empty the filter and start a recovery cleaning procedure • Empty the filter media and replace with adequate depth media to cover the nozzles
• Filter media siltation at the end of the filtration cycle	• High content of fine silt in source water	• Check the SDI; if >30 for long periods, apply periodic dredging of the intake area • If the problem persists, add pretreatment using lamella clarification or microscreens
• Excessive amount of backwash water (>5%)	• Excessive length of backwash • Excessively high solid content in source water	• Improve upstream treatment (coagulation and settling). • Reduce backwash cycle length

8.5.4.1 Iron-Coagulant Overdosing

Iron-coagulant overdosing is the most commonly observed problem in saline water pretreatment systems with surface water intakes applying granular media filters. Good operations' practices usually involve frequent parallel measurement of SDI of the source water, filtered water, and the RO-feed water after cartridge filtration and comparison of the test results to identify changes in the SDI value and appearance of the SDI pads.

A practical criterion for well-operating pretreatment system is the difference of SDI_{15} of the filtered water entering and exiting the cartridge filters. If the pretreatment system is working well, this difference is usually less than 0.2 units (typical accuracy of the SDI test), which means that the cartridge filters are not retaining solids because they have already been adequately removed by the pretreatment system.

If the difference between the SDI_{15} values of the cartridge filter inlet and outlet is higher than 0.5 units, and SDI_{15} of the filtered water exceeds 3, this is an indication that the pretreatment system could be further optimized. If the SDI level in the filtered water frequently exceeds 5, then the pretreatment system operation has to be improved by reevaluation of the source-water quality and the applied chemical-conditioning strategy.

The SDI pads used for measurement of the silt density index of the feed and source waters are visually inspected for color, odor, and retention/accumulation of solids, and coagulant after every test. These pads are typically light creamy or off-white if the granular media filters perform well. If the pads are slightly yellow, this is typically an indication of iron and/or colloidal organics in the source water. Iron could originate from overdosing of coagulant (if ferric chloride or sulfate is used) and/or from corrosion along the path of the pretreatment system flow.

Colloidal organics could originate from the natural organic matter (NOM) in the source water, overdosing of an organic polymer or antiscalant or bio growth on the intake facilities (e.g., wells).

Reddish brown discoloration of the SDI pads used for filter effluent testing is a clear indication of overdosing of iron-based coagulant, and this challenge would need to be troubleshooted by reduction of the coagulant dosage by at least 50%. The optimum coagulant dosage, as well as the need to add flocculant and to adjust pH of the source water, should be determined based on jar testing and on the past operational experience with similar source-water quality.

Gray discoloration of the SDI pads is indicative of the presence of carbon fines and usually occurs if GAC filtration layer is incorporated into the filter media or granular activated filter is used to abate source water with high organic loading.

If solid particles are observed on the SDI pads from the filtered water test prior to the cartridge filters, check the condition of the underdrain system—there may be breaches of its integrity (i.e., broken nozzles, cracks on the bottom of the filter cells, or other defects that may cause small amounts of filtration media to be conveyed with the filtered water). If solid particles are observed on the SDI pads from the test of sample of filter effluent, then check the condition of the cartridge filters—it is likely that some of the individual cartridges may have collapsed or are short-circuiting feed water and allowing particles in the filtered water to escape being captured by the cartridge filters.

If dark black discoloration is observed on the cartridge filters, this problem is most likely caused by high content of manganese impurities in the ferric chloride coagulant. To address this challenge, operators can either switch to ferric sulfate or identify alternative source of ferric chloride supply, which has lower manganese content.

8.5.4.2 Sand and Anthracite Media Intermixing

Under normal operational conditions of well-designed dual-media (anthracite and sand) downflow filters, the anthracite and sand media stay separated into two distinctive filtration layers during filter operations. If intermixing of the two media occurs (Fig. 8.7), the propagation of anthracite into sand often results in significant reduction of the length of the filtration cycle (typically from 24 h or more down to 2–6 h).

Practical experience shows that one of the most common reasons for the intermixing of the sand and anthracite media in filter cells is the use of overly heavy anthracite. In general, two types of anthracite products are available on the market—a lighter product (referred to as "Welsh type") anthracite that has a specific gravity of 1400–1450 kg/m^3 and is suitable for granular media filtration; and a heavier ("Pennsylvania type") anthracite with specific gravity of 1600–1650 kg/m^3, which is often used as construction material. While the heavier anthracite is less costly and could sometimes be selected based on lower overall media cost, when used as a top layer over a typical sand filtration media of specific gravity of 2500–2650 kg/m^3, this anthracite sinks into the sand media and the two media completely intermix within several filter cycles of their installation. Media intermixing is irreversible (i.e., sand and

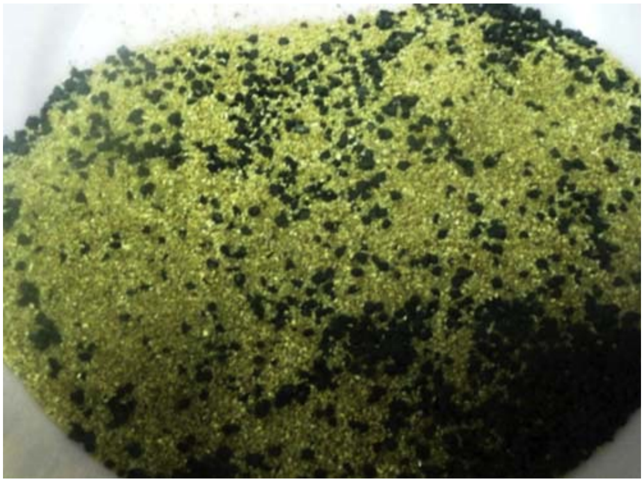

FIGURE 8.7 Intermixing of sand and anthracite media.

anthracite cannot be separated easily) and the only practical solution is to replace both the anthracite and the sand media.

This operation's challenge underlines the critical importance of the selection of the specific weight of anthracite. As a general rule of thumb, the specific weight of the anthracite media should be more than 1000 kg/m^3 lighter than the specific weight of the underlying filtration sand media. In addition, when anthracite is replaced or topped off periodically, it is important to closely match the specific weight of the original and new commercial anthracite products. Using overly light product would have the disadvantage of elevated media expansion and excessive loss of media during the filter backwash cycle.

Another common reason for intermixing of sand and anthracite media is the mismatch of the grain size ranges of the sand and anthracite media. Such intermixing will occur even if the specific weight difference of sand and anthracite is within the range specified above. As a practical rule of thumb, the coarsest size of anthracite in its size range should not exceed five times the finest size of the underlying sand media. For example, if the selected anthracite has a media-size range of 0.5–2.0 mm and sand has a range of 0.15–0.90 mm, the two medias are very likely to intermix because the largest size anthracite grains in the anthracite size range (i.e., 2.0 mm) are over 13 times larger than the smallest size in the sand size range (i.e. 0.15 mm), i.e., 2.0 mm/0.25 mm = 13.3. This particular challenge underlines the necessity to use both anthracite and sand media of low uniformity coefficient (preferably 1.4 or lower), which would allow to avoid the significant discrepancy of media sizes and associated media propagation effect.

8.5.4.3 Filter-Media Solidification and Mud Ball Formation

One of the common reasons for poor filter performance and reduction of the length of the filter cycle over time is the low backwash velocity of the filtration media. Such low backwash velocity could be caused by:

1. Inadequate filter backwash rate due to backwash pump-capacity constraints, capacity limitation of the storage tank for backwash water of the tank storing spent filter backwash, or due to operator error,

2. Inadequate depth of the gravel packing layer (for filters with nozzle underdrain),
3. Plugged nozzles or underdrain blocks,
4. Hardening of the media due to calcium carbonate buildup, and
5. Plugging of the gravel packing layer with sand due to incompatibility of gravel and sand sizes.

Filter cells should be backwashed at a rate that is 3–5 times higher than the average filtration rate to provide complete removal of the solids accumulated in the filtration media over the length of the entire filtration cycle (24–36 h) (Qasim et al., 2000). The lower end of backwash rate range (i.e., approximately 3 times the filtration rate) is applied during periods of low source water turbidity (i.e., turbidity < 2 NTU), while the maximum rate (5 times the filtration rate) is used for periods of heavy solids and organic loads (turbidity > 15–20 NTU).

Under average turbidity conditions the optimum backwash rate is usually 3.0–4.0 times the average filtration rate (i.e., backwash rate of 30–40 m^3/m^2 h for filtration rate of 10 m^3/m^2 h). If the filter backwash rate is inadequate, solids not removed with the backwash water will begin to accumulate in the filter cells and occupy a portion of the filtration media, which in turn will gradually reduce the length of the filtration cycle and deteriorate filtered water quality. In addition, balls of mud, coagulant, silt, and other solids will begin forming inside the media and the media grains could begin solidifying in some portions of the bottom, and such solidification will create zones with little to no filtration and backwashing.

In most conventional filter-cell underdrains using nozzles, a gravel packing layer of adequate depth and size would need to be provided to maintain sufficient and uniform backwashing of the filter media. As a rule of thumb, the depth of the gravel packing layer should extend at least one to two times the height of the nozzles. For example, for typical 40-mm nozzles widely used in filter underdrain systems, the gravel pack layer depth should be 80–120 mm. In addition, the size of the pebbles of the top gravel layer that is in contact with the filtration sand layer should be at least two times smaller than the higher end of the sand media size. For example, if the sand media has size of 0.6–0.8 mm, then the size of the underlying gravel should be less than 1.6 mm (2×0.8 mm).

8.5.4.4 Filter-Media "Cratering"

Filter-media cratering (e.g., creation of indentions on the surface of the filter media, which resemble small craters) is usually caused by broken filter nozzles or filter underdrain blocks. Loss of filtration material through the broken nozzles creates indentation on the surface of the filter media above the impacted underdrain, which is clearly visible on the surface of open gravity filters.

8.5.4.5 Filter-Media Mounding

Plugged filter nozzles or inadequate thickness of the underdrain gravel pack covering the nozzles tend to cause more intense vertical flow around the impacted area during backwash, which lifts the filter surface media in this location and forms a mound of media (Fig. 8.8). In such cases, the underdrain gravel pack has to be increased in depth and reinstalled. Otherwise, this condition will result in a permanent reduction of the length of filtration cycles, in extended maturation period and in deterioration of the filtrate water quality over time.

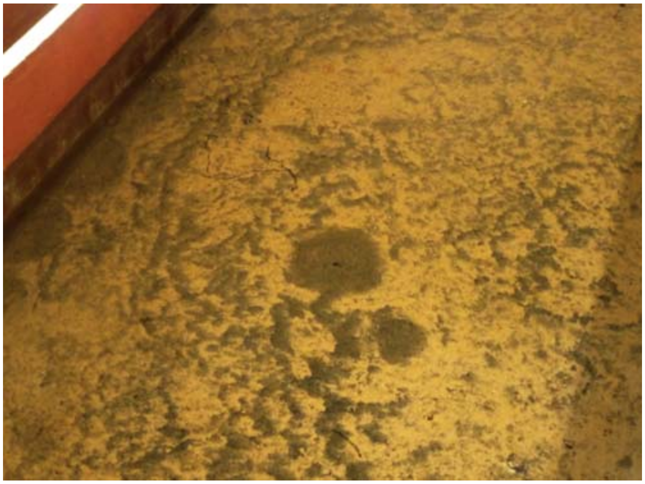

FIGURE 8.8 Filter media mounding.

8.5.4.6 High Content of Fine Silt in the Source Water

Use of intake lagoons could result in increase of fine silt content in source water over time because silt naturally contained in the water tends to accumulate at the bottom of the lagoon and as it exceeds certain threshold depth it tends to migrate into the intake pump wells and to be conveyed to the desalination plant's pretreatment system. Typically, this effect of migration of fine silt from the intake source water can be monitored by measuring the SDI_5 of the source water (or $SDI_{2.5}$ if silt content increase is so high that SDI_5 cannot be measured).

When the intake lagoon depth is such that there is at least 3–5 m (10–16 ft) from the bottom of the lagoon sediments to the entrance of the intake pump wells, this buffer distance prevents fine silt from entering the intake, and the SDI_5 of the source water is usually between 10 and 20. As sediments accumulate at the bottom of the lagoon over time, some of them are conveyed to the intake and typically result in a slow increase in source water SDI. Usually, when SDI_5 exceeds 25–30, the performance of the desalination plant granular filtration system would begin to be impacted because the fine silt would begin to accumulate on the surface of the filter media and decrease its ability to filter the source water, which in turn would reduce the length of the filtration cycle.

The most cost-effective troubleshooting measure in this case is the periodic dredging of the bottom of the intake area when the SDI_5 of the source water exceeds 30, if SDI increase is caused by siltation. Sometimes, similar SDI increase could also be caused by algal blooms or by strong winds that resuspend bottom sediments and silt. However, well-designed granular media filters would usually be able to handle such SDI increase up to a level of 30 and algal content of up to 20,000 cells/L.

Filter surface siltation of the anthracite or pumice media could also be triggered by naturally occurring events that cause resuspension of large quantities of fine silt and residuals from the bottom of the ocean floor such as high intensity winds, storms, or waves especially if the plant has an open onshore intake or shallow offshore intake located in the tidal zone (e.g., 200–300 m offshore). Besides dredging the intake area, alternative solutions of this treatment challenge is the construction of pretreatment step such as lamella settler clarification or grit separation by microscreens. Additional information on granular media filter performance monitoring and troubleshooting is presented in the following publications (Beverly, 2005; Voutchkov, 2014).

8.6 SOURCE WATER PRETREATMENT PRIOR TO GRANULAR MEDIA FILTRATION

Most saline water particles and microorganisms have a slightly negative charge, which has to be neutralized by coagulation. In addition, these neutralized particles would need to be agglomerated in larger flocks that can be effectively retained within the filter media. Therefore, source-water conditioning by coagulation and subsequent flocculation are necessary prior to granular media filtration. Coagulation and flocculation processes and design criteria associated with source-water conditioning prior to filtration are discussed in Chapter 6.

Source water may need to undergo additional pretreatment prior to granular filtration (sand removal, sedimentation, DAF) depending on its quality. Alternative pretreatment processes and configurations are presented in Chapter 7.

8.7 PLANNING AND DESIGN CONSIDERATIONS

The design criteria presented herein should be used as guidelines only—media size, depth and configuration, especially for medium and large desalination plants, are recommended to be selected based on pilot testing for the site-specific conditions and water quality associated with the project for a period that encompasses worst-case scenario water quality (i.e., significant rain events of intensity higher than 15 mm, dredging near the intake area, algal bloom events, periods of strong seasonal winds, etc.).

8.7.1 Single-Stage Dual-Media Filters

8.7.1.1 Gravity Filters—Key Design Criteria

Key design criteria for single-stage dual-media gravity filters for medium- and large-size desalination plants are presented below:

Filter type	Dual-media, down-flow
Backwash	Air—water
Average filter cell run length	24 h
Flow distribution to Individual cells	Pipe (if concrete channel used, channel depth should be tapered to keep velocity in the distribution channel above 2 m/sec at all times).
Number of filter cells	8—18
Filter cell width	3—8 m (10—26 ft)

Filter cell depth	4.5–7.5/15–25 ft
Filter cell length-to-width ratio	2:1–4:1 (typically 3:1)
Individual filter cell area	25–100 m^2 (270–1100 ft^2)
Maximum water depth above filter bed	2.5 m/8 ft (should be equal or slightly higher than filter bed headloss, which usually is 1.8–2.4 m/6–8 ft).
FILTRATION RATE (AT DESALINATION PLANT INTAKE DESIGN FLOW)	
With all filters in service	8–10 m^3/m^2 h (3–4 gpm/ft^2)
With two filters out of service	15 m^3/m^2 h (6 gpm/ft^2)
FILTER MEDIA	
Top Layer: Anthracite or Pumice	
Anthracite/pumice layer—depth	1.0–1.8 m (2.6–6.0 ft)—for deep bed filters/high turbidity waters
Anthracite/pumice layer—depth	0.4–0.8 m for shallow bed filters—used for waters with low turbidity (<5NTU) and low organic content (TOC < 2 mg/L)
Anthracite/pumice—effective size	0.8–2 mm (typical 1.2 mm)
Anthracite/pumice—uniformity coefficient	1.3–1.7 (preferable < 1.4)
Anthracite—specific gravity	1.5–1.6 tons/m^3
Anthracite—bulk density	0.80–0.85 tons/m^3
Pumice—specific gravity	1.1–1.2 tons/m^3
Pumice—bulk density	0.40–0.55 tons/m^3
Bottom Layer: Sand	
Sand layer—depth	0.8–2.0 m for deep bed filters
Sand layer—depth	0.4–0.6 m—for shallow bed filters
Sand—effective size	0.4–0.6 mm
Sand—uniformity coefficient	<1.4
Sand—specific gravity	2.65 tons/m^3
Sand—bulk density	1.5–1.9 tons/m^3
Air–Water Filter Backwash System	
Maximum backwash rate	55 m^3/m^2 h (22 gpm/ft^2)
Average backwash rate	40–45 m^3/m^2 h (16–18 gpm/ft^2)
Duration (total air + water)	15–30 min (includes filter cell draining & fill up).

8.7.1.2 *Pressure Filters—Key Design Criteria*

The main design criteria for single-stage dual-media pressure filters in small- and medium-size desalination plants are very similar to these of gravity filters. Design criteria by which pressure filters differ from gravity filters are presented below:

Number of filter vessels	6–20
Filter vessel diameter	1.2–6 m/4–20 ft (typically 3 m/10 ft)
Filter vessel length	2.5–15 m/8–50 ft (typically 6 m/20 ft)
Depth of filter bed	0.6–0.9 m (2–3 ft)
FILTRATION RATE (AT DESALINATION PLANT INTAKE DESIGN FLOW)	
With all filters in service	12–25 m^3/m^2 h (5–10 gpm/ft^2)
With two filters out of service	30 m^3/m^2 h (12 gpm/ft^2)
HEADLOSS ACROSS THE FILTER VESSEL	
Total headloss across the filter	15–30 m/45–90 ft (avg. 20/65 ft)
Net headloss available for filtration	7.5–15 m (25–50 ft)

8.7.2 Two-Stage Filters

8.7.2.1 *Key Design Criteria*

Key design criteria for the first, coarse media, stage of two-stage media filtration system are as follows:

Filter type	Single- or dual-media, down-flow, air–water backwash
Average filter cell run length	24–48 h
FILTRATION RATE (AT DESALINATION PLANT INTAKE DESIGN FLOW)	
With all filters in service	12–25 m^3/m^2 h (5–10 gpm/ft^2)
With two filters out of service	30 m^3/m^2 h (12 gpm/ft^2)
ANTHRACITE OR SAND FILTER MEDIA	
Anthracite layer—depth	0.4–1.0 m (1.3–3.3 ft)
Anthracite—effective size	1.0–2.0 mm (typical 1.5 mm)
Anthracite—uniformity coefficient	<1.5
Sand layer—depth	0.4–1.0 m (1.3–3.3 ft)
Sand—effective size	0.4– 0.6 mm
Sand—uniformity coefficient	<1.5
AIR–WATER FILTER BACKWASH SYSTEM	
Maximum backwash rate	60 m^3/m^2 h (25 gpm/ft^2)

Average backwash rate	45–55 m^3/m^2 h (18–22 gpm/ft^2)
Duration (total air & water)	20–30 min (includes filter cell draining & fill up).

All other filter design parameters are the same as those of single-stage dual-media filters, which were described in the previous section. It should be noted that usually the second stage (polishing filter) is typically designed as a dual-media shallow filter, unless enhanced organics removal is needed.

8.7.3 Design Examples

The design examples provided below are developed for a hypothetical 50,000 m^3/day (13.2 MGD) seawater desalination plant designed for a recovery of 43%. The pretreatment filtration system has to be designed to produce the same quality of filtered water presented in Table 8.1.

8.7.3.1 Example of Single-Stage Dual-Media Gravity Filter

In this example, the source water quality has turbidity of 0.3–10 NTU (TSS of 0.5–15 mg/L) with occasional spikes of turbidity to 15 NTU (TSS = 20 mg/L) during rain events. The source water SDI_5 is in a range of 6–12. Maximum algal count is <20,000 cells/L and total hydrocarbon levels are below 0.04 mg/L at all times. This water quality is typical for an open ocean intake with medium depth (8–10 m/26–33 ft from the water surface).

Since open-ocean water does not contain elevated content of silica, iron, and manganese, these fouling compounds will not have impact on filter design. Because of the relatively high level of turbidity in the water, the backwash volume is expected to be 5% of the intake source water quality. Since the plant is designed for a total 43% recovery, the total volume of filtered water that will need to be produced for the RO system operations is 50,000 m^3/day/0.43 = 116,279 m^3/day. In addition, the filtration system will have to be designed to produce backwash water for the filters of volume, which is approximately 5% of the source water flow. As a result, the total plant intake flow for which the filters will need to be designed is 116,279 m^3/day × 1.05 = 122,093 m^3/day.

Table 8.4 presents a summary of the key design criteria for the dual-media (sand and anthracite) gravity filtration system for seawater pretreatment. The system is designed for air–water backwash and filtration cycles of 24–48 h depending on the source-water turbidity. The desalination plant source water is preconditioned using ferric chloride for coagulation and polymer for flocculation. The chemical source water-conditioning system also includes addition of sulfuric acid to maintain optimum pH for the coagulation process. The filtration system is designed with rinse to waste provisions, which allows to discharge the first 10–15 min of the flow produced of individual filtration cells immediately after their backwash to avoid sending this higher turbidity water to the RO-membrane system.

8.7.3.2 Example of Single-Stage Dual-Media Pressure Filter

Pressure filters are less costly than gravity filters in terms of construction expenditures. However, they apply significantly higher pressures to the algae in the source water and therefore, as single-stage filters they are preferable to be used when the source water is collected by very

TABLE 8.4 Example of Single-Stage Dual Granular Media Gravity Filtration System for 50,000 m^3/day (13.2 MGD) Desalination Plant

Component/Parameter	Specifications/Design Criteria
FEED WATER	
• Design flow rate, m^3/day (MGD)	122,093 (32.3 MGD)
• Turbidity, NTU	0.3–10 (max 20)
• SDI_5	6–10 (max 12)
DESIGN CHEMICAL DOSAGES	
• Ferric chloride, mg/L	10 (0.5–30 mg/L)
• Polymer, mg/L	0.25 (0.0–1.0 mg/L)
• Sulfuric acid, mg/L: Target pH: 7.6	8 (0–30 mg/L)
FILTER CELLS	
• Number	12
• Width, m (ft)	4 (13.2 ft)
• Length, m	15 (49.2 ft)
• Filter cell area	60 m^2 (649.44 ft^2)
• Total filter cell depth	6.0 m (19.7 ft)
• Total filter media depth	1.8 m (5.3 ft)
• Water depth above the filter bed	2.0 m (6.6 ft)
TOP-LAYER FILTER MEDIA	
• Type	Anthracite
• Depth	1.0 m (3.3 ft)
• Effective size	1.0 mm
• Uniformity coefficient	1.3
• Specific density, tons/m^3	1.55
BOTTOM-LAYER FILTER MEDIA	
• Type	Sand
• Depth	0.8 m (3.3 ft)
• Effective size	0.6 mm
• Uniformity coefficient	1.3
• Specific density, tons/m^3	2.65
FILTER PERFORMANCE PARAMETERS	
• Average surface loading rate	7.1 m^3/m^2 h (2.9 gpm/ft^2)
• Maximum surface loading rate with 2 units out of service	8.5 m^3/m^2 h (3.5 gpm/ft^2)
• Filtration cycle	24–48 h
BACKWASH RATE	
• Average	20 m^3/m^2 h (8 gpm/ft^2)
• Maximum	50 m^3/m^2 h (20 gpm/ft^2)

deep open-ocean intake (typically deeper than 15 m/50 ft), which collects water with limited algal content or by subsurface intakes (i.e., beach wells), which have already prefiltered algae contained in the water via slow-sand filtration.

In this example, the source-water quality has turbidity of 0.2–2 NTU (TSS of 0.5–5 mg/L) with consistent source-water quality, which is typically not affected significantly by algal blooms or rain events. The source-water SDI_5 is in a range of 3–6. Maximum algal count is <1000 cells/L and total hydrocarbon levels are below 0.04 mg/L at all times.

For the purposes of this example, it is assumed that the source water does not contain elevated content of silica, iron, and manganese and, therefore, these fouling compounds will not have impact on filter design. Because of the relatively low level of turbidity in the water, the average backwash volume is expected to be only 3% of the intake source-water quality. Since the plant is designed for a total 43% recovery, the total volume of filtered water that will need to be produced for the RO system operations is 50,000 m^3/day/0.43 = 116,279 m^3/day. In addition, the filtration system will have to be designed to produce backwash water for the filters of volume, which is approximately 3% of the source water flow. As a result, the total plant intake flow for which the filters will need to be designed is: 116,279 m^3/day × 1.03 = 119,767 m^3/day (31.6 MGD).

Key design criteria for the dual-media (sand and anthracite) pressure filtration system for seawater pretreatment are provided in Table 8.5. Similar to the previous example, the source water is preconditioned using ferric chloride for coagulation and polymer for flocculation. However, the chemical dosages are smaller and reflective of the better quality source water. The chemical source water conditioning system also includes addition of sulfuric acid to maintain optimum pH for the coagulation process.

8.7.3.3 Example of Two-Stage Gravity/Pressure Filter System

Two-stage filtration systems are usually applied when the source water is collected by a relatively shallow open intake (4–8 m/13–26 ft) from the water surface or by onshore open intake, which collects water from entire depth of the water column. In this case, the purpose of the first-stage granular gravity media filters is to remove the larger-size particles captured from the surface water column, such as large algae, silt, and solids.

Typically, the first-stage filters produce filtrate of turbidity below 5 NTU and SDI_5 in a range of 6–8 or lower. The main reason why the first-stage filters are selected to be gravity- rather than pressure-driven is to minimize the breakage of algae biomass and release of organics associated with it. The gravity filters operate at hydrostatic pressure, which practically eliminates such breakage by not allowing the algae to penetrate deep into the filters and by gently retaining them on the surface of the top layer of the filtration media.

The second-stage filters receive filtrate from the first stage and are designed to polish this filtrate to levels acceptable for its processing through the RO system (see Table 8.1). Because most of the algal mass in the source water is already removed by the coarser first-stage filters, the possibility for algal breakage and release of easily biodegradable organics in the second-stage filters is reduced significantly. This allows to design the second-stage filters as pressure-driven units and to benefit from the lower costs associated with the use of this type of filters.

The source water in this example is of worse quality than that of the previous two examples and has turbidity of 2–30 NTU (TSS of 5–50 mg/L) with occasional turbidity spikes of up to 50 NTU during the most severe period of algal blooms occurring periodically in the intake area.

TABLE 8.5 Example of Single-Stage Dual Granular-Media Pressure Filtration System for 50,000 m^3/day (13.2 MGD) Desalination Plant

Component/Parameter	Specifications/Design Criteria
FEED WATER	
• Design flow rate, m^3/day (MGD)	119,767 (31.6 MGD)
• Turbidity, NTU	0.2–2 (max 5)
• SDI_5	3–6 (max 8)
DESIGN CHEMICAL DOSAGES	
• Ferric chloride, mg/L	1 (0.5–5 mg/L)
• Polymer, mg/L	0.15 (0.0–0.5 mg/L)
• Sulfuric acid, mg/L: Target pH: 6.7	8 (0–30 mg/L)
FILTER CELLS	
• Number	10
• Diameter, m (ft)	3.5 (11.5 ft)
• Length, m	12 (39.4 ft)
• Filter vessel area	42 m^2 (452 ft^2)
• Total filter media depth	1.2 m (4.0 ft)
TOP-LAYER FILTER MEDIA	
• Type	Pumice
• Depth	0.6 m (3.3 ft)
• Effective size	0.6 mm
• Uniformity coefficient	1.3
• Specific density, tons/m^3	1.15
BOTTOM-LAYER FILTER MEDIA	
• Type	Sand
• Depth	0.6 m (3.3 ft)
• Effective size	0.4 mm
• Uniformity coefficient	1.3
• Specific density, tons/m^3	2.65
FILTER PERFORMANCE PARAMETERS	
• Average surface loading rate	11.9 m^3/m^2 h (4.9 gpm/ft^2)
• Maximum surface loading rate with 2 units out of service	14.9 m^3/m^2 h (6.1 gpm/ft^2)
• Filtration cycle	24–48 h
BACKWASH RATE	
• Average	35 m^3/m^2 h (1 gpm/ft^2)
• Maximum	60 m^3/m^2 h (25 gpm/ft^2)

The source water SDI_5 is in a range of 10–16. Maximum algal count is <40,000 cells/L and total hydrocarbon level is below 0.10 mg/L at all times.

Similar to the other two examples, in this example, it is also assumed that the source water does not contain elevated contents of silica, iron, and manganese—therefore, these fouling compounds do not have impact on filter design. Because the pretreatment system consists of two-stage filtration, the overall volume of backwash water is expected to increase, with the first-stage filter daily backwash water volume equal to 6% of the total daily plant intake flow and the second-stage pressure filters using 4% of the intake flow.

Taking under consideration that the plant is designed for the same 43% recovery, the total volume of filtered water that will need to be produced for the RO-system operations is still 50,000 m^3/day/0.43 = 116,279 m^3/day. However, to accommodate the 10% (6% for first-stage filters + 4% for second-stage filters) of additional flow for backwashing of the two stages of the pretreatment filters, the intake source water volume will be 116,279 m^3/day × 1.1 = 127,907 m^3/day (33.8 MGD).

In this case as with the other two examples, the source water is preconditioned using ferric chloride for coagulation, polymer for flocculation, and sulfuric acid for pH adjustment to optimize the use of coagulant. While provisions for addition of coagulant are usually incorporated ahead of each of the two filtration stages, most of the particle precipitation, coagulation, and flocculation occur in the coagulation/flocculation chambers upstream of the first-stage filters. Therefore, coagulant addition upstream of the second-stage filters is not typically practiced. However, facilities for in-line coagulation and flocculation of the feed to the second-stage filters are sometimes installed, because for certain periods of the year, if the water quality is very good, the first-stage filtration can be bypassed. In this case, coagulant and flocculant are fed directly into the inlet to the second-stage filters or coagulation, and flocculation facilities are installed upstream of the fist-stage filters, and bypass line is provided from these facilities directly to the feed of the second-stage filters.

It should be pointed out that two-stage filtration systems are not very commonly used because in most existing desalination projects, the location and depth of the intake are typically selected such that the collected source water is of good quality and requires single-stage filtration only. However, if the intake for the desalination plant is built in a shallow water body (e.g., the Persian Gulf) or the RO desalination plant is colocated with thermal desalination plant and/or power plant with onshore intake, then the use of two-stage filtration system is warranted.

An alternative solution for very large desalination plants [400,000 m^3/day (106 MGD) or more] would be to construct deep tunnels under the ocean bottom, which extend outside of the tidal zone at locations where intake can be built at depth of 15 m (50 ft) or more below the water surface. Experience with construction of such intakes in Australia indicates that the costs for such structures could be much higher than the costs for construction of second-stage pretreatment system. In such cases, two-stage filtration systems could become a more attractive and cost-effective solution to minimize the biofouling potential of the saline source water than deeper intakes.

Tables 8.6 and 8.7 present key design criteria of the first- and second-stage of a two-stage pretreatment system of 50,000 m^3/day plant with shallow or onshore intake.

8.8 CONSTRUCTION COSTS OF GRANULAR MEDIA FILTRATION SYSTEMS

Fig. 8.9 depicts construction costs for the year 2017 for granular media gravity and pressure filters as a function of the desalination plant intake flow they pretreat. As seen from this figure,

TABLE 8.6 Example of First-Stage Gravity Filters of Two-Stage Filtration System for 50,000 m^3/day (13.2 MGD) Desalination Plant

Component/Parameter	Specifications/Design Criteria
FEED WATER	
• Design flow rate, m^3/day (MGD)	127,907 (33.8 MGD)
• Turbidity, NTU	2–30 (max 50)
• SDI_5	8–12 (max 16)
DESIGN CHEMICAL DOSAGES	
• Ferric chloride, mg/L	15 (2–40 mg/L)
• Polymer, mg/L	0.25 (0.25–1.5 mg/L)
• Sulfuric acid, mg/L: Target pH: 6.7	8 (0–30 mg/L)
FILTER CELLS	
• Number	10
• Width, m (ft)	4 (13.2 ft)
• Length, m	15 (49.2 ft)
• Filter cell area	60 m^2 (649.44 ft^2)
• Total filter cell depth	5.4 m (19.7 ft)
• Total filter media depth	1.2 m (5.3 ft)
• Water depth above the filter bed	2.0 m (6.6 ft)
TOP-LAYER FILTER MEDIA	
• Type	Anthracite
• Depth	0.6 m (3.3 ft)
• Effective size	1.4 mm
• Uniformity coefficient	1.4
• Specific density, tons/m^3	1.55
BOTTOM-LAYER FILTER MEDIA	
• Type	Sand
• Depth	0.6 m (3.3 ft)
• Effective size	0.8 mm
• Uniformity coefficient	1.4
• Specific density, tons/m^3	2.65
FILTER PERFORMANCE PARAMETERS	
• Average surface loading rate	8.9 m^3/m^2 h (3.6 gpm/ft^2)
• Maximum surface loading rate with 2 units out of service	11.1 m^3/m^2 h (4.5 gpm/ft^2)
• Filtration cycle	24–48 h
BACKWASH RATE	
• Average	27 m^3/m^2 h (11 gpm/ft^2)
• Maximum	55 m^3/m^2 h (23 gpm/ft^2)

TABLE 8.7 Example of Second-Stage Pressure Filters of Two-Stage Filtration System for 50,000 m^3/day (13.2 MGD) Desalination Plant

Component/Parameter	Specifications/Design Criteria
FEED WATER	
• Design flow rate, m^3/day (MGD)	120,930 (32 MGD)
• Turbidity, NTU	0.2–5 (max 10)
• SDI_5	4–8 (max 10)
DESIGN CHEMICAL DOSAGES	
• Ferric Chloride, mg/L	0.0 (0.0–8 mg/L)
• Polymer, mg/L	0.0 (0.0–0.5 mg/L)
• Sulfuric acid, mg/L: Target pH: 6.7	0.0 (0.0–30 mg/L)
FILTER CELLS	
• Number	10
• Diameter, m (ft)	3.5 (11.5 ft)
• Length, m	12 (39.4 ft)
• Filter vessel area	42 m^2 (452 ft^2)
• Total filter media depth	1.2 m (4.0 ft)
TOP-LAYER FILTER MEDIA	
• Type	Sand
• Depth	0.6 m (3.3 ft)
• Effective size	0.6 mm
• Uniformity coefficient	1.3
• Specific density, tons/m^3	2.65
BOTTOM-LAYER FILTER MEDIA	
• Type	Garnet
• Depth	0.6 m (3.3 ft)
• Effective size	0.3 mm
• Uniformity coefficient	1.3
• Specific density, tons/m^3	4.10
FILTER PERFORMANCE PARAMETERS	
• Average surface loading rate	12.0 m^3/m^2 h (4.9 gpm/ft^2)
• Maximum surface loading rate with 2 units out of service	15.0 m^3/m^2 h (6.1 gpm/ft^2)
• Filtration cycle	24–48 h
BACKWASH RATE	
• Average	35 m^3/m^2 h (14 gpm/sq ft)
• Maximum	60 m^3/m^2 h (25 gpm/sq ft)

pressure filters have relatively lower cost than gravity filters for the same daily volume of pretreated saline source water.

For example, in the case of 50,000 m^3/day (13.2 MGD) single-stage gravity filtration system presented in Table 8.4, the plant intake flow is 122,093 m^3/day (32.3 MGD). For this flow, the estimated construction cost of this pretreatment system is US$ 8.9 million.

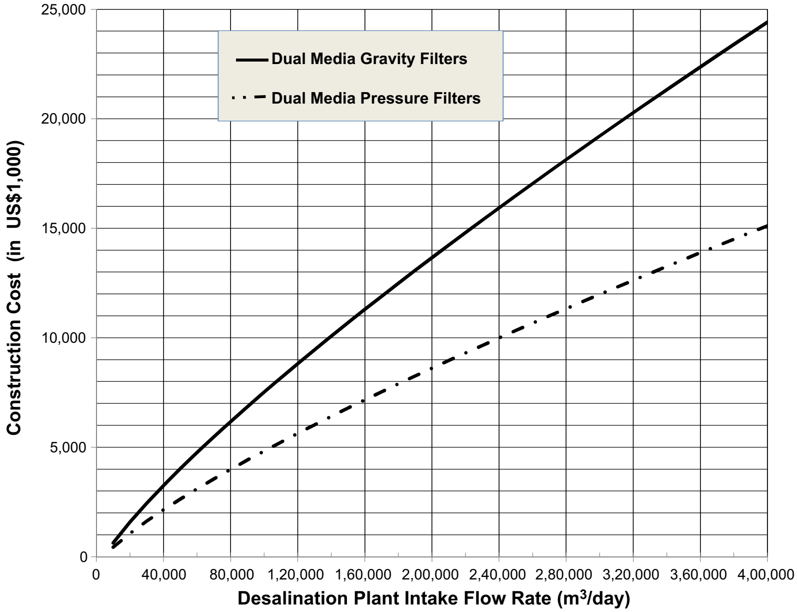

FIGURE 8.9 Construction costs of dial-media gravity- and pressure filters.

For the example of pressure-driven filtration system (Table 8.5), which has a feed flow of 119,767 m^3/day, the estimated construction cost of this system is US$ 5.6 million.

Similarly, for the example of two-stage gravity/pressure granular media filtration pretreatment system for RO plant with recovery of 43%, the construction cost of the first-stage filters that have feed flow of 127,907 MGD (33.8 MGD) is US$ 9.0 million, and the cost of the second-stage pressure filters (at feed flow of 120,930 m^3/day) is US$ 5.7 million. As a result, the total filtration pretreatment cost for this plant is US$ 14.7 million. Comparison of the costs of these three examples, underlines an important fact, i.e., the construction cost of the pretreatment system for desalination plant of the same fresh water production capacity can vary in a wide range (in this case, between US$ 5.6 and US$ 14.7 million) depending on the saline source water quality.

References

AWWA, 2007. Manual of Water Supply Practices M46. Reverse Osmosis and Nanofiltration, second ed. American Water Works Association, Denver, Colorado.

Beverly, R.P., 2005. Filter Troubleshooting and Design Handbook. American Water Works Association, Denver, Colorado.

Kawamura, S., 2000. Integrated Design and Operation of Water Treatment Facilities, second ed. John Wiley & Sons, Inc.

Qasim, S.R., Motley, E., Zhu, G., 2000. Water Works Engineering—Planning, Design & Operation. Prentice Hall PTR, Upper Saddle River, New Jersey.

Voutchkov, N., 2014. Desalination Engineering: Operation and Maintenance. McGraw Hill, New York.

Wilf, M., Awerbuch, L., Bartels, C., Mickley, M., Pearce, G., Voutchkov, N., 2007. The Guidebook to Membrane Desalination Technology: Reverse Osmosis, Nanofiltration and Hybrid Systems Process Design, Applications and Economics. Balaban Publishers, Rehovot.

CHAPTER

9

Membrane Filtration

OUTLINE

Pretreatment for Reverse Osmosis Desalination
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9.1 INTRODUCTION

Particulate, colloidal inorganic, and some of the solid and colloidal organic foulants contained in the saline source water can be removed successfully using microfiltration (MF) or ultrafiltration (UF) membrane pretreatment. Fig. 9.1 depicts a general schematic of water desalination plant with membrane pretreatment. As indicated on this figure, saline water pretreatment includes several key components: (1) coarse and fine screens similar to those used for plants with conventional pretreatment; (2) microscreens to remove fine particulates and sharp objects from the water that could damage the membranes; and (3) UF or MF membrane system.

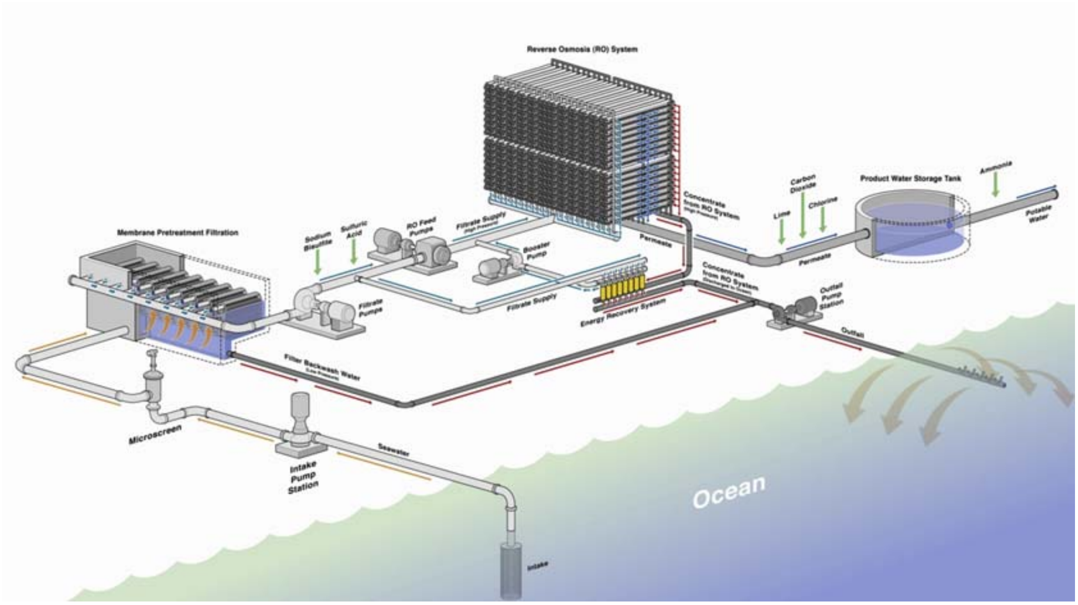

FIGURE 9.1 General schematic of desalination plant with membrane pretreatment.

Depending on the force (pressure or vacuum) driving the filtration process, membrane pretreatment systems are classified as pressure driven (pressurized) and vacuum driven (submerged). Based on the size of membrane pores, the membrane systems used for pretreatment are classified as MF and UF, with pore sizes of 0.04–0.1 μm and 0.01–0.02 μm, respectively. Although earlier generations of MF elements had pore sizes of 0.1–0.2 μm, at present the difference between MF and UF element pores is only a factor of 2–3. Table 9.1 provides a list of some of the large-size seawater desalination plants with membrane pretreatment in operation at present.

One of the largest pressure membrane pretreatment systems in the world is located in Binningup, Australia, at the 300,000 m^3/day (79 MGD) Southern seawater desalination plant. The largest desalination plant with vacuum-driven membrane pretreatment is the 300,000 m^3/day (79 MGD) Adelaide desalination facility in Australia (Fig. 9.2).

TABLE 9.1 Large Desalination Plants With Membrane Pretreatment

Desalination Plant Location and Capacity	Pretreatment System Type, Configuration, and Supplier	Hydraulic Loading Rate	Notes
Shuwaikh, Kuwait 350,000 m^3/day (92 MGD)	Pressure ultrafiltration (UF) Norit/ Pentair	60–77.5 L/m^2 h	Shallow intake
Adelaide seawater reverse osmosis (SWRO) desalination plant, Australia—300,000 m^3/day (79 MGD)	Submerged UF membranes Memcor-Evoqua	52–64 L/m^2 h	Deep open intake
Southern seawater desalination project—300,000 m^3/day (79 MGD)	Pressure UF membranes Memcor-Evoqua	40–50 L/m^2 h	Shallow open intake
Quigdao SWRO plant, China—232,000 m^3/day (61 MGD)	Pressure UF system Norit-Pentair	80–93 L/m^2 h Avg.—85 L/m^2 h	Open intake
Palm Jumeira, UAE, two SWRO plants @ 96,000 m^3/day each	Pressure UF system Norit-Pentair	60–80 L/m^2 h Avg.—85 L/m^2 h	Open intake
Bekton desalination plant, London, UK—150,000 m^3/day (40 MGD)	Pressure UF system Norit-Pentair	40–70 L/m^2 h	Saline river intake
Fukuoka SWRO plant, Japan—96,000 m^3/day (25 MGD)	Pressure UF membranes Hydranautics	60–80 L/m^2 h	Subsurface intake
Kindasa SWRO plant, Saudi Arabia—90,000 m^3/day (24 MGD)	Dual media pressure filtration followed by pressure UF Hydranautics	Dual media filtration rate—15–20 m^3/m^2 h UF flux—80–100 L/m^2 h	Near shore open intake in industrial port
Yu-Huan SWRO plant, China—36,000 m^3/day (10 MGD)	Submerged UF membranes GE Zenon	40–60 L/m^2 h	Open intake

FIGURE 9.2 Ultrafiltration (UF) pretreatment system of the Adelaide desalination plant, Australia.

Although at present less than 10% of all existing desalination plants worldwide have UF or MF pretreatment, application of membrane filtration for saline water pretreatment is gaining a wider acceptance over the past 10 years (Busch et al., 2009; Lazaredes and Broom, 2011). The number of medium and large desalination plants with membrane pretreatment has increased from less than a half a dozen in 2002 to over 40 in 2011 (Gasia-Brush et al., 2011).

Practical experience at the 40,000 m^3/day (10.6 MGD) Addur seawater desalination plant in Bahrain (Burashid and Hussain, 2004) and the 20,000 m^3/day (5.3 MGD) Sohar RO1 desalination plant (Shahid and Al Sadi, 2015) shows that membrane pretreatment alone may not always provide competitive solution, especially for challenging saline source waters of high, organic content, and biofouling potential.

A number of desalination plants with relatively shallow intakes and source water of high turbidity, algal content, and organic loads (i.e., Shuwaikh, Bekton, Kindasa) have additional pretreatment steps before the membrane filtration system to cope with these source water quality challenges. However, in most existing desalination installations worldwide, membrane filtration is the only pretreatment system before reverse osmosis (RO) desalination.

9.2 THE MEMBRANE FILTRATION PROCESS

All membrane pretreatment processes operate in cycles that include the following four activities: (1) processing, (2) backwash, (3) cleaning, and (4) integrity testing. The four operational modes are typically monitored and controlled by a programmable logic controller.

9.2.1 Source Water Processing (Filtration)

The membrane filtration of the saline source water takes place during the processing phase. Depending on the specific membrane product and the membrane configuration, the filtration process can either occur in direct-flow or in cross-flow mode. In direct flow mode, all of the source water passes through the membranes. In cross-flow operations, only a portion of the source flow (typically 90%–95%) passes through the membranes, whereas the remaining flow (reject) travels along the feed side of the membranes and its movement along the membrane surface generates shearing velocity that evacuates the solids removed from the saline water out of the membrane.

Usually in a cross-flow mode, a portion of the reject stream is recirculated back to the feed system. Cross-flow pattern in such elements is similar to that in RO membrane elements. The key benefit of such flow pattern is that membranes can be operated continuously. The main problem with cross-flow elements, however, is that they have relatively lower packing density that limits their productivity and requires significant energy expenditure to maintain flow tangential to the membrane surface. Therefore, the newest and most commonly used UF and MF membrane elements available on the market are designed to operate in a direct-flow configuration. Direct-flow membranes, however, cannot be operated continuously because the solids in the source water rejected by the membranes are accumulated on the membrane surface and have to be removed periodically via intermittent backwash.

The two most important membrane performance parameters associated with the filtration cycle of membrane pretreatment systems are membrane flux and trans-membrane pressure (TMP).

9.2.1.1 *Membrane Flux*

The membrane flux is the volume of filtered water produced by a unit of membrane area. This parameter is most commonly measured in liters per square meter per hour (L/m^2 h) and also often referred to as Lmh or lmh, and gallons per day per square foot (gfd). The two flux measures relate as follows: 1.0 gfd = 1.705 lmh. Typically, pretreatment systems of desalination plants are designed for flux of 40–80 lmh (24–47 gfd; see Table 9.1).

Accumulation of solids on the surface of the membranes and in the membrane pores (membrane fouling rate) increases with the increase of the membrane flux. To maintain a reasonably long membrane cycle (30 min or more), the operational flux has to be selected such that the fouling rate on the membranes is reasonably low and within the time of one filtration cycle, the pressure loss created by the solids accumulated on the membranes stays below the maximum pressure that the membrane feed pumps are designed to deliver. Such flux is also referred to as a sustainable flux. Usually, the higher the solids content and fouling potential of the source water, the lower the sustainable flux for the same filtration cycle length.

9.2.1.2 *Trans-membrane Pressure*

TMP is the difference between the feed pressure and the filtrate pressure of the pretreatment system. This pressure drives the flow through the membranes and, therefore, it is directly related to the membrane flux. The TMP also has an impact on membrane fouling and filtration cycle length.

For most UF and MF membrane systems used for saline water pretreatment, the TMP is usually reported in bars or pounds per square inch (psi). Sometimes, TMP is reported in kilopascals (kPa) with 1 bar = 100 kPa. Typically, pretreatment systems operate at TMP between 0.2 and 1.0 bar (2.8–14.2 psi). Pressure-driven systems can operate at TMPs higher than 1 bar, whereas the theoretical maximum operating vacuum of the vacuum driven (submerged systems) is 1 bar. However, due to the potential for collapsing of the membrane fibers by excessive vacuum, the practical maximum TMP of submerged membrane pretreatment systems is usually limited to 0.7–0.8 bars (10–11 psi).

Depending on the strength of their membranes, pressure systems can operate at TMP of up to 2.5 bars (35 psi). The higher the durability and flexibility of the membrane fibers, the higher the maximum pressure the membrane fibers can handle. Based on this and other criteria the maximum TMP varies by the supplier.

9.2.1.3 Membrane Permeability

Similar to RO membranes, the third critical performance parameter associated with the filtration cycle is the membrane permeability. This parameter is defined as the ratio between the membrane flux and the TMP. Membrane permeability is measured in lmh/bar or gfd/psi (1 gfd/psi = 25 lmh/bar). Most MF and UF membrane elements used for saline water pretreatment operate at membrane permeability of 75–500 lmh/bar (5–20 gfd/psi).

9.2.2 Membrane Backwash

During processing mode, solids filtered out of the source water accumulate on the feed side of the membrane surface. These solids are periodically removed out of the filtration system by backwashing of the membranes with filtered water or concentrate. Backwash is usually triggered by timer and occurs every 20–120 min for approximately 30–60 s. Backwash could also be initiated when the TMP reaches a certain maximum threshold, beyond which the membrane system cannot perform at target flux and filtered water quality. If the threshold TMP is exceeded, typically the membrane system production capacity (flux) is decreased, the filtered water quality deteriorates, and the membranes could be exposed to irreversible and sometimes permanent fouling. However, most often in practice the pretreatment membrane backwash is triggered by preset time. This time is adjusted based on the amount of solids in the source water and its fouling potential.

For saline source waters with high content of turbidity, silt, and dissolved organics, the filtration cycle could be as short as 20–30 min. For source waters with very low organic and solids content such as these collected via intake wells or deep open intakes, the filtration cycle could extend beyond 60 min and backwash may not be needed for periods of several hours.

Membrane backwash is a multistep process, which for most commercially available UF and MF systems applies a combination of filtered water and air in a sequence and at rates designed to maximize the removal of particulates that have accumulated in the membrane system during the processing cycle. Backwash plays a very important role in the normal operation of membrane systems because membranes have significantly smaller volume/capacity available to store solids within the filtration vessels than granular media filters. This smaller solid's retention capacity of membrane systems is the main reason why membrane modules have to be backwashed 30–50 times more frequently than granular media filter cells (i.e., typically once every 30 min vs. one every 24 h).

Air and water backwash is mainly intended to remove particulates out of the membrane pretreatment system and does not involve the use of any cleaning chemicals. However, over time membrane surface would also accumulate organic deposits and biofilm. This type of membrane fouling is controlled by chemically enhanced backwash (CEB) also referred to as maintenance wash, which is typically practiced once or two times per day.

During CEB, the membranes are soaked for several minutes in chlorine and sometimes other cleaning chemicals (acids, alkali, or sodium bisulfite) and are than backwashed. The needed chemical dosages are a function of the predominant type and amount of foulants in the source water and of the type of membrane material and configuration.

Periodic membrane backwash and CEB do not completely eliminate membrane fouling and, therefore, the TMP needed to produce filtered saline water of target volume (flux) and quality increases over time. Once the TMP reaches a certain preset maximum level (typically 0.7–0.8 bar for submerged systems and 1.5–2.5 bars for pressurized systems), the membrane modules have to be taken offline and cleaned with chemicals that aim to reduce the TMP back to a reasonable level. Such a type of deep membrane cleaning aiming at the recovery of the membrane permeability, which cannot be achieved by CEB, is referred to as clean-in-place (CIP).

Membrane CIP is typically needed every 1–3 months and is performed using a combination of low-pH solution of citric or sulfuric acid followed by a high-pH solution of sodium hydroxide and sodium hypochlorite. The cleaning chemicals are recirculated through the membranes for a period of 4–24 h and then the membranes are flushed and returned back to normal operation. Depending on the nature of the fouling, sometimes other cleaning chemicals (biocides) are used to address specific fouling compounds (i.e., oil and grease, excessive biogrowth, etc.).

It should be pointed out that UF and MF pretreatment units have CIP system separate from that of the RO membrane racks, despite the fact that many components of the two CIP systems are similar.

Microbial biofouling is one of the most difficult types of fouling to remove from the UF and MF membranes and usually recovery of membrane permeability from such fouling requires very high dosage of chlorine (e.g., over 500 mg/L). If the membranes used for pretreatment of saline water of high biofouling potential have low chemical resistance and durability, and they can be damaged at chlorine dosages over 150 mg/L, while appropriate cleaning of the membranes requires chlorine dosage of 500 mg/L, then such membranes will lose permeability quickly and they will need to be cleaned in place very frequently (e.g., once every 1–2 weeks). For comparison, if more costly but more durable membranes that can withstand chlorine dosages of 500 mg/L or more are used to pretreat highly fouling source water, the membrane CIP frequency would typically be reduced to once every several months.

Typically, the lowest cost UF and MF membrane products available on the market also have the lowest strength and chemical resistance and therefore, the integrity of such membranes can be easily damaged by exposure to high dosages of membrane cleaning chemicals (oxidants, acids or bases). Therefore, such UF or MF membranes should only be used for pretreatment of saline source water of low fouling potential.

9.2.3 Membrane Cleaning

Typically, inorganic foulants such as iron and manganese solids and colloids accumulated on the membrane surface are cleaned with citric, sulfuric acid, or hydrochloric acid. Fouling caused mainly by organic materials is treated by base such as sodium hydroxide, whereas biological and algal fouling is cleaned using disinfectants such as sodium hypochlorite or peracetic acid, and surfactants. In all cases, the CIP is completed with heated cleaning solutions and RO permeate or filtrate. Using RO permeate for CIP cleaning is preferable.

9.2.4 Membrane Integrity Testing

All membrane pretreatment systems are equipped with integrity testing features that allow detecting occasional loss of membrane filtration capability caused by breaks or punctures of the membrane fibers or leafs, cracks of the membrane modules, piping, and connectors, and other problems that could occur during membrane production, installation, or operation. The most widely used membrane system integrity test is a pressure-hold/visual test, which is performed while the membrane pretreatment module/rack is offline.

During the pressure hold test, water is purged from the system using filtrate and then air is applied under pressure of 0.3–1.0 bar (4.2–14.2 psi) and decay of air pressure is monitored over time. Typically, the membrane module integrity is adequate when the pressure loss over 5-min period is less than 10% of the initial pressure applied to the membranes. The pressure decay test procedures vary for the various commercially available membrane products and configurations and, therefore, the membrane integrity testing system and conditions have to be coordinated with the membrane system supplier/equipment manufacturer.

Besides pressure-hold test, other offline membrane integrity tests used are vacuum hold test, bubble point test, diffusive air flow tests, etc. (AWWA, 2005). In addition, membrane integrity is monitored online by particle passage counting, filter effluent turbidity measurements of the individual membrane modules, or acoustic sensing.

The most popular method for online membrane integrity monitoring is the continuous effluent turbidity measurement of the individual membrane modules (trains) comprising the pretreatment system. Usually, the breach of integrity of a given train/module in a system is identified by comparing the filtered effluent turbidity of the train/module to that of the average turbidity from all modules and the turbidity of the other membrane pretreatment racks.

9.3 KEY FILTRATION SYSTEM COMPONENTS

9.3.1 Filter Vessels/Modules

MF and UF filtration membranes are configured in individual functional filtration units referred to as modules. Most commonly used membrane module configurations are hollow fiber, tubular, flat sheet, and spiral. The membrane modules are contained in housings, shells, or cassettes, which are assembled into larger membrane filtration system components, vessels and racks.

9.3.2 Membrane Filtration Media

9.3.2.1 Membrane Materials

Membranes used for saline water pretreatment are typically made of polyethersulfone (PES), polyvinylidene difluoride (PVDF), or polysulfone (PS). All of the membrane products made of these materials are hydrophilic. The PES is the most hydrophilic of all of these materials. Hydrophilic materials have two key advantages: (1) they do wet easily which makes them more permeable for a given pore size; and (2) they have higher resistance to attachment

TABLE 9.2 Materials of Membrane Filtration/Ultrafiltration (MF/UF) Membrane Products Used for Saline Water Pretreatment

Membrane Manufacturer	Type of Membrane	Membrane Material	Direction of Flow
X-flow—Pentair	Pressure-driven UF	Polyethersulfone (PES)	Inside-out
Memcor—Evoqua	Pressure- and vacuum-driven MF	Polyvinylidene difluoride (PVDF)	Outside-in
Hydranautics—Nitto Denko	Pressure-driven UF	PES	Inside-out
Hyflux	Pressure-driven UF	PES and PVDF	Outside-in
GE Zenon	Vacuum-driven UF	PVDF	Outside-in
Dow	Pressure-driven UF	PVDF	Outside-in
Toray	UF and MF	PVDF	Outside-in
Pall/Asahi	Pressure-driven MF	PVDF	Outside-in
Inge (BASF)	Pressure-driven UF	PES	Inside-out
Koch	Pressure-driven UF	Polysulfone (PS) and PES	Inside-out

of organic materials on their surface (i.e., to biofouling). Table 9.2 presents summary of key membrane suppliers and the type and material of their membrane products.

Overview of the information in Table 9.2 indicates that the most commonly used membrane materials are PVDF and PES. Pearce (2011) provides a detailed information of the benefits of alternative membrane materials. In general, PES has a higher permeability but lower durability and chlorine resistance. Conversely, PVDF membranes have higher strength and flexibility. Such material features are very important if the membranes will be exposed to high pressures/pressure surges and have to withstand significant mechanical stress. These membrane characteristics, especially material strength and flexibility, are of critical importance if the membranes will be cleaned by air backwash, which typically creates a significant stress on the membrane fibers and potting service. This is also one of the main reasons why PES membrane systems are backwashed with water only. Accidental release of large amounts of air in the feed to PES systems could cause fiber breakages and potting interface challenges.

Conditions of elevated fiber breakage could occur, for example, in pretreatment system/RO system configuration without break tank/s between the intake pump station, the pretreatment system, and the RO feed piping. In this configuration of direct coupling of the pretreatment and RO systems, the desalination plant intake pumps convey source water through the membrane pretreatment system into the suction of the booster pump of the RO feed system. Because the required booster pump suction pressure is usually in a range of 2.0—3.5 bars, the membrane pretreatment system has to be designed to convey water at this pressure. This direct coupling configuration results in the feed pressure to the pretreatment system in a range of 2.7—5.0 bars, which is relatively close to the maximum pressures that many of the commercially available membrane pretreatment systems are designed to operate at (i.e., around 6—8 bars).

If a hydraulic surge is triggered by the intake pumps and the surge protection of the pipeline connecting the intake pump station to the feed of the membrane pretreatment system is inadequate, this surge could create pressures along the water flow path well above two times the operating pressure (i.e., over 5 bars), which the membrane material and potting interface may not be able to withstand, if their strength is inadequate. As a result, current practice shows that use of PES pretreatment membranes with low-pressure rating (e.g., 3 bars) in seawater reverse osmosis (SWRO) plants with direct coupling configuration experiences frequent fiber breakages, loss of membrane integrity, and short useful life.

In general, PVDF membrane systems are better suited to handle pressure surges and have enhanced durability and chlorine resistance as compared to PES membranes. However, this type of membrane product is also more expensive and has lower permeability and caustic resistance than PES. In addition, PES membranes have narrower pore size distribution, which could be beneficial in terms of the filtered water quality.

Practical experience to date shows that PVDF membranes are more suitable than PES membranes for the treatment of saline source waters of high biofouling potential such as these of the Persian Gulf and Red Sea. However, PVDF membranes are more costly and despite the obvious disadvantages of PES membranes for such highly fouling waters, to date PES membranes have found wider application because of their lower costs, especially in countries with procurement regulations where equipment and membrane selection are based on the lowest capital costs only and the quality of the products is not factored in the evaluation process (e.g., design-build project procurement).

9.3.2.2 Membrane Geometries

The most widely used membranes in saline water pretreatment have hollow fiber, tubular, and spiral wound geometry. Hollow fiber membranes typically consist of several hundred to several thousand membrane fibers bounded at each end by epoxy or urethane resin and encased in individual modules. Typically, the internal diameter of the membrane fibers is 0.4–1.5 mm.

Depending on the membrane manufacturer, the hollow-fiber (capillary) membrane elements may be operated in an inside-out or outside-in flow pattern. Inside-out mode of operation provides a better control of flow and more uniform flow distribution than outside-in flow pattern. However, outside-in flow pattern usually results in lower headlosses through the module and operation under this pattern is less sensitive to the amount of solids in the saline source water.

Tubular membranes have inner tube diameters, which are an order of magnitude larger than that of hollow fiber membranes (i.e., 1.0–2.5 cm). The individual membrane tubes are placed inside fiberglass-reinforced plastic or stainless steel tube and the two ends of the tube are sealed with gasket or other clamp-type device. The typical flow pattern for these membranes is inside-out—i.e., the saline source water is introduced into the tube lumens under pressure and it flows through the walls of the membrane tubes into the outside shell of the module.

The key advantages of hollow fiber membrane elements are as follows: (1) high surface area to volume (packing density), which allows the reduction of the overall footprint of the filtration system; (2) easier backwash; (3) filtration at lower feed pressure—TMP is typically 0.2–1.0 bar; (4) lower pressure drop across the membrane modules (0.1–1.0 bar).

Tubular membranes have the following advantages: (1) large-channel diameters allow them to treat waters of higher solid content compared to hollow-fiber membranes operating in an inside-out mode (this advantage is not significant if an outside-in hollow-fiber membrane is compared with inside-out tubular membrane); (2) these membranes can be operated at approximately two times higher cross-flow velocity, which is beneficial in terms of biofouling control.

9.3.3 Service Facilities and Equipment

All membrane pretreatment systems have three types of service support facilities and equipment: (1) backwash system, (2) CIP system, and (3) cleaning chemical feed system. The backwash system typically includes filtered water storage tank, backwash pumps, and (depending on the membrane type) air compressors for air backwash. Some membrane pretreatment systems (i.e., Pentair's, Seaguard) use only water backwash.

The CIP system for the membrane pretreatment facility has very similar configuration to that of the RO CIP system. Sometimes, the same CIP system is used for both pretreatment and RO membrane cleaning, which is not desirable, especially for larger desalination plants because of the added operational complexity. The chemical feed system usually includes acid, base, sodium hypochlorite, and sodium bisulfite storage and feed systems to service the CEB and CIP membrane cleaning activities. In addition, some of the membrane pretreatment systems are designed for enhanced performance by addition of conditioning chemicals to the feed water (coagulants, flocculants, powdered activated carbon, and acid). However, most saline water pretreatment systems are designed to operate without source water conditioning.

9.4 FILTER TYPES AND CONFIGURATIONS

Depending on the type of the driving filtration force, membrane pretreatment filters are divided in two categories—pressure driven (pressurized) and vacuum driven (submerged).

9.4.1 Pressurized Membrane Systems

Pressurized UF and MF systems consist of membrane elements installed in pressure vessels, which are grouped in racks (trains), similar to these of RO systems (see Fig. 9.3). At present, practically all key membrane suppliers offer pressurized UF and MF systems. Depending on the direction of the feed flow through the membranes, the pressure-driven systems are divided into outside-in [also referred to as pressure-driven outside (PDO) feed] systems and inside-out [or pressure-driven inside (PDI) feed]. In PDO systems, the feed source water is distributed around the filter fibers and after passing through the membranes, the filtered water is collected through the fiber lumen. In PDI systems, the source water is fed into the filter lumens and is collected on the outside of the fibers.

In general, PDO systems are more difficult to clean only with water backwash and require air backwash to achieve the same level of productivity recovery as PDI. In addition, PDO either operate at higher feed pressures for the same design flux as PDI systems or at lower fluxes at the same design TMP.

FIGURE 9.3 Pressurized membrane filtration system with vertical membrane elements.

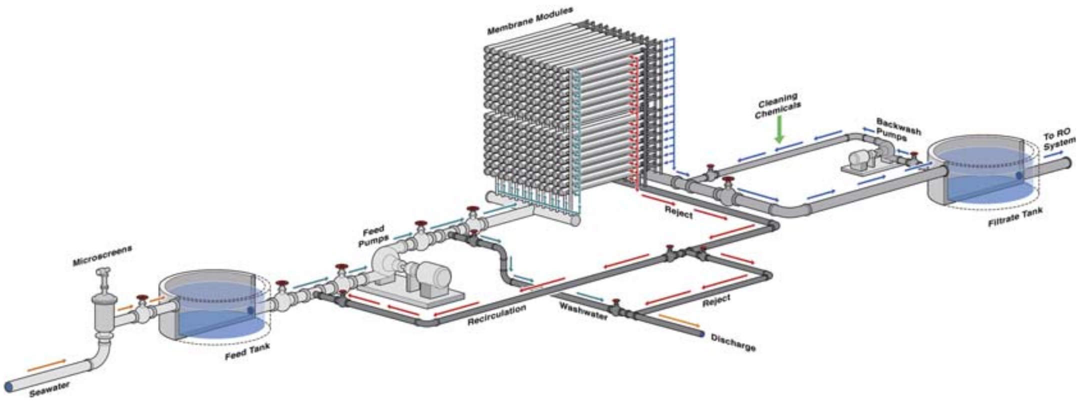

FIGURE 9.4 Schematic of pressurized membrane pretreatment system with horizontal elements.

A general schematic indicating the key components of a pressurized system is presented on Fig. 9.4.

As shown in Fig. 9.4, the source water conveyed by the intake pumps passes through a microscreen into a wet well from where this water is pumped into the UF system. The filtered water is collected from the system and directed into a storage tank, from where it is pumped into the RO system.

9.4.2 Submerged Membrane Systems

Submerged filtration UF and MF systems consist of membrane modules installed in open tanks (Fig. 9.5).

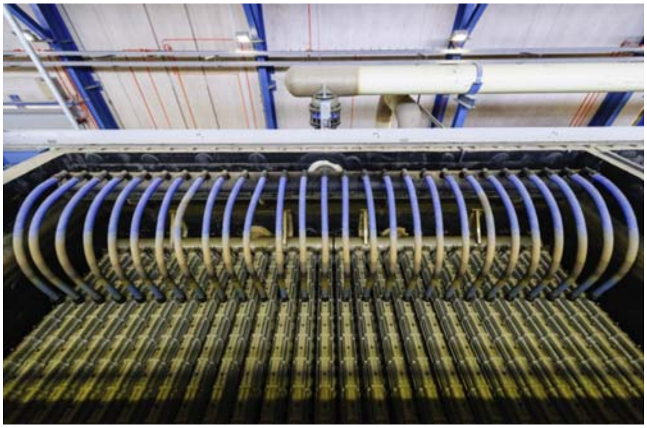

FIGURE 9.5 Submerged membrane filtration (MF) system.

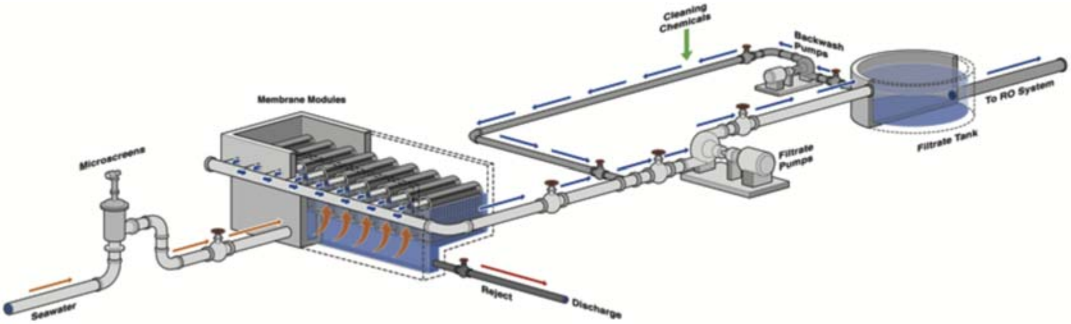

FIGURE 9.6 General schematic of submerged membrane pretreatment system.

A general schematic indicating the key components of submerged pretreatment system is presented in Fig. 9.6.

All submerged systems are outside–in systems where the filtered water is conveyed into the fiber lumens by the vacuum applied on the lumens. The membrane modules are typically installed in concrete or metal tanks and designed so that they can be removed relatively easily for inspection. Each of the tanks can be operated individually and taken out of service for cleaning, inspection, and maintenance. Usually, the tanks are open to the air and can be installed under a light shed for the direct protection of the membranes and equipment from sunlight. In some existing plants, the submerged source water pretreatment system is installed in a building.

9.4.3 Comparison of Pressurized and Submerged Systems

As mentioned previously, membrane systems can be divided in two main groups depending on the type of membrane elements they apply: pressure-driven (pressurized) and submerged. Pressurized membrane systems use membrane elements installed in pressure vessels or housings and the membrane separation process in these systems is driven by of 0.2–2.5 bar of pressure. Submerged systems use membrane modules/cassettes, which are immersed in tanks and operate under a slight negative pressure (vacuum) of typically 0.2–0.8 bar. The following issues are recommended to be considered when choosing between submerged and pressurized type of membrane pretreatment system.

9.4.3.1 Handling Source Water Quality Variations

Submerged membrane systems are usually more advantageous when treating saline source water of variable quality in terms of turbidity, such as intake surface waters that experience frequent turbidity fluctuations of 20 NTU or more. Because these membrane systems are located in tanks (vessels) with relatively large retention volume, they can equalize the source water solids load in the tanks and thereby can reduce the impact of water quality fluctuations on pretreatment system performance. Because shallow open intakes often yield source water with wide turbidity fluctuations, submerged pretreatment systems are usually more suitable for such applications.

Pressurized membrane systems have limited capacity to retain solids because the individual membrane elements are located in a tight membrane vessel of a very small solid retention volume. Therefore, if a pressurized system is exposed to a large amount of solids, the membrane elements and vessels would fill up with solids very quickly, which in turn would trigger frequent membrane backwash and result in destabilization of the membrane system performance. If a pressurized membrane system is overloaded in terms of solids, this system will have to be derated to its capacity to hold solids. Otherwise, such systems will experience very frequent backwashes and ultimately interruption of normal operations.

To address this deficiency of the pressurized membrane systems, some membrane manufacturers offer membrane modules with adjustable fiber density, which allows customizing membrane system design to the more challenging water quality. Typically, these customized membrane elements have fewer fibers and more empty space within the membrane element and thereby provide more volume to retain higher influent solids loads. However, this customization is usually at the expense of the installation of more membrane elements and overall enlarges the size and costs of the membrane system.

Typically, the tanks in which submerged membrane elements are installed provide a minimum hydraulic retention time of 10–15 min and have an order of magnitude higher volume and capacity available to handle saline source water of elevated turbidity and to temporary store solids. This renders submerged membrane pretreatment systems more suitable for high-turbidity water applications. The aeration scouring, which submerged systems typically apply for backwashing, also improves their tolerance to high solid loads. In addition, because submerged systems usually operate at lower TMP, their rate of membrane fouling is lower and they have more stable operation during transient solid load conditions.

Pressurized membrane pretreatment systems, however, are often more suitable for cold saline source waters (i.e., for saline waters of minimum monthly average temperature of 15°C or less). Productivity of submerged systems is more sensitive to source water temperature/viscosity. The maximum trans-membrane operational pressure available for submerged membrane systems is limited to one atmosphere of vacuum, although in practical terms such systems operate at lower maximum TMP (0.7–0.8 bar).

Usually, pressure-driven membrane systems are very suitable and cost competitive for source waters collected by deep open intakes and subsurface wells because such waters have relatively limited turbidity fluctuation span. Because pressure-driven systems can operate at higher fluxes for the same TMP, these pretreatment systems are designed more aggressively and are more cost competitive in such applications.

9.4.3.2 Pretreatment System Footprint

If the source water quality is of high-solids content, which limits the design flux of the pressure systems, submerged membrane pretreatment systems are usually more space efficient than pressurized systems because they allow the installation of more membrane surface area per unit facility footprint area. Significant space reduction is achieved by the fact that the submerged membrane elements do not need to be installed in individual membrane vessels. In addition, the submerged systems typically have only one pipeline system for permeate collection. The distribution of the feed water and the collection and evacuation of the spent filter backwash water are completed at the tank level, which allows the system design to be simplified and the number of valves, piping, and auxiliary service facilities to be reduced.

The smaller footprint of submerged membrane pretreatment systems renders them more beneficial for large-size desalination plants. Typically, submerged membrane systems occupy 10%–20% less space than pressurized membrane installations for pretreatment systems with the same design flux. However, for colder source water of very high quality, which is typical for desalination plants with deep open intakes or subsurface intakes, because the pressurized systems can be designed at 20%–30% higher fluxes, pressure systems would yield pretreatment installations of smaller overall site footprint.

Submerged pretreatment systems have a clear advantage for retrofitting existing granular media filters into membrane pretreatment filters. Submerged membrane tank modules could be installed in the existing granular media filter cells, filter backwash tanks, disinfection contact tanks, or other existing structures, thereby reducing the construction costs for the retrofit. Typically, a conservatively designed granular media filter structure could house submerged membrane filtration system of 1.5–2 times higher production capacity than that of the original filtration system applying only moderate structural modifications.

9.4.3.3 Equipment and Construction Costs and Energy Requirements

Depending on the size of the system and the intake of saline water quality, the site-specific conditions may favor the use of either pressurized or submerged type membrane pretreatment system. Pressurized systems are typically very cost competitive for small and medium size installations because they can be manufactured and assembled in a factory off-site and shipped as packaged installations without the need of significant site preparation or construction of separate structures.

As pressure-driven pretreatment technology evolves, the construction costs of these membrane systems are reduced, larger individual modules are becoming available on the market and most-recent projects indicate that pressure-driven membrane systems are becoming very cost-competitive for all plant sizes and currently they are the most commonly used type of membrane pretreatment systems in seawater desalination plants (Busch et al., 2009).

Equipment and construction costs of larger-size plants with more challenging saline source water quality may be lower when using submerged membrane systems, especially for plant retrofits. An exception to this rule-of-thumb is the treatment of low-temperature source waters.

Because for the same water quality, submerged systems of the same type (MF or UF) usually operate at lower pressure, the total power use for these systems is slightly lower. Typically, submerged systems may use 10%–30% less energy than pressurized systems for water sources of medium to high turbidity and temperature between 18 and 35°C.

9.4.3.4 Commoditization Potential

Currently, both submerged and pressurized membrane systems differ by the type and size of the individual membrane elements; the configuration of the membrane modules; the type of membrane element backwash; and the type of membrane integrity testing method. However, the submerged systems are easier to standardize due to their simplified configuration.

The lack of membrane system unification, standardization, and commoditization results in a significant dependence of the desalination plant owner on the membrane manufacturer supplying their system to continue to provide membrane elements for the system and to improve their existing technology to stay competitive and match the performance of other membrane manufacturers in the future. As a result, the owner of the desalination plant takes the risk of the membrane technology they use at the time of plant construction to become obsolete and out-of-date in the near future due to the accelerated dynamics in the development of new membrane technologies and products or due to the original system manufacturer exiting the membrane market.

The inherent risks associated with the incompatibility of the membrane technologies available today can be partially mitigated by selecting membrane pretreatment system that can be designed to accommodate the replacement of this system with at least one other existing system/membrane elements of similar type. From this point of view, conservatively designed submerged membrane systems offer a better opportunity to handle future changes.

Currently, the submerged systems available on the market have many more similarities than differences compared to pressurized systems. Typically, all existing submerged systems use similar size and depth tanks, which house their membrane modules/cassettes, and have comparable type of membrane CIP and backwash systems. The submerged systems could be designed around the use of a particular membrane technology, but because of their similarities their tanks and auxiliary facilities could be sized to accommodate the replacement of the initially selected membrane system with an alternative membrane system from other manufacturer, if needed in the future. For comparison, pressurized systems are more difficult to commoditize because of the major differences in terms of size, diameter, type of pressure vessels and modules, and type and size of backwashing system.

Despite the lower technical difficulties to commoditize submerged membrane systems, in recent years, the membrane manufacturing industry has embarked on creation of commoditized products for pressure-driven membrane technologies because of the overall higher productivity and competitiveness of pressure-driven pretreatment systems. It is most likely that the future commoditized MF/UF membrane systems for saline water pretreatment would employ outside-in, PVDF, and air–water backwashed membrane modules.

In a parallel track, several companies including Suez, Veolia, H_2O Innovation, and Wigen Water have developed and offer commercialized MF/UF systems, which can accommodate membrane modules from several different manufacturers such as PALL, Dow, GE Zenon, Toray, and Hydranautics. Such membrane pretreatment systems are referred to as "Universal MF/UF Systems" because they allow the owner of the desalination plant/membrane pretreatment system to use interchangeably and to procure competitively membranes from a number of well-established membrane manufacturers with comparable products. It should be pointed out that the capital costs for the universal membrane systems are usually 20%–30% higher than these for a membrane system tailored to a specific membrane product. However, the additional capital costs are likely to be compensated by operation and maintenance savings from membrane replacement.

9.5 FILTER PERFORMANCE

9.5.1 Removal of Solids

MF and UF membrane systems have proven to be very effective for turbidity removal as well as the removal of nonsoluble and colloidal organics contained in the source water. Turbidity can be lowered consistently below 0.1 NTU and filter effluent SDI_{15} levels are usually below 3 over 90% of the time. The suspended solids removal efficiency of membrane pretreatment systems is not sensitive to variations in turbidity and silt content of the saline source water. However, the amount of particulate foulants in the source water may impact the cleaning frequency and length of the filtration cycle.

Typically, membrane pretreatment systems can handle saline source waters with turbidity lower than 50 NTU and SDI_5 lower than 16. If the saline source water turbidity and/or silt content exceed these thresholds, an upstream pretreatment system such as lamella settler or dissolved air flotation clarifier would need to be installed to reduce the feed water turbidity and/or silt below the threshold levels.

9.5.2 Removal of Organics

Membrane pretreatment does not remove a significant amount of dissolved organics and aquatic microorganisms, which typically cause RO membrane biofouling. Because of the very short water retention time of the membrane pretreatment systems, they do not provide measurable biofiltration effect, unless designed as membrane bioreactors. For comparison, granular media filters, depending on their configuration, loading rate, and depth, could remove 20%–40% of the soluble organics contained in the source water. In addition, UF and MF membrane filtration systems have very limited oil and grease removal efficiency and usually do not provide adequate protection of the downstream RO membranes from the destructive impact of these contaminants.

9.5.3 Removal of Microorganisms

9.5.3.1 Algae

UF and MF membranes can remove practically all algae from the saline source water. Their operation will typically not be affected by mild or moderate algal blooms, when the content of algae in the saline source water is less than 20,000 cells/L. Depending on the membrane type, product, and operating TMP, higher algal content could cause varying degrees of reduction of membrane productivity and could result in shorter filtration cycles and higher CEB and CIP membrane cleaning frequencies.

As indicated previously, the operation of the membrane pretreatment systems causes breakage of the cells of the algae contained in the saline source water that results in the release of easily biodegradable organics in the filtered water, which may cause accelerated biofouling of the downstream RO membranes if the organic content exceeds certain threshold (e.g., total organic carbon (TOC) > 2 mg/L).

9.5.3.2 Bacteria and Viruses

Both MF and UF systems can remove four or more logs of pathogens such as *Giardia* and *Cryptosporidium*. Usually, typical UF membrane elements with a pore size of 0.01−0.02 μm can remove over 4 logs (99.99%) of viruses. MF elements with pore sizes of 0.03 μm or less can achieve 3 logs virus removal. Older generation MF membranes (pore openings of 0.1−0.2 μm) do not provide effective virus removal.

In general, both MF and UF membranes do not remove marine bacteria completely and cannot be considered effective barriers for preventing biofouling of the downstream RO elements.

9.6 PLANNING AND DESIGN CONSIDERATIONS

To date, UF membranes have found wider application for saline water pretreatment than MF membranes. Results from a comparative MF/UF study (Kumar et al., 2006) indicate that "tight" (20 kDa) UF membranes can produce filtrate of lower SWRO membrane fouling potential than 0.1 μm MF membranes. For comparison, these MF membranes produce filter effluent of water quality similar to that of 100 kDa UF membranes.

Under conditions, where large amounts of silt particles of size comparable to the MF membrane pore are brought into suspension by naval ship traffic or ocean bottom dredging near the area of the intake, the silt particles may lodge into the MF membrane pores during the filtration process and ultimately may cause irreversible MF membrane fouling. Because UF membrane pores are smaller and fiber membrane structure differs from that of MF membranes, typically UF membrane pretreatment systems do not face this problem. Potential problem of this nature can usually be identified by side-by-side pilot testing of MF and UF membrane systems during periods of elevated silt content.

The two most important parameters associated with the design of any membrane pretreatment system are the design flux and feed water recovery. Membrane flux determines the amount of total membrane area and modules/elements needed to produce a certain volume of saline water. Feed water recovery indicates that the fraction of the saline source water that is converted into filtrate suitable for saline water desalination. A number of factors can impact the selection of these two parameters and ultimately can influence the size and configuration of the membrane pretreatment system. These factors are discussed below.

9.6.1 Source Water Turbidity

Saline source water quality has a significant impact on the configuration and design of the membrane pretreatment system. Saline water of higher turbidity will result in the need for system design around lower membrane flux and will usually yield lower overall recovery due to higher backwashing frequency requirements. Because the decrease of design membrane flux results in a proportional increase in the total needed membrane surface area (i.e., requires additional membrane modules and equipment) and lowering system recovery means that additional saline water would need to be collected to produce the same volume of filtered water, depending on the solids content of the source saline water, it may be more economical to remove a portion of the solids before membrane pretreatment by employing upstream sedimentation, dissolved air flotation or granular media filtration.

Typically, saline source water of annual average turbidity lower than 20 NTU, maximum turbidity of 50 NTU or less and SDI_5 lower than 16 could be treated economically without upstream solids removal. If the saline source water turbidity and/or silt content are above these thresholds or if the saline water intake experiences frequent and extensive algal blooms and/or turbidity spikes, then the saline water solids removal by DAF, sedimentation, or coarse media filtration may be warranted and economical.

The most prudent approach to determine the effect of high turbidity and silt on the design of the membrane pretreatment system is to complete pilot testing during times of the year when the source water has the highest level of solids (e.g., conditions of algal blooms, frequent rain events, heavy ship traffic, or winds and/or currents carrying high amount of sand, silt or seaweeds in the plant intake). If rain- and algal bloom-related saline source water turbidity spikes occur during different periods of the year, pilot testing should encompass both of these periods.

9.6.2 Organic Content of the Source Water

Similar to turbidity, high organic content of the saline source water would result in the need to design and operate the MF/UF system at lower design flux and sometimes lower recovery. Depending on the nature of the organic compounds in the saline water, pretreatment membranes can experience biofouling similar to that of SWRO membranes. If the source water organics consist mainly of natural organic matter that could be coagulated easily, such as humic acids, then coagulation, and flocculation upstream of the membrane filtration may result in a significant improvement of membrane flux and performance.

It should be underlined that MF and UF membranes are not very effective in removing dissolved organics and marine bacteria associated with algal bloom events, even if the saline source water is conditioned with coagulant. During algal blooms, usually over 80% of the organics contained in the source water are dissolved and cannot be removed by coagulant addition. Therefore, if the saline source water is exposed to frequent and intensive algal bloom events when the source water TOC is higher than 2 mg/L for prolonged periods of time (i.e., week or more), membrane pretreatment system would need to be designed for conservatively low flux (<65 lmh for pressure-driven MF/UF systems and <40 lmh for vacuum-driven systems). Under such conditions, MF/UF system recovery should be reduced to 90% compared to typical recoveries of 93%–95%.

TABLE 9.3 Temperature Correction Factor

Temperature (°C)	Flux Correction (% Increase/Decrease)
5	55
10	30
15	15
20	0
25	−10

The most prudent approach to determine the effect of algal bloom related biofouling on the design of the membrane pretreatment system is to complete pilot testing during the time of the year, when algal blooms are most likely to occur and are at their highest intensity.

9.6.3 Temperature

Saline water viscosity increases with the decrease of temperature. Viscosity affects the membrane ability to produce filtered water as more pressure (or vacuum for submerged systems) is required to overcome resistance associated with flow across the membrane surface area and through the membrane pores when operating at a constant flux (i.e., producing the same filtered flow). Typically, average design membrane flux is established for average annual temperature, flow, and turbidity and is then adjusted for the minimum monthly average temperature using the correction factor shown in Table 9.3.

For example, if the flux determined for an average annual conditions is 80 lmh, and the average annual temperature is 20°C, but the minimum monthly average temperature is 15°C, then the design flux should be reduced by approximately 15% (i.e., down to 68 lmh) for the membrane pretreatment system to be able to produce the same filtrate flow during all months of the year at approximately the same recovery and power demand.

The correction factor presented in Table 9.3 is a "rule of thumb" based on practical experience and it may vary from one membrane product to another. Most membrane manufacturers have recommended capacity compensation factors for their system and should be consulted when such a factor is selected for the site-specific conditions of a given project. The most prudent approach to determine the effect of temperature/viscosity increase on the design of the membrane pretreatment system is to complete pilot testing during the coldest month of the year.

9.6.4 Experience With Existing Installations

At present, less than 10% of the desalination plants worldwide have membrane pretreatment and this type of pretreatment had modest success in terms of performance benefits and cost efficiency. Although the use of UF and MF membranes for saline water pretreatment to date is fairly limited, over 20 years of such full-scale experience exists for fresh water filtration applications. An operations survey completed at 10 fresh drinking water treatment plants in the United States (Atassi et. al, 2007), which employ MF or UF membrane filtration systems, has identified a number of challenges these plants have experienced during their startup, acceptance testing, and full-scale operations (Table 9.4). Although the MF and UF systems were used for fresh water treatment, most of the lessons learned from these applications are relevant to saline water applications as well.

Overview of the information presented in Table 9.4 points out that for the use of MF/UF systems for pretreatment to be successful, such systems have to be designed carefully and conservatively and have to be operated by highly trained staff that often is not readily available in many desalination plants worldwide. Because MF/UF membranes are uncommoditized proprietary products, the plant owner and operators will have to rely on the membrane manufacturer and supplier to provide expert guidance and technical support, which typically are not necessary if conventional granular media filtration is used for pretreatment.

9.7 OVERVIEW OF MEMBRANE PRODUCTS USED FOR SALINE WATER PRETREATMENT

9.7.1 X-Flow (Pentair)

X-Flow (Pentair), formerly known as Norit, has two UF membrane products, which are specifically developed for membrane pretreatment of saline water, Seaguard and Seaflex (see Figs. 9.7 and 9.8). Both products are available in three types of modules, which differ by the size of their filtration area per module—64, 55, and 40 m^2, respectively. To date, the X-Flow UF membrane product that has found the widest full-scale application for seawater pretreatment is the Seaguard 40 module.

The main difference between the Seaguard and Seaflex modules is their maximum TMP and the configuration of the racks in which they are installed. Seaguard UF membranes can withstand higher maximum TMP (8 bars vs. 3 bars for Seaflex modules), which makes them more suitable for desalination plants with UF/RO direct coupling configuration, where the saline water from the intake is pumped directly through the UF membrane system and into the suction header of the high-pressure RO feed pumps. This type of UF-RO system configuration allows to eliminate the need for an interim tank and pump station between the pretreatment and RO systems and results in capital cost savings. However, its successful use also requires the pretreatment membranes to be able to withstand higher working pressures because they have to carry through not only the pressure needed for normal operation of the UF membranes (0.1−0.8 bar) but also the suction pressure needed for the high-pressure pumps of the RO system (2.0−3.5 bars). In addition, these membranes have to be able to withstand elevated pressures caused by hydraulic surges.

Seaguard UF membrane system racks have horizontal configuration (similar to that of RO racks) with two to four membranes per module (vessel) and up to 120 modules per rack. For comparison, Seaflex UF system racks have vertical configuration with only one membrane module and can accommodate a maximum of 18 modules per rack—these racks are also referred to as X-line Racks. The main reason of the difference of the two configurations is that the horizontal configuration is less sensitive to hydraulic surges/pressure spikes and

TABLE 9.4 Ultrafiltration/Membrane Filtration (UF/MF) System Survey: Lessons Learned (Atassi et al., 2007)

Utility No.	Primary Membrane System Problems	Root Cause of Problem	Lessons Learned
1.	• Unable to meet design capacity	• Lower achievable flux than projected. • Pilot testing did not address extreme WQ conditions.	• Pilot test during extreme water quality (WQ) conditions. • Use conservative safety factor when up-scaling pilot testing results.
2.	• High clean-in-place frequencies; • Excessive downtime and operation and maintenance (O&M) costs.	• Excessive membrane fouling. • Pilot testing did not address extreme WQ conditions.	• Pilot test during extreme WQ conditions. • Additional pretreatment may be needed to address extreme WQ conditions.
3.	• Unable to meet design capacity.	• Undersized membrane ancillary support systems.	• Ancillary support systems can be a significant bottleneck if undersized.
4.	• Higher membrane replacement costs.	• Lower than projected membrane life. • Potential membrane fouling and lack of previous data by suppliers.	• Additional pretreatment may be needed to obtain the useful membrane life indicated by membrane supplier.
5.	• Excessive downtime and maintenance. • Lower than projected WQ.	• Excessive fiber breakage. • Fouling or WQ putting higher stress on the fibers than expected.	• Lack of experience with use of membranes for given WQ may require change in membrane chemistry and durability
6.	• Unable to meet design capacity.	• Higher than anticipated downtime. • Manufacturer missed to include valve opening/closing time for integrity tests. • More membrane capacity needed to be installed.	• Complete thorough review of the downtime for all MF/UF system operational steps under worst case operations scenario.
7.	• Higher than expected O&M costs.	• More frequent chemical cleaning needed than initially projected.	• Pilot test during extreme WQ conditions. • Use conservative safety factor when up-scaling pilot testing results.
8.	• Excessive system downtime.	• Failures in membrane potting. • System not handling water pressure/potting materials not tested previously.	• Never use membrane that has components or materials that have never been tested previously!
9.	• Difficult system operation.	• Insufficient system training for staff.	• Plan for additional staff training beyond the minimum offered by the manufacturer.
10.	• Excessive downtime. • Failure to meet product WQ targets.	• Frequently failing membrane integrity testing. • Air leaking form gaskets and valves.	• Make sure that replacement of failed caskets, valves, and seals is included in the manufacturer membrane system warrantee.

FIGURE 9.7 Pentair X-flow Seaguard ultrafiltration pretreatment system.

FIGURE 9.8 Pentair X-Flow Seaflex ultrafiltration pretreatment system.

is, therefore, more advantageous for directly coupled UF-RO systems than the vertical configuration of the UF racks. However, the vertical configuration offers more compact design and smaller footprint, which ultimately results in lower overall capital costs. Therefore, this system is usually more attractive for large desalination plants with an interim tank and pump-station between the UF and RO systems.

All modules within the Seaflex X-line racks are fed from a common source water line. The Seaflex modules are directly attached to the feed lines and are not contained in a separate vessel compared to the Seaguard membrane modules, which are installed in horizontal housings (vessels) and their filtrate tubes are connected with plastic interconnectors and O-rings similar to these used for interconnecting the RO membranes in their pressure vessels.

Both the Seaguard and Seaflex membrane pretreatment systems employ UF PES membranes with inside-out flow pattern and pore size of 0.020–0.025 μm. The horizontal Seaguard systems have some similarities with RO systems in terms of vessel diameter and general configuration. However, these systems have only four membrane elements (instead of seven or eight

TABLE 9.5 Pentair X-Flow Pretreatment Systems

Parameter	Seaguard 64 and 55	Seaguard 40	Seaflex 64 and 55	Seaflex 40
Module configuration	Horizontal	Horizontal	Vertical	Vertical
Number of modules per pressure vessel (housing)	2–4	2–4	1 (no separate housing)	1 (no separate housing)
Module length	1.54 m (60-in.)	1.54 m (60-in.)	1.54 m (60-in.)	1.54 m (60-in.)
Module diameter	220 mm (8.7-in.)	200 mm (8.0-in.)	220 mm (8.7-in.)	200 mm (8.0-in.)
Membrane area	64 & 55 m^2 (689 and 592 ft^2)	40 m^2 (430 ft^2)	64 and 55 m^2 (689 and 592 ft^2)	40 m^2 (430 ft^2)
Fiber inner diameter, mm	0.8	0.8	0.8	0.8
Fiber outer diameter, mm	1.3	1.3	1.3	1.3
Maximum feed pressure	8 bars (116 psi)	8 bars (116 psi)	3 bars (43 psi)	3 bars (43 psi)
pH operating range	2–12	2–12	2–12	2–12
Weight per module (dry)	34 kg (75 lbs)	25 kg (55 lbs)	34 kg (75 lbs)	25 kg (55 lbs)
Suitable for inline configuration	Yes	Yes	No	No

used in RO vessels) and the UF vessels are fed from both ends with a dead end at the center of the vessel. This configuration is driven by the fact that because of the high fluxes at which these systems operate, flow distribution within the vessels is even worse than that in RO systems and the fourth element in sequential UF configuration will be used very inefficiently (i.e., will have very low productivity). To address the uneven hydraulic distribution within the vessels (housings) X-Flow/Pentair have incorporated bypass tubes within the vessels.

The horizontal (Seaguard) design allows for a more efficient use of the height of existing buildings because a number of vessels can be added vertically at the same basic footprint. This system also has fewer valves and overall shorter interconnecting piping.

The vertical (Seaflex) configuration has a better flow distribution within its single vessel, which allows the system to be designed at 10%–12% higher flux at comparable membrane surface area. The vertical membrane configuration also allows an easier and better membrane cleaning and flushing. Table 9.5 provides a summary of the key design parameters for these systems.

9.7.2 Memcor (Evoqua)

Memcor (Evoqua) have both vacuum- and pressure-driven membrane systems, which are divided into two lines of products, "Xpress" or "X" for small package systems and "Component", "C", engineered systems designed to serve plants with a production capacity of over 4000 m^3/day (1.1 MGD). Both the "X" and the "C" modules are available in pressure driven (XP and CP) and submerged configurations (XS and CS). All membranes are made of PVDF

TABLE 9.6 Memcor (Evoqua) Membrane Pretreatment Systems

Parameter	Submerged CS-10V	Pressurized CP-L10 V	Pressurized CP-L20 V
Membrane elements (modules)	S10 V	L10 V	L20 V
Module configuration	Vertical	Vertical	Vertical
Number of modules per unit (maximum)	900	960	960
Module length	1.19 m (47-in.)	1.19 m (47-in.)	1.80 m (71-in.)
Module diameter	130 mm (5.2-in.)	150 mm (5.9-in.)	120 mm (4.7-in.)
Membrane area	27.9 m^2(300 ft^2)	55 m^2(252 ft^2)	38.1 m^2(410 ft^2)
Fiber inner diameter, mm	0.5	0.25	0.25
Fiber outer diameter, mm	0.8	0.5	0.5
Maximum trans-membrane pressure	1.2 bar (17 psi)	1.5 bar (21 psi)	1.5 bar (21 psi)
pH operating range	1.0–10.0	1.0–13.0	1.0–13.0
Weight per module	6.5 kg (14 lbs)	6.5 kg (14 lbs)	9 kg (20 lbs)
Suitable for pump-through configuration	No	Yes	Yes

and their fibers have pore sizes of 0.04 μm. Memcor membranes are classified as UF type and their pore size is adequately small to remove up to 3 logs (99.9%) of viruses. Table 9.6 provides a summary of the key performance parameters of the Memcor membrane elements that have found application for seawater pretreatment. Besides the UF membranes listed in Table 9.6, Memcor offers a number of other UF elements.

Both submerged and pressurized Memcor (Evoqua) units are designed for combination of air and water backwash. Submerged systems have slightly higher backwash volume (reject flow), of 7%–8% compared to pressure-driven systems, which generate backwash volume of 4%–5% of the total intake flow.

The Southern Seawater Desalination Plant in Perth, Australia, with total fresh water production capacity of 300,000 m^3/day (80 MGD) uses pressure-driven Memcor CP 960 membrane racks (see third column of Table 9.6). These racks have 912 modules each. The desalination plant consists of two 150,000 m^3/day (40 MGD) plants, which share common intake and outfall. Each plant has two sets of four duty and one standby racks (e.g., a total of 20 CP 960 racks for the entire plant). The membranes are designed for a flux of 65 lmh (38 gfd) and produce filtered water of SDI_{15} lower than 3 for over 95% of the time.

The 300,000 m^3/day (80 MGD) Adelaide desalination plant in Australia uses submerged Memcor CS membrane pretreatment system and consists of two 150,000 m^3/day (40 MGD) plants with common intake and outfall. The pretreatment system for each of the two 150,000 m^3/day plants has 14 Memcor CS 900 units (see first column of Table 9.6) deigned at a flux of 52 lmh (30 gfd) and does use coagulants.

FIGURE 9.9 HYDRAcap 40, 60, and Max 80 ultrafiltration (UF) membrane modules.

9.7.3 Hydranautics (Nitto-Denko)

This membrane manufacturer offers pretreatment membranes which are low-pressure UF modules (see Fig. 9.9) that have found application at several large seawater desalination plants such as the 90,000 m^3/day (24 MGD) Kindasa SWRO desalination plant in Saudi Arabia and the 96,000 m^3/day (25 MGD) Fukuoka desalination plant in Japan. HYDRAcap 60 is the Hydranautics UF membrane module that has been used most widely to date.

In 2011, Hydranautics has introduced a new UF membrane of over three times higher surface area (105 m^2 vs. 30 m^2) and productivity than the baseline HYDRAcap 60 model—the HYDRAcap MAX 80. This membrane is made of more durable material, PVDF, and can handle water of very high turbidity (up to 80 NTU). The new membrane module has a pore size of 0.08 μm and outside-in flow pattern. For comparison, the HYDRAcap 60 is inside-out type of UF membrane.

The key membrane performance parameters for the baseline HYDRAcap 60 membrane modules are presented below:

Membrane Type	**Ultrafiltration, Pressure Driven, Inside-Out**
Two widely used models	Hydracap 60/60-LD
Typical operating pressure	0.2–0.5 bar
Maximum feed pressure	3.0 bars
Trans-membrane pressure that triggers backwash	(Typically 1.1–1.4 bar)
Filtration cycle length	15–60 min
Backwash duration	30–60 s
Membrane material	Hydrophilic polyethersulfone (PES)
Nominal pore size	0.08
Module diameter	225 mm (8.8-in.)
Module length	1000 and 1500 mm (40-in. and 60-in.)
Membrane module filtration area	30 m^2 (1000-mm module) and 46 m^2 (1500-mm module)
Design flux	60–100 L/m^2 h
Number of modules needed to produce 1000 m^3/day of filtrate	10–16 HYDRAcap 60 1500-mm modules

9.7.4 General Electric Water and Process Technologies

General Electric Water and Process Technologies (now part of Suez) offers both submerged and pressurized ZeeWeed UF pretreatment systems, with hollow-fiber membrane modules, which is operated in an outside-in-mode. The UF fibers of the submerged modules are combined into bundles, which are installed in standard-size modules (see Fig. 9.10). In the case of the pressurized UF membranes (ZW 1500), the membrane fibers are installed in vertical polyvinyl chloride housings. To date, the main model that has found full-scale application for pretreatment of seawater is the submerged ZW1000 PVDF membrane and its modified versions (see Table 9.7). Up to 96 membrane modules can be installed in a membrane cassette, and these cassettes are immersed into tanks fed with source water. Air scouring system is usually installed in the tanks to loosen and release the solids retained within the hollow fiber bundles during the backwash cycle of system operation.

The main difference between the versions of the original ZW1000 is the total membrane cassette surface area and reduced internal fiber diameter. The V2 version has a smaller surface area compared to the other versions because it is designed to handle higher solid loads

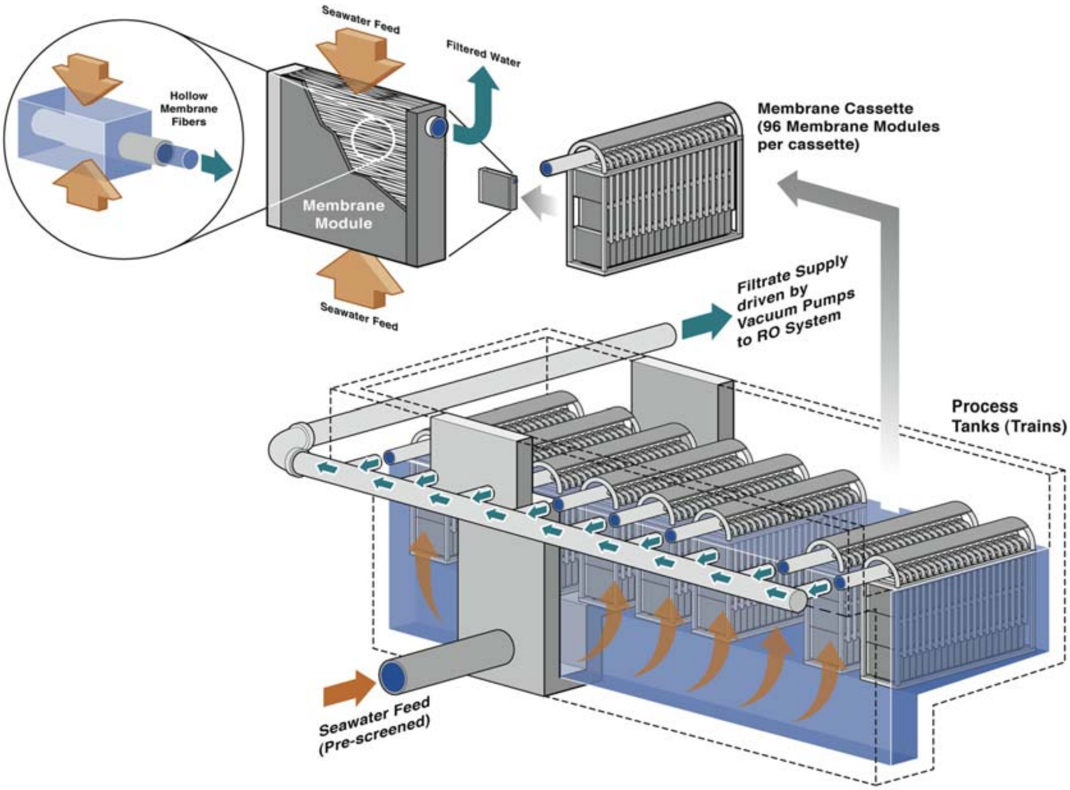

FIGURE 9.10 GE ZeeWeed vacuum-driven ultrafiltration (UF) membrane system.

by having relatively smaller fiber packing density—the product design is tailored for wastewater applications. The V3 version of ZW1000 has higher inner diameter fibers than the V2 version (0.5 mm vs. 0.4 mm), which allows to reduce TMP and to increase membrane productivity. These performance benefits make the ZW1000-V3 membranes likely the most suitable product for waters of lower solids content such as seawater applications with relatively deeper open intakes.

GE Zenon also offers pressurized, outside-in membrane product referred to as ZW1500. This module has a significantly higher fiber length than the ZW1000 elements (1.7 vs. 0.6 m) and is designed to operate at a maximum TMP of 2.75 bars. This pressure module has a diameter of 180 mm (7 ft) and a total membrane area of 55.7 m^2 (600 ft^2)· Table 9.7 presents the key performance parameters of the ZW1000, V2, and 3 and ZW1500 elements, which are typically offered for desalination plant pretreatment applications.

9.7.5 Other UF and MF Membrane Products

Other commercially available membrane products used for pilot and full-scale desalination projects include the following: Hyflux and Inge pressure-driven PES UF membranes;

TABLE 9.7 GE ZW Seawater Pretreatment Membranes

Parameter/Membrane Module	ZW1000-V2	ZW1000-V3	ZW1500
Module configuration	Horizontal installed in cassettes	Horizontal installed in cassettes	Vertical installed in polyvinyl chloride housing
Number of modules per cassette/rack (maximum)	96	96	192
Cassette width/module diameter	2.11 m (6.9 ft)	1.82 m (6.0-ft)	18 cm (7 in.)
Cassette height/module length	2.54 m (8.3 ft)	2.60 m (8.5 ft)	1.75 m (69 in.)
Membrane area	37.2 m^2(400 sq ft)	46.5 and 55.2 m^2 (500 and 594 sq ft)	55.7 m^2 (600 sq ft)
Fiber inner diameter, mm	0.4	0.5	0.47
Fiber outer diameter, mm	0.7	0.8	0.95
Maximum feed pressure	0.69 bar (9.7 psi)	0.69 bar (9.7 psi)	3.8 bars (55 psi)
pH operating range	2.0–10.0	2.0–10.0	2.0–10.0
Weight per cassette/module	1007 kg (2200 lbs)	1007 kg (2200 lbs)	22 kg (50 lbs)
Suitable for pump-through configuration	No	No	No

TABLE 9.8 Toray, Pall, and Dow Seawater Pretreatment Systems

Parameter/Manufacturer	Toray	Pall	Dow
Model of membrane element (module)	HFU–2020N	UNA620A	SPF-2860
Module configuration	Vertical	Vertical	Vertical
Nominal pore size	0.01 μm	0.10 μm	0.03 μm
Number of modules rack (maximum)	75	75	75
Module diameter	22.0 cm (8.5 in.)	16.5 cm (6.5 in.)	22.6 cm (8.9 in.)
Module length	1.78 m (70 in.)	2.22 m (87.2 in.)	1.82 m (71.7 in.)
Membrane area	71.1 m^2 (775 ft^2)	49.4 m^2 (538 ft^2)	50.4 m^2 (549 ft^2)
Maximum feed pressure	5.0 bars (72 psi)	5.0 bars (72 psi)	6.5 bars (94 psi)
pH operating range	2.0–13.0	2.0–12.0	2.0–12.0
Suitable for pump-through configuration	Yes	Yes	Yes

Toray pressure-driven modified PVDF UF and MF membranes; Dow pressure-driven UF PVDF membranes; Koch pressure-driven PS UF and Pall/Asahi Aria pressure-driven MF membranes. Key performance parameters of the Toray, Pall, and Dow membrane elements that have found applications for pretreatment of seawater are shown in Table 9.8.

9.8 DESIGN EXAMPLES

9.8.1 Submerged UF Pretreatment System

This design example illustrates the determination of the configuration of vacuum-driven UF membrane pretreatment system for 500,000 m^3/day (13.2 MGD) seawater desalination plant designed for 43% RO system recovery and 5% reject flow (i.e., 95% UF system recovery). In addition, the membrane pretreatment system has 100 μm disk filters upstream of the UF system. The source water turbidity varies between 0.3 and 10 NTU [total suspended solids (TSS) of 5–15 mg/L] with occasional spikes of up to 15 NTU (TSS = 20 mg/L). Maximum algal count in the source water is 20,000 cells/mL and the hydrocarbon levels are below 0.04 mg/L. The pretreatment system is designed to operate without addition of coagulant, flocculant, and without pH adjustment of the source water flow.

The pretreatment system will need to be designed to treat a total of 97,920 m^3/day [(50,000 m^3/day/43%) × 1.05 = 122,093 m^3/day]. Key design parameters of the system are as follows:

SOURCE WATER	
Total flow	= 122,093 m^3/day (32.3 MGD)
Temperature (average annual/minimum monthly average)	= 20°C/15°C
Turbidity (average annual/daily maximum)	= 1.5 NTU/10 NTU
Algal count (average annual/daily maximum)	= 200/20,000 cells/mL
TARGET QUALITY OF PRETREATED SALINE WATER	
Turbidity (average/maximum)	= 0.05 NTU/0.3 NTU
SDI_{15}	<3 (95% of the time)/<5 at all times
VACUUM DRIVEN UF MEMBRANE PRETREATMENT SYSTEM—CALCULATION OF KEY DESIGN PARAMETERS	
Membrane module	ZW1000-V3
Average flux (at average annual temperature of 20°C)	= 40 L/m^3 h (based on pilot test)
Temperature correction factor for min. monthly average temp of 15°C	= 1.15
Design flux @ minimum monthly avg. temperature of 15°C	= 45 L/m^3 h/1.15 = 39.1 L/m^3 h
Total membrane area required = [(122,093 m^3/day × 1000)/(39.1 L/m^3 h × 24 h)]	= 130,108 m^2
Number of membrane modules @ 55.2 m^2/module	= 130,108/55.2 = 2357 modules
Number of membrane cassettes @ 48 modules/cassette	= 2357/48 = 49 cassettes

Each tank is designed to house one spare cassette—i.e., the tank structure dimensions are determined for six cassettes (5 + 1). The final UF system configuration includes 10 tanks sized to house six cassettes per tank with five cassettes installed per tank along with connections to sixth cassette. With one tank in backwash and one tank out of service for cleaning, the plant operating flux will be 122,093 m^3/day × 1000 L/(10 Tanks × 5 Cassettes × 48 modules × 55.2 m^2/module × 24 h) = 38.4 lmh (22.5 gfd). This flux is within the acceptable range of 30−45 L/m^2 h (18−26 gfd), determined by pilot testing.

9.8.2 Pressure-Driven UF Pretreatment System

This example illustrates the key design steps and criteria of pressure-driven X-Flow (Pentair) Seaguard UF pretreatment system for the same size desalination plant (50,000 m^3/day/13.2 MGD) and source water quality as described in Section 9.8.1. The pressure-driven UF pretreatment system has the same recovery rate (95%) and the same disk filters upstream of it and, therefore, has the same design feed flow (122,093 m^3/day, 32.3 MGD). Similar to the vacuum-driven system in the example earlier, this pretreatment system is designed to produce filtered water of target quality without the need to condition the source water with chemicals, i.e., no coagulant, flocculant, and acid are added to the feed water.

SOURCE WATER	
Total feed flow	= 122,093 m^3/day (32.4 MGD)
Temperature (average annual/minimum monthly average)	= 20°C/15°C
Turbidity (average annual/daily maximum)	= 1.5 NTU/10 NTU
Algal count (average annual/daily maximum)	= 200/20,000 cells/mL
TARGET QUALITY OF PRETREATED SALINE WATER	
Turbidity (average/maximum)	= 0.05 NTU/0.3 NTU
SDI_{15}	<3 (95% of the time)/<5 at all times
PRESSURE-DRIVEN SEAGUARD UF MEMBRANE PRETREATMENT SYSTEM—CALCULATION OF KEY DESIGN PARAMETERS	
Membrane module (8-in. element)	= SEAGUARD 40
Average flux (at average annual temperature of 20°C)	= 65 lmh (based on pilot test)
Temperature correction factor for min. monthly average temp. of 15°C	= 1.15
Design flux at minimum monthly avg. temperature of 15°C	= 65 L/m^3 h/1.15 = 56.5 L/m^3 h
Total membrane area required	= [(122,093 m^3/day × 1000)/(56.5 L/m^3 h × 24 h)] = 90,039 m^2
Number of membrane modules at 40 m^2/module	= 90,039 m^2/40 m^2 = 2251 modules
Number of membrane vessels at 2 modules/vessel	= 2251/2 = 1125 vessels

As indicated previously, the vessels are configured in racks. The proposed design would have 10 racks with 120 vessels (240 modules) per rack for a total of 1200 vessels/2400

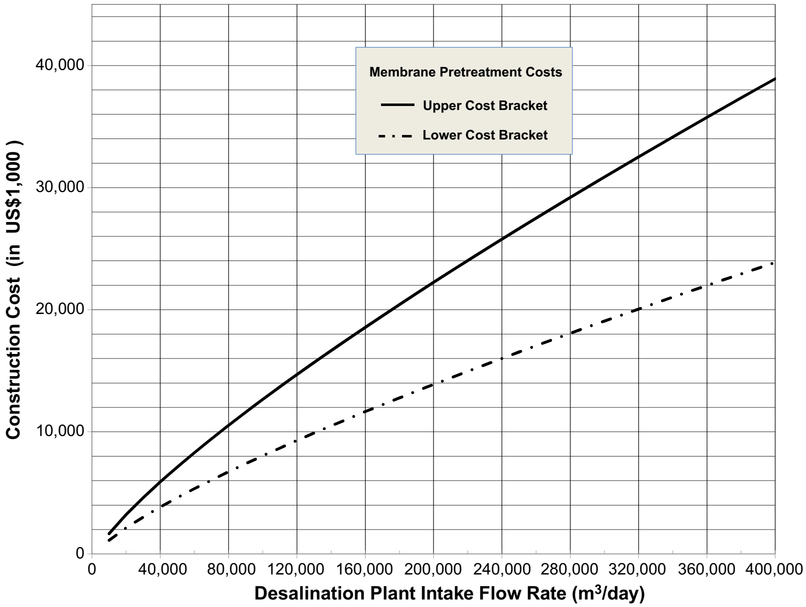

FIGURE 9.11 Construction costs of membrane pretreatment systems.

modules for the entire plant. This configuration incorporates 7% of safety factor (2400/2251 = 7%). With one UF train in backwash and one train out of service for cleaning, the plant operating flux will be 122,093 m^3/day × 1000 L/(8 UF trains × 120 vessels × two modules × 40 m^2/module × 24 h) = 66.2 L/m^3 h (38.8 gfd), which is within the acceptable range of 50–75 L/m^2 h (30–44 gfd) determined by pilot testing.

9.9 CONSTRUCTION COSTS OF MEMBRANE PRETREATMENT SYSTEM

The construction costs for membrane pretreatment are difficult to determine based on existing projects because of the relatively limited track record of this type of pretreatment compared to granular media filtration and also because of the diversity of membrane products and configurations presently available on the market. Therefore, rather than a single cost curve, Fig. 9.11 presents a range of probable year 2017 costs of membrane pretreatment for desalination plants as a function of the plant intake flow rate.

For the example of 50,000 m^3/day desalination plant described in the previous section of this chapter, which has an intake flow of 122,093 m^3/day (32.3 MGD), the construction cost in US$ of the intake pretreatment system in the year 2017 is estimated in a range of US$9.4–14.9 million (average of US$12.1 million) based on cost curves depicted in Fig. 9.11.

References

American Water Works Association (AWWA), 2007. Manual of Water Supply Practices, M46, Reverse Osmosis and Nanofiltration, second ed., pp. 205–206 Appendix C.

Atassi, A.M., White, M., Rago, L.M., 2007. Membrane failure headlines: facts behind the hype regarding initial problems experienced at low pressure membrane installations. In: Proceedings of 2007 AWWA Membrane Technology Conference.

Burashid, K., Hussain, A.R., 2004. Seawater RO plant operation and maintenance experience: addur desalination plant operation assessment. Desalination 165, 11–22.

Busch, M., Chu, R., Rosenberg, S., 2009. Novel trends in dual membrane systems for seawater desalination: minimum primary pretreatment and low environmental impact treatment schemes. In: Proceedings of IDA World Congress, the Palm-Dubai, UAE, November 7-12, 2009 (Ref: IDAWC/DB09-019).

Gasia-Brush, E., Busch, M., Garica-Molina, V., Kolbe, U., 2011. Dual membrane systems in seawater desalination: drivers for selection and field experiences. In: Proceedings of IDA World Congress, Perth, Australia, September 4-9, 2011 (Ref: IDAWC/PER11-244).

Kumar, S.A., Adam, S., Pearce, W., 2006. Investigation of Seawater Reverse Osmosis Fouling and Its Relationship to Pre-treatment Type, Environmental Science and Technology, vol. 40. ACS Publications, pp. 2037–2044.

Lazaredes, H., Broom, G., 2011. Review of the pre-treatment selection/design for Australia's big six desalination projects. In: Proceedings of IDA World Congress, Perth, Australia, September 4-9, 2011 (Ref: IDAWC/PER11-244).

Pearce, G.K., 2011. UF/MF Membrane Water Treatment – Principles and Design, Water Treatment Academy. TechnoBiz Communications Co., Ltd, Bangkok, Thailand.

Shahid, K., Al Sadi, A., May–June, 2015. Algae: A Tide of Trouble, Desalination and Water Reuse. Taylor & Francis, pp. 35–37.

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CHAPTER

10

Comparison of Granular Media and Membrane Pretreatment

OUTLINE

10.1 INTRODUCTION

Membrane filtration technologies have a number of advantages compared to conventional granular media filtration systems. Granular media filtration, however, is a well understood and widely used pretreatment technology with a proven track record, which has a number of features that may render it cost competitive under many circumstances. Therefore, the selection of filtration technology for pretreatment of saline water should be based on a thorough life-cycle cost–benefit analysis.

Side-by-side pilot testing of the two types of pretreatment systems is also highly recommended to develop background system performance information for an objective

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technology evaluation and selection. The following issues have to be taken into consideration when selecting between granular media and membrane pretreatment filtration for a specific application.

10.2 EFFECT OF SOURCE WATER QUALITY ON PERFORMANCE

Membrane filtration has a wider spectrum of particle removal capabilities than conventional media filtration. Because particulate separation process is based on filtration through a membrane with a fairly uniform pore size, particulate removal efficiency is higher and more consistent than the more randomly porous granular media filtration bed. Single or dual-media filters usually have a lower removal efficiency in terms of raw source water organics in suspended form, disinfection byproduct precursors, fine particles, silt, and pathogens.

Membrane filtration technologies are less prone to upsets caused by seasonal changes in source water turbidity, color, pathogen contamination, and size and type of water particles because their primary treatment mechanism is a mechanical particle removal through fine-pore membranes (Pearce, 2011). Therefore, the upstream chemical coagulation and flocculation of the source water particles is of a lesser importance for their consistent and efficient performance.

In contrast, pretreatment performance of granular media filtration systems is very dependent on how efficient chemical coagulation and flocculation of the source water are ahead of the filtration process. Therefore, for applications where intake source water quality experiences significant seasonal variations and presents a challenge in terms of high level of pathogens and elevated concentration of fine particles and particulate organics, membrane filtration technologies are likely to offer performance benefits. However, if the source water for the desalination plant is collected from an open intake located far from the tidally influenced zone and at adequate depth to be exposed to only limited seasonal variations (typically 10 m/33 ft or deeper), granular media filtration may offer a very cost-effective pretreatment alternative to membrane filtration.

Source water temperature is a very important factor when selecting pretreatment system. Application of vacuum-driven membrane pretreatment systems is usually less cost effective than conventional granular media filtration for source water of temperature lower than 15°C (59°C) because the productivity (flux) of vacuum-driven membrane filtration is dramatically reduced by the significant increase in viscosity of source water at low temperature (AWWA, 2007).

Another condition under which the use of sand media filtration may have certain additional benefits is when a source water is exposed to sudden and unpredictable changes of specific contaminants such as very high or low pH, chemical spills, oil and grease spills, frequent exposures to very high source water temperature, or contaminants that may damage the membrane filtration (MF) or ultrafiltration (UF) pretreatment membranes irreversibly, if they are used for this application. If the membrane elements are permanently damaged, the cost of their replacement could be significant, especially for large reverse osmosis (RO) desalination plants.

As indicated previously, source water naturally contains sharp particles, which could damage MF and UF membranes upon contact (Ransome et al., 2015). To remove these sharp source water particles, the RO plant intake system should incorporate a microscreening system of mesh size of 120 μm or less ahead of the membrane pretreatment system (see Chapter 5 for more details). Performance and reliability of conventional granular media pretreatment systems are not sensitive to the content of sharp objects in the source water and do not require elaborate and costly microscreening ahead of the filters. Typically, mechanical traveling screens of 3–10 mm openings provide an adequate protection of conventional granular media pretreatment systems.

A particular challenge for seawater pretreatment systems with open intakes could be the content of barnacle plankton in the source water of size small enough to pass through the microscreening system and to grow to adult barnacles on the walls of the wet well of the pumps feeding the pretreatment system. When some of the barnacles enter the membrane pretreatment feed pumps, the shells are broken by the pumps into small sharp particles, which are then pumped (or drawn by vacuum) against the membrane fibers causing occasional punctures and over time resulting in the loss of membrane integrity and performance. Typically, such challenges can be addressed by matching the size of the microscreens upstream of the membrane pretreatment system with the size of the smallest barnacle (or other shellfish) plankton species that may occur in the saline source water. This issue warrants a detailed year around investigation as the size and type of plankton species contained in the source water typically change seasonally. If such a challenge is likely to occur for a particular project, it may be prudent to consider the installation of cartridge filters downstream of the membrane pretreatment system, which would be able to capture particles and other impurities that may pass through the punctured pretreatment fibers.

Because installation and operation of a microscreening system downstream of source water intake screens is only needed if membrane filtration is used for pretreatment, the cost for microscreening of the source water should be taken under consideration when comparing conventional and membrane filtration pretreatment.

On the other hand, the use of membrane pretreatment would eliminate the need and costs for installation and operation of cartridge filter system ahead of the RO feed pumps. Cartridge filters are needed when granular filtration system is used for pretreatment to protect the downstream RO membranes from damage caused by fine sand particles, which may be conveyed occasionally with the pretreated water. The size of MF and UF membranes is usually several orders of magnitude smaller than that of cartridge filters, and therefore, if the pretreatment membrane integrity is maintained intact, cartridge filtration would not be necessary. However, if the membrane pretreatment system employs low-quality MF or UF membranes, which have limited durability and could easily break during hydraulic pressure surges or get punctured by shells due to inadequate prescreening of the source water, the use of cartridge filters is desirable and recommended.

Occurrence of frequent and prolonged algal blooms in the area of the source water intake is another important factor to consider when selecting the type and configuration of source water pretreatment system. As indicated previously (see Chapter 7), many of the marine microalgae, which grow excessively during algal blooms, cannot withstand external pressure of more than 0.3–0.6 bars (4–8 psi) and their cells could break when exposed to pressure or vacuum-driven MF or UF filtration.

When algal cells break, they release easily biodegradable organic compounds, which could trigger accelerated growth and formation of biofilm of marine bacteria on the RO membranes. In turn, the accelerated biofilm formation could cause heavy fouling of the RO membranes and could result in a significant reduction of desalination plant production capacity within several weeks, and sometimes days, after the beginning of the algal bloom event. In such source water conditions, gravity downflow granular media filtration may be preferable over membrane pretreatment because it allows removing the algae from the source water with a minimum breakage of their cells. Therefore, gravity granular media filters have found much wider application than UF or MF filters for pretreatment of saline source waters originating from shallow intakes, which are exposed to high-intensity algal bloom events.

Conversely, if the source water used for desalination originates from subsurface intake (e.g., beach wells), this water does not contain algae and its pretreatment using pressure-driven granular media filters or MF or UF membranes is usually more cost competitive than conventional gravity granular media filtration. In such conditions, pretreatment often could be completed adequately by using cartridge filters only.

10.3 SURFACE AREA REQUIREMENTS

Membrane pretreatment is typically more space efficient compared to granular media filtration. The smaller footprint benefits of membrane filtration are usually of greater importance when the site available for building new or upgrading existing desalination plant is very limited or where the cost of new land acquisition is very high. Typically, the footprint of conventional single-stage dual media filtration system is 30–50 $m^2/1000\ m^3$ day (1200–2100 ft^2/MGD) of desalination plant production capacity.

The space benefits of membrane filtration are more significant for high-turbidity source waters where two-stage granular media filtration may be required to achieve comparable performance to a single-stage membrane pretreatment system. For example, difficult to treat source water, which necessitates the granular media filtration system to be designed for surface loading rates of less than 10 m^3/m^2 h (4 gpm/ft^2) or where a two-stage granular media filtration is needed to produce comparable filter effluent, the membrane filtration systems may have up to 80% smaller footprint.

Under good to moderate source water quality conditions, when granular media filters are designed at a surface loading rate of 8.5–12.0 m^3/m^2 h (3.5–5.0 gpm/ft^2), their surface is approximately 40%–50% larger than that of an ultra- or microfiltration systems producing similar filtered water quality. For better-than-average source water quality (silt density index—$SDI_{15} < 4$) where granular media filters can perform adequately at surface loading rates of 15–20 m^3/m^2 h (6–8 gpm/ft^2), the total footprint difference is usually only 20%–30% to the benefit of membrane pretreatment.

10.4 QUANTITY AND QUALITY OF GENERATED RESIDUALS

Conventional and membrane pretreatment systems differ significantly by the type, quality, and quantity the residuals generated during the filtration process (Table 10.1). Typically,

TABLE 10.1 Comparison of Waste Streams From Granular Media and Membrane Pretreatment

Waste Stream (Percent of Plant Intake Volume)	Granular Media Filtration	Membrane Filtration
Intake bar screens wash-water	0.1–0.2	0.1–0.2
Microscreen wash water	None (not needed)	0.5–1.5
Spent filter backwash water (reject)	3.0–6.0	5.0–10.0
Chemically enhanced backwash	None (not needed)	0.2–0.4
Spent membrane cleaning chemicals	None (not needed)	0.03–0.05
Total (% of feed volume)	**3.10–6.20**	**5.83–12.15**

granular media filtration systems generate only one large liquid waste stream—spent filter backwash. The volume of this stream in a well-designed and operated desalination plant varies between 3% and 6% of the total plant intake source water volume. In addition to the particulate solids and colloids that are contained in the saline source water, this waste stream also contains coagulant (typically iron salt) and may have flocculant (polymer). Usually, for every kilogram of total suspended solids (TSS) in the saline source water, approximately 0.8 kg of coagulant is added for its cost-effective conditioning. As a result, coagulant addition nearly doubles the amount of solids generated by the pretreatment system.

Membrane pretreatment systems generate two large liquid residual streams: (1) spent membrane backwash water (reject) and (2) membrane cleaning solution from daily chemically enhanced backwash (CEB). The volume of the spent membrane filter backwash water is typically 5%–10% of the plant intake source volume—i.e., approximately two times larger than the spent filter backwash water volume of granular media pretreatment systems. In addition, they generate waste stream from the backwash of the microscreens.

The difference in total liquid residual volume generated by membrane pretreatment systems is even larger, taking into account that the microscreens needed to protect the membrane pretreatment filters will be a source of an additional waste discharge from their intermittent cleaning. Although conventional traveling fine bar screens use 0.1%–0.2% of the intake source water for cleaning, the microscreens generate waste screen-wash volume, which equals to 0.5%–1.5% of the intake flow. The relatively larger waste stream volume of the membrane pretreatment system would require proportionally larger intake source water volume, which in turn would result in increased size and construction costs for the desalination plant intake facilities, and pump station, and in higher operation and maintenance (O&M) costs for source water pumping to the pretreatment facilities.

In addition to the daily membrane washing and monthly membrane cleaning, cost competitive design and operation of membrane pretreatment systems require daily CEB of the membranes using a large dosage of chlorine (typically 200–1,000 mg/L) and strong base or/and acid over a short period of time. This performance enhancing CEB adds to the volume of the waste streams generated at the RO membrane plant and to the overall cost of source water pretreatment. The daily volume of waste stream generated during CEB is usually 0.2%–0.4% of the volume of the intake source water.

Another waste stream that is associated only with membrane pretreatment is generated during the periodic chemical cleaning of the pretreatment membranes. Extended off-line chemical cleaning, often referred to as clean-in-place (CIP), during which membranes are soaked in a solution of hydrochloric and/or citric acid, sodium hydroxide, biocides, and surfactants, is critical for maintaining steady-state membrane performance and productivity, and such cleaning is usually needed once every 1–3 months. The CIP cleaning generates additional waste stream that is 0.03%–0.05% of the source water volume.

One key advantage of membrane pretreatment systems is that the waste filter backwash generated by them contains less or often no source water conditioning chemicals (e.g., coagulant, flocculant, and acid), and therefore, it is more environmentally friendly compared to the waste filter backwash stream generated by conventional granular media filtration pretreatment facilities. This benefit stems from the fact that typically coagulant dosage for source water pretreatment by membrane filtration is two to three times lower than that for granular media filtration.

In some cases, source water may not need to be conditioned with coagulant before membrane pretreatment and this spent filter backwash water could be disposed along with the RO concentrate without further treatment. For comparison, due to the high content of iron, the spent filter backwash from granular media filtration pretreatment would need to be treated by sedimentation and the settled solids would need to be dewatered and disposed to sanitary landfill. Otherwise, the high content of iron salt in the backwash water will cause the desalination plant discharge to have a red color every time, when a pretreatment filter is backwashed and the backwash is discharged with the plant concentrate. Fig. 10.1 depicts the reddish discoloration of the desalination plant discharge caused by the iron coagulant contained in the spent backwash water generated from granular media filtration system. After the addition to the saline source water, the iron coagulant (ferric chloride or ferric sulfate) is converted to ferric hydroxide, which is also commonly known as "rust" and has red color.

The waste streams generated during the CEB and the CIP membrane cleaning should be pretreated onsite in a neutralization tank before discharge. The additional treatment and disposal costs of the waste membrane cleaning chemicals should be taken under consideration when comparing membrane and granular media pretreatment systems.

FIGURE 10.1 Discharge of untreated backwash water from granular media filters.

10.5 CHEMICAL USE

Typically, the cost of chemical conditioning of source water for granular media filtration is in a range of 4%–6% of the total annual O&M costs for production of desalinated water. Granular media pretreatment systems use 50%–100% more source water conditioning chemicals (iron salts and sometimes polymer) than membrane pretreatment systems for the removal of particulate and colloidal foulants (Prihasto et al., 2009). Practical experience to date shows that most membrane pretreatment systems do not use coagulants or only apply them intermittently during periods of severe algal blooms or heavy rains.

Granular media pretreatment systems do not use any chemicals for filtration media cleaning (except for an occasional addition of chlorine). In contrast, membrane pretreatment systems use a significant quantity of membrane cleaning chemicals for CEB and CIP, which in terms of total annual chemical costs may be comparable to the total costs of source water conditioning chemicals used by granular media filters. The cost of these cleaning chemicals should be considered in the cost–benefit analysis of the plant pretreatment system.

Another factor that should be accounted for in the overall plant chemical use and cost analysis is the RO system cleaning frequency, and therefore, the RO membrane cleaning costs. These costs may be reduced by using membrane pretreatment due to the typically better solids and silt removal efficiency of this type of pretreatment. However, if microbial fouling (e.g., biofouling) is the predominant type of RO fouling that occurs at a given desalination plant, membrane filtration typically does not offer any significant advantages to granular media pretreatment and in some cases may accelerate the biofouling rate due to enhanced algal cell breakage/release of easily biodegradable organics under the high pressure/vacuum needed for filtration.

Under such circumstances, the accelerated biofouling caused by the breakage of algal cells by the membrane pretreatment system would result in more frequent (rather than reduced) RO membrane cleaning and elevated CIP chemical costs as compared to the use of gravity granular media filtration system for source water pretreatment.

A significant difference between the two types of pretreatment systems is the amount of chlorine used for filtration media maintenance. To control pretreatment filtration rate and RO biofouling, granular media filters are occasionally fed with chlorinated source water, which contains 1.5–5.0 mg/L of chlorine. This, so-called "shock" chlorination is typically completed once per month for 4–6 h at a time.

For comparison, chemically enhanced backwash with chlorine dosages of 20–200 mg/L is performed on all pretreatment membranes a minimum of once per day for a period of 20–30 min. Because the CEB process of most membrane pretreatment systems involves air–water backwash, some of the chlorine is stripped into the surrounding air and can cause corrosion of nearby equipment and unprotected structures. Therefore, the use of protective filter cell structure and equipment coatings and of suitable corrosion resistant materials is of critical importance.

In addition, some of the applied chlorine is soaked into the pretreatment membranes and they may leach chlorine into the RO system feed for 20–40 min after CEB. Therefore, dechlorination of the filtered source water with sodium bisulfate after CEB cleaning is very important and the additional cost of dechlorination chemical should be taken into consideration when comparing granular media and membrane source water pretreatment.

10.6 POWER USE

Granular media pretreatment systems use limited amount of power to separate particulates from the source water. As mentioned previously, large RO desalination plants typically include single-stage gravity granular media filtration pretreatment process, which has minimum power requirements—typically less than 0.05 kWh/m^3 (0.2 kWh/1000 gal). On the other hand, depending on the type of the membrane system (pressure or vacuum-driven), membrane systems use approximately four to six times more power—0.2 to 0.4 kWh/m^3 (0.75–1.5 kWh/1000 gal) to remove particulates from the source water compared to gravity granular media filters. More power is not only used to create a flow-driving pressure through the membranes, but also for membrane backwash and source water pumping. The total power use has to be taken into consideration when completing a life-cycle cost comparison of conventional versus membrane pretreatment system for a given application.

Although the power demand difference holds true for comparison of single-stage gravity granular media filters and single-stage pressure or vacuum-driven UF or MF filters, this difference is negligible if pressure granular media filters are used for pretreatment. Single-stage pressure granular media filters operate at comparable feed pressures and power demand to membrane pretreatment systems. However, two-stage pressure filtration systems typically use more electricity than a single-stage MF or UF system producing comparable filtered water quality.

For example, a comparative cost analysis between a two-stage pressure filtration system and a single-stage pressure-driven membrane pretreatment system completed during the planning phase of the Perth II SWRO Project in Australia reveals that the two-stage pressure filter system would use 20% more electricity than the membrane pretreatment system (Molina et al., 2009). For this project, the need to consider two-stage granular media pretreatment was driven by the significant fluctuation in source water turbidity (5–50 mg/L of TSS) attributed to the relatively shallow plant intake.

10.7 ECONOMY OF SCALE

Membrane and granular media pretreatment systems may yield different economies of scale depending on the water treatment plant capacity. Usually, both technologies have a comparable economy of scale for plant capacity of up to 40,000 m^3/day (10.6 MGD). For desalination plants with a capacity of 40,000–200,000 m^3/day (10.6–52.8 MGD), the granular media filtration systems typically yield higher economy of scale benefits. The anticipated economy of scale reduction of construction costs for membrane pretreatment capacity increase from 40,000 to 200,000 m^3/day (10.6–52.8 MGD) is in a range of 3%–5% only. For comparison, granular media filtration pretreatment systems in the same capacity range may yield 8%–10% economy of scale-related construction cost benefits.

The main reason for the smaller economy-of-scale benefits of membrane pretreatment technologies for large-capacity desalination plants is the maximum size of membrane modules currently available on the market. Typically, depending on the manufacturer and the membrane technology, the largest membrane modules available today are between 2,000 and 8,000 m^3/day (0.5–2.1 MGD) of water production capacity, although recently some immersed membrane system manufacturers offer membrane modules of up to 20,000 m^3/day (5.3 MGD) of production capacity. In comparison, the maximum size of the individual granular media filter cells can reach 32,000 m^3/day (8.5 MGD) or more, thereby allowing the higher overall construction cost reduction due to the fewer filter cells, less service equipment, and piping.

One of the current source water pretreatment trends worldwide is to use membrane technologies for large-plant applications. As the number and type of large plant membrane application opportunities increase in the future, it is likely that the membrane manufacturers will develop larger-size individual membrane modules, which in turn would improve membrane pretreatment system economy of scale and competitiveness.

10.8 FILTRATION MEDIA REPLACEMENT COSTS

Well-operating granular media filters lose 5%–10% of filter media every 5 years, which has to be replaced to maintain consistent performance. The costs of granular media replacement are usually well predictable and relatively low. At present, membrane element useful life typically varies in a range of 5–7 years. Assuming 5 years of average pretreatment membrane useful life, approximately 20% of the membrane elements would need to be replaced per year to maintain pretreatment system production capacity and performance. Taking under consideration that the annual costs for replacement of MF/UF membranes are comparable to the annual costs for replacement of RO elements, the use of membrane pretreatment would result in an order-of-magnitude higher initial and annual expenditures for media replacement than granular media filtration.

Additional factor that may contribute to the need for more frequent replacement of membrane elements is the failure of membrane element integrity. Typically, the main reason triggering the need for early membrane element replacement is the loss of integrity rather than the loss of production capacity. The limited track record of long-term use of membrane systems and the uncertainty related to the factors triggering the need for their replacement have to be taken under consideration when selecting between granular media and membrane pretreatment for RO desalination plants. The risk of loss of membrane integrity should be handled accordingly in the membrane element useful life warranty provided by the membrane manufacturer/supplier. Upon request, most MF or UF membrane suppliers are willing to guarantee maximum percent of fiber breakage per year, which should be below 0.1% of the total number of membrane fibers.

As indicated previously, in most cases, the use of membrane pretreatment would result in filtered source water of lower particular and colloidal fouling capacity. As a result theoretically, the use of membrane pretreatment instead of granular filtration should reduce the frequency of RO membrane cleaning and replacement. This observation holds true especially for source seawaters of low-microbial fouling potential. However, because of the limited full-scale performance track record to verify this assumption and the fact that for seawaters with high-microbial fouling potential membrane pretreatment would make very little difference in terms of frequency of RO cleaning, at present most RO membrane suppliers are reluctant to provide warrantees for longer useful life or lower cleaning frequency of the RO membranes downstream of the membrane pretreatment system. As a result, this potential benefit of membrane pretreatment cannot be easily accounted for in an actual cost–benefit analysis for full-scale desalination projects. As indicated previously, practical experience at desalination plants with open intakes and frequent algal bloom events shows the opposite trend—RO membranes downstream of UF or MF pretreatment systems are actually cleaned more frequently than these downstream of gravity granular media filtration systems due to the negative impact of the pressurization of the source water on the algal cells contained in it—algal cell breakage and release of easily biodegradable cell substances that accelerate RO membrane fouling.

10.9 COMMODITIZATION

Granular media filtration systems are fully commoditized and the same quality of filtration media can be obtained by a number of manufacturers at a competitive market price. For comparison, membrane pretreatment systems are not commoditized—all UF and MF membrane manufacturers offer their own design, size, and configuration of membrane elements/modules, and pretreatment system racks. The membrane pretreatment systems also differ by the type, concentration, and volume of membrane cleaning chemicals needed for CEB and CIP, by filter backwash rate, type, and sequence, and by their membrane integrity testing method.

The absence of product uniformity and commoditization in the membrane market at present is an indication of the fact that membrane filtration is a fast-growing field of the water equipment industry and carries some benefits and disadvantages. The availability of multiple membrane suppliers and systems allows to better accommodate the site-specific needs of a given membrane application, thereby increasing the potential for use of membrane source water pretreatment. In addition, the lack of commoditization of the MF and UF membrane market along with the increase in membrane applications in recent years, spurs the interest of many manufacturers, which traditionally do not produce membranes to enter the membrane market with new products. This, in turn, results in increased competition and in accelerated development of new membrane technologies, products, and equipment.

Fifteen years ago, there were less than half-a-dozen of membrane manufacturers, which offered MF and UF membranes and membrane systems to the municipal market and to the desalination industry. This number has increased dramatically over the past 5 years and today practically all large and many medium-size equipment and plastics manufacturers offer their own unique MF or UF membrane system.

The absence of standardization of membrane size, vessels, and configuration, however, also has a number of disadvantages that may hinder the use of membrane pretreatment, especially for large desalination plants. As the membrane market gets oversaturated with manufacturers offering similar membrane products, the market growth is likely to exceed the demand, which would trigger the exit of some of the current membrane manufacturers form the market. As a result, the manufacturers exciting the membrane market would no longer produce membrane elements and provide maintenance and technical support for their existing systems. Because their system configuration, membrane elements, and vessel type are unique, the owners of such membrane systems will have to invest significant funds and efforts to modify their membrane installations to accommodate alternative membrane equipment.

The current diversity of membrane element sizes and configurations and the lack of standardization and commoditization may have a number of disadvantages for the membrane plant owner in the long run. If an existing membrane manufacturer discontinues the production of membrane elements or a given type (e.g., abandon production of submersible systems in favor of pressure membrane systems), the submersible membrane pretreatment system owner would incur additional costs to procure and install a new pretreatment system because the other available membrane systems would be incompatible with the owner's existing system. While replacing/retrofitting the existing pretreatment system to accommodate new membranes, the desalination plant owner would likely face reduced plant production capacity due to the downtime needed for membrane system replacement and the fact that the productivity of old membrane elements, which cannot be replaced with alternative membrane product when needed, will decrease over time.

The membrane plant owner is likely to also incur additional costs to train their staff in operating and maintaining the new membrane pretreatment system. In addition, the owner may experience a potential increase in unit membrane element and vessel costs over time because the membrane elements would have to be purchased from a sole-source manufacturer rather than to be competitively procured at market price and warrantee conditions.

Installation of nonstandardized membrane elements and vessels limits the desalination plant owner opportunities to benefit from the use of new and improved membrane pretreatment technologies, and elements, which might be readily available in the future. In the current highly diversified membrane technology market, membrane plant owner would be very dependent on the commitment of the pretreatment membrane manufacturer, whose system they use, to excel in their existing technology and to develop competitive and compatible membrane elements and technologies in the future.

An example that illustrates the concerns discussed earlier was observed in the seawater desalination membrane market approximately 20 years ago, when one of the key manufacturers of hollow-fiber RO membrane elements, the DuPont's subsidiary—Permasep, decided to exit the market for these membranes. In the nineties, Permasep had a dominant portion of the membrane source water desalination market supplying hollow-fiber RO membrane elements to several thousand membrane installations worldwide. The hollow-fiber membrane elements and vessels used by Permasep were different from these used by other hollow-fiber membrane manufacturers and incompatible with these of other manufacturers offering spiral-wound desalination membranes. Permasep's exit of the membrane source water desalination market triggered the need for significant modifications and expenditures by the desalination plant owners using their RO membranes to accommodate the necessary changes.

Standardization of membrane systems, elements, and vessels has another significant advantage to the owner of the desalination facility, which has been proven by the evolution of the RO membrane market—the significant reduction of membrane costs. Currently, RO desalination membranes and vessels produced by various manufacturers are standardized in size, configuration, and performance and can be used interchangeably. The commoditization of the RO desalination market over the past 20 years contributed to the two- to three-fold reduction of RO membrane element costs, which on the other hand spurred the development of new large size RO desalination plants worldwide.

Another, often forgotten, benefit of membrane technology unification is the potential reduction of the cost of membrane plant funding, and therefore, of the overall cost of water production. The capital cost of a given desalination project consists of two key elements—(1) cost of construction and (2) cost of capital needed to develop the project and finance this construction. Because the cost of capital is typically 20%–30% of the overall project costs, using commoditized membrane pretreatment systems could yield cost benefits sometimes higher than the savings that may result from implementation of new and unique advanced technologies or equipment.

A membrane pretreatment system, which can accommodate a number of different membrane elements, vessels and equipment is considered a lower investment risk and, lower cost-of-capital system. Therefore, considering all other conditions being equal, the cost of capital (e.g., bond interest rate or equity return expectation) for funding a project using standardized membranes or well-proven conventional granular media pretreatment system would typically be lower compared to that for funding desalination plant that uses a unique membrane pretreatment system configuration and elements, which cannot be supplied competitively from alternative manufacturers.

Although a new advanced membrane pretreatment system that has unique features may yield appreciable near-term construction and operation cost savings, these savings may be compromised over the useful life of the project, which is typically 25–30 years, if the pretreatment system design is not flexible enough to accommodate the benefits of future membrane technologies, especially taking under consideration that the UF and MF membrane technologies are in an exponential stage of development today and new or improved competitive products and systems are available almost every year.

Based on the current status and diversity of the micro- and ultrafiltration technologies, a sound approach toward reducing risks associated with the funding and implementation of a membrane pretreatment system is to design the system configuration in such a manner that would accommodate the replacement of this system/membrane elements with at least one other existing system/membrane elements of similar type. For example, if the preliminary engineering analysis and subsequent pilot testing indicate that a submersible vacuum-driven type of membrane pretreatment system is more suitable for a particular application, this desalination plant should be designed to accommodate at least one or two other submersible membrane systems currently available on the market. The additional construction and installation cost expenditures to provide flexible pretreatment system configuration that allows future modifications and use of alternative suppliers of the same type of membrane elements at minimal expenditure or replacement are very likely to be compensated by lowering the funding costs (costs of capital) for the project and by minimizing the overall life-cycle costs of the membrane plant. Similarly, as indicated in Chapter 9, the project owner could consider the use of "Universal" MF/UF system that allows accommodating different UF membrane modules from a number of alternative manufacturers. Example of such a system (the Spectrum Universal UF Rack from Wigen Water Technologies) is illustrated in Fig. 10.2.

This rack can accommodate UF membranes from different manufacturers including Toray, Pall, GE/Zenon, Dow, Memstar, and Siemens (Evoqua).

FIGURE 10.2 Universal ultrafiltration rack that accommodates membranes from several manufacturers.

10.10 WATER PRODUCTION COSTS

At present, the overall cost of production of desalinated water using the membrane pretreatment is typically 5%–15% higher than that for fresh water produced by desalination plants with conventional source water pretreatment. In some cases, such as conditions when the source water quality is highly variable and/or the cost and availability of land are at premium, membrane pretreatment may be more cost advantageous. Also, when source water quality is fairly high, the membrane pretreatment system could be designed quite aggressively and could have a clear capital and life-cycle cost advantage compared with conventional granular media filtration.

Key factors that are often underestimated or omitted when comparing granular media and membrane pretreatment systems are (1) the additional capital and O&M costs of the microscreening system needed to protect the integrity of the pretreatment membranes; (2) the actual chemical costs and frequency of pretreatment membrane cleaning and chemically enhanced backwash; (3) the useful life and replacement costs of the pretreatment membranes—most analyses assume 5 years—whereas actual operational data show that membranes need to be replaced in approximately 3 years due to the loss of integrity; (4) the 5%–10% higher cost of project financing associated with the use of membrane pretreatment because of the long-term risk associated with the use of technology of limited full-scale track record and commoditization, especially for large-scale desalination plants.

Table 10.2 presents an example of the cost comparison of conventional gravity dual-media filtration system and UF vacuum-driven pretreatment system for seawater reverse osmosis (SWRO) desalination plant with a production capacity of 50,000 m^3/day (13.2 MGD). This example assumes conventional pretreatment that consists of single-stage dual media filters and membrane pretreatment employing vacuum-driven UF membranes. The conventional pretreatment system is assumed to use 5 mg/L of ferric chloride for source water coagulation, whereas the membrane pretreatment system is designed to operate without coagulant addition. As a result, the desalination plant with conventional pretreatment incorporates a solid handling system for treatment of spent filter backwash water, whereas the membrane pretreatment system does not include solids handling facilities, and the spent membrane wash water is assumed to be disposed to the ocean with the RO system concentrate without further treatment. This assumption represents a best-case solids handling scenario for desalination plant systems with membrane pretreatment.

Other assumptions in this example that favor membrane pretreatment are (1) relatively high land costs of the desalination plant site; (2) 12.5% higher design flux (and therefore smaller size) of the RO system using membrane pretreatment; (3) avoidance of cartridge filter installation upstream of the RO system with membrane pretreatment; (4) relatively long useful life of the membrane pretreatment filters (5 years); and (5) reduction of RO membrane cleaning costs due to membrane pretreatment, which typically would not be the case if the main type of RO membranes experience is microbial biofouling.

Review of Table 10.2 indicates that both capital and O&M costs for conventional pretreatment are lower than these for membrane pretreatment. To determine the amortized value of the capital costs, these costs are divided by a capital recovery factor (CRF). The CRF is a function of the interest rate of the capital and the number of years over which the investment is repaid. The CRF can be calculated using the following formula:

$$\mathrm{CRF} = \{(1+\mathrm{i})^{\mathrm{n}} - 1\}/\{\mathrm{i}\,(1+\mathrm{i})^{\mathrm{n}}\} \tag{10.1}$$

where n is the period of repayment of capital expenditures and i is the interest rate of the amortized investment.

In this example, the total construction costs for both plants are amortized using CRF estimated for the interest rate of 5.7% over a period of 20 years. Applying Formula (10.1) for these conditions results in CRF of 11.752. The capital cost component of the seawater desalination costs will be US\$0.48/m^3 (US\$1.82/1,000 gal) for a plant with conventional pretreatment and US\$0.50/m^3 (US\$1.89/1,000 gal) for plant with membrane pretreatment. These costs are determined by dividing the construction costs in Table 10.2 by the CRF and by the annual production capacity of the plant. For example, for the plant with conventional pretreatment system, these costs are calculated as follows: US\$102,200,000/(11.752 × 50,000 m^3/day × 365 days) = US\$0.48/m^3.

The annual O&M costs for the conventional pretreatment system for this example are estimated at US\$9,470,000/year (see Table 10.2). When converted to the O&M cost of water component of the desalination plant, these costs are US\$9,470,000 per year/(50,000 m^3/day × 365 days) = US\$0.52/m^3 (US\$1.96/1,000 gal). Similarly, the O&M costs for the desalination plant with membrane pretreatment system are calculated as US\$0.53/m^3 (US\$1.82/1,000 gal). Based on the capital and O&M cost estimates presented earlier, the

TABLE 10.2 Cost Comparison of 50,000 m^3/day Seawater Reverse Osmosis (SWRO) Desalination Plants With Granular Media and Membrane Pretreatment

Item	Granular Media Pretreatment 2017 US$ (in 1,000)	Membrane Pretreatment 2017 US$ (in 1,000)
CAPITAL COSTS		
Open ocean intake	2,760	2,900
Intake pump station	1,350	1,450
Coarse and fine screens	690	750
Microscreens	0	1,650
Coagulation/flocculation system	470	0
Cartridge filters	1,320	0
Source water chlorination system	210	220
Pretreatment membrane cleaning system	0	990
Filter tanks (excluding media/membranes)	4,890	3,490
Filtration media (sand/anthracite or ultrafiltration membranes)	550	4,280
Membrane pretreatment system—service equipment	0	2,480
Filter backwash system	520	880
Dechlorination system	110	190
Land costs	1,380	990
SWRO system	35,320	30,910
Posttreatment system	2020	2020
Solids handling facilities	1,520	100
Discharge outfall	2,530	2,690
Other facilities and systems	9,910	9,910
Engineering and construction management	10,140	12,950
Start up and commissioning	2,060	2,410
Other costs	24,450	26,840
Total capital costs	**US$102,200**	**US$108,100**
Amortized capital costs US$/$m^3$ (US$/1000 gal)	*US$0.48 (US$1.82)*	*US$0.50 (US$1.89)*

(*Continued*)

TABLE 10.2 Cost Comparison of 50,000 m^3/day Seawater Reverse Osmosis (SWRO) Desalination Plants With Granular Media and Membrane Pretreatment—cont'd

Item	Granular Media Pretreatment 2017 US$ (in 1,000)	Membrane Pretreatment 2017 US$ (in 1,000)
Operation and Maintenance (O&M) Costs	**2017 US$/year (in 1,000)**	**2017 US$/year (in 1,000)**
Labor	580	690
Chemicals for coagulation/flocculation	310	0
Chemicals for pretreatment membrane cleaning	0	260
Chemicals for chemically enhanced backwash of pretreatment membranes	0	170
Chemicals for reverse osmosis (RO) membrane cleaning	170	110
Other chemicals	140	160
Microscreen maintenance and spare parts	0	40
Cartridge filter replacement	180	0
Pretreatment membrane replacement	0	310
RO membrane replacement	890	630
Granular media addition	30	0
Other maintenance and spare part costs	940	1100
Solids handling and sludge disposal	410	0
Disposal of spent membrane cleaning solution to sewer	40	110
Power use for seawater pretreatment	60	420
Power use by RO and other systems	4,590	4,590
Other O&M costs	1,130	1,130
Total annual O&M costs	**US$9,470**	**US$9,720**
Annual O&M costs, US/m^3(US$/1,000 gal)	*US$0.52 (1.96)*	*US$0.53 (2.00)*
Water production cost, US$/$m^3$ (US$1,000 gal)	**US$1.00 (3.78)**	**US$1.03 (3.89)**

total water production costs for RO plant with conventional and membrane pretreatment are US$1.00/$m^3$ and US$1.03/$m^3$ (US$3.78 and 3.89 US$/1000 gal), respectively.

For the example shown in Table 10.2, the cost of seawater desalination using membrane pretreatment is slightly higher (3%) even when the water quality and site-specific conditions favor the use of this type of pretreatment. The main items where the construction costs of the two systems differ significantly are the costs of filtration media, the pretreatment system, the RO system, and the solids handling facilities. The intake costs for the desalination plant with membrane pretreatment system are higher because this system would require the collection of 8% more source seawater than the conventional pretreatment system. As explained previously, this additional intake water is needed for the washing of the microscreens and the backwash water of the pretreatment membranes.

The costs associated with the RO system are lower for the membrane pretreatment system because this system is designed at a 12.5% higher flux (15.3 vs. 13.6 L/m^2 h). The high design flux allowance for RO systems with membrane pretreatment stems from the expectation that membrane filtration would provide superior pretreatment. Explanation for the cost differences of the other items could be found in the previous sections of this chapter.

The main differences of the O&M costs of the two desalination plants are related to the higher use of power for the membrane pretreatment process (0.35 vs. 0.04 kWh/m^3) and to the pretreatment system maintenance and membrane replacement costs. It should be pointed out that depending on the applied membrane pretreatment technology, the annual cost for replacement of pretreatment membranes could be comparable to these for replacement of RO membrane elements. On the other hand, the use of membrane pretreatment is expected to eliminate or significantly minimize the sludge disposal costs and to decrease RO membrane replacement rate, cleaning frequency, and costs.

Although the design assumptions used in Table 10.2 favor membrane pretreatment, in many cases, not all of the benefits of this type of pretreatment may be applicable to the site-specific conditions of the a given RO project and the cost of water difference between membrane and conventional pretreatment could exceed 10% in favor of conventional pretreatment. As membrane filtration technologies evolve and next generations of membrane products are more closely tailored to fit the specific challenges of saline water pretreatment, it is very likely that membrane pretreatment would become cost competitive for the majority of saline source water conditions.

10.11 CONCLUDING REMARKS

Granular media filtration is currently the dominating technology for saline water pretreatment worldwide. However, since year 2000, membrane pretreatment has been gaining wider acceptance due to its superior removal of particulate and colloidal foulants, simplicity of plant operation, performance stability, and more compact footprint. In addition, membrane pretreatment is easier to monitor, operate, and automate. These advantages are likely to propel membrane filtration into the pretreatment technology of choice in the future.

At present, two key impediments for the wider application of MF and UF membrane technology are the lack of compatibility between various membrane products and configurations—as well as the relatively higher costs of the membrane pretreatment systems. These challenges are likely to be solved in the near future by the convergence and commoditization of various UF and MF membrane technologies. Many of the key membrane manufacturers such as Dow, Toray, Hydranautics, Pall, GE Water and Power Technologies, and Memcor (Evoqua) are currently working on the development of commoditized products and systems, which will have pressure-driven UF membranes made of polyvinylidene difluoride (PVDF) material, outside-in filtration direction, and air–water backwash. These products are projected to be commercially available by the year 2025.

Meanwhile, a number of companies, such as Suez, Veolia, H_2O Innovation, and Wigen Water are already offering Universal (Open Platform) UF/MF rack systems, which can accommodate the latest models of pressure-driven UF membrane products made of PVDF material from most of the key membrane manufacturers of products and services for the desalination industry (Guibert, 2014; Sweizbin, 2017).

Another challenge that faces the wider use of membrane pretreatment is the limited durability, loss of integrity, and relatively short useful life of the membranes. The limited pretreatment membrane lifespan has also adverse effect on the costs for their replacement. Currently, the annual MF/UF seawater membrane replacement costs are typically comparable to the annual replacement costs of the SWRO membranes they are installed to protect. These challenges are expected to be addressed with the development of alternative ceramic membrane materials or other plastic materials, which are less sensitive to integrity failures from sharp objects and biofouling (Kang et al., 2016).

To date, the main emphasis of the research and development in the field of membrane pretreatment technology has been on the effective removal of particulate and colloidal fouling. However, worldwide experience with seawater desalination over the last two decades indicates that abating microbial fouling is often the most challenging aspect of saline water pretreatment. Therefore, it is very likely that the paradigm of membrane pretreatment technology in the future will change from focus on removal of particulate and colloidal foulants from saline water to emphasis on removal of soluble easily biodegradable organics in membrane biological reactors, similar to these used in wastewater treatment. As a result, a new generation of bioreactor-based membrane pretreatment technologies is likely to emerge, develop, and excel before this type of pretreatment receives widespread acceptance in desalination plants.

References

AWWA, 2007. Manual of Water Supply Practices M46, Reverse Osmosis and Nanofiltration, second ed. American Water Works Association, Denver, Colorado.

Guibert, D., 2014. MF/UF developments: universal rack design and process convergence. In: Proceedings of AMTA Technology Transfer Workshop, Ann Arbor, Michigan, May 13, 2014.

Kang, J.S., Sung, S.C., Lee, J.J., Kim, H.S., 2016. Application of ceramic membrane for seawater desalination pretreatment. Desalination and Water Treatment 5 (55), 26700–26705. Taylor & Francis.

Molina, G.V., Marzal, M.A., Hoehn, K.U., 2009. Designing membrane systems for coming future: Perth II desalination plant. In: Proceedings of IDA World Congress, Dubai, November 7–12, 2009.

Pearce, C., 2011. UF/MF Membrane Water Treatment Principles and Design. Water Treatment Academy, TechnoBiz Communications.

Prihasto, N., Qi-Feng, L., Seung-Hyun, K., 2009. Pre-treatment strategies for seawater desalination by reverse osmosis system. Desalination 249, 308–316. Elsevier.

Ransome, T.R., Ferguson, M., de Miguel, A., Sibma, S., 2015. Operation of a large scale RO plant with ultrafiltration pretreatment: challenges in early operation. In: Proceedings of IDA World Congress on Desalination and Reuse, 2015, San Diego, CA, USA. Ref: IDAWC15_Ransome_51524.

Sweizbin, J., 2017. A concept for open platform membrane system. In: Proceedings of 2017 Membrane Technology Conference & Exposition, February 13–17, Long Beach, California.

CHAPTER

11

Guidelines for Pretreatment System Selection

OUTLINE

11.1 INTRODUCTION

As indicated in Chapter 2, saline source water contains five key types of foulants that depending on their concentration could have a significant impact on desalination plant operations: (1) particulate, (2) colloidal, (3) mineral, (4) natural organic matter, and (5) microbial. The purpose of pretreatment is to reduce the content of all types of foulants contained in the

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source water to levels, which allow for reliable, efficient, and cost-effective desalination by reverse osmosis (RO). However, the selection of pretreatment system configuration is mainly based on the concentrations of the particulate, colloidal, natural organic matter, and microbial foulants because they govern the design of the most widely used pretreatment technologies at present. Mineral scaling is not considered a factor determining the selection of pretreatment system configuration because all saline source waters have scaling compounds at levels that could influence RO system performance and, therefore, all desalination plants are equipped with antiscalant addition systems.

In the pretreatment methodology presented herein, the amount of particulate foulants is quantified by the levels of turbidity and silt density index of the saline source water; colloidal fouling is reflected by the content of total hydrocarbons (THC) and iron and manganese in the water; and natural organic matter and microbial fouling are quantified by total organic carbon concentration. In addition, microbial fouling caused by algal blooms is taken into account by the total concentration of algal cells in the saline source water, which as shown in Table 2.8, defines the severity of the algal blooms that may occur in the intake area.

11.2 PRETREATMENT SELECTION GUIDELINES

The most viable membrane pretreatment configurations and technologies recommended for the site-specific conditions of a given desalination project are classified based on the type of the desalination plant intake and source water quality.

11.2.1 Plants With Subsurface Intake

As indicated in Chapter 4, subsurface intakes (e.g., beach wells, horizontal directionally drilled wells, infiltration galleries) usually collect saline water of superior quality than open intakes because their water is naturally prefiltered though the bottom sediments. As a result, typical saline source water collected by subsurface intakes will have the water quality shown in Table 11.1.

TABLE 11.1 Typical Source Water Quality of Plants With Subsurface Intakes

Source Water Quality Parameter	Value
Turbidity, NTU	<0.5
SDI_{15}	<3.0
Total organic carbon, mg/L	<1.0
Total hydrocarbons, mg/L	<0.01
Total iron in reduced form, mg/L	<0.05
Total manganese in reduced form, mg/L	<0.02
Algae, cells/L	0.0

SDI, silt density index.

Such water quality already complies with the target pretreated water quality listed in Table 8.1 and, therefore, it is directly suitable for processing through the RO system without further pretreatment except for the addition of antiscalant. Therefore, the alternative pretreatment systems, which are most viable for this source water quality are:

- Cartridge filters only;
- Cartridge filters and single-stage dual media pressure granular media filters with a loading rate of 16–25 m^3/m^2 h (7–10 gpm/ft^2);
- Pressure-driven ultrafiltration/microfiltration (UF/MF) filters with a design flux of 100–120 Lmh (60–70 gfd).

In most well intakes collecting saline water, which is unaffected by fresh water runoff or fresh water aquifers, cartridge filtration is sufficient and no additional pretreatment (except for the addition of antiscalant) is needed (Voutchkov, 2010). In this case, cartridge filters mainly serve as a protection device to capture small sand or silt particles, which may be released to the source water stream when individual wells are turned on as well as to capture rust released from the intake wells or other equipment upstream of the RO membranes.

Combination of cartridge filtration and single-stage single or dual media pressure filtration is typically used only when the source water quality is expected to change seasonally or if the coastal aquifer is unconfined and hydraulically connected to potential sources of contamination, which could be immobilized over the useful life of the desalination plant. Under such conditions, the pretreatment filters are often run without the addition of coagulant of flocculant.

In addition, pressure filters are sometimes installed to reduce the frequency of RO system clean-in-place (CIP) frequency and enhance the useful life of the RO membranes. Practical experience shows that the installation of granular media pressure filtration system versus cartridge filters only allows to increase the time between two CIP cleanings of the RO system from once every 6 months to once every 12 months and to extend the useful life of the RO membranes from 7 to 10 years. Such trade-offs between additional capital costs for installation of pretreatment filters and reduced operation and maintenance (O&M) costs for RO membrane cleaning and replacement are warranted when the cost of cleaning chemicals is relatively high or the desalination plant is at a remote location and chemical delivery would be difficult and costly.

Use of pressure-driven UF or MF pretreatment filters instead of pressure-driven granular media filters for this application is usually more costly and is only practiced if periodically the source water contains contaminants such as iron and manganese in reduced form, which are easier to remove with membrane instead of granular media filters.

It should be pointed out that the saline source water of some desalination plants with subsurface intakes contains iron and manganese in reduced form above their threshold levels of 0.05 and 0.02 mg/L, respectively. Such conditions typically occur when the subsurface intake of the desalination plant is located near the estuary of a large river entering the saline water body (e.g., ocean, sea). In this case, to protect the RO membranes against colloidal fouling, the iron and manganese will have to be oxidized (typically by chlorination or permanganate addition), precipitated, and removed by filtration. In this case, the preferred pretreatment option is UF or MF filtration. Alternatively, pressure granular media filtration with special greensand media has to be used to remove iron and manganese below the threshold levels indicated earlier.

11.2.2 Plants With Open Intake and Good Water Quality

Usually, desalination plants with deep open intakes (e.g., intakes with depth of 15 m (50 ft) or more below the water surface) produce excellent source water quality (see Table 11.2).

Although the source water quality produced by deep open intakes is good, such saline waters usually require pretreatment mainly aiming at the reduction of particulate fouling. Biofouling and colloidal fouling potential of such waters are relatively low and typically hydrocarbons cannot reach such depth at quantities that can create RO membrane fouling problems.

Practical experience shows that the most suitable pretreatment for desalination plants with deep intakes and good water quality are:

- Single-stage pressure granular dual media filters with a loading rate of 14–20 m^3/m^2 h (6–8 gpm/ft^2) and cartridge filters;
- Pressure-driven UF/MF filters with a design flux of 70–80 Lmh (41–47 gfd);
- Vacuum-driven UF/MF filters with a design flux of 50–60 Lmh (29–35 gfd).

The most common pretreatment configuration for desalination plants with water quality shown in Table 11.2 to date has been a combination of pressure-driven granular media filters and cartridge filters. Although gravity granular media filters could be used instead of pressure filters, they are usually costlier for this water quality and do not offer the advantage of producing filtered water with a significantly lower biofouling potential because source water at depths higher than 15 m (50 ft) does not contain such large amounts of algae as that collected by shallow intakes.

It should be pointed out that not all deep open intakes do always yield source water quality as good as that presented in Table 11.2. If the open intake is located on the pathway of an underwater current, near the point of confluence with a river, in the tidal zone or in an area with frequent naval traffic (e.g., port or ship channel), the source water column is mixed frequently and the surface source water contamination is spread through the entire water column depth, which in turn results in the deterioration of the source water quality collected by

TABLE 11.2 Typical Source Water Quality of Plants With Deep Open Intakes

Source Water Quality Parameter	Value
Turbidity, NTU	0.5–10.0
SDI_5	8–10
Total organic carbon, mg/L	0.5–1.0
Total hydrocarbons, mg/L	<0.01
Total iron in reduced form, mg/L	<0.05
Total manganese in reduced form, mg/L	<0.02
Algae, cells/L	100–20,000

SDI, silt density index.

the desalination plant intake. Therefore, even when the desalination plant has a deep intake, the source water quality in the vicinity of the intake has to be investigated year-around to capture worst-case scenario water conditions, which ultimately should be used for the selection of the most viable pretreatment system configuration.

11.2.3 Plants With Open Intake Exposed to Moderate Algal Blooms

This type of desalination plants is most common worldwide—such plants usually have medium depth intakes where source water quality is periodically influenced by moderate algal blooms. The typical source water quality for such desalination plants is shown in Table 11.3.

The most efficient and cost-effective pretreatment alternatives for desalination plants with water quality influenced by moderate algal blooms are:

- Single-stage gravity granular dual media filters with a loading rate of 8–10 m^3/m^2 h (3–4 gpm/ft^2) and cartridge filters;
- Pressure-driven UF/MF filters with a design flux of 50–60 Lmh (29–35 gfd);
- Vacuum-driven UF/MF filters with a design flux of 30–40 Lmh (18–23 gfd).

Although all three pretreatment configurations listed earlier have found full-scale implementation worldwide, at present over 90% of the desalination plants with this source water quality use combination of gravity granular media filters and cartridge filters. Because of their relatively low-design loading rate, vacuum-driven UF or MF filters are usually the least cost competitive of the three pretreatment alternatives.

11.2.4 Plants With Open Intake Exposed to High-Intensity Algal Blooms

Desalination plants located in the equatorial and tropical zones of the world are most commonly exposed to high-intensity algal blooms and heavy rains, which could last for prolonged periods of time—2–4 months per year. Such natural events have a profound impact on the source water quality of the desalination plants and their successful management involves

TABLE 11.3 Typical Source Water Quality of Plants With Intakes Exposed to Moderate Algal Blooms

Source Water Quality Parameter	Value
Turbidity, NTU	2–20
SDI_5	10–18
Total organic carbon, mg/L	0.5–1.5
Total hydrocarbons, mg/L	<0.02
Total iron in reduced form, mg/L	<0.05
Total manganese in reduced form, mg/L	<0.02
Algae, cells/L	50–40,000

SDI, silt density index.

TABLE 11.4 Typical Source Water Quality of Plants With Intakes Exposed to High Intensity Algal Blooms

Source Water Quality Parameter	Value
Turbidity, NTU	2–30
$SDI_{2.5}$	12–24
Total organic carbon, mg/L	0.5–8.0
Total hydrocarbons, mg/L	<0.02
Total iron in reduced form, mg/L	<0.05
Total manganese in reduced form, mg/L	<0.02
Algae, cells/L	100–60,000

SDI, silt density index.

the use of more complex, often multistage pretreatment systems. Table 11.4 presents the typical water quality of desalination plants, where intakes are of shallow (6–8 m/20–26 ft) or medium depth (8–15 m/26–50 ft) and exposed to high-intensity algal blooms.

Because of the high particulate and biofouling potentials of saline source waters exposed to high-intensity algal blooms or prolonged seasonal rain events (e.g., monsoons), the successful pretreatment of such waters requires the use of one of the following types of two-stage filtration systems:

- Gravity granular dual media filters with a loading rate of 8–12 m^3/m^2 h (3–5 gpm/ft^2); pressure granular dual media filters with a loading rate of 14–20 m^3/m^2 h (6–8 gpm/ft^2) and cartridge filters;
- Dissolved air flotation (DAF) clarifiers, single-stage gravity granular dual media filters with a loading rate of 6–8 m^3/m^2 h (2.5–3.5 gpm/ft^2) and cartridge filters;
- Gravity granular dual media filters with a loading rate of 10–12 m^3/m^2 h (4–5 gpm/ft^2) and pressure-driven UF/MF filters with a design flux of 70–80 Lmh (41–47 gfd);
- Gravity granular dual media filters with loading rate of 10–12 m^3/m^2 h (4–5 gpm/ft^2) and vacuum-driven UF/MF filters with a design flux of 50–60 Lmh (29–35 gfd).
- DAF clarifiers and pressure-driven UF/MF filters with a design flux of 50–60 Lmh (29–35 gfd);
- DAF clarifiers and vacuum-driven UF/MF filters with a design flux of 30–40 Lmh (18–23 gfd).

The most efficient and cost-effective configuration for the site-specific conditions of a given desalination project should be established based on lifecycle cost analysis of these alternatives. The pretreatment configuration that has found the widest application for such source waters at present is a combination of dual media gravity filters and dual media pressure filters followed by cartridge filters.

The selection of DAF or gravity granular media filters as a first pretreatment stage will depend mainly on three factors: (1) length of periods when the source water turbidity exceeds

10 NTU; (2) the predominant size of algae in the source water; and (3) the content of THC in the water.

It should be pointed out that DAF clarifiers do not operate well when treating source waters with turbidity lower than 10 NTU. If the year-around source water quality analysis for a specific project shows that periods of turbidity concentration over 10 NTU last less than 2 weeks per year, then gravity granular media filters will be a better choice for the first-stage pretreatment of the source water because they provide a much higher reduction of particulate and organic foulants at lower source water turbidities.

Over 10 years of experience with using DAF for seawater pretreatment in equatorial and tropical zones of the world shows that gravity granular media filters provide better removal of microalgae (e.g., algae with size $< 20\ \mu m$) than DAF clarifiers. Therefore, the selection of one technology over the other should be based on the algal profiles completed in the intake area during the season of intensive algal blooms.

If the desalination plant intake is located in industrial or recreational ports, which are exposed to frequent oil spills, the use of DAF as a first pretreatment stage would be more advantageous than the construction of gravity granular media filters because DAF can provide a significantly higher THC removal efficiency than filtration (95%–99% vs. 10%–15%). Usually, granular media and UF/MF filters cannot provide effective protection of the RO membranes against colloidal fouling caused by oil spills when the THC concentration in the source water exceeds 0.1 mg/L.

11.2.5 Plants With Open Intake Exposed to Severe Algal Blooms

Prior experience shows that some enclosed saline water bodies, such as the Persian Gulf and Red Sea, could experience severe algal blooms during which the algal content in the seawater exceeds 60,000 algal cells/L. Such algal bloom for example occurred in the Persian Gulf in the winter of 2008/2009, when the algal content in the seawater exceeded 1 million cells/L and continued for over 4 months. During this algal bloom event, which has a repeatability of once every 10 years, the total organic carbon in the seawater exceeded 8 mg/L and turbidity varied between 20 and 30 NTU. The severe algal bloom was caused by microalgae *Cochlodinium polykrikoides*. Their cells have sizes between 5 and 20 μm and are poorly removed by DAF clarifiers. This algal bloom caused the shutdown of all SWRO desalination plants in the Persian Gulf, which used membrane pretreatment and resulted in derating of the fresh water production capacity by 20%–30% of most of SWRO desalination plants with single-stage granular media filtration pretreatment for a period of 2–3 months.

Severe algal blooms do not occur only in the Middle East. Such events are observed frequently in Australia, Southern California, Chile, South Africa, and other parts of the world. The saline source water quality reflecting typical severe algal bloom events is shown in Table 11.5.

As with RO desalination plants with open intakes exposed to high-intensity algal blooms, pretreatment in this case also includes two-stage system. However, instead of dual media (anthracite and sand) filters with normal filtration bed depth of 1.4–1.6 m (4.6–5.3 ft)—trimedia filters (anthracite, sand and garnet) with a total depth of 2.0 m (6.6 ft) or more are recommended to be used to provide enhanced biofiltration and thereby to reduce down to

TABLE 11.5 Typical Source Water Quality of Plants With Intakes Exposed to Severe Algal Blooms

Source Water Quality Parameter	Value
Turbidity, NTU	5–40
$SDI_{2.5}$	16–30
Total organic carbon, mg/L	0.5–16.0
Total hydrocarbons, mg/L	<0.02
Total iron in reduced form, mg/L	<0.05
Total manganese in reduced form, mg/L	<0.02
Algae, cells/L	>60,000

SDI, silt density index.

acceptable levels the biofouling potential of the source water. The recommended pretreatment alternatives are as follows:

- Gravity granular trimedia filters with loading rate of 8–10 m^3/m^2 h (3–4 gpm/ft^2), pressure granular dual media filters with a loading rate of 14–16 m^3/m^2 h (6–7 gpm/ft^2), and cartridge filters;
- DAF clarifiers, single-stage gravity granular trimedia filters with a loading rate of 6–8 m^3/m^2 h (2.5–3.5 gpm/ft^2), and cartridge filters;
- Gravity granular trimedia filters with a loading rate of 10–12 m^3/m^2 h (4–5 gpm/ft^2) and pressure-driven UF/MF filters with a design flux of 60–70 Lmh (35–41 gfd);
- Gravity granular trimedia filters with a loading rate of 10–12 m^3/m^2 h (4–5 gpm/ft^2) and vacuum-driven UF/MF filters with a design flux of 45–55 Lmh (26–32 gfd);
- DAF clarifiers and pressure-driven UF/MF filters with a design flux of 40–50 Lmh (23–29 gfd);
- DAF clarifiers and vacuum-driven UF/MF filters with a design flux of 25–35 Lmh (15–20 gfd).

To date, the most widely used pretreatment system configurations for desalination plants with saline source waters experiencing severe algal blooms are (1) combination of DAF clarifiers and deep gravity media filters and (2) of DAF clarifiers and pressure driven UF or MF membrane filters.

11.2.6 Plants With Open Intake Exposed to High Hydrocarbon Concentrations

Natural saline source waters typically do not contain hydrocarbons, oil, or grease. These colloidal foulants originate from manmade activities and are observed in industrial or recreational ports, ship channels, along oil tanker routes, or near wastewater treatment plant discharges. Content of THC over 0.02 mg/L in the water fed to the RO membranes will cause their irreversible fouling and destruction. Therefore, plants exposed to high THC concentrations will have to be equipped with robust technologies for the removal of hydrocarbons.

The most popular pretreatment configuration at present is the use of two-stage system, the first stage of which includes DAF clarifiers and the second stage consists of granular media or MF/UF membrane filters. Such pretreatment can handle up to 1 mg/L of THC. If the source water contains higher THC levels, the desalination plant will need to be shut down until the intake area is decontaminated.

Another pretreatment alternative applied at the 300,000 m^3/day Adelaide desalination plant in Australia is feeding powdered activated carbon (PAC) into the source seawater upstream of the plant's disk filters for the duration of the oil spill event. PAC of grain size smaller than the microscreen openings is fed upstream of the microscreens in order to facilitate thorough mixing and adsorption of the THC by the PAC before its removal by the plant's vacuum-driven membrane filtration system. The key advantage of this pretreatment approach is its cost efficiency. The construction and O&M costs for the PAC feed system are less than 10% of the cost of construction and operation of DAF. Taking under consideration that the Adelaide desalination plant has a very deep intake (22 m/73 ft) and that the amount of THCs that could reach the intake at this depth is very small, the use of PAC instead of DAF is much more cost effective and provides equal protection of the RO membranes.

11.3 ADDITIONAL CONSIDERATIONS FOR SELECTION OF PRETREATMENT

Intake type and associated source water quality are the key technical factors used for the selection of pretreatment configuration of the desalination plant. However, depending on the site-specific circumstances, other nontechnical factors may influence the pretreatment selection process. Such factors include environmental project permitting, plant site footprint requirements, waste stream quantity and quality, coagulant dose, chemical costs, energy costs, and economy of scale (Jackangelo et al., 2017).

11.3.1 Permitting Issues

Environmental permitting for desalination plants is controlled by various governmental entities responsible for enforcement of the country's environmental regulations. Regulatory requirements associated with the desalination plant pretreatment system mainly focus on the discharge of the waste streams generated during the pretreatment process such as intake screenings, spent filter backwash, filter-to-waste water discharged during the maturation of the filtration cells, and spent MF/UF membrane cleaning chemicals. These requirements vary significantly worldwide (Mickley and Voutchkov, 2016). However, two key approaches used in the environmental regulations of many countries are to either allow direct discharge of the pretreatment waste streams along with the plant concentrate or to require treatment of the pretreatment waste streams by sedimentation prior to discharge of the clarified stream after blending with the desalination plant concentrate. In such cases, the solids separated from the pretreatment waste stream are dewatered by centrifuges or belt filter presses and are disposed as sludge to a landfill after their dewatering.

Numerous regulatory limitations for turbidity of the discharge and/or direct specific permitting requirements for treatment of the concentrate usually impact the selection of pretreatment. As indicated in Chapter 10, membrane pretreatment generates approximately two times less solids and therefore it allows for easier compliance with the turbidity limitations in the desalination plant discharge permit. In addition, use of membrane pretreatment does not discolor the discharge area and therefore often regulatory agencies allow spent filter backwash water from membrane pretreatment to be discharged directly without pretreatment. The latest regulations in most developed countries require that the backwash water from conventional granular media pretreatment systems using coagulant to be treated before discharge to remove the discoloration of the discharge and reduce the discharge turbidity.

The 2015 California Ocean Plan, in the United States, has requirements for reduction of impingement and entrainment of marine organisms and, therefore, it encourages the use of subsurface intake systems (wells) instead of open intakes. Such regulations have a significant impact on the selection of pretreatment system because if intake wells are selected, they would likely result in more simplified pretreatment system that consists of cartridge filters only or combination of cartridge filters and pressure-driven granular filtration system. It should be pointed out, however, that some subsurface intakes may yield source waters of high content of iron and/or manganese in a reduced form, which would require a special type of granular media pretreatment filters (e.g., greensand filters) or conservatively designed oxidation and membrane pretreatment system to remove the iron and manganese (typically to below 0.1 mg/L) so that these compounds do not cause fouling of the downstream RO membranes.

11.3.2 Impact of Project Procurement on Pretreatment System Selection

Practical experience to date shows that the type of project procurement—design–bid–build (DBB), design–build (DB), or design–build–operate (DBO)—has significant impact on the selection of a pretreatment system. When the designer and contractor are not responsible for the plant operation (e.g., under DBB or DB methods of project delivery), the designer may prefer to select the lowest capital cost granular media pressure-driven pretreatment system or to choose the use of membrane pretreatment with a very aggressive design flux.

Since performance guarantees and acceptance testing typically last only 1 month, the lowest-cost contractor with aggressive pretreatment system design could arrange the plant testing to occur during nonalgal bloom periods when the plant is not exposed to challenging performance and to pass the acceptance testing. This procurement approach is to the detriment of the project owner and operator, who will face project performance constraints during the period of challenging source water conditions (e.g., algal bloom season, monsoon season).

Another approach for project procurement is when the same contractor is responsible for design, construction, and operation of the plant. Under such procurement conditions (e.g., DBO), the designer may select a robust and conservative pretreatment system, which is capable of operating reliably and cost effectively under all conditions, including challenging water quality events such as algal blooms.

References

Jackangelo, J., Voutchkov, N., Badruzzaman, M., Wienrich, L.A., 2017. Pretreatment for Seawater Reverse Osmosis: Existing Plant Performance and Selection Guidance, Water Environment and Reuse Foundation, WRRF-14–07.

Mickley, M., Voutchkov, N., 2016. Database of Permitting Practices for Seawater Concentrate Disposal, Water Environment and Reuse Foundation, WRRF-13–07.

Voutchkov, N., 2010. Considerations for selection of seawater desalination system. Desalination 261, 354–364. Elsevier.

CHAPTER

12

Reverse Osmosis System Design and Pretreatment

OUTLINE

Pretreatment for Reverse Osmosis Desalination

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12.1 OVERVIEW OF TYPICAL SWRO DESALINATION SYSTEM

Fig. 12.1 depicts typical configuration of large SWRO system. Filtered water that is produced by the desalination plant's pretreatment system is conveyed by transfer pumps from a filtrate water storage tank through cartridge filters and into the suction pipe of the high-pressure RO feed pumps. As indicated in Chapter 5, the main purpose of the cartridge filters is to protect the RO membranes from damage. Cartridge filters are typically used only for desalination plants with granular media filtration systems. Pretreatment systems employing MF or UF membranes typically do not have cartridge filters.

The high-pressure feed pumps are designed to deliver the pretreated water to the RO membranes at pressures required for membrane separation of the freshwater from the salts, which typically are 55–85 bars. The actual required feed pressure is site-specific and is mainly determined by the source water salinity, temperature, permeability of the selected SWRO membranes, and the configuration of the RO system.

The RO-membrane elements are installed in pressure vessels, which usually house 6–8 elements per vessel (see Fig. 12.2). Multiple pressure vessels are arranged on support structures (racks), which form RO trains. Each RO train is typically designed to produce between 10% and 20% of the total amount of the membrane desalination product water flow. Fig. 12.1 depicts one RO train with its key components—filter effluent transfer pump, cartridge filter, high-pressure pump, and membrane rack with vessels.

12.2 SWRO MEMBRANE ELEMENTS—KEY TYPES AND PRETREATMENT CONSIDERATIONS

The "engine" of every SWRO plant is the RO-membrane element. The two types of materials used to manufacture SWRO membrane elements today are (1) polyamide, and (2) cellulose acetate and its derivatives. RO membranes made of polyamide are also referred to as

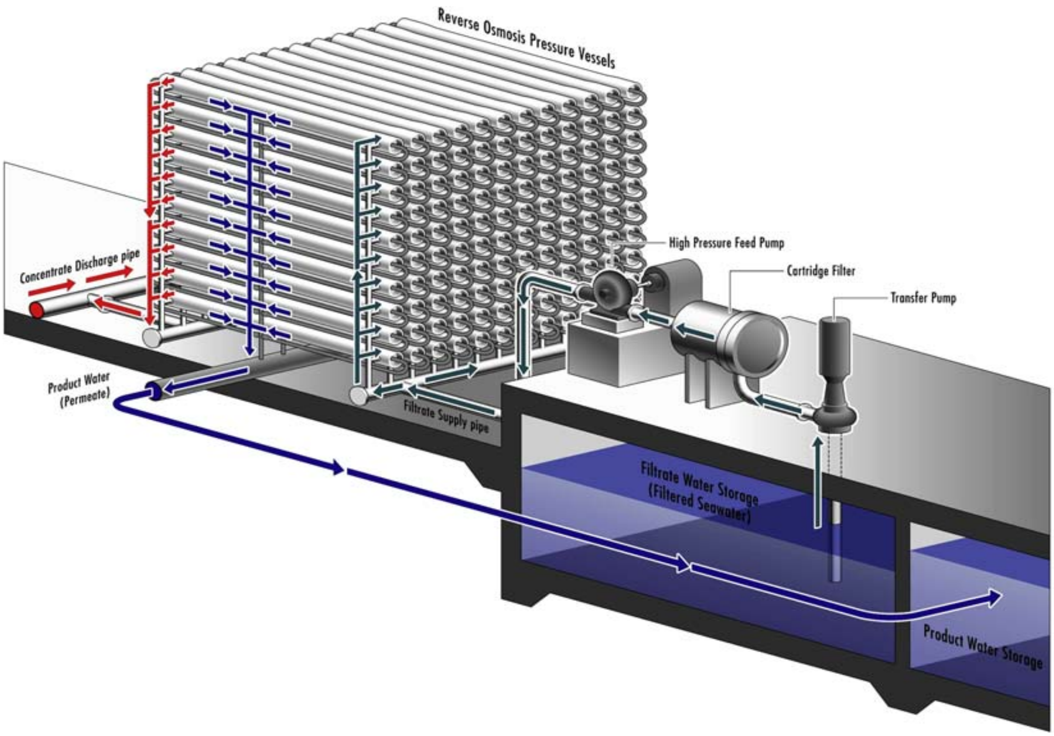

FIGURE 12.1 RO-membrane system configuration.

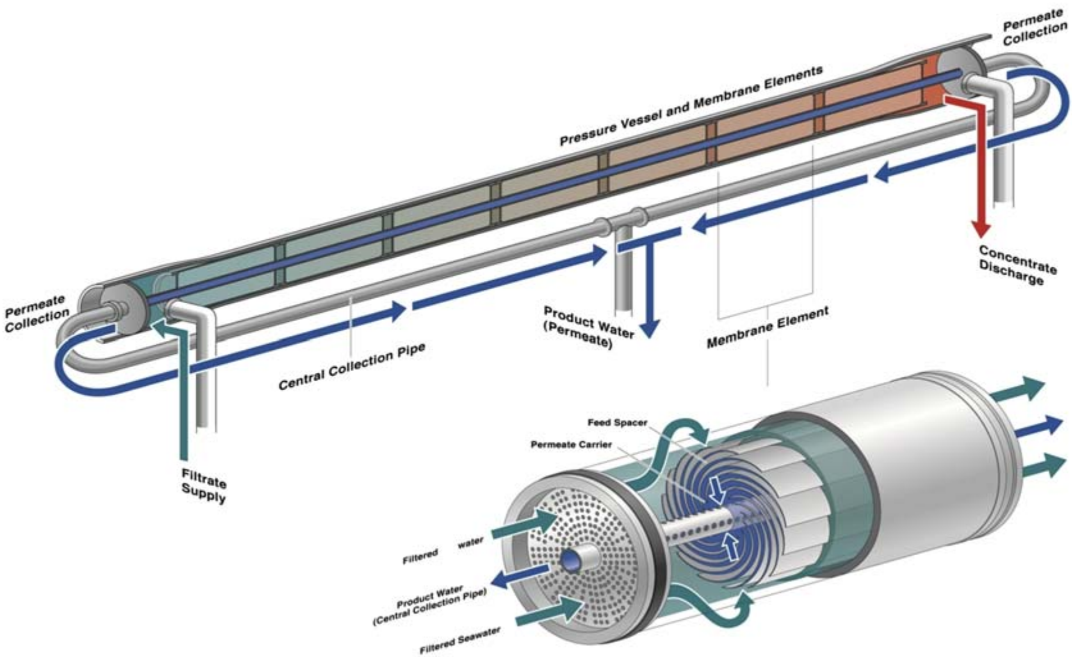

FIGURE 12.2 Spiral-wound RO-membrane element and vessel.

TABLE 12.1 Comparison of Thin-Film Composite and Cellulose Acetate SWRO Membranes

Parameter	Thin-Film Composite Membranes	Cellulose Acetate Membranes
Salt rejection, %	Higher (>99.5%)	Lower (up to 95%)
Net driving pressure, bars	Lower (10–15 bars)	Higher (15–30 bars)
Surface charge	Anionic (limits use of cationic pretreatment coagulants)	Neutral (no limitations on pretreatment coagulants)
Chlorine tolerance	Poor (up to 1000 ppm-h) feed dechlorination needed	Good continuous feed of 1–2 ppm of chlorine is acceptable
Cleaning frequency	High (weeks to months)	Lower (months to years)
Pretreatment requirements	High (SDI <4)	Lower (SDI <5)
Organics removal	High	Relatively lower
Biogrowth on membrane surface	May cause performance problems	Limited—not a cause of performance problems
pH tolerance	High (1–13)	Limited (4–6)

thin-film composite membranes. Key advantages and disadvantages of the thin-film composite and cellulose acetate SWRO membranes in terms of their sensitivity to seawater pretreatment are presented in Table 12.1.

Mainly because of their higher membrane rejection and lower operating pressures, the thin-film composite membranes are the preferred choice for most SWRO-membrane installations today. Exceptions are applications in the Middle East, where the source water is rich in organics, and the cellulose acetate membranes offer significant benefits in terms of reduced membrane biofouling, cleaning, and pretreatment needs.

The most widely used thin-film composite SWRO-membrane elements consist of two membrane sheets glued together at their three sides and opened at the fourth side, similar to an envelope and, therefore, they are also referred to as membrane envelopes. The opened side of the membrane envelopes is attached to the central permeate collection tube with small openings that collect the permeate from the membrane envelopes. All membrane envelopes are spirally wound around the central permeate tube (Fig. 12.2). The outside surface of the membrane envelopes, which retains the source water minerals, has microscopic molecular porous structure that can reject particulate and dissolved solids, microorganisms, and other impurities of size smaller than 200 Da. This polyamide film on the surface of the membrane envelopes that actually separates the salts from the water is usually less than 0.2-μm thin, and in order to withstand the high pressure required for salt separation, it is supported by a second thicker membrane layer, which is typically made of higher-porosity polysulfone material and has several orders of magnitude larger membrane openings.

The membrane envelopes are separated by approximately 0.71-mm (28-mil) thick feed spacer, which forms feed channels and facilitates the conveyance of the feed-concentrate stream along the length of the membrane elements. Some RO-membrane elements have wider distance between the envelopes (0.87-mm/34-mil). The wider the feed channel between the envelopes, the more foulants it could accumulate before the membranes need cleaning. Therefore, the RO elements with wider (34-mil) spacers are usually preferred for saline source waters with high fouling potential and less conservatively designed pretreatment filtration systems.

The most widely used commercially available SWRO-membrane elements have diameter of 20 cm (8-in.), length of 100 cm (40-in.), and produce 11−14 m^3/day of permeate. As shown in Fig. 12.2, the SWRO membrane elements are connected in series inside the pressure vessel. Typically, one pressure vessel houses from six to eight SWRO-membrane elements. Most RO-membrane systems are designed with seven elements per vessel.

A recent design trend for RO feed waters that have very low turbidity (<0.1 NTU), silt content ($SDI_{15} < 3$), and organic concentration (TOC < 1.0 mg/L) at all times is to use eight elements per vessel. Such high RO feed-water quality can be achieved either if the source water originates from well-designed subsurface intake or if it is collected by open intake and is pretreated via conservatively designed single or two-stage filtration system. Typically, the higher the source seawater total dissolved salts (TDS) and fouling potential, and the lower the design membrane flux the fewer elements are used per vessel. SWRO systems designed around lower salinity feed water (e.g., TDS < 35 ppt), with low fouling potential are suitable for eight membrane elements per vessel configuration. Such SWRO systems can be designed for relatively high average permeate flux of 15 lmh/9 gfd or more.

In the majority of existing seawater RO installations, the entire pretreated seawater flow is fed at the front end of the membrane vessel and collected at the rear end. However, most recent RO-system designs incorporate permeate flow collection from both front and back ends of the membrane vessel as shown in Fig. 12.2. Removal of some of the permeate from the front end is practiced when the desired concentrations of TDS, chloride, and boron in the desalinated water are targeted to be below 250, 100, and 0.75 mg/L, respectively, because the permeate collected from the front end of the membrane vessel (usually the first two to three elements) already meets these water quality targets.

If left in the RO-membrane vessel, the high-quality permeate generated from the front-end membrane elements will blend with the permeate form the remaining membranes in the central collection pipe of the RO vessel, whose permeate has inferior water quality, and then the entire permeate volume produced in the RO vessel will need to be retreated in a second RO pass to meet the same product water target. The ability to collect permeate from both ends also allows to achieve a better control over the water quality of permeate produced by the desalination system and, to some extend, to control the fouling of the RO system membranes.

Commercially available membrane RO elements are of standardized diameters and length, and salt rejection efficiency. Standard membrane elements have limitations with respect to a number of performance parameters such as: feed water temperature (45°C/104°F); pH (minimum of 2 and maximum of 12); silt density index (less than 4); chlorine content (not tolerant to chlorine in measurable amounts); and feed water pressure (maximum of 83−100 bars). Some producers offer 16- and 18-in. spiral-wound SWRO- and BWRO-membrane elements but they have found very limited application in large-scale installations due to their relatively higher costs and elevated fouling propensity.

The ratio between the volume of the product water produced by the membrane desalination system and the volume of the source water used for its production is commonly defined as recovery and is presented in percent of the plant RO-system feed water volume.

The maximum recovery that can be achieved by a given pressure-driven membrane desalination system mainly depends on the source water salinity and is limited by the magnitude of the osmotic pressure to be overcome by the RO system high-pressure feed pumps, and by the fouling and scaling potential of the source seawater. It should be pointed out that higher the salinity and fouling potential of the saline source water, the lower the target design recovery should be. For example, relatively lower (33–35 ppt) salinity of the Pacific Ocean allows the SWRO systems treating such water to be designed for recovery of 50%, if the seawater has relatively low fouling potential. For the Red Sea and Persian Gulf seawaters, which have salinities of 42–46 ppt, the sustainable design SWRO system recovery is 40%–45%. For desalination plants in the Middle East exposed to severe algal blooms, the sustainable design recovery is even lower—36%–38%. Higher recoveries (40%–45%) could be achieved, however, if the desalination plant is equipped with more robust pretreatment as that recommended in Chapter 11 for such seawaters.

As indicated in previously, scaling occurs when the minerals left behind on the rejection side of the RO membrane are concentrated to a level at which they begin to form precipitates (crystalline compounds) which in turn plug the membrane surface and interfere with freshwater transport through the membrane. Typically, seawater desalination plants can only turn 40%–60% of the source water into low-salinity permeate.

Membrane performance tends to naturally deteriorate over time due to combination of material wear-and-tear and irreversible fouling of the membrane elements. Typically, membrane elements have to be replaced every 5–7 years to maintain their performance in terms of water quality and power demand for salt separation.

Improvements of membrane element polymer chemistry and production process have made the membranes more durable and have extended their useful life. Use of conservatively designed dual or tri-media granular filtration systems and MF or UF-filtration pretreatment prior to RO desalination would allow to extend the membrane useful life beyond 7 years and further beyond. On the other hand, practical experience shows that MF- and UF-membrane pretreatment systems designed for very high fluxes (80 lmh or more) usually result in accelerated aging of the downstream SWRO membrane elements and reduction of their useful life below 3 years.

12.2.1 Classification of Thin-Film Composite SWRO Membranes

As indicated previously, 8-in. thin-film composite SWRO membranes are the most widely used types of membranes at present. The three key performance parameters of all RO membranes are: salt rejection, flux/productivity, and operating pressure. Currently, there are a number of commercially available SWRO membrane elements designed with special features allowing to optimize their performance around one or more of these three key performance parameters. Commercially available RO and NF elements at present can be classified in the following key groups:

1. Standard Rejection;
2. High Rejection;
3. High Productivity (or Low-Energy); and
4. High Pressure.

12.2.2 Standard Rejection SWRO Membrane Elements

Standard rejection membrane elements are designed to remove up to 99.6% of the salts from the source seawater. These membrane elements are most widely used at present and have found applications in variety of RO-system configurations. Such SWRO membrane elements have spacer of 28-mil and are suitable for seawater with low- and moderate fouling potential and well-deigned pretreatment system (see Chapter 11 for design guidelines).

12.2.3 High-Rejection SWRO Membrane Elements

High-rejection membrane elements are designed with tighter membrane structure allowing to increase the mass of rejected ions, and to better reject smaller size ions, such as boron for example. The higher rejection membrane capabilities of 99.75%–99.85% come at a price—10%–20% higher operating pressure. In general, these membrane elements are also more prone to fouling as compared to standard rejection SWRO membrane elements and their use requires more elaborate seawater pretreatment in terms of particulate, colloidal, and microbial foulants.

12.2.4 High-Productivity (Low-Energy) Membrane Elements

High-productivity membrane elements are designed with features to yield more product water per membrane element. These features are: (1) higher surface area and (2) denser membrane packing. Increasing active membrane envelope surface area allows gaining significant productivity for the same size (diameter) membrane element. Higher productivity of the membrane elements is obtained by using higher permeability membranes with wider molecular channels. Therefore, the salt rejection of these elements is lower than that of standard and high rejection SWRO membranes. The high productivity elements have a standard yield of 9000–12,000 gallons per day (gpd) and salt rejection of 99.2%–99.4%.

Increasing membrane size/diameter can also increase the total active surface area of a membrane element. Although 20-cm (8-in.) SWRO membrane elements are still the "standard" size and are most widely used in large full-scale applications, larger 16- and 18-in. size SWRO-membrane elements are currently available.

Another alternative for improving membrane productivity is increasing the number of membrane leafs packed into the same size (diameter) membrane. This is accomplished either by the use of thinner feed channel spacers, or by improving the element construction. Using thinner feed spacers typically increases the membrane pressure drop. As a result, higher productivity membrane elements using this approach also have higher operational pressure requirements for the same salt rejection level and flux.

Denser membrane-leaf packing makes membranes also more susceptible to fouling and their use requires high-quality source water and more elaborate pretreatment. To address this issue, the newest high-productivity membrane elements actually use wider spacer to compensate for the increased fouling potential and pressure.

The dynamics of the high-productivity (or low energy) membrane element development is illustrated by an example of the development of seawater membranes. In the second half of 1990s the typical 20-cm (8-in.) SWRO-membrane element had a standard productivity of

5000–6000 gpd at a salt rejection of 99.6%. In 2003, several membrane manufacturers introduced high-productivity seawater membrane elements that are capable of producing 7500 gpd at salt rejection of 99.75%. Just 1 year later, even higher productivity (9000 gpd at 99.7% rejection) seawater membrane elements were released on the market.

Some of the newest high-productivity SWRO-membrane elements have unit production capacity of 12,000–16,000 gpd, provide flexibility and choice, and allow to trade productivity and pressure/power costs. The same water product quality goals can be achieved either by (1) reducing the system footprint/construction costs by designing the system at higher productivity or by (2) reducing the system overall power demand by using more membrane elements, designing the system at lower flux and recovery, and taking advantage of newest energy-recovery technologies that further minimize energy use if the system is operated at lower (35%–45%) recovery.

12.2.5 High-Pressure SWRO Elements

The main purpose of this type of SWRO elements is to produce freshwater from concentrated seawater with salinity of 50,000–60,000 mg/L and are used to maximize water recovery form a given source water volume. While a standard membrane element can only allow to recover up to 50% of the source seawater, the high-pressure SWRO are suitable to obtain recoveries of 60% and higher.

The high-pressure membrane elements are specifically designed to operate at 20%–40% higher pressure than that of the other types of membrane elements listed above and to treat high-salinity concentrate produced by the first stage of a two-stage SWRO system. While these high-pressure elements are commercially available from a number of manufacturers, they have not found widespread use, because they also typically have a higher fouling rate and elevated cleaning costs as well as reduced useful life. It should be pointed out that higher the feed pressure of the RO elements, the higher the fouling rate, and the faster the increase of transmembrane pressure for the same saline source water.

12.3 INTERNALLY STAGED MEMBRANE CONFIGURATION—FOULING IMPLICATIONS

Ideally, redistributing and evening out the feed pressure and flux of all seven RO elements in the vessel to near-equal level can achieve the most energy-efficient desalination process with lowest fouling within the RO vessels. A widely used membrane configuration design allows to achieve such more even flux distribution by combining three different models of membranes with different permeability within the same vessel instead of using the same model of RO elements throughout the vessel (which is a typical configuration for conventional SWRO systems). This membrane vessel configuration was developed by Dow Filmtec (Mickols et al., 2005) and is known as Inter-Stage Design (ISD) (Fig. 12.3).

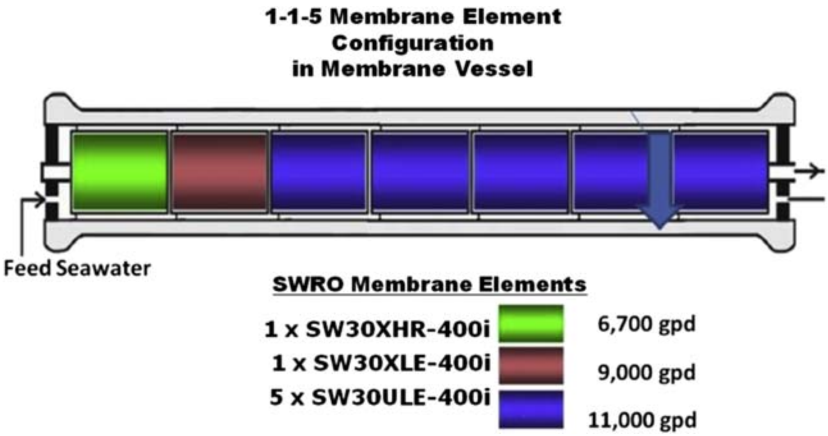

FIGURE 12.3 Internally staged membrane configuration.

The first (lead) element (Fig. 12.3), which receives the entire seawater feed flow of the vessel, is a low-permeability/high salt-rejection element (Dow Filmtec SW30 XHR-400*i*). Because of its low permeability, this element produces only 10%–14% (instead of 25%) of the permeate flow produced by the entire vessel and, thereby, preserving the feed energy for more effective separation by the downstream RO-membrane elements in the vessel.

The second RO element in the pressure vessel is of a standard (average) rejection permeability and salt rejection (Dow Filmtec SW30 XLE-400*i*), and produces approximately 14%–16% of the total flow, while the remaining five elements in the vessel are of the same high-permeability/low-rejection model (Dow Filmtec SW30 ULE-400*i*). This 1–1–5 combination of low-permeability/high-rejection and high-permeability/low-rejection elements results in a more even distribution of flux and pressure along the vessel typically, which yields 5%–10% energy savings and reduces the fouling rate of all membrane elements.

The more even distribution of flux achieved by the hybrid membrane element configuration not only results in lower overall energy use but also reduces the fouling of the first element as compared to standard configuration with all of the RO membranes in the vessels being of the same model. Therefore, the use of inter-stage membrane configuration would impose less stringent requirements on the upstream pretreatment system as compared to conventional RO membrane configuration using the same model of elements in all positions within the vessels.

12.4 ALTERNATIVE SWRO-MEMBRANE SYSTEMS AND PRETREATMENT

The SWRO-system configurations that are most widely applied at present include: single-pass treatment, where the source water is processed by reverse osmosis only once (Fig. 12.4); and two-pass RO treatment, where the seawater is first processed through a SWRO system and then permeate produced by this system is reprocessed by brackish RO membranes (see Figs. 12.5 and 12.6).

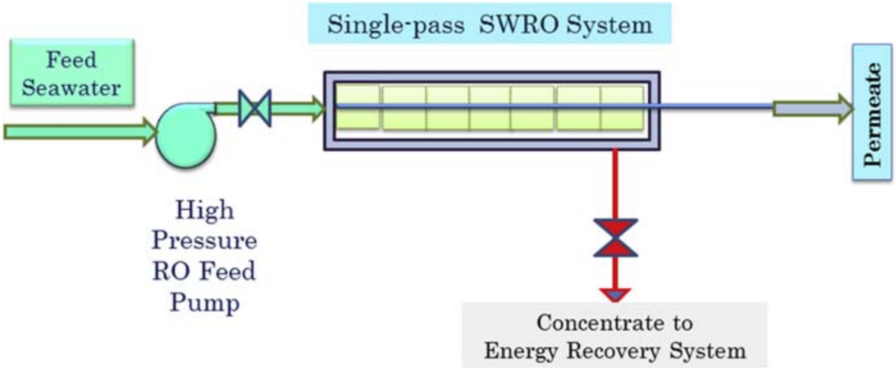

FIGURE 12.4 Single-pass SWRO system.

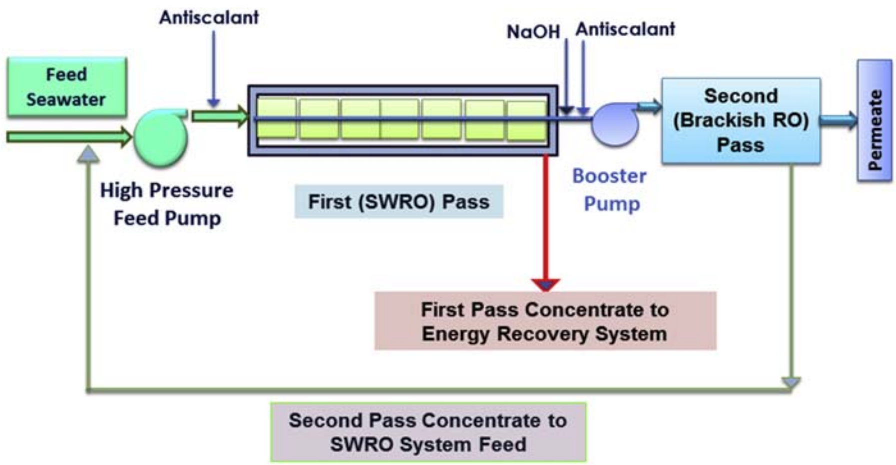

FIGURE 12.5 Conventional full two-pass SWRO system.

12.4.1 Single-Pass SWRO Systems

Single-stage SWRO systems (Fig. 12.4) are designed to produce desalinated seawater (permeate) in one step using only a single set of RO trains operating in parallel. In general, between 800 and 900 SWRO-membrane elements installed in 100–150 vessels are needed to produce 10,000 m^3/day (2.6 MGD) of permeate suitable for potable use in a single-stage SWRO system.

Under a typical single-stage SWRO-system configuration, each RO train has a dedicated system of transfer pump for pretreated seawater followed by a high-pressure RO feed pump. The high-pressure feed pump motor/operation is coupled with that of energy-recovery equipment.

Single-stage SWRO systems are widely used for production of drinking water. However, these systems have found limited industrial application mainly because of the water-quality limitations of the produced permeate. Even if using the highest-rejection RO-membrane elements commercially available today (nominal minimum rejection of 99.85%), the single-stage

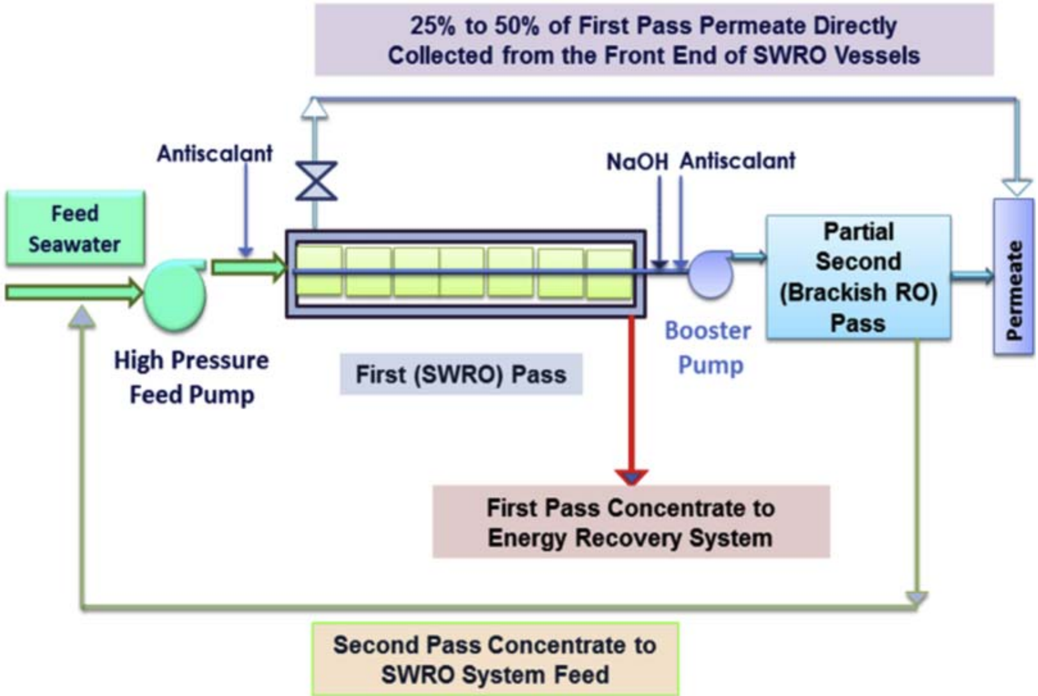

FIGURE 12.6 Split-partial two-pass SWRO system.

SWRO desalination systems typically cannot consistently yield permeate with TDS concentration lower than 200 mg/L, chloride level of less than 100 mg/L, and boron concentration lower than 0.5 mg/L, especially when source water temperatures exceed 18–20°C (64–68°F).

If enhanced boron removal is needed in such systems, high boron rejection membranes are used, and/or sodium hydroxide and antiscalant is added to the RO system feed water to increase pH to 8.8 or more, which in turn improves boron rejection.

Single-pass RO systems require robust upstream pretreatment because they produce desalinated water in single step and do not have the ability to reprocess the permeate if their water quality is impacted by accelerated RO-membrane fouling caused by poor pretreatment system performance. It is important to note that inadequate removal of foulants by the pretreatment system not only impacts negatively the RO-system production capacity but also results in deterioration of membrane rejection and permeate water quality because fouling exacerbates the concentrate polarization effect on the membrane surface, which in turn increases salt transfer through the membranes.

12.4.2 Two-Pass SWRO Systems

Two-pass SWRO systems are typically used when either the source seawater salinity is relatively high (i.e., higher than 35,000 mg/L) and/or the product water-quality requirements are very stringent. For example, if high salinity/high-temperature source water (such as Red Sea and Persian Gulf seawater for example) is used in combination with standard-rejection (99.6%) SWRO membranes, then single-stage SWRO systems may not be able to produce permeate suitable for drinking water use. In this case, two-pass SWRO systems have proven to be a very efficient and cost-effective configuration for potable water production. RO systems with two or more passes are also widely used for production of high-purity industrial water.

The two-pass SWRO systems typically consist of a combination of a single-pass SWRO system and a single or multiple-pass brackish water RO (BWRO) system connected in series (Figs. 12.5 and 12.6). Permeate from the SWRO system (i.e., first pass) is directed for further treatment to the BWRO system (i.e., second pass) to produce a finished water of low mineral content. The concentrate from the pass-two BWRO system is returned to the feed of the first-pass SWRO system to maximize the overall desalination system production capacity and efficiency. Two-pass SWRO systems are classified in two main groups: conventional full two-pass systems and split-partial two-pass systems.

12.4.2.1 Conventional Full-Two Pass SWRO Systems

In conventional full two-pass SWRO membrane systems (see Fig. 12.5), the source seawater is first treated by a set of SWRO membrane trains (referred to as first RO pass) and then the entire volume of desalinated water from the first pass is processed through a second set of brackish water desalination membrane trains (Greenlee et al., 2009). If enhanced boron removal is needed, sodium hydroxide and antiscalant are added to the feed water to the second RO pass to increase pH and improve boron rejection.

Conventional two-pass RO systems are usually exposed to accelerated fouling during severe and moderate algal blooms and the most common impact of inadequate pretreatment is the reduction of RO-system recovery and productivity due to rapid increase in transmembrane pressure and differential pressure.

12.4.2.2 Split-Partial Two-Pass SWRO Systems

In split-partial two-pass systems the second RO pass typically processes only a portion (50%–75%) of permeate generated by the first pass. The rest of the low salinity permeate is produced by the front (feed) SWRO elements of the first pass. This low-salinity permeate is collected from the front end of the permeate tube, and without additional desalination, it is directly blended with permeate produced by the second RO pass (see Fig. 12.6).

As indicated in Fig. 12.6, the second-pass concentrate is returned to the feed of the first-RO-system pass. When the desalination system is designed for enhanced boron removal, this concentrate has pH of 9.5–11 and potentially could cause precipitation of calcium carbonate on the membranes (e.g., mineral scaling). In order to avoid this challenge, typically antiscalant is added to the feed to the second pass (brackish RO) system. Long-term experience with such configuration indicates that this solution is very effective in preventing scaling of the first-pass RO membranes by the recycled second-pass concentrate. Because boron level in the front permeate stream is usually between 0.25 and 0.50 mg/L, no additional treatment of this stream is needed even if the plant is designed and operated for enhanced boron removal.

While the recycling of the second-pass concentrate returns a small portion of the source water salinity and, therefore, it slightly increases the salinity of the seawater fed to the first RO pass, the energy-use associated with this incremental salinity increase is significantly smaller than the energy savings of processing the entire volume of the first-pass permeate through the second pass. Under the split partial two pass RO system configuration the volume of permeate pumped to the second RO-pass and the size of this pass are typically 25%–50% smaller than the volume pumped to the second RO-pass under conventional once-trough operation. Since pumping energy is directly proportional to flow, the energy costs for the second-pass feed pumps are reduced proportionally, i.e., with 25%–50%. For SWRO system operating at 45% recovery, such savings will amount to 14%–22% of the energy of the first-pass RO pump.

The concentrate returned from the second pass carries only 1%–2% of additional salinity to the first-pass RO feed, which reduces the energy benefit from such recovery proportionally, i.e., by 1%–2% only. As a result, the overall energy savings of the use of split partial two-pass RO system as compared to conventional two-pass RO system are between 12% and 20%. Practical experience with large SWRO desalination plants indicates that the average total RO system life-cycle cost savings associated with applying such SWRO system configuration are typically between 14% and 16%.

At present, most new SWRO desalination systems are designed with split partial second pass configuration because this configuration allows reducing the size of the second-pass RO system and the overall freshwater production costs. It should be pointed out that split-partial second RO-pass systems can be configured in several alternatives, which may involve the use of the same or different type of membrane elements within the first-pass SWRO system.

Example, of plant with partial-second pass configuration is the 95,000 m^3/day (25 MGD) Tampa Bay seawater desalination facility in Florida. The second pass at this facility is designed to treat up to 30% of the permeate produced by the first-pass SWRO system as needed in order to maintain the concentration of chlorides in the plant-product water always below 100 mg/L.

The partial second pass at the Tampa Bay water seawater desalination plant was installed to provide operational flexibility and to accommodate the wide fluctuations of source water salinity (16–32 ppt) and temperature (18–40°C/64–104°F). Typically, the product water quality target chloride concentration of 100 mg/L at this plant is achieved by only operating the first pass of the system. However, when source water TDS concentration exceeds 28 ppt and/or the source water temperature exceeds 35°C (95°F), the second pass is activated to maintain adequate product water quality. The percent of first-pass permeate directed for additional treatment through the second pass is a function of the actual combination of source water TDS and temperature and is adjusted based on the plant product water chloride level.

Conventional full two-pass RO systems and split partial second-pass systems are equally vulnerable to accelerated fouling due to underperforming pretreatment especially during periods of high-intensity and severe algal blooms. However, the product water quality of partial second-pass RO systems is usually more affected by algal bloom events than that of full two-pass systems, because these systems have more flexibility to handle the permeate water quality impacts from fouling.

12.4.2.3 Product Water Quality of Single- and Two-Pass SWRO Systems

Tables 12.2 through 12.6 present summary of the range of permeate water quality produced by typical single-pass and partial two-pass seawater desalination systems processing

TABLE 12.2 Reverse Osmosis Permeate-Water Quality, Seawater Source—Pacific Ocean

Water-Quality Parameter	Pacific Ocean Source Seawater Quality	Permeate Water Quality	
		Single-Pass SWRO System	Split Partial Two-Pass RO System
Temperature, °C	14–28	15–29	16–30
pH	8.0	6.3–7.2	7.6–7.8
Ca^{2+}, mg/L	358	0.6–1.1	0.2–0.5
Mg^{2+}, mg/L	1,720	1.8–2.8	0.07–0.10
Na^{+}, mg/L	9,900	78–134	9–20
K^{+}, mg/L	600	3.0–6.0	0.43–0.60
$CO_3{}^{2-}$, mg/L	2.0	0.0	0.0
$HCO_3{}^{-}$, mg/L	170	1.8–2.5	0.4–0.7
$SO_4{}^{2-}$, mg/L	2,570	2.6–5.3	0.7–1.3
Cl^{-}, mg/L	18,100	130–195	13–20
F^{-}, mg/L	2.1	0.9–1.2	0.7–0.9
$NO_3{}^{-}$, mg/L	0.00	0.00	0.00
B^{-}, mg/L	4.5	0.7–1.2	0.3–0.5
Br^{-}, mg/L	73	0.6–0.9	0.2–0.4
TDS, mg/L	33,500	220–350	25–45

five different seawater sources: Pacific Ocean water; Atlantic Ocean water; Mediterranean seawater; Persian Gulf seawater; and Red Sea water.

Comparative analysis of the water-quality information presented in Tables 12.2–12.6 indicates that split partial two-pass RO systems usually produce significantly better water quality than single-pass systems, especially in terms of TDS, sodium, chloride, and boron. It should be pointed out that permeate water quality of split-partial second-pass systems is less sensitive to changes in the fouling potential of the source seawater or diminished performance efficiency of the upstream pretreatment system.

12.4.3 Two-Stage SWRO Systems

Two-stage SWRO membrane systems are mainly used to maximize the overall desalination plant recovery and reduce the volume of concentrate discharged by the desalination plant. A general schematic of a two-stage RO system is shown in Fig. 12.7. In these SWRO systems, typically the entire volume of the concentrate generated by the first-stage SWRO system is directed to a second-stage SWRO system for further treatment and enhanced recovery. Permeate from both systems is blended prior to final use.

TABLE 12.3 Reverse Osmosis Permeate-Water Quality, Seawater Source—Atlantic Ocean

		Permeate Water Quality	
Water-Quality Parameter	**Atlantic Ocean-Source Seawater Quality**	**Single-Pass SWRO System**	**Split Partial Two-Pass RO System**
Temperature, °C	16–30	17–31	18–32
pH	8.0	6.3–7.2	7.7–8.0
Ca^{2+}, mg/L	410	0.7–1.4	0.3–0.5
Mg^{2+}, mg/L	1,302	1.6–2.4	0.35–0.8
Na^{+}, mg/L	10,506	83–160	11–25
K^{+}, mg/L	390	2.4–4.5	0.35–0.50
CO_3^{2-}, mg/L	2.0	0.0	0.0
HCO_3^{-}, mg/L	145	1.4–2.0	0.5–0.8
SO_4^{2-}, mg/L	2,720	2.4–5.8	1.1–1.2
Cl^{-}, mg/L	19,440	146–220	20–34
F^{-}, mg/L	2.5	1.0–1.6	0.8–1.2
NO_3^{-}, mg/L	0.00	0.00	0.00
B^{-}, mg/L	4.5	0.7–1.2	0.3–0.5
Br^{-}, mg/L	78	0.8–1.1	0.3–0.5
TDS, mg/L	35,000	240–400	35–65

The main advantage of such SWRO system configuration is that it allows achieving very high level of use (recovery) of the available source seawater and the energy used by the first-stage RO system. For example, while a single-stage SWRO system configuration typically allows turning 35%–50% of the source seawater to potable water, the two-stage SWRO system recovery may reach 60%–65%. Designing the SWRO plant around higher recovery allows minimizing the size of the plant intake and pretreatment facilities, and the capital expenditures for their construction and operation.

However, because of the high salinity of the first-stage concentrate (typically above 55,000 mg/L), the practical implementation of two-stage SWRO systems requires use of high-pressure SWRO membranes, membrane vessels, piping, and auxiliary equipment that can withstand and perform well at very high pressures (up to 98 bars/1400 psi). The cost of this equipment is usually higher than the cost of the same size and type of equipment built to operate at more "standard" pressures (i.e., below 70 bars/1000 psi). Therefore, the viability of using two-stage SWRO system has to be carefully assessed based on a comprehensive life-cycle cost analysis for the site specific conditions of a given project. To date, two-stage

TABLE 12.4 Reverse Osmosis Permeate-Water Quality, Seawater Source—Mediterranean Sea

Water-Quality Parameter	Mediterranean Source-Seawater Quality	Permeate Water Quality: Single-Pass SWRO System	Permeate Water Quality: Split Partial Two-Pass RO System
Temperature, °C	16–28	17–29	18–30
pH	8.1	6.3–7.2	7.9–8.1
Ca^{2+}, mg/L	480	1.0–2.0	0.35–0.45
Mg^{2+}, mg/L	1,558	1.9–2.8	0.5–1.0
Na^{+}, mg/L	12,200	98–196	15–34
K^{+}, mg/L	480	3.0–5.5	0.8–1.8
CO_3^{2-}, mg/L	5.6	0.0	0.0
HCO_3^{-}, mg/L	160	1.7–2.4	0.5–0.8
SO_4^{2-}, mg/L	3,190	2.9–6.3	1.4–2.95
Cl^{-}, mg/L	22,340	169–260	25–52
F^{-}, mg/L	1.4	0.7–1.1	0.5–0.8
NO_3^{-}, mg/L	0.00	0.00	0.00
B^{-}, mg/L	5.0	0.9–1.5	0.4–0.6
Br^{-}, mg/L	80	0.9–1.3	0.35–0.6
TDS, mg/L	40,500	280–480	45–95

systems have been mainly used to upgrade the capacity and improve energy use of existing conventional single-stage SWRO plants.

Sometimes, two-stage SWRO systems are designed to be operated with lower feed pressure to the first stage, because of the highly fouling nature of the feed water. Since operation at lower feed pressure allows to reduce fouling of the first-stage/pass elements—this configuration could be beneficial if the pretreatment system is not very robust and the water quality from the SWRO system is of high alluvial organic or particular content.

12.4.5 Hybrid SWRO Systems With Multiple Passes and Stages

The two-pass and two-stage RO system configurations may be combined to achieve an optimum plant design and tailor desalination plant operation to the site-specific water-source water quality conditions and product water-quality goals.

An example of a full-scale two-pass/two-stage SWRO system application is the 170,500 m^3/day Fujairah seawater desalination plant (Sanz et al., 2007). A general treatment process schematic of this plant is depicted in Fig. 12.8. The plant uses Gulf of Oman source seawater.

TABLE 12.5 Reverse Osmosis Permeate Water Quality Seawater Source — Red Sea

Water-Quality Parameter	Red Sea Source Seawater Quality	Permeate Water Quality: Single Pass SWRO System	Permeate Water Quality: Split Partial Two Pass RO System
Temperature, °C	22–33	23–34	24–35
pH	7.0–8.0	6.8–7.8	7.6–8.0
Ca^{2+}, mg/L	500	1.1–2.1	0.5–0.7
Mg^{2+}, mg/L	1,540	1.8–3.4	0.7–1.0
Na^{+}, mg/L	13,300	142–220	20–38
K^{+}, mg/L	489	3.2–6.5	1.2–1.8
CO_3^{2-}, mg/L	2.4	0.0	0.0
HCO_3^{-}, mg/L	142.4	1.4–2.0	0.5–1.0
SO_4^{2-}, mg/L	3,100	2.8–6.2	1.9–2.6
Cl^{-}, mg/L	22,840	195–276	29–58
F^{-}, mg/L	0.9	0.5–0.7	0.3–0.5
NO_3^{-}, mg/L	0.00	0.00	0.00
B^{-}, mg/L	5.3	1.2–1.7	0.45–0.80
Br^{-}, mg/L	80	1.0–1.4	0.45–0.60
TDS, mg/L	42,000	350–520	55–105

The first pass of the Fujairah plant consists of 17 duty and 1 standby RO trains using standard rejection SWRO membranes producing permeate of TDS concentration of 400–500 mg/L. The overall recovery of the first-pass SWRO system is 43%. The second pass has 8 BWRO trains with a total recovery rate of 90%. The second-pass BWRO trains have two stages and treat approximately 80% of the first-pass permeate to TDS concentration of 10–20 mg/L. The rest (i.e., 20%) of the first-pass permeate is blended with the second-pass permeate to produce finished water of TDS concentration of less than 120 mg/L. Concentrate produced by the second-pass RO system has salinity lower than that of the source seawater and is recycled to the feed of the first-pass RO system (see Fig. 12.8). Despite of its complexity, the two-pass/two-stage RO system in Fujairah performs very reliably.

An example of a two-pass/two-stage SWRO plant is the 146,000 m^3/day (38 MGD) Point Lisas facility in Trinidad (Fig. 12.9). This plant produces high-quality desalinated seawater of TDS concentration of 85 mg/L or less, which is predominantly used for industrial applications. The first pass of the Point Lisas SWRO system consists of six two-stage RO units. Each of the first-stage RO trains uses SWRO membranes and is coupled with Pelton wheel energy-recovery device. The second-stage trains of the first pass are equipped with brackish water RO elements.

TABLE 12.6 Reverse Osmosis Permeate-Water Quality Seawater Source—Persian Gulf

		Permeate-Water Quality	
Water-Quality Parameter	Persian Gulf Source Seawater Quality	Single Pass SWRO System	Split Partial Two Pass RO System
Temperature, °C	18–35	19–36	20–37
pH	6.0–7.0	5.1–6.0	5.1–6.0
Ca^{2+}, mg/L	570	1.4–2.6	0.6–0.8
Mg^{2+}, mg/L	1,600	2.0–3.6	0.9–1.3
Na^{+}, mg/L	14,100	142–228	25–45
K^{+}, mg/L	530	4.3–6.8	1.5–2.2
$CO_3{}^{2-}$, mg/L	4.2	0.0	0.0
$HCO_3{}^{-}$, mg/L	155	1.8–2.3	0.6–0.9
$SO_4{}^{2-}$, mg/L	3,300	3.1–6.5	2.1–3.2
Cl^{-}, mg/L	24,650	222–305	37.5–64
F^{-}, mg/L	1.5	0.9–1.2	0.5–0.8
$NO_3{}^{-}$, mg/L	0.00	0.00	0.00
B^{-}, mg/L	6.3	1.3–2.5	0.7–1.0
Br^{-}, mg/L	83	1.2–1.5	0.60–0.80
TDS, mg/L	45,000	380–520	70–120

The entire volume of permeate from the first pass of the Point Lisas SWRO system is further treated in a second-pass RO system to meet the final product water-quality specifications. The second-pass system also consists of two stages—each equipped with BWRO membranes. The Point Lisas seawater desalination plant has the same number of first-pass and second-pass RO membrane trains.

Because of the high fouling potential of its source seawater, this two-pass/two-stage configuration has proven to be beneficial. The high-pressure pumps of the first stage are designed for 80% lower feed pressure than these of typical single-stage system, which allows to significantly reduce the fouling on the RO elements of this pass. The permeate from the first pass is treated through a second pass system and an interstage booster pump is installed to deliver the remaining 20% of the feed pressure needed to produce water similar to that in a single-pass RO. Overall, this configuration produces better water quality, higher recovery, and results in less membrane fouling than a typical single-pass system for the same source water quality.

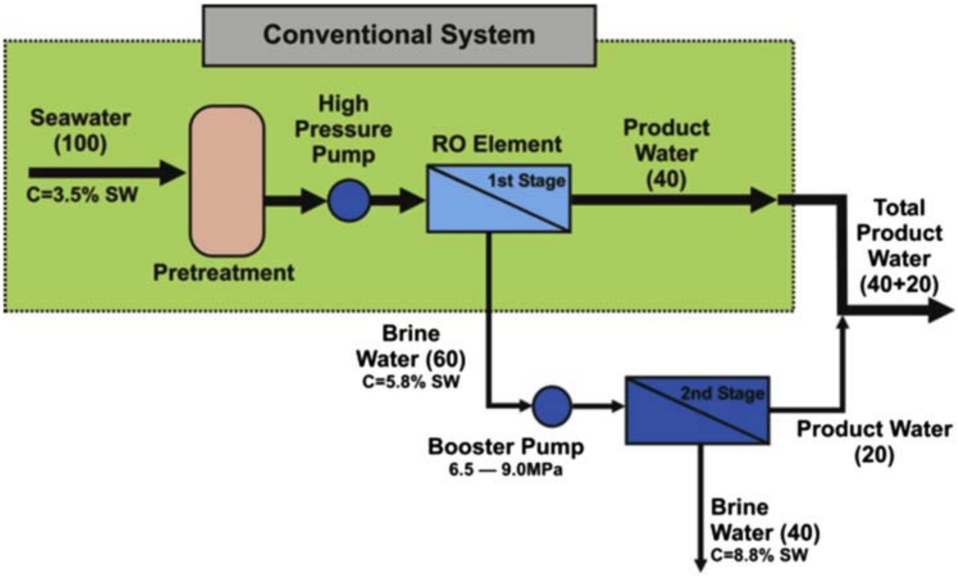

FIGURE 12.7 Two-stage SWRO system.

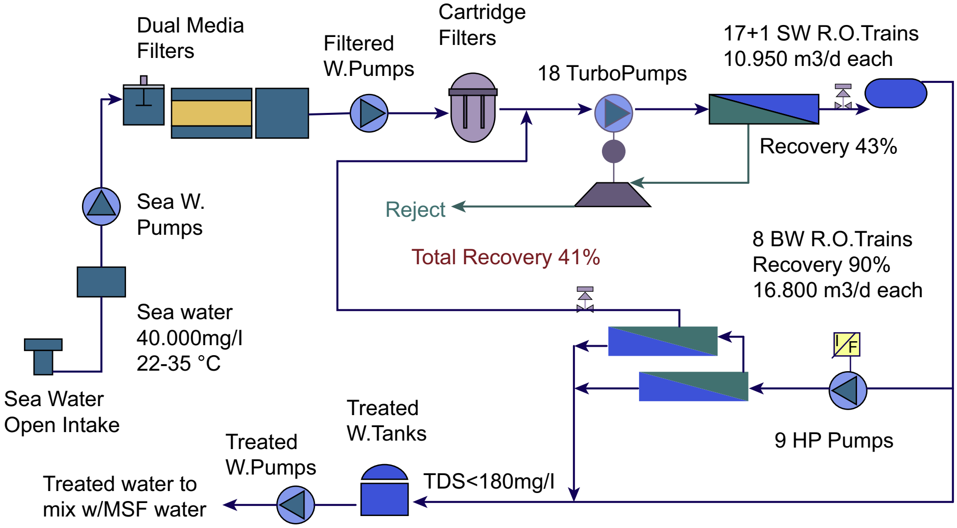

FIGURE 12.8 Schematic of the Fujairah seawater desalination plant.

12.4.6 Three-Center/Four-Stage RO-System Configuration

As indicated previously, a typical conventional SWRO system is configured in individual equally sized RO trains each of which is serviced by a separate transfer pump, cartridge filter vessel, high-pressure feed pump, and energy-recovery equipment dedicated to this RO train.

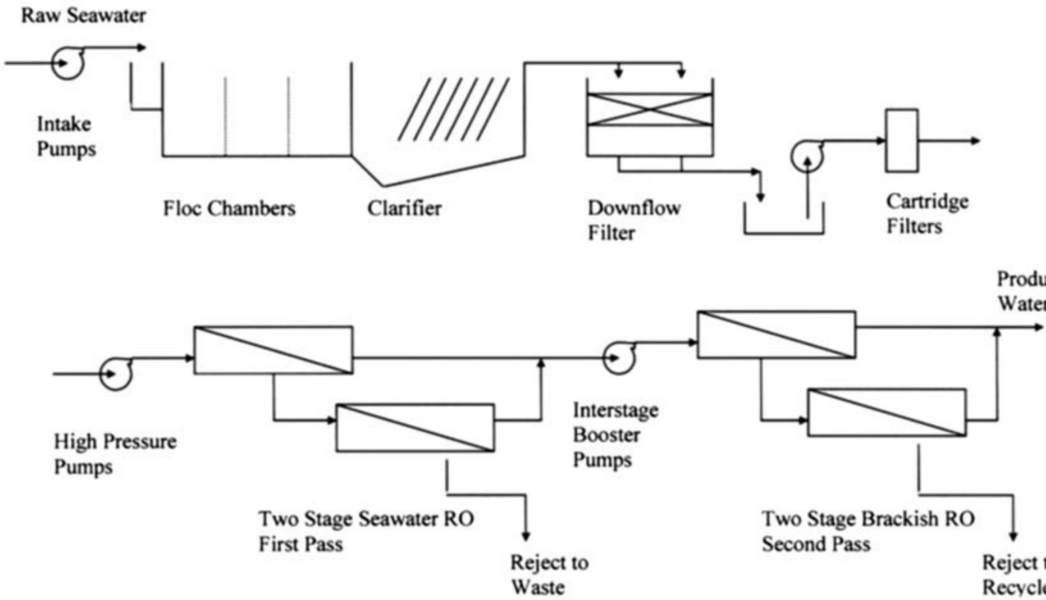

FIGURE 12.9 Point Lisas seawater desalination plant schematic.

The size of the individual RO trains depends on the overall production capacity of the SWRO plant and typically varies between 2000 m^3/day (0.5 MGD) and 21,000 m^3/day (5.5 MGD). The main advantage of this RO train-based configuration is that it is modular and allows for relatively easy flow distribution and service of the individual trains. Since typically the size of the individual RO train does not exceed 10%–20% of the total plant production capacity, train shut down for maintenance (membrane cleaning and replacement, and equipment service) is handled either by using a standby RO train or by temporary increase of the production capacity of the RO trains remaining in service.

The RO-train based configuration is very suitable when the SWRO plant is designed and intended to operate at a constant production output. At present, most of the existing large municipal SWRO plants worldwide are designed to supplement existing conventional water supply sources rather than to be the primary or the only source of water supply for a given area. Therefore, the operation of these SWRO plants does not need to have the flexibility to follow the actual diurnal and monthly fluctuations in product water demand and most of the existing plants are designed to operate at constant production capacity.

In the future, the SWRO is likely to become a prime rather than a supplemental source of water supply for many coastal communities pressured by population growth, especially for large urbanized or industrial areas with limited traditional local sources of freshwater supply (i.e., groundwater or river, or lake water). The SWRO systems servicing such areas have to be designed to have the operational flexibility of matching desalination plant production with the product water-demand fluctuations.

Shift of the SWRO plant operational paradigm from constant to variable production flow requires a change of the typical SWRO configuration from one that is most suitable for constant

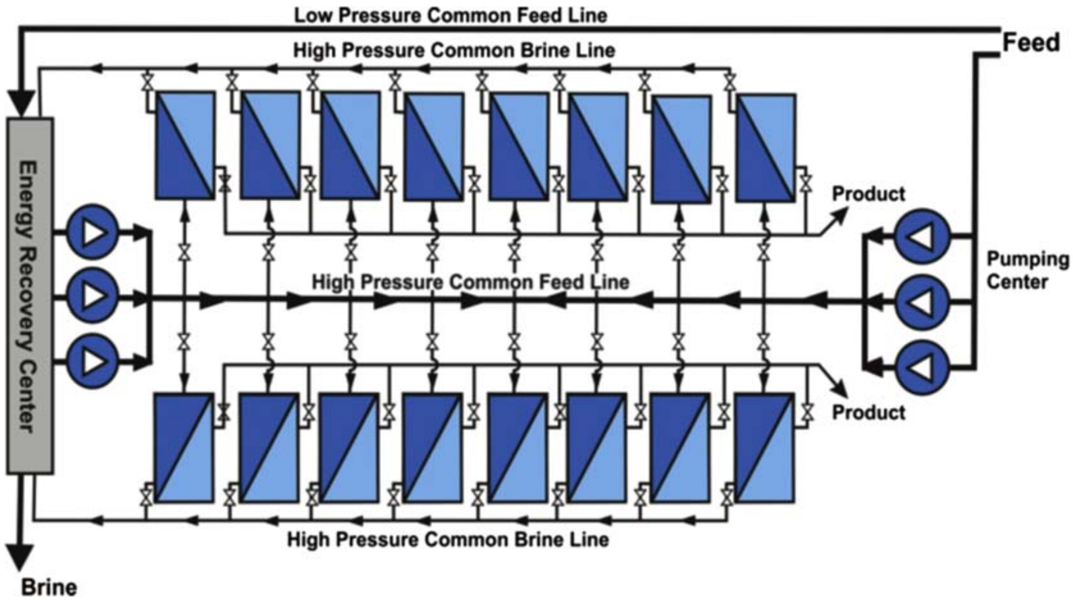

FIGURE 12.10 Three-center SWRO system configuration.

production output to one that is most cost-effective for delivery of varying permeate production flow. A response to such shift of the desalination plant operational paradigm is the three-center RO system configuration used for the first time at the 330,000 m^3/day Ashkelon desalination plant in Israel (see Fig. 12.10). Under this configuration, the RO-membrane vessels, high-pressure pumps, and energy-recovery equipment are no longer separated in individual RO trains, but are rather combined in three functional centers—a high-pressure RO feed-pumping center, a membrane center, and an energy-recovery center (Liberman, 2002). The three functional centers are interconnected via service piping.

The RO feed-pumping center includes only a few large-capacity high-pressure pumps that deliver seawater to the RO-membrane center. The main benefit of using few large-capacity high-pressure pumps rather than a large number of small capacity units is the gain in overall pumping efficiency.

Typically, the smaller the ratio between the pressure and the flow delivered by a given pump, the better the pump efficiency and the “flatter” the pump curve (i.e., the pump efficiency is less dependent of the variation of the delivered flow). Therefore, pump efficiency can be improved by either reducing the pressure delivered by the pump or by increasing pump flow. Since the pump-operating pressure decrease is limited by the RO-system target salt separation performance, the main approach to improve pump efficiency is to increase unit pump flow. While a conventional size high-pressure RO-feed pump of small capacity would typically have maximum total energy use efficiency of 80%−85%, the use of 10 times larger size pump may allow to increase the pump efficiency to 88%−92%, especially for large SWRO plants. This beneficial feature of the three-center design is very valuable in the case of systems delivering varying flow.

While in a conventional RO train design the membrane vessels are typically grouped in 100−200 units per train and in 2−20 RO trains, the membrane center configuration contains two to four times larger number of RO-vessel groups (banks) and a smaller number of membrane vessels per bank. Under this configuration the individual vessel banks are directly connected to the high-pressure pump-feed lines and can be taken off service one at a time for membrane replacement and cleaning.

Although the feed-water distribution piping for such membrane center configuration is more elaborate and costly than that use of individual RO trains that contain two to three times more vessels per train, what is lost in capital expenditure is gained in overall system performance reliability and availability. A reliability analysis completed for a 95,000 m^3/day (25 MGD) SWRO plant (Liberman and Wilf, 2005) indicates that the optimum number of vessels per bank for this scenario is 54 and number of RO banks per plant is 20. A typical RO train-based configuration would include two to four times more (108–216) vessels per RO train and two to four times less (5–10) RO trains. According to this analysis, the use of the three-center configuration instead of the conventional RO-train-based approach allows to improve RO-system availability from 92% to 96% (avg. 95%) to 98%, which is a significant benefit in terms of additional amount of water delivered to the customers and improvement in water supply reliability.

The centralized energy-recovery system included in the three-center configuration (Fig. 12.10) uses high-efficiency pressure exchanger-based energy-recovery technology. The proposed configuration allows to improve the overall energy-recovery efficiency of the RO system and to reduce system power, equipment, and construction costs. While typically, the energy recovery efficiency of the conventional Pelton wheel systems drops significantly when the overall SWRO plant recovery is reduced below average design conditions, the energy-recovery efficiency of the pressure-exchanger systems improves with the reduction of the plant recovery rate. This allows operating the SWRO plant cost-effectively while delivering variable product water flow. For example, if the SWRO plant output has to be reduced by 40% to accommodate low diurnal demand, a SWRO system with RO train-based configuration has to shut down 40% of its trains and if this low demand persists, it has to flush these trains to prepare them for a the next start-up.

An RO system with three-center configuration would only need to lower its overall recovery to achieve the same reduction of the diurnal demand. Although temporary operation at lower recovery would result in elevated costs for pumping and pretreatment of larger volumes of source water, these extra operational expenses are typically compensated for by the improved energy-recovery efficiency that results from operating the SWRO the system at lower water recovery ratio. In addition to providing flexibility in operating cost-effectively the desalination plant at varying production flows, the three-center design yields low-fouling RO-plant operations of very high availability factor.

It is important to point out that the RO-desalination system used at the Ashkelon desalination plant, as that of a number of other large desalination plants, (e.g., Hadera, and Sorek SWRO plants in Israel, and the Carlsbad desalination plant in California) is a four-stage SWRO system depicted in Fig. 12.11 (Gorenflo et al., 2007).

Table 12.7 presents a summary of the source-water quality and the design and actual product-water quality of the Ashkelon SWRO plant in Israel, which was the first facility where such reverse osmosis configuration was used and, therefore, has the longest track record of successful performance (Dreizin et al., 2008).

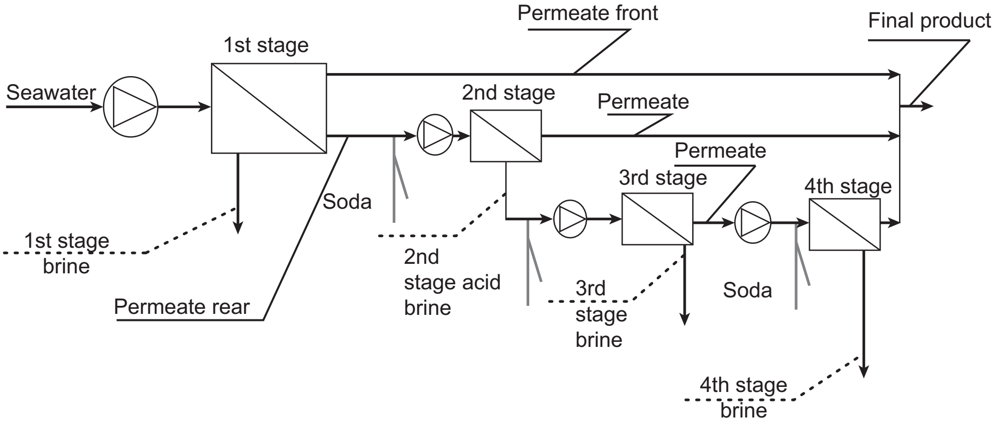

FIGURE 12.11 Ashkelon four-stage SWRO system.

TABLE 12.7 Reverse Osmosis Permeate Water-Quality at Ashkelon Seawater Desalination Plant

Water-Quality Parameter	Mediterranean Source Seawater Quality	Freshwater Quality	
		Permeate	Finished Water
Temperature, °C	16–28	17–29	18–30
pH	8.1	7.5–8.5	8–8.5
Ca^{2+}, mg/L	483	0.2–0.5	90–110
Mg^{2+}, mg/L	1,557	0.5–0.8	0.5–0.8
Na^{+}, mg/L	12,200	6–30	30–39
K^{+}, mg/L	481	1.5–1.8	1.5–1.8
HCO_3^{-}, mg/L	162	0.6–0.8	45–50
SO_4^{2-}, mg/L	3,186	1.0–1.4	1.0–1.8
Cl^{-}, mg/L	22,599	10–15	10–15
F^{-}, mg/L	1.4	0.1–0.2	0.8–1.0
NO_3^{-}, mg/L	0.00	0.00	0.00
B^{-}, mg/L	5.3	0.2–0.3	0.2–0.3
Br^{-}, mg/L	80	0.2–0.3	0.2–0.3
TDS, mg/L	40,679	20–40	180–220

Actual experience at the desalination plants listed above indicates that the four-stage configuration provides very reliable performance and it is not as sensitive to changes in source-water quality (salinity, temperature and biofouling potential) as are conventional two-pass RO systems. All of the referenced desalination plants with four-stage RO systems have open intakes periodically impacted by moderate algal blooms, the source water of which is successfully pretreated by conventional gravity granular media filtration systems of medium depth of 1.4–1.6 m (4.6–5.3 ft).

12.5 ALTERNATIVES FOR CONTROL OF MICROBIAL FOULING

As discussed in Chapter 2, marine microorganisms that naturally occur in the source seawater and their biological waste products deposited on the surface of the membrane elements can decrease RO system productivity over time. For the microbial fouling (biofouling) to occur the following three key factors need to be in place: (1) the source seawater would need to contain metabolically active marine organisms that can produce biofilm that allows them to attach to the membranes and to colonize the membrane surface over time; (2) the source water will need to have relatively large concentration of easily biodegradable organics which serve as a food source for these organisms that can sustain their growth on the membrane surface; (3) favorable hydrodynamic conditions within the membrane elements that facilitate microbial attachment and biofilm accumulation. Of these three factors, however, the most important factor governing the occurrence and extend of microbial fouling is the content of easily biodegradable organics in the source seawater. It is important to note that marine microorganisms inhabit the source seawater at all times, but unless they are "activated" by the presence of large amount of biodegradable organics, they would typically behave as particles and would pass through the SWRO membranes and exit the treatment process with the concentrate without causing measurable biofouling and significant impact on RO membrane fresh water production capacity.

Taking under consideration these three factors, the microbial fouling process could be controlled by a combination of one or more of the following methods: (1) reduction of biodegradable organics in the seawater; (2) reduction of bacterial content in the source seawater; and (3) creation of hydrodynamic conditions in the membrane elements that do not allow the active bacteria to attach to the membrane surface and form biofilm. The alternatives available to control biofouling based on the three biofouling control methods are discussed below.

12.5.1 Reduction of Bacterial Food Sources in Source Water

Microbiological fouling can be effectively controlled by reduction of source-water constituents that accelerate microorganism growth. Since food content is the most important factor for microbial fouling, controlling this content is typically the most efficient biofouling protection strategy.

As indicated in Chapters 6, 7, and 8, controlling the content of easily biodegradable organics could be achieved by reduction of their content ahead of the RO-membrane system in the pretreatment facilities. The organic content of the source seawater could be reduced effectively by one or more of the following approaches: (1) biodegradation upstream of the SWRO system; (2) coagulation; (3) adsorption of organics upstream of the SWRO system; and (4) gentle removal of algal biomass from the source water by DAF and/or gravity filtration.

12.5.1.1 Biodegradation of Organics Upstream of the RO System

Source-water organics biodegradation and removal upstream of the RO system are the most efficient of the four approaches listed above, because the total content of organics could be reduced below the threshold level of 1 mg/L of TOC, which the rate of membrane biofouling slows down significantly.

Biodegradation could be achieved either using deep-gravity granular media filters designed to be operated as biofilters, or in a membrane filtration system configured to function as a membrane bioreactor (MBR) rather than as a particulate removal system. At present, biofiltration in gravity granular media filters is a well-understood process, and this pretreatment technology is widely used and viable. For comparison, the development of MBR technology for seawater pretreatment is in its infancy and at present is not practiced in full-scale desalination plants.

The biodegradation approach for control of RO-membrane biofouling is based on the fact that if the easily biodegradable organics contained in the saline source water are consumed by the microorganisms in the pretreatment system, then this food source would no longer be available for bacteria to grow on the RO-membrane surface and to form sustainable biofilm, and therefore to cause biofouling.

It should be pointed out that biodegradation of seawater organics practically always occurs in conventional granular media filtration systems, but unless these filters are designed to provide adequate contact time and favorable conditions for steady biogrowth, the removal of easily biodegradable organics in gravity filters is typically lower than 30% (measured at TOC reduction). Gravity filters designed for controlled biofiltration could yield organics removal of over 60%, which in the case of moderate or low-intensity algal blooms will be adequate to sustain stable and reliable performance of the RO system.

For comparison, UF and MF membrane pretreatment systems yield significantly lower organic removal efficiency (10%–15% of TOC reduction) because they are designed for a very short contact time and for frequent and vigorous agitation of the membrane surface, which does not allow significant biomass to be formed and maintained in the pretreatment membrane modules.

12.5.1.2 Coagulation

As discussed in Chapter 6, coagulation allows to enhance the removal of particulate organic matter in the saline-source water by agglomeration of the organic particles and their subsequent removal by sedimentation and/or filtration. Removal of source-seawater organics is usually between 5% and 20% (measured as reduction of the TOC concentration). In most cases however, such removal would not allow to bring the concentration of TOC in the feed water to the RO system below 1 mg/L and, therefore, to prevent accelerated biofouling of the RO membranes. Coagulant addition could be beneficial for both granular media and membrane filtration, especially if the organics in the source water mainly consist of NOM.

However, coagulant addition is not likely to enhance the removal of dissolved organics such as these that typically occur during algal blooms.

The MF and UF systems typically remove only organics in particulate form. Because during algal blooms over 85% of the organics in the source water are usually in dissolved form, existing MF and UF systems currently available on the market are ineffective for controlling RO-membrane biofouling. Practical experience from full-scale installations shows that these systems could provide higher removal of organics by enhanced coagulation, but only if the main type of organics in the saline source water is NOM. In most SWRO desalination plants with open intakes, however, NOM contributes only very small portion of the organic substances in the water—the main source of organics are algae, dead and living bacteria, and exopolymer products excreted from bacteria.

12.5.1.3 Adsorption of Organics by Activated Carbon

Activated carbon adsorption could yield 10%–30% of TOC removal from the saline source water, and when combined with granular media filtration, it could enhance removal of organics in the pretreatment system to over 40%. The addition of a layer of 0.3–0.5 m (1.0–1.5 ft) of granular activated carbon (GAC) on the top of the anthracite in gravity media filters (often referenced in practice as installation of "carbon cap") usually is an effective measure for reduction of dissolved organics during moderate- and high-intensity algal blooms to levels that prevent the occurrence of excessive biofouling of the downstream RO membranes.

It should be pointed out, however, that the organics absorption capacity of typical carbon cap is limited to 3 to 4 months only. After this period, the GAC layer have to be removed from the surface of the gravity media filters and sent for regeneration and replaced with a new GAC. Because the installation, replacement, and regeneration of GAC are relatively costly, the use of GAC-carbon caps has found limited application in desalination plants, to date.

If membrane pretreatment is used, then enhanced removal of dissolved organics in the source water could be achieved by introduction of powdered activated carbon (PAC) to the feed-water upstream of the desalination plant's microscreens. The size of the PAC should be such that it can successfully pass through the microscreens. Practical experience at the Adelaide desalination plant shows that using the microscreens for mixing of the saline source water and the PAC is very successful and is preferable to feeding the PAC directly upstream of the MF/UF pretreatment system. In both cases, the PAC with the dissolved organics absorbed by it will be retained on the pretreatment membranes and removed with the membrane backwash water.

The most suitable dosage for PAC addition is between 0.5 and 1.0 mg/L of PAC per each mg/L of TOC contained in the source water. Because this dosage is relatively high and the PAC cannot be recovered, regenerated, and reused in the treatment process, the use of this biofouling control method has found very limited application in full-scale desalination plants.

12.5.1.4 Gentle Removal of Algal Biomass Form Source Water

Since algal biomass caries significant amount of biodegradable organics, its gentle removal by DAF or slow downflow gravity-driven granular media filtration could be a very efficient measure for controlling the release of organics in the source water and subsequent SWRO biofouling. This approach is very cost-effective when the source seawater intake is exposed to frequent and extensive algal blooms. Algal biomass removal could reduce the content of source water organics with over 50%, by preventing the release of the organics contained in the algal cells into the seawater fed to the SWRO system.

12.5.2 Reduction of the Bacterial Content of Source Water

The concentration of the microorganisms in the source water can be effectively reduced by: (1) their exposure to strong oxidants (e.g., disinfectants) or UV light; or (2) by depriving the microorganisms from oxygen by applying strong reducing agents such as sodium bisulfite. These methods are briefly described below.

12.5.2.1 Bacterial Growth Control by Microbial Oxidation and UV Inactivation

Several oxidants such as chlorine, chlorine dioxide, and chloramines can be used for surface source-water microbiological growth control. Microbial control by adding biocides to the source seawater is of controversy and focus of research at this time. This is because some membrane desalination plants have had significant microbial fouling problems after chlorination or other means of microbial control—even possibly worse than if no chemical biocides were used. It should be noted that continuous chlorination and dechlorination upstream of the SWRO membranes can increase bioactivity and associated membrane biofouling by increasing the content of assailable organic compounds in the source water. Some desalination plants have experienced permanent damage of their SWRO membranes by exposure to the chemical oxidant when the dechlorination chemical system failed.

Chlorination is the most popular process for biofouling control practiced at present. This process was discussed in detail in Chapter 6. Ultraviolet (UV) inactivation of aquatic microorganisms is an alternative method for membrane biofouling control. However, in some facilities, microbial regrowth after UV treatment negated the benefits, so its use should be evaluated very carefully. UV irradiation method is power-intensive and, therefore, usually less cost-effective than chlorination/dechlorination. The cost-effectiveness of microbial UV inactivation is dependent on the source water quality and the design on the UV system. If the source water has high levels of turbidity, the UV dosage could be relatively high and biofouling control ineffective. For optimum performance, it is recommended that the turbidity of the saline source-water fed to the UV unit not to exceed 10 NTU. The best location of the UV system would be between the cartridge filter and the RO membranes. Because of space constraints, however, that is not usually possible, so as an alternate, installing the UV system just prior to the cartridge filters is acceptable.

12.5.2.2 Bacterial Growth Control by Reduction of Oxygen

Seawater collected by open intakes is usually rich in oxygen, which stimulates biological growth. Many marine bacteria use oxygen for bioassimilation of the dissolved organics in the ambient ocean water. Therefore, removal of oxygen by reducing chemicals such as sodium bisulfite hinders their growth and ultimately suppresses the biofouling on the membranes (Wilf et al., 2007).

While inactivation of biofouling microorganisms by oxygen reduction is effective over short periods of time (several days to 1 week), in the long term, continuous addition of reducing chemical does not yield effective control. The reason behind this challenge is the fact that most marine bacteria are not strictly aerobic and they can also bioassimilate organics in the source water under anaerobic conditions.

While the bacterial growth in anaerobic environment is of slower rate than that under aerobic conditions, such biogrowth would also cause accelerated biofouling of the membranes. Therefore, control of bacterial growth on the RO-membrane surface by continuous addition of reducing chemical (e.g., sodium bisulfite) is not sustainable long-term solution to biofouling challenges caused by algal blooms—such blooms usually last for 4–6 weeks or more at a time.

12.5.3 Creation of Favorable Hydrodynamic Conditions in the SWRO Membrane Elements

As indicated in Chapter 2, SWRO biofouling could be controlled by creating turbulent hydrodynamic conditions on the surface of the SWRO membranes, which would prevent active marine bacteria from attaching to the surface and developing biofilm.

Fundamental research in this area indicates that such conditions could be created if the SWRO membrane systems are operated at relatively low recoveries (30%–35% vs. 40%–50%) and low permeate membrane flux, as well as by designing the membrane feed/concentrate spacer geometry such that it induces turbulent flow and high scouring (shear) velocity on the membrane surface (Winters et al., 2007). The work in this field is ongoing and holds significant potential.

References

Dreizin, Y., Tenne, A., Hoffman, D., 2008. Integrating large scale seawater desalination plants within Israel's water supply system. Desalination 220, 132–149. Elsevier.

Gorenflo, A., Brusilovsky, M., Faigon, M., Liberman, B., 2007. High pH operation in seawater reverse osmosis permeate: first results from the World's largest SWRO plant in Ashkelon. Desalination 203, 82–90. Elsevier.

Greenlee, L.F., Lawler, D.F., Freeman, B.D., Marrot, B., Moulin, P., 2009. Reverse osmosis desalination: water sources, technology, and today's challenges. Water Research 43, 2317–2348. Elsevier.

Liberman, B., December 2002. The importance of energy recovery devices in reverse osmosis desalination. In: The future of Desalination in Texas, vol. 2. Texas Water Development Board. Technical Papers, Case Studies Desalination Technology Resources (Report # 363).

Liberman, B., Wilf, M., 2005. Evolution of configuration of RO seawater desalination elements. In: Proceedings of IDA World Congress – Singapore, September 11–16. Ref: SP05-059.

Mickols, W., Busch, M., Maeda, Y., Tonner, J., 2005. A novel design approach for seawater plants. In: Proceedings of IDA World Congress-Singapore, September 11–16, pp. 2–25. Ref: SP05-052.

Sanz, M.A., Bonnelye, V., Cremer, G., 2007. Fujairah reverse osmosis plant: 2 years of operation. Desalination 203, 91–99. Elsevier.

Wilf, M., Awerbuch, L., Bartels, C., Mickley, M., Pearce, G., Voutchkov, N., 2007. The Guidebook to Membrane Desalination Technology. Balaban Desalination Publications, L'Aquila, Italy.

Winters, H.S., Fane, A., Nicolaisen, B., 2007. Microbial membrane fouling; microbial cake enhanced osmotic pressure (CEOP) or thermodynamic restriction?. In: Proceedings of IDA World Congress-Masspalomas, Gran Canaria-Spain, October 21–26, pp. 1–15. Ref: IDAWC/MP07-267.

Glossary

Aerobic Containing oxygen.
Ambient Surrounding of background.
Ambient seawater Seawater in the open ocean used for desalination.
Anaerobic Not containing oxygen.
Anthropogenic Caused by human activity.
Antiscalant A chemical added to the saline source water fed to a membrane system, e.g., RO membranes, which inhibits or prevents precipitation of minerals on the membrane surface.
Aquifer A natural underground layer, often of sand or gravel, that contains water.
Backwash The process of solids' removal from a filter media or membrane surface by applying air and/or clean water in the opposing flow direction.
Bactericide A chemical capable of destroying bacteria.
Biocide A chemical used to inactivate microorganisms (e.g., chlorine).
Biofouling Presence and growth of microorganisms and their secretations on the membranes surface.
Biomass Organic matter of vegetable or animal origin.
Brackish water Water with a total dissolved solids concentration less than 15,000 mg/L.
Brine (concentrate) Water with high which is a product of the desalination process salinity, usually in excess of 2000 mg/L for brackish water and over 37,000 mg/L for seawater.
Brine seal Rubber or plastic device of special design, which is installed on the outerside of the membranes near their feed side, which seals the space between the membrane elements and the pressure vessel to prevent the bypassing of feed water/concentrate around the elements.
Coagulation Pretreatment process used in some desalination plants, which consists of addition of salts (e.g., ferric chloride) to source water in order to improve suspended solids agglomeration and their more efficient removal by further treatment, i.e., sedimentation or filtration.
Colloids Suspended solids with a diameter of less than 1 μm that cannot be removed by sedimentation alone.
Colocation Location of a desalinated plant with an existing power generation facility with potential connection of desalination plant intake and/or outfall to the cooling water discharge.
Concentrate See brine.
Concentrate management The handling and disposal or reuse of concentrated liquid generated by the desalination system during the salt separation process.
Concentration polarization Phenomenon in which solutes form a dense, more concentrated layer next to the membrane surface and restricts flow through the membrane.
Desalination A process that removes dissolved solids (salts) from saline source water.
Diffuser Offshore end portion of outfall, which consists of discharge ports configured to maximize the mixing of the desalination plant discharge with the ambient receiving waters.
Dissolved air flotation (DAF) The removal of particulate matter by attaching the particles to rising air bubbles in a specially designed flotation tank.
Double-pass or Two-pass RO system RO system that consists of two sets of RO trains configured in series, in which permeate from the first set is processed through the second set of RO trains.
Dual work exchanger energy recovery (DWEER) An energy recovery device used in RO-desalination plants, which represents an isobaric type pump for direct conversion of concentrate pressure into RO feed-water pressure via piston.
Ecosystem A community of living organisms interacting with each other and with nonliving components of the environment they inhabit.
Effective size The media grain diameter at which 10% of the media by weight is smaller than this diameter, as determined by sieve analysis.
Feed water The influent water that is fed into a treatment process/system.

Filtrate The purified water that is produced by the granular media or membrane filtration.
Filtration The removal of particulate matter in source water by passing through a granular medium such as sand and/or anthracite of ultrafiltration or microfiltration membranes.
Flux The rate of water flow across a unit of membrane surface area expressed in liters per hour per square meter (L/m^2 h or Lmh).
Fouling The accumulation of contaminants on the membrane surface (e.g., MF, UF, NF, or RO membranes) resulting in loss of performance.
Hardness Concentration of calcium and magnesium slats in water.
Hydraulic turbo booster (HTB) A energy recovery device, also known as *turbocharger*, used in RO-desalination plant that consists of turbine and centrifugal pump connected on the same shaft.
Inert Matter that neither dissolves in water nor reacts chemically with water or other substances.
Injection zone Geological formation receiving desalination plant discharge via deep injection well.
Inorganic Description of all matter that does not originate from living organisms (animals, plants, bacteria, etc.), also referred as mineral.
Intake The facility through which source water is collected to produce fresh water in desalination plant.
Interstage design (ISD) An RO-membrane configuration design to achieve a more even flux distribution that combines three different models of membranes with different permeability within the same vessel.
Ion An atom or group of atoms/molecules that has a positive charge (cation) or negative charge (anion) as a result of having lost or gained electrons.
Langelier saturation index (LSI) A parameter indicating the tendency of a water solution to precipitate or dissolve calcium carbonate.
Membrane A device, usually made of an organic polymer, that allows the passage of water and certain constituents, but rejects others above a certain physical size or molecular weight.
Membrane element An individual membrane unit of standard size and performance.
Membrane system A complete system of membrane elements, pumps, piping, and other equipment that can treat feed water and produce clean water (permeate).
Microfiltration Filtration through membranes of pore size between 0.1 and 0.5 μm.
Near-shore discharge Disposal on the desalination plant waste streams through structure (channel, pipe, weir, etc.) located on the shore or within several hundred meters from the shore in the tidal zone.
Net driving pressure (NDP) See transmembrane pressure.
Offshore discharge Disposal of desalination plant waste streams via long outfall structure extending beyond the tidal zone.
Open intake Intake collecting source water directly from the water column of surface water body.
Organic Description of matter that includes both natural and man-made molecules containing carbon and hydrogen; all organisms living in water are made up of organic molecules.
Osmosis The naturally occurring transport of water or other solvent through a semipermeable membrane from a less concentrated solution to a more concentrated solution.
Osmotic pressure A pressure applied on the surface of semipermeable membrane as a result of the naturally occurring transport of water from the side of the membrane of lower salinity to the side of the membrane with higher salinity.
Pelton wheel An energy recovery technology used in RO plants consisting of enclosed turbine in which concentrate is applied through a high-velocity nozzle onto spoon-shaped buckets located on the periphery of the wheel.
Permeability The capacity of membrane material to transit flow, also named flux.
Permeate (sometimes called filtrate or product water) The purified water produced during membrane filtration.
Pigging The practice of using pipeline inspection gauges or "pigs" to perform various maintenance operations on a pipeline.
Pressure exchanger (PX) An energy recovery technology used in RO-desalination plants that consists of individual fiberglass vessels connected to common feed and concentrate manifolds, each of which contains a ceramic rotor with a number of cylinders, i.e., rotor chambers.
Pressure filtration Filtration aided by imposing pressure across an enclosed filter vessel.
Pressure vessel A housing containing membranes in a preset configuration that operates under pressure; for RO systems, pressure vessels are plastic or metal tube-shaped devices that house 6–8 RO elements.

Pretreatment Process that includes one or more source water treatment technologies (e.g., screening, coagulation, sedimentation, filtration, chemical addition etc.) that aim to remove foulants from the saline source water prior to RO separation in order to protect the membranes and improve desalination plant performance.

Product water Low-salinity (fresh) water usually with TDS concentration of 500 mg/L or less produced by the desalination plant and suitable for distribution system delivery. For the desalination plant permeate to be converted to product water, it has to be disinfected and conditioned for corrosion and predetermined water quality requirements.

Recovery The ratio of desalinated low-salinity water (filtrate or permeate) flow to feed water flow of filtration system.

Reverse osmosis Pressure driven movement of water through a semipermeable membrane from the side of the membrane with more concentrated solution to that of a less concentrated solution.

Salinity The concentration of total dissolved solids in water.

Salt passage The ratio of the concentration of salt/s (ions) in permeate and the concentration of the same salts (ions) in the feed seawater; typically, salt passage is expressed as a percentage of the feed water concentration of salts.

Salt rejection The ratio of salts (ions) removed (rejected) by the RO membrane to the same salts (ions) in the saline source water; salt rejection is equal to 100% minus the salt passage.

Scale Mineral deposits formed on the surface of membrane and/or membrane matrix as a result of concentration (saturation) of the mineral/s to a level at which they form insoluble amorphous or crystalline solids.

Scale inhibitor See antiscalant.

Scaling Process of scale formation on the surface or in the matrix of RO membrane.

Semipermeable membrane A membrane that has structure that allows small molecules, such as water, to pass while rejecting a large portion of the salts contained in the feed water.

Silt A sedimentary material consisting of very fine particles intermediate in size between sand and clay.

Silt density index (SDI) A dimensionless parameter widely used to quantify the potential of seawater or brackish water to cause particulate and colloidal fouling of RO membranes.

Specific permeability (flux) The capacity of membrane material to transit flow also named *specific flux*; expressed as the membrane flux normalized for temperature and pressure, in liters per square meter per hour per bar (Lmh/bar).

Spiral-wound element An RO or NF membrane element that consists of membrane leafs wound around a central permeate collection tube and including feed and permeate spacers, antitelescoping devices and a brine seal.

Stage A set of pressure vessels installed and operated in parallel.

Subsurface intake Intake located below the ground surface collecting source water from groundwater aquifer. Examples of subsurface intakes are vertical, horizontal, and slant wells and infiltration galleries.

Suspended solids Particulate solids suspended in the water.

Thin-film composite membranes (TFC) Semipermeable membranes manufactured principally for use in water desalination systems that can be considered as a molecular sieve constructed in the form of a film from two or more layered materials.

Total dissolved solids (salinity) A measure of the total mass of all dissolved solids contained in the water.

Total suspended solids The concentration of filterable particles in water (retained on a 0.45 μm filter) and reported by volume.

Train (rack, skid) A membrane system that consists of rack housing a number of pressure vessels that have a common feed, permeate, and concentrate piping and control equipment, and can be operated independently; an RO system or MF or UF membrane system consists of multiple trains operating in parallel.

Transmembrane pressure (TMP) The driving force that transmits product water through the membrane.

Troubleshooting A process of diagnosing and correcting the source of a problem.

Turbidity A measure of concentration of suspended solids in water, which is determined by the amount of light scattered by these solids.

Ultrafiltration Filtration through membranes of pore size between 0.01 and 0.05 μm.

Uniformity coefficient The ratio of the 60th percentile media grain diameter to the effective size of the filer media.

Variable-frequency drive (VFD) A type of adjustable-speed drive used in electromechanical devices to control motor speed and torque by varying motor input frequency and voltage.

Viscosity A tendency of fluid to resist flow (movement) as a result of molecular attraction (cohesion).

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Pretreatment for Reverse Osmosis Desalination

Nikolay Voutchkov

Pretreatment for Reverse Osmosis Desalination is a comprehensive reference on all key existing and emerging seawater pretreatment technologies used for desalination. The book focuses on reverse osmosis membrane desalination, which at present is the most widely applied technology for production of fresh drinking water from highly saline water sources (brackish water and seawater). Each chapter contains examples illustrating various pretreatment technologies and their practical implementation.

Students and researchers in the subject, chemical engineers, scientists in regulatory agencies, and other municipal water supply professionals will find this book useful.

Key Features

- Provides an in-depth overview of key theoretical concepts associated with desalination pretreatment
- Gives insight into the latest trends in membrane separation technology
- Incorporates analytical methods and guidelines for monitoring pretreatment systems

Nikolay Voutchkov has over 25 years of experience in the field of desalination and water reuse as an independent technical advisor to public utilities implementing large desalination projects in Australia, United States, and the Middle East and to private companies and investors involved in the development of advanced membrane technologies. For over 11 years, he served as a Chief Technology Officer and Corporate Technical Director for Poseidon Resources, a private company involved in the development of the largest seawater desalination projects in the United States. He has received awards from the WateReuse Research Association, the International Desalination Association, the International Water Association, and the American Academy of Environmental Engineers.

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