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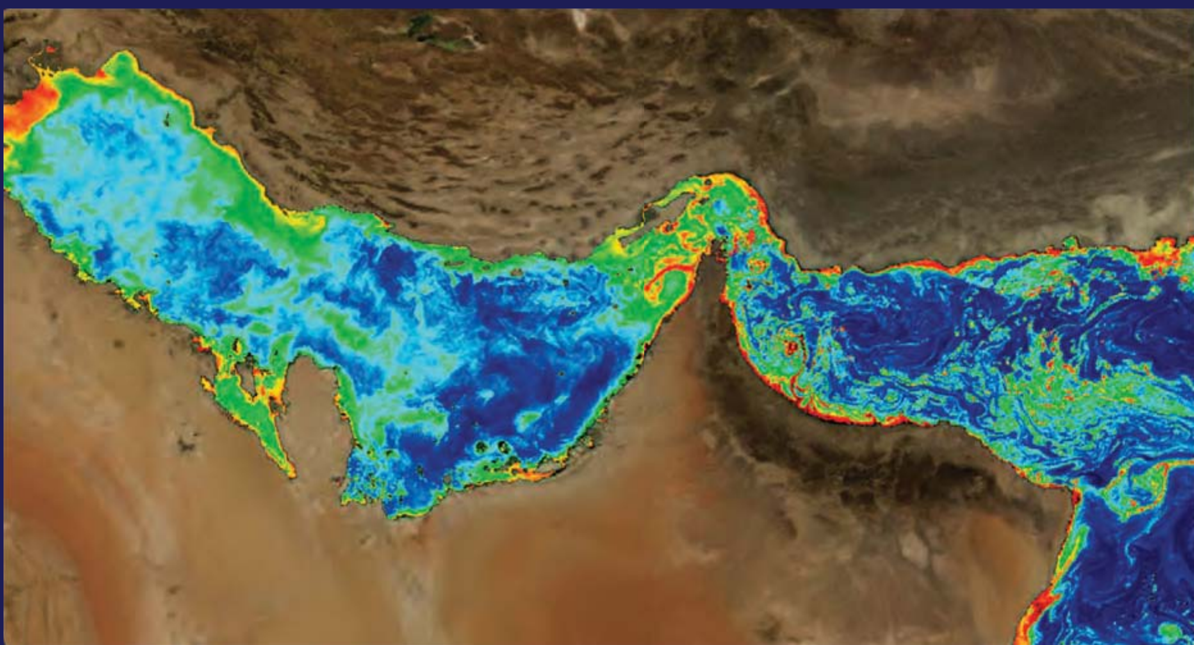
Harmful Algal Blooms (HABs) and Desalination: A Guide to Impacts, Monitoring, and Management



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Harmful Algal Blooms (HABs) and Desalination: A Guide to Impacts, Monitoring and Management

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9 ALGAL BIOMASS PRETREATMENT IN SEAWATER REVERSE OSMOSIS

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9.1 INTRODUCTION

Harmful algal blooms (HABs) can result in a substantial increase in the organic and solids load in the seawater feed to be treated at a desalination plant. In this chapter, the removal of this material is addressed in the context of the multi-barrier treatment process for seawater reverse osmosis (SWRO) as presented in Chapter 8 on risk management for HAB events. While this chapter covers removal of non-toxic material, Chapter 10 builds upon these principles and discusses the mechanisms and effectiveness for each barrier with respect to toxin removal. This chapter covers only the main barriers used in the SWRO desalination plants for HAB bloom risk mitigation, though the authors acknowledge that other niche treatment barriers exist in SWRO systems. The treatment processes discussed here are chlorination and dechlorination, dissolved air flotation (DAF), granular media filtration (GMF), microscreens for microfiltration/ultrafiltration (MF/UF), MF/UF itself, cartridge filtration and SWRO. Coagulation is discussed in general terms and then more specifically for DAF, GMF, and MF/UF pretreatments. Each treatment process is broken down into a discussion of how the process works and then how HAB cells affect the process operation. Importantly, the chapter deals with how upstream actions can detrimentally affect downstream treatment processes with respect to algal blooms.

In particular, this chapter discusses removal mechanisms for algal organic matter (AOM) and how operational actions can prevent detrimental effects of AOM. As discussed in Chapter 2, the chemical composition of AOM usually includes proteins, polysaccharides, nucleic acids, lipids, and other dissolved organic substances. AOM compounds typically cover a wide size spectrum, ranging from less than 1 nm to more than 1 mm. Based on their size cut-off, GMF and MF/UF are expected to remove only part of high molecular weight AOM (as shown in Chapter 2, Figure 2.2). SWRO is expected to achieve complete removal of AOM, but will suffer from fouling issues if AOM is not removed upstream.

9.2 CHLORINATION IN SWRO

9.2.1 Overview

The disinfection properties of chlorine have been known for many years and routine use of chlorine in water treatment processes began in the early 1900s; however, in the last decade water chlorination has received criticism due to its production of disinfection byproducts.

In order to prevent marine growth, such as molluscs, in seawater intakes for both SWRO and thermal plants, biocides such as chlorine, ozone, potassium permanganate, and hydrogen peroxide can be used. The most widely used among them is chlorine, usually applied in one of three forms: 1) chlorine gas; 2) calcium hypochlorite; and 3) sodium hypochlorite, with the latter being the most typical (Bahamdan et al. 1999).

Chlorination in SWRO desalination plants is typically applied either at regular intervals (intermittent/shock chlorination) or continuously (rare) and dosed at the intake structure with additional dosing points downstream to maintain the required chlorine residual. Residual concentration of chlorine in the feedwater is typically kept at 0.2 - 4 mg/L (Agus et al. 2009). For intermittent chlorination, the duration of the dose may last from 30 to 90 minutes and is typically undertaken every 1 to 2 weeks at a random time (Ferguson et al. 2011). Often this is undertaken more frequently, up to daily.

The continuous chlorination method has suffered from criticism that it causes biofouling of the reverse osmosis (RO) process unit. Appelgate et al. (1989) reported that chlorine degrades humic acids and high molecular weight compounds present in coastal seawater to smaller molecules that can be assimilated by bacteria. The chlorine suppresses bacterial activity, but when the sodium metabisulfite (SMBS) is added to remove chlorine prior to the RO, the surviving bacteria quickly take advantage of the nutrients generated by the degradation of larger molecules and enter into a cycle of enhanced growth. The significant increase in the biomass of bacteria after dechlorination causes slime development of biofilm on the surfaces of pipes and RO membranes (Winters 1995). To overcome the problem, alternate disinfectants have been used such as chlorine dioxide, chloramine and copper sulfate (Winters and Isquith 1995) although these have not been widely adopted. Plants that practice continuous chlorination now typically use very low doses (0.1 to 0.3 mg/L).

Shock chlorination, also referred to as intermittent chlorination, is undertaken to control growth of marine life on pipelines and equipment that are constantly in contact with seawater. Both continuous and shock chlorination inhibit but do not fully prevent the growth of marine life so the presence of common oceanic foulants such as molluscan shells and barnacle deposits is unavoidable. The growth rate can be controlled to manageable levels, however. The effectiveness of shock chlorination can also be limited by the withdrawal of marine creatures into their protective shells and their re-emergence when chlorine has dissipated, especially when chlorine addition is undertaken at regular times rather than random intervals. Shock chlorination of seawater collected by open ocean intakes will result in an increased load of particulates on all solid removal processes due to the varied size range of solids introduced. The suspended solids will be in the micron size range, such as finer silts through to larger particles such as molluscan shells, byssal threads, and smaller debris like molluscan tissues and larvae. The impact on a membrane filtration system can be minimized by installing strainers or disk filters that would remove the bulk of the solids load created by the chlorination process (Ferguson et al. 2011).

As an alternative to using chlorine, a low dose of chlorine dioxide can be compatible with polyamide RO membranes, under certain conditions and to a certain extent. When generating chlorine dioxide onsite, it can be contaminated with chlorine and thus dechlorination is still necessary (Dow Water and Process Solutions 2015).

Intake chlorination can also produce carcinogenic compounds like trihalomethanes and haloacetic acids, as well as other inorganic disinfection byproducts (DBPs) such as chlorite, chlorate, bromate, and nitrogenous DBPs (N-DBPs). The latter generally form in much smaller amounts than chlorinated DBPs, but have been a growing concern over the past decade because of their greater health risk. High levels of nitrogen-containing compounds in AOM, which increases dramatically during a bloom, can lead to the formation of significant quantities of N-DBPs (Le Roux et al. 2015).

As bromide is present in seawater, hypobromus acid (HOBr) formation is favored over hypochlorus acid (HOCl). HOBr has been found to react much more rapidly than HOCl with organic (and inorganic) compounds. It is also reported that HOBr has been found to be 25 times stronger than HOCl in its halogen substitution. Therefore, further reaction through a series of oxidations and reductions leads to the formation of the carcinogenic species bromate. Where bromate becomes an issue, sodium hypochlorite generated from seawater should be ceased and other alternatives used such as chlorine gas or calcium hypochlorite (Al-Rasheed et al. 2009).

The concentration of the above DBPs in the final permeate/distillate should be controlled to meet guideline levels prescribed by the WHO or other local agencies. A study by Le Roux et al. (2013) showed that desalinated water produced by thermal MSF plants and SWRO plants is drinkable and poses no threat to human health with respect to DBPs.

Destabilizing and inactivating algal cells through chlorination, which lyses the cells through breakdown of the cell wall, can theoretically prevent biofouling on pre-treatment membranes and RO if a suitable treatment strategy is employed, such as by using coagulation during pretreatment to increase the percentage of AOM removed. Similar outcomes are observed when other oxidative agents (i.e. ozone, permanganate and non-oxidizing biocides) are used instead of chlorination (Heng et al. 2008).

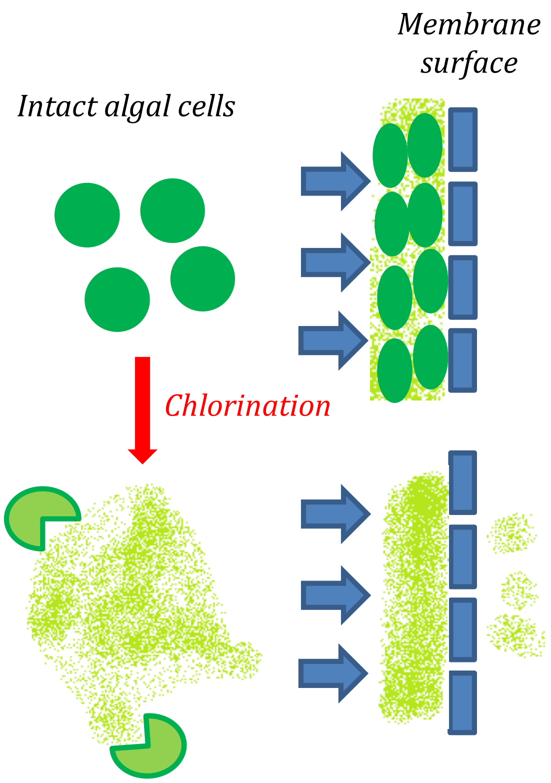
9.2.2 Chlorination during a HAB

In addition to the applied concentration of chlorine, cell destruction efficiency is also dependent on the duration of exposure and depends largely upon the cell wall thickness and type. Thicker cell walls will take longer to degrade, therefore both the chlorine dose and the residence time in the intake are important. Typical residence times for SWRO intakes are highly variable (5-60 mins) and can depend upon factors such as the length of intake tunnels and pump wells. Previous research using the marine dinoflagellate *Prorocentrum*, which has a thick cellulose cell wall, showed that doses of 2-3 mg/L of chlorine could achieve complete algal cell lysis within 24 hours (Resosudarmo et al. 2017). Total cell lysis also occurred in the process. Even non-lethal doses of 0.1 – 0.5 mg/L were found to reduce the photosynthetic ability of marine algae significantly without a high degree of cell lysis. While the mechanism of chlorination of HAB cells is still being investigated, one study by Azanza et al. (2001) showed that cells of the dinoflagellate *Pyrodinium bahamense* were degraded via rupturing of the thecal plates and the release of mucilage.

There are also other benefits to chlorination beyond the destruction of algal cells during HABs. The efficiency of coagulation-based processes (such as prior to DAF, GMF or UF) increases significantly when combined with low doses of chlorine (0.1 - 0.5 mg/L). Previous research has shown that up to 96% of algal cells can be removed when both processes are combined (Shen et al. 2011). Chlorination was also found to reduce fouling of an outside-in UF membrane based pretreatment systems (Xu et al. 2014). Once a dose above 1.5 mg/L was applied, further increase in residual chlorine levels did not result in additional fouling improvement.

The negative effects of chlorination are typically observed when excessive doses are applied to the feed seawater during a HAB. High levels of chemical stress can result in the lysis of algae cells and release of intracellular organic matter (IOM) comprising potential foulants, taste and odor compounds and/or toxins. While the value will vary depending upon the algal species in a particular bloom, due in part to differences in cell wall composition, previous research with the marine alga *Tetraselmis suecica* showed that significant cell lysis occurred at chlorine levels exceeding 5 mg/L (Resosudarmo et al. 2017). Another study using the freshwater *Microcystis aeruginosa* cyanobacterium demonstrated that chlorine doses as low as 0.8 mg/L were sufficient to induce cell lysis (Ma et al. 2012). In both studies, exceeding the tolerable levels of residual chlorine resulted in large amounts of IOM release. In the case of *Microcystis*, cell lysis also resulted in toxin release.

Analysis of the IOM released from *Tetraselmis suecica* cell lysis found that the majority was small enough to pass easily through UF pretreatment membranes and potentially increase RO biofouling, as depicted in Figure 9.1 (Resosudarmo et al. 2017). Other studies have pointed out that AOM released by algae during a bloom can also contribute to long-term permeability decline of pretreatment membranes and reduced cleaning effectiveness, as discussed in Chapter 2 (Heng et al. 2008; Hung and Liu 2006). High molecular weight AOM comprises protein and polysaccharide compounds, including transparent exopolymer particles (TEP) and their precursors as discussed in Chapter 2.



- Intact algal cells prevent IOM from passing through the UF membrane pores.
- Mucilage surrounding the algal cells is retained.
- Rapid transmembrane pressure increase due to compressible fouling layer.

- Released IOM can now transmit through the membrane pores.
- Mucilage now attaches to the membrane surface, preventing efficient cleaning
- Reduced short-term fouling

Figure 9.1. Proposed effect of algal cell lysis on UF membrane fouling and rejection. Modified from Resosudarmo et al. 2017.

It is therefore paramount that the correct residual chlorine dose is applied during algal blooms. While this can be difficult due to the high number of existing algal species responsible for blooms, it is important to note that lower chlorine doses combined with longer exposure times will lead to less IOM release during treatment.

When long exposure is not feasible, it is possible to use a chlorine dose between 3-5 mg/L, demonstrated to be sufficient to achieve destruction of many types of algae (Junli et al. 1997). This also indicates that applying chlorination in a continuous manner during the pretreatment stages may be more beneficial towards overall plant performance, as opposed to shock chlorination at high concentrations. Continuous chlorination has, however, been extensively shown to cause SWRO biofouling due to breakdown of organics into assimilable organic carbon, which is more easily used by biofouling bacteria as a nutrient source (Dow Water and Process Solutions 2015). Resosudarmo et al. (2017) showed that the increase in chlorine dose greatly increased the amounts of fouling compounds such as biopolymers, building blocks, low molecular weight (LMW) acids and LMW neutrals, with the greatest increase being for the LMW acids (Figure 9.2). While biopolymers (mostly macro polysaccharide-like and protein-like molecules) only appear to increase a small amount, they have been identified as major foulants affecting membrane filterability (Zheng et al. 2010) and thus a small increase has a major impact on fouling potential. One option to remove IOM may be the addition of powdered activated carbon (PAC) to pretreatment systems as it has been shown to limit the transmission of IOM to downstream RO membranes (Huang et al. 2015).

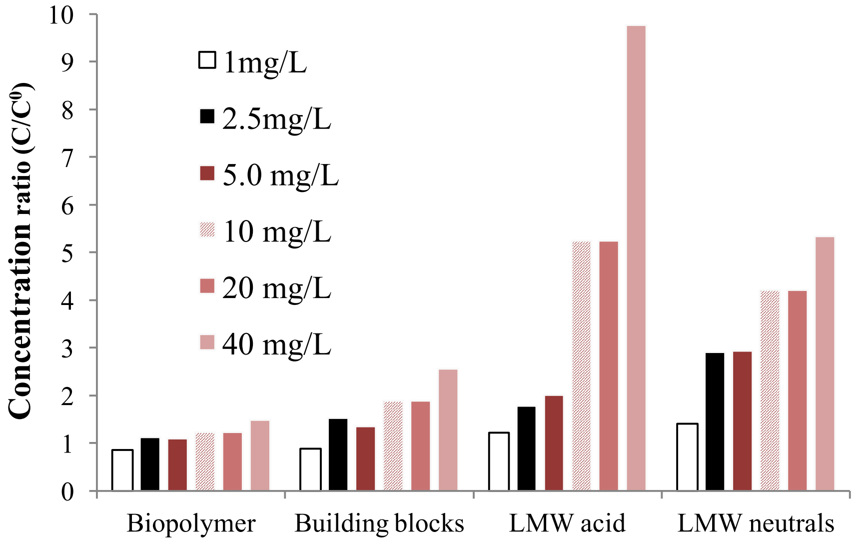


Figure 9.2. Effect of excessive chlorination on *Tetraselmis suecica* algal cells, showing a significant increase in the low molecular weight (LMW) fractions. C/C^0 is the concentration of compound in the chlorinated solution divided by the concentration in the sample prior to chlorination, thus normalizing the conditions to the baseline concentration. Modified from Resosudarmo et al. 2017.

Biofouling of SWRO membranes remains a significant detriment to successful plant operation despite the use of chlorination. The bacteria responsible for forming biofilms can survive chlorine addition through several possible mechanisms. Chlorine-resistant bacteria may become individually encapsulated in response to chlorine and be protected from its biocide effects or chlorine may promote bacterial aggregation, both of which would protect cells from biocides (Appelgate et al. 1989). The formation of bacterial aggregates is a defense mechanism in which only the outer cells are impacted by chlorine. The formation of bacterial aggregates, while it decreases the number of colony forming units (cfu), enhances the attachment of the aggregated bacteria to a surface to initiate biofilm formation (Mir et al. 1997).

Much of the organics formed during HABs act as chemical conditioning agents that modify the RO membrane surface to allow for bacterial attachment. Therefore, there is an increase in fouling potential during HABs and the use of increased concentrations of chlorine to cope with these blooms will only enhance bacterial aggregation and mucoid development. This in turn increases the rate of biofouling of the RO membranes. It is well understood that aggregated bacterial cells are more capable of tolerating environmental stress, and that survival of cells in aggregates promotes a highly clustered spatial distribution of bacteria on surfaces.

Another aspect often overlooked is the potential negative impact on the integrity of pre-treatment UF/MF membranes. While these membranes are assumed to be resistant, previous research has shown that prolonged exposure to high chlorine levels can result in accelerated membrane ageing (Regula et al. 2013). Membranes exposed to high levels of chlorine were found to have lower permeability, hydrophilicity, and tensile strength. Therefore, pretreatment membranes situated in plants frequently exposed to marine algal blooms and high chlorine levels may have shorter than expected lifetimes.

9.2.3 Summary

Chlorination of the intake can lyse HAB cells, but this may complicate downstream processes if not managed correctly. Shock chlorination leads to more aggressive lysis of HAB cells and subsequent coagulation pretreatment processes may not remove all AOM prior to the RO stage, causing biofouling. A strategy for avoiding cell lysis is to avoid shock chlorination during a HAB. A low continuous dose (0.1-0.2 mg/L) of hypochlorite may be a better approach to minimize the lysis of algal cells, while releasing some AOM to assist coagulation. More research is required to ascertain the best approach. Care should be taken when undertaking this approach to ensure RO biofouling is not inadvertently occurring. Further complications in choosing a chlorination strategy when HAB species are toxic are discussed in Chapter 10.

9.3 DECHLORINATION IN SWRO

Since polyamide RO membranes are susceptible to oxidative degradation from chlorine, dechlorination of the RO feed water upstream of the RO membranes is necessary. This is achieved by adding a reducing agent - typically SMBS. In theory, 1.34 mg of SMBS will remove 1.0 mg of free chlorine. In practice, however, 3.0 mg of SMBS is normally used to ensure complete dechlorination of 1.0 mg of chlorine (Dow Water and Process Solutions 2015). To the authors' knowledge, no references were available at the time of publication showing the direct impact of SMBS on HABs, but SMBS may lyse HAB cells; however, if pretreatment is operated efficiently, very few HAB cells will be present entering the RO. Given SMBS is routinely used to preserve RO elements for long term storage (Dow Water and Process Solutions, 2015), the direct effect of SMBS on the surface of the RO membranes may prevent biofouling to some degree.

9.4 COAGULATION FOR DAF, GMF, AND UF PRETREATMENT

9.4.1 Overview

The coagulation process is critical for removal of HAB cells and thus it is important to optimize coagulant dose to obtain a high removal of particulates and AOM (via particle destabilization and agglomeration or adsorption). As discussed in Chapter 2, high molecular weight AOM such as biopolymers, particularly very sticky TEP, have been identified as the main cause of membrane fouling rather than the algal cells themselves.

Critical to the downstream process is achieving a low iron residual to prevent iron fouling of MF/UF and RO. This section examines process parameters that may be optimized for each unit process downstream of the coagulant dosing point. For GMF/DAF/MF/UF, process conditions such as optimization of pH, coagulant dose, mixing speed, and coagulation time are discussed as well as associated operational parameters such as GMF filter rates, MF/UF flux and DAF recycle rates. This section discusses optimization of backwash frequency, and when this does not yield improved filtration conditions, how cleaning can be best undertaken (such as in the case of MF/UF). These coagulation/flocculation strategies are illustrated by the use of supporting research on simulated algal blooms to isolate and elucidate precise mechanisms for scale-up and use in plant scenarios.

A coagulant (typically a hydrolyzed metal salt) with the opposite charge to a suspended colloid is added to raw water to overcome the repulsive charge and "destabilize" a suspension.

In a colloidal suspension, particles will settle very slowly or not at all because the particles carry the same surface charges that mutually repel each other; the coagulant accelerates this particle settling process by neutralizing particle charge. For example, when colloidal particles in source seawater typically are negatively charged, ferric chloride is added as a coagulant to create positively charged ions (specifically cationic ferric chloride hydrolysis products), which attract and ultimately neutralize the charge of the seawater's suspended solid particles. Once the repulsive charges are neutralized, the van der Waals force agglomerates the particles and form micro floc. Conversely, flocculation involves the process of clumping the small, destabilized micro flocs together into larger aggregates so that they can be more easily separated from the water. Flocculation is a physical process and does not involve the neutralization of charge. Coagulation may be used in conjunction with flocculation to assist with suspended solids separation.

Ferric salts, ferric chloride, and ferric sulfate, are the best choice for seawater coagulation (Edzwald and Haarhoff 2011). When ferric chloride is introduced in the seawater, both of them form ferric hydroxide, which is a large, positively charged molecule that attracts and coagulates predominantly negatively charged seawater suspended solids particles. While aluminum sulfate and polyaluminum chlorides (PACls) have been studied extensively at laboratory and pilot-scale in seawater RO pretreatment (Gabelich et al. 2006), they are not used in full-scale applications, primarily due to the relatively high solubility of aluminum, which may result in carryover and accumulation on RO membranes leading to aluminum hydroxide fouling (Gabelich et al. 2005). Ferric chloride is less soluble over a wider pH range, resulting in lower residual dissolved iron in RO feed water and less fouling problems. Further, ferric hydroxide has a high ratio of cationic charge to total mass (Jamaly et al. 2014) that makes hydrolysis products more reactive and adsorptive with emulsified and semi-emulsified organic matter; e.g. oil and grease, natural and synthetic organic matter. The settled sludge volume of the ferric hydroxide formed from ferric chloride is reportedly 30–60% that of sulfate based coagulants (e.g. $Fe_2(SO_4)_3$). Additionally, the sludge developed from ferric chloride is generally much more dewaterable (CWT 2004).

The most important process parameters for coagulation are mixing intensity (G), flocculation shear rate (product of mixing intensity and time, G x t), pH, and temperature. Camp and Stein (1943) developed the basic theory of power input for mixing and defined G as the mean velocity gradient, which is proportional to the square root of power dissipated per unit volume of liquid. Upon coagulant addition, hydrolysis is instantaneous and the reactions that lead to the formation of ferric hydroxide occur in the order of seconds. As such, mixing has to ensure that the coagulant is fully dispersed within the liquid in the shortest time possible. Flocculation occurs by particle collision through thermally induced Brownian motion, stirring, or differential settling.

Notably, coagulation is electric-force driven attraction of the negatively charged particles of the source water by the positively charged molecules of the coagulant, (which in the case of ferric salts is a trivalent metal salt), which neutralizes the charge of suspended, colloidal, and dissolved materials so these solids no longer repulse each other and subsequently are removed by aggregation followed by sedimentation or flotation. The mechanisms of destabilization for the negative particle charge of the suspended solids naturally occurring in seawater were summarized previously by Crittenden et al. (2012):

1) compression of the electrical double layer;
2) adsorption and charge neutralization;
3) adsorption and inter-particle bridging; and
4) enmeshment in a precipitate or "sweep floc".

It is unlikely that significant changes in ionic strength would occur due to coagulant addition. Therefore, the compression of the double layer should not be the dominant mechanism in the coagulation process, especially for seawater coagulation. Adsorption and inter-particle bridging usually happen when nonionic polymers and high-molecular-weight polymers are added. In the latter two mechanisms, surface charge plays an important role in dictating the speed and effectiveness of the formation of larger particles. The speed and effectiveness of the second mechanism – formation of larger particles by physical contact (enmesh/catch or sweep flocculation) –depends mainly on the number of particles in the source water (i.e., the turbidity/total suspended solids (TSS), concentration of the seawater) and their nature (Edzward and Haarhoff 2011).

The dose of coagulant is determined by the content of solids in the source water, the electrical charge of the solid particles, and desired removal, the concentration of algae and particles, in addition to temperature, pH, alkalinity, and salinity, and other factors. In general, the higher the negative electrical charge of the particles in the source water and the less algae the seawater contains, the lower the coagulant dose needed. In addition, the higher the content of mineral particles in the seawater (i.e. the higher the turbidity/TSS concentration) the more coagulant will be needed to engage these particles in the formation of larger flocs.

If the source water particles have a strong negative charge, the dominating mechanism for large floc formation is electric attraction – therefore, relatively low doses of coagulant could achieve high coagulation effect. If the source particles do not have a strong charge, then the dose of coagulant will mainly be driven by the content of mineral suspended solids and natural organic matter (NOM) in the source water and the time the source water particles and coagulant will have to get in physical contact with each other – i.e. to collide with each other and stick together to form a larger floc. Recent research suggests that the dominant mechanism for coagulation of the marine dinoflagellate *Prorocentrum minimum* was through sweeping flocs, with charge neutralization constituting a critical step (Zhu et al. 2014). Therefore, operators often incorrectly assume that if they add more coagulant to poorly coagulating waters (i.e. waters containing particles with low or non-existent negative charge) they will improve the coagulation and filtration process.

During hydrolysis of salts, a complex of polynuclear, positively charged species are formed in a matter of seconds. During flocculation, colloidal particles and some fraction of dissolved AOM may attach to the floc body, and are eventually retained in downstream DAF, GMF, and UF processes.

Coagulation and flocculation are critical for DAF, GMF, and UF. During the DAF process, compressed air is introduced into a recycle stream of clarified water, is dissolved, and subsequently generates 10–100 µm bubbles when released through dispersion headers into a DAF tank. Coagulated particles, such as algae, attach to the bubbles and float to the top of the water column where they are mechanically or hydraulically removed. Two other removal mechanisms are adsorption, where the particles stick to the media surface, and biological removal, where soluble contaminants are removed through biological metabolism. GMF typically accumulates materials larger than 10 µm (Ripperger et al. 2012), which may include algal cells and large AOM (See Chapter 2, Figure 2.2). Since the size of some algae in seawater can be smaller than this threshold, coagulation to increase algal cell size is of critical importance to improve the removal efficiency of GMF. Unlike GMF, UF removes particles through physical straining only.

Coagulant addition is accomplished ahead of the SWRO pretreatment sedimentation tanks, dissolved air flotation units, GMF or MF/UF. The optimum coagulant dose is pH dependent and should be established on site through jar or pilot testing to provide site-specific conditions that will be encountered during plant operation. 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 a temperature of 10^oC is 8.2, while for source water temperature of 35^oC, the optimum pH decreases down to 7.4 (Edzwald and Haarhoff 2012). Other factors influencing pH adjustment are salinity and alkalinity. The acidity constants $K1^{sw}$ and $K2^{sw}$ of carbonic acid are a function of salinity and temperature. For example, $K1^{sw}$ for seawater salinity of 35,000 ppm is $10^{-5.99}$ at 10°C, and $10^{-5.76}$ at 35°C. These differ greatly from fresh water constants (ionic strength approaching 0 M), which are $10^{-6.46}$ and $10^{-6.34}$. Seawater alkalinity is mainly contributed by carbonate and bicarbonate in addition to borate.

The mechanisms proposed for coagulation of NOM are a chemical phase change or precipitation by complexation with soluble metal species for pH < 6 and adsorption to and/or enmeshment in metal hydroxide precipitates for pH > 6 (Dennett et al. 1996). In seawater, dissolved metal speciation is affected by the high ionic strength, and optimum pH values for organic matter complexation are higher than those reported in freshwater (Duan et al. 2002; Edzwald and Haarhoff 2011). For lower temperature seawater (< $20^{\circ}C$), coagulation pH of 6.5 - 7 should be effective to maximize the availability of $Fe(OH)_2^+$; however, Henderson et al. (2008a) demonstrated that AOM is characteristically different from NOM and therefore existing knowledge on NOM coagulation may not be adequate to explain on AOM fouling potential and removal in UF systems.

The formation of agglomerates in part comes about due to the negative charges (that naturally occur on the surfaces of particles, including algae, in the untreated water) becoming overcome by the addition of coagulants and sometimes polymers that neutralize surface charges, encouraging closer contact and subsequent agglomeration. Similarly, colloidal and dissolved organic matter, such as that produced by algae, can interact with coagulants, undergoing a phase change as they grow to form larger particles. In freshwater, both ferric and alum coagulants are commonly applied due to the production of positively-charged hydrolysis products (Duan and Gregory 2003); however, in seawater applications, ferric salts are the coagulant of choice as previously discussed (Edzwald and Haahoff 2011). Hence, much of the research conducted to date has focused on ferric chloride, although other ferric coagulants have been considered.

Algal cells are covered with AOM produced during metabolic activities. The charge of those molecules is likely influenced by H+/OH− ions, providing opportunities to use pH adjustment to control coagulation and optimize coagulant doses. Since cells are negatively charged, adding hydrogen ions may neutralize the negatively charged functional groups (e.g. phosphate and carboxyl). At pH lower than 5.5, cells can lyse under stress, releasing intracellular substances that may not be fully removed by downstream pretreatment and may contribute to RO membrane fouling (Zhu et al. 2014).

9.4.2 Type of coagulation feed systems

Chemical conditioning of the source seawater includes three key components: chemical feed system, coagulation, and flocculation tanks. The purpose of coagulation tanks is to achieve accelerated mixing of the coagulant with the source water and to neutralize the electric charge of the source water particles and colloids. Subsequent agglomeration of the coagulated particles into larger and easy to remove flocs is completed in flocculation tanks.

While coagulation is a relatively rapid chemical reaction, flocculation is much slower sand typically requires longer contact time and mixing conditions. Therefore, coagulation and flocculation system design requirements differ. The mixing intensity is defined by velocity gradient G (1/s). The required mixing energy expressed as G × t (t, retention time, s) is typically 4,000 - 20,000 for rapid mixing, and 30,000 - 80,000 for flocculation.

Several process conditions can be applied for coagulation prior to UF/MF, DAF, or GMF (as illustrated in Figure 9.3).

- Rapid mixing and no flocculation, i.e. inline coagulation;
- Rapid mixing and flocculation;
- Rapid mixing, flocculation and sedimentation (not common before MF/UF); and
- Rapid mixing, flocculation and flotation.

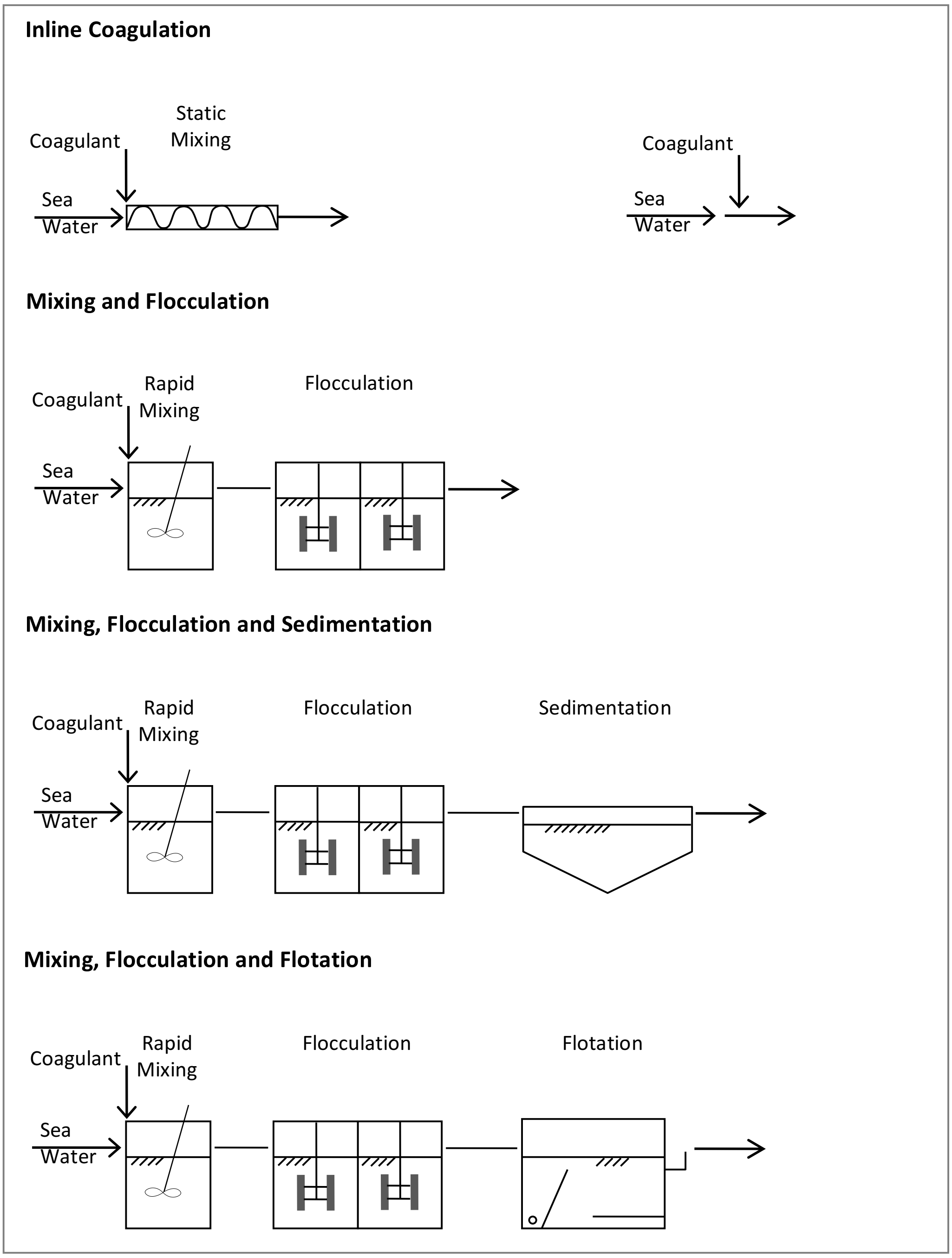


Figure 9.3. Schematic presentation of various process conditions for coagulant application with or without solids separation processes prior to MF/UF systems (Tabatabai 2014).

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 (i.e. to facilitate efficient coagulation). The two coagulant mixing systems most widely used in desalination plants are: in-line static mixers (Figure 9.4) and mechanical (flash) mixers installed in coagulation tanks (Figure 9.5).

Figure 9.4. In-line static mixer at the Adelaide SWRO desalination plant prior to UF pretreatment. Approximate flow through this mixer is 6,250 m^3/h.

Figure 9.5. Flash mixers in a coagulation tank.

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 to 2.4 m/s and are designed to operate in plug-flow hydraulics in order to provide uniform mixing within the entire pipe cross section.

Mechanical flash mixing systems consist of a coagulation tank with one or more mechanical mixers and chambers. The coagulation tank is typically designed for mixing energy G x t = 4,000 to 6,000. This type of mixing usually provides a more reliable and consistent coagulation, especially for desalination plants designed for significant differences in minimum and maximum plant production (i.e. more than 1:10).

9.4.3 Coagulation operational considerations

The increase of the coagulant dose 2-3 fold (usually applicable for clarification) is a common practice during algal bloom events and often results in deterioration rather than improvement of the downstream clarification and/or filtration process, if there are no clarification processes such as sedimentation and DAF in front of GMF and UF. Overdosing results in an excessive quantity of coagulant, which could have the undesirable effect of increasing the stability of colloidal particles and accelerating dispersion of colloids This is due to reversal of surface charge, more specifically, formation of high density positive charges on the colloids' surface and mutual electrostatic repulsion. As a result, higher silt density index (SDI) and turbidity values may be found following filtration compared to before coagulation. In addition, as the content of coagulant particles is significantly larger than that of naturally occurring suspended solids, a lot of excessive unreacted coagulant remains in the conditioned seawater. The larger flocs, coupled with inadequate mixing of coagulant with source seawater and presence of unreacted coagulant, result in accelerated clogging of the pretreatment filtration media and downstream cartridge filters (Figure 9.6) and often cause heavy fouling of the RO membranes (Figure 9.7) during algal bloom events.

Figure 9.6. Coagulant accumulation on cartridge filters due to overdosing. Photo: Voutchkov 2013.

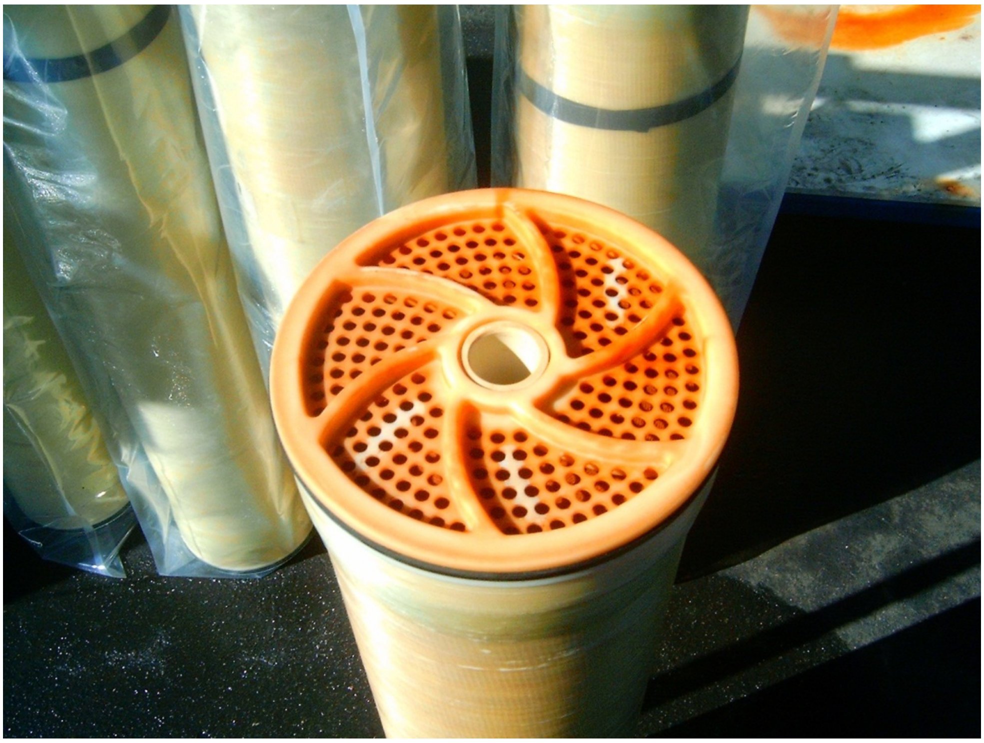


Figure 9.7. Coagulant residue on the RO membrane feed due to overdosing. Photo: Voutchkov 2013.

The effect of overdosing of ferric coagulant on the SDI can be recognized by visually inspecting the SDI test membranes. In Figure 9.8, the first two SDI test membranes are discolored as a result of coagulant overdosing, resulting in a feedwater with higher fouling potential. Hence, they were measured at the 5-minute interval (SDI_5 reading of 16.2 and 16.3) compared to the third SDI membrane showing no discoloration and measured after 15 minutes (SDI_{15}).

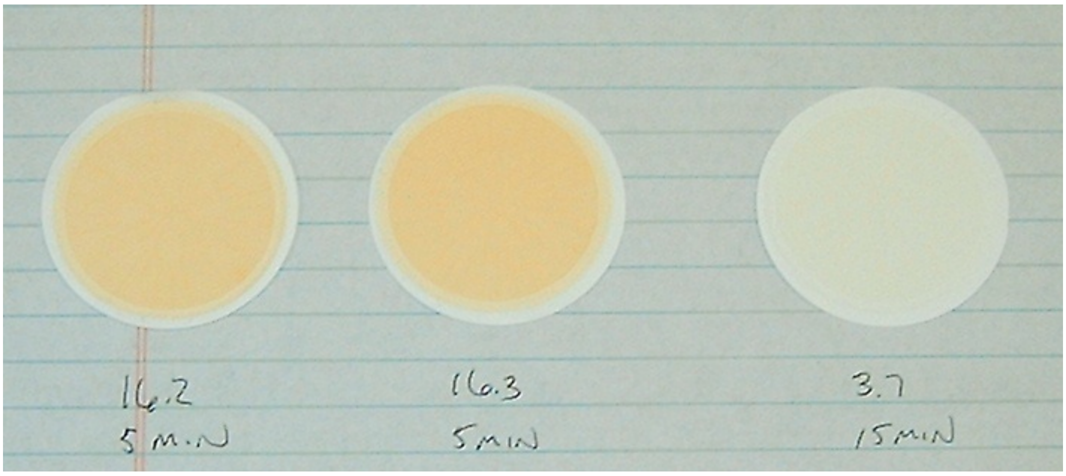


Figure 9.8. Iron accumulation on the first two SDI test membranes due to Coagulant Overdosing compared to the SDI membrane on the right hand side. SDI measurements and time interval are below the SDI membranes.

In such situations, a significant improvement in RO feedwater SDI during an algal bloom event can be attained by simply reducing the coagulant feed dose or in 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. Coagulation optimization for algal bloom is further discussed in Sections 9.4.4, 9.4.5 and 9.4.6. Jar testing is recommended each time an algal bloom occurs that causes the turbidity of the raw seawater to exceed 5 NTU or TSS to exceed 10 mg/L.

9.4.4 Coagulation and flocculation for DAF pretreatment

Good coagulation chemistry is essential to obtain favorable attachment of algal cells to bubbles generated in the DAF pretreatment system. Coagulation chemistry is the most important operating control variable affecting flotation performance. Without coagulation, the algal cells and other particles carry a negative charge. Since bubbles are also negatively charged, resultant bubble attachment is poor. Good coagulation chemistry depends upon using an appropriate coagulant dose and adjustment to a suitable pH. Optimum coagulation conditions are those of coagulant dose and pH that produce flocs with charge within an optimal operating window close to neutral, as measured using zeta potential, as this minimizes the electrostatic barrier to contact resulting from surface charges (Henderson et al., 2008b). This produces flocs with relatively high hydrophobicity, which minimizes metal hydroxide precipitates. Metal coagulant hydroxides are hydrophilic, and therefore, when sweep flocculation mechanisms dominate, bubble-particle attachment efficiency is compromised. Charge neutralization conditions are therefore preferred in DAF, but can be difficult to achieve, particularly in freshwater conditions, due to the narrow operating window (Henderson et al. 2008c). These conditions cause high bubble attachment efficiency. The increased ionic strength of seawater means that the zeta potential is not as extreme as in freshwater systems due to electrical double layer compression that should in fact make coagulation easier.

In a DAF system, removal of *Prorocentrum minimum* increased by ~5-10% in a study by Zhu et al. (2014) at 60 and 50 mg/L ferric chloride doses, respectively when pH was adjusted to

6.4 and 6.3. Adjustment of pH maintained algal removal at over 90% even with 30 mg/L of ferric chloride.

The formation of agglomerates in part comes about due to the negative charges that naturally occur on the surfaces of the particles (or algal cells) and bubbles in the untreated water becoming overcome by the addition of coagulants and sometimes polymers which neutralize the surface charges encouraging close contact and bonding.

The need for effective surface charge neutralization and coagulation, whilst not peculiar to DAF, requires the chemistry to be optimized both in terms of the coagulant dose and pH. Ease of adjusting the pH by a mineral acid is dependent on the "buffer intensity". This term can be simply defined as the resistance of pH to change. The buffer intensity arises from the ionic strength of both inorganic carbon and alkalinity or borate in freshwater or seawater, respectively.

The resistance is greater in seawater and increases as the water temperature falls, resulting potentially in higher doses of chemicals (e.g. sulfuric acid or hydrochloric acid when downstream RO necessitates low sulfates) to achieve the optimal coagulation pH. Dose rates should be calculated post jar testing.

As noted by Edzwald (2010), floc sizes in the range of 25–50 μm (pin point flocs) were of the optimal size to achieve high floc-bubble collision efficiency and for separation of floc bubble aggregates. The majority of marine bloom-forming algae (dinoflagellates, diatoms, and cyanobacteria), fall into this size.

9.4.5 Coagulation for GMF pretreatment

Coagulation combined with granular media filtration is the most commonly used method for seawater pretreatment at present. The severity of algal blooms in the area of the intake, as well as the size and charge of the algae cells most commonly occurring during algal blooms, have a significant impact on the sizing of these facilities and their efficient operation.

Conventional (or GMF) pretreatment technology is based upon removal of suspended solids and some organics through coagulation and flocculation. The process is well established and in the majority of cases is capable of producing SWRO feedwater of the required quality with respect to suspended solids, SDI and turbidity and is thus very effective as a pretreatment. The quality of the treated water varies significantly with quality of the raw water; however, deterioration of raw water quality will affect operation of the filtration system. In conventional pretreatment systems, coagulation is mainly applied to improve surface loading rates and ensure that product water quality meets the requirements of RO membrane manufacturers in terms of turbidity and SDI_{15}. Coagulant dose in conventional SWRO pretreatment systems may range from 0.5 to 10 mg Fe/L, although in some cases doses as high as 20 mg Fe/L have been reported during poor water quality events (Lattemann 2010; Edzwald and Haarhoff 2011).

Use of coagulants is critical for the effective and consistent performance of GMF pretreatment filtration systems; however, if the source water contains low turbidity (< 0.5 NTU) and the prevailing size of particles is less than 5 μm (which is common for deep intakes with low algal content), coagulant addition does not yield a significant improvement in the GMF process. In this case, the addition of a minimal amount of coagulant (i.e. 0.5 mg/L or less) or even no coagulant addition is viable. In such conditions, it is critical to have a prolonged period of coagulation and flocculation (i.e. coagulation and flocculation times of 10 minutes or more), because for these particles, the main mechanism for floc formation is physical contact rather than charge attraction (Voutchkov 2013).

9.4.6 Coagulation for MF/UF pretreatment

In contrast, to GMF and DAF, MF/UF systems do not rely on coagulation to enhance permeate quality in terms of turbidity and SDI as particles as fine as 0.04 µm (MF membranes) or 0.01 µm (UF membranes) can be removed without coagulation. UF membranes are generally preferred over MF in SWRO pretreatment due to better removal of particulate/colloidal organics, silt, and pathogens from seawater owing to their smaller pore size (Voutchkov 2009). Operating without coagulant addition offers many advantages – reducing process complexity and costs. Operators give preference to systems that require no coagulant or if this is not achievable, minimum amounts of coagulant. Minimizing or eliminating coagulant addition, while maintaining stable process performance and high permeate quality, can be achieved by optimizing process parameters or applying alternative coagulation process conditions. Operating without coagulant also avoids potential environmental impacts associated with the use and disposal of pretreatment chemicals such as coagulants, coagulant aids, and others. (WHO 2007). In areas with more stringent legislation on brine discharge, such as Europe, Australia, and the USA, coagulant-rich waste streams require extensive treatment and handling prior to discharge, which add a significant cost component to the overall pretreatment process. In such cases backwash water containing coagulant is treated separately (e.g. by gravity settling in lamella plate sedimentation tanks). Supernatant can be either disposed with RO concentrate or recycled at the head of the pretreatment. The coagulant-rich sludge retained in the sedimentation tank is often dewatered onsite and transported to sludge treatment facilities or landfills (WHO 2007).

Potential impacts on the environment associated with the use and disposal of pretreatment chemicals such as coagulants, coagulant aids, and others increase process complexity (WHO 2007). Coagulants and coagulant aids (high molecular weight organics, e.g. partially hydrolyzed polyacrylamide) present in spent backwash water are typically discharged to the ocean without treatment (Lattemann 2010). Ferric chloride has very low toxicity for marine organisms; however, discharge may cause an intense discoloration of the reject stream (red discoloration of the concentrate), which may increase turbidity and reduce light penetration, or could bury sessile benthic organisms at the discharge site (Lattemann and Höpner 2008).

In some cases, coagulation is required upstream of MF/UF seasonally or in response to poor water quality events to avoid excessive fouling. For example, coagulation may be necessary if the source water contains NOM particles with strong negative charge that could be coagulated easily and removed via filtration; 2) during heavy algal blooms (for AOM removal); or 3) during oil spill events. Coagulant dosing prior to MF/UF filtration can greatly reduce AOM passing through the pretreatment membranes, in particular UF membranes, which can promote biofouling in the RO (see Chapter 2). Moreover, coagulation can reduce pore blocking and/or surface attachment by sticky particles such as biopolymers produced during an algal bloom, enhancing cake filtration. This will reduce non-backwashable fouling of MF/UF membranes and pressure increase (Guigui et al. 2002; Choi and Dempsey 2004; Schurer et al. 2013). If the source water is consistently high in suspended solids and organics, then additional pretreatment steps may be applied upstream of the MF/UF to reduce loading onto the membranes and to achieve higher membrane fluxes.

Coagulation is commonly applied using an inline mode in MF/UF systems for SWRO pretreatment and sometimes with DAF ahead of the MF/UF. Inline coagulation is the application of a coagulant without removal of flocs through a clarification step. Inline coagulation may also be characterized by the absence of a flocculation chamber, as large floc size for enhanced settling is not a requirement in UF systems. The absence of flocculation and clarification steps in the overall process scheme results in lower investment cost for inline coagulation as compared to conventional schemes that consist of coagulation/flocculation/sedimentation/flotation. In most inline coagulation applications, mixing is achieved in a static mixer or the wet well of the UF feed pump station. Flocculation may occur during the mixing step (simultaneously with coagulant dispersion and hydrolysis), in the piping network that carries the coagulated feedwater to the membranes, or within the UF capillaries.

As discussed previously, Henderson et al. (2008a) demonstrated that AOM is characteristically different from NOM and therefore existing knowledge on NOM coagulation may not be adequate to explain the effect of coagulation on AOM fouling potential and removal in MF/UF systems. Tabatabai (2014) therefore carried out a series of laboratory-scale experiments to optimize AOM removal by pressure-driven inside–out UF (150 kDa) membranes using inline coagulation and to compare removal with conventional pretreatment. In these experiments, AOM was harvested from the marine diatom species *Chaetoceros affinis* to simulate seawater bloom conditions, creating feed solutions at a concentration of 0.5 mg C/L as biopolymers in synthetic seawater. The microalga *Chaetoceros* was selected as it is known to produce large quantities of extracellular polysaccharides throughout their growth cycle (Watt 1968; Dam and Drapeau 1995; Myklestad 1995), making it a suitable choice for laboratory-scale production of AOM in short-term experiments.

In this series of experiments, the optimal ferric coagulant dose and pH for removal of biopolymer $TEP_{0.4}$ (a fraction of biopolymer), and dissolved organic carbon (DOC) were investigated (Figure 9.9). (See Chapter 5 for information on tests to measure biopolymers, TEP, modified fouling index - UF (MFI-UF) and high resolution liquid chromatography – organic carbon detection (LC-OCD)). As expected, more AOM was removed with increasing ferric coagulant dose to a point at which there were diminished returns, which defines the optimum dose (10 mg Fe/L). Two pH values were investigated, a pH typical of seawater (8), and seawater acidified to a pH of 5, due to the two coagulation mechanisms explained previously (Section 9.4.1). Laboratory-scale results from Tabatabai (2014) demonstrated that pH did not greatly affect coagulation efficiency of AOM in terms of removal of biopolymers (Figure 9.10).

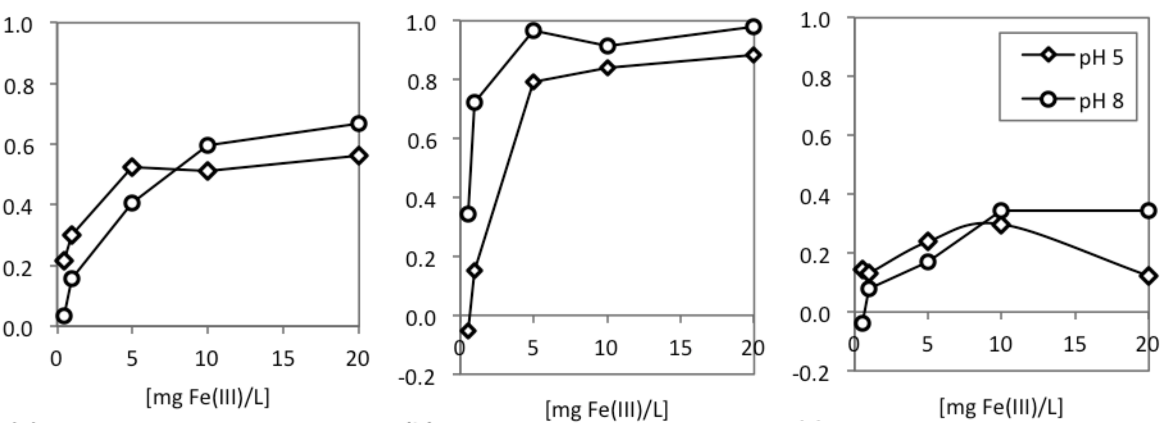


Figure 9.9. Removal rates of biopolymers (left panel), $TEP_{0.4}$ (settled samples; center panel;) and DOC as a function of coagulant dose and pH (right panel). Figure: Tabatabai 2014.

Mixing intensity is also a factor for consideration in maximizing AOM removal and thus reducing fouling on the RO membrane. In addition to AOM removal, the MFI-UF (see Chapter 5, Section 5.5.2) enabled the effect of varying process parameters such as mixing intensity and flocculation time on the fouling potential of the RO feedwater to be assessed. At a coagulant dose of 1 mg Fe(III)/L (Figure 9.10), mixing intensity of 1100 s^{-1} resulted in substantially lower MFI-UF values than for 100 s^{-1}. Mixing time, however, did not affect filterability of coagulated AOM flocs, as no difference was observed in MFI-UF values for 20 s versus 240 s of mixing time.

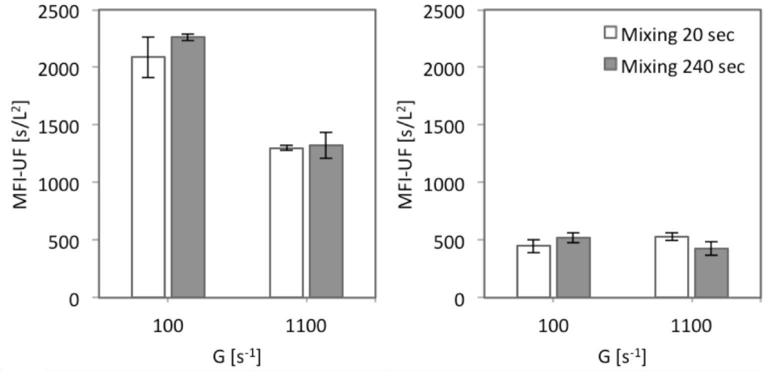


Figure 9.10. Effect of mixing intensity (G) and mixing time on fouling potential of coagulated AOM for 1 mg Fe(III)/L (left panel) and 5 mg Fe(III)/L (right panel). Figures: Tabatabai 2014.

The mode of coagulant dosing was investigated by a variety of pretreatment steps in another series of experiments using *Chaetoceros affinis*, as described previously (Tabatabai 2014). High resolution LC-OCD was employed to investigate removal of biopolymer fractions to ascertain the best coagulant dosing mode.

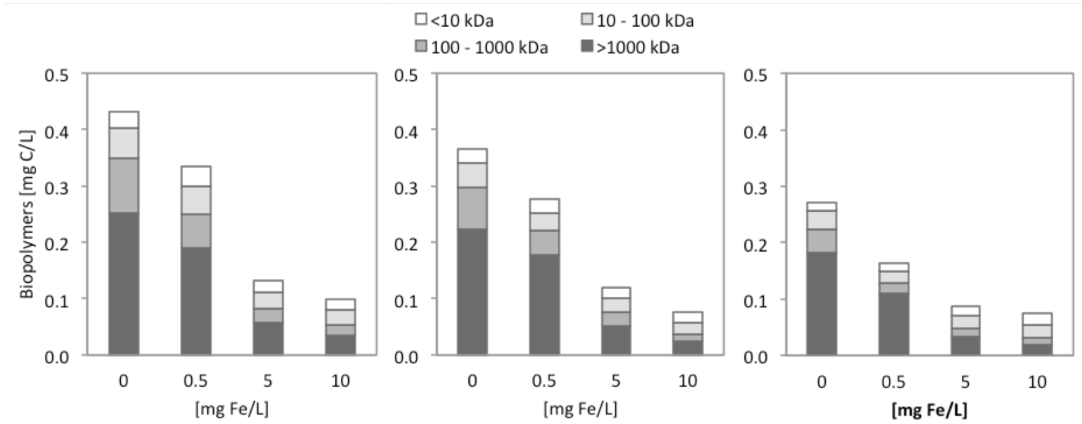


Figure 9.11. Biopolymer concentrations for fractions of different molecular weight as a function of coagulant dose for (a) Mode A-coagulation(coag)/flocculation(flocc)/sedimentation(sed), (b) Mode B - coag/flocc/sed/0.45 µum, and (c) Mode C-inline coag/UF. Figures: Tabatabai 2014.

Higher MWCO biopolymer fractions of the AOM were well removed using inline coagulation with UF (150 kDa) and also in the experiments designed to simulate coagulation with flocculation coupled to different levels of pretreatment (i.e. sedimentation and sedimentation followed by 0.45-µm-membrane filtration (Figure 9.11).

As mentioned previously, coagulation can enhance cake filtration on the UF membrane thereby, reducing non-backwashable fouling and pressure development in MF/UF systems. The effect of coagulation on fouling propensity and removal of AOM in pressure driven inside–out UF membranes from the laboratory-scale experiments of Tabatabai (2014) and the findings from the Jacobahaven demonstration plant (see Case Study 11.10) where *Chaetoceros* was also found are further discussed in the following sections.

a) Fouling potential and compressibility of AOM. Coagulation can reduce transmembrane pressure (TMP) increase during filtration of algae-laden feedwater in pressure driven inside-out (PDI) UF membranes. When coagulant is dosed to seawater containing high concentrations of AOM, iron reacts with the biopolymers, changing the properties in such a way that the fouling potential of AOM is improved (i.e. lower MFI-UF values; Figure 9.12

top) and the AOM cake/gel layer becomes less compressible (i.e. more linear TMP curves, Figure 9.12 bottom). This effect was demonstrated for a synthetic seawater solution of AOM harvested from *C. affinis* to simulate algal bloom conditions. Inline coagulation was performed at different coagulant doses prior to filtration through PDI UF membranes with nominal MWCO of 150 kDa.

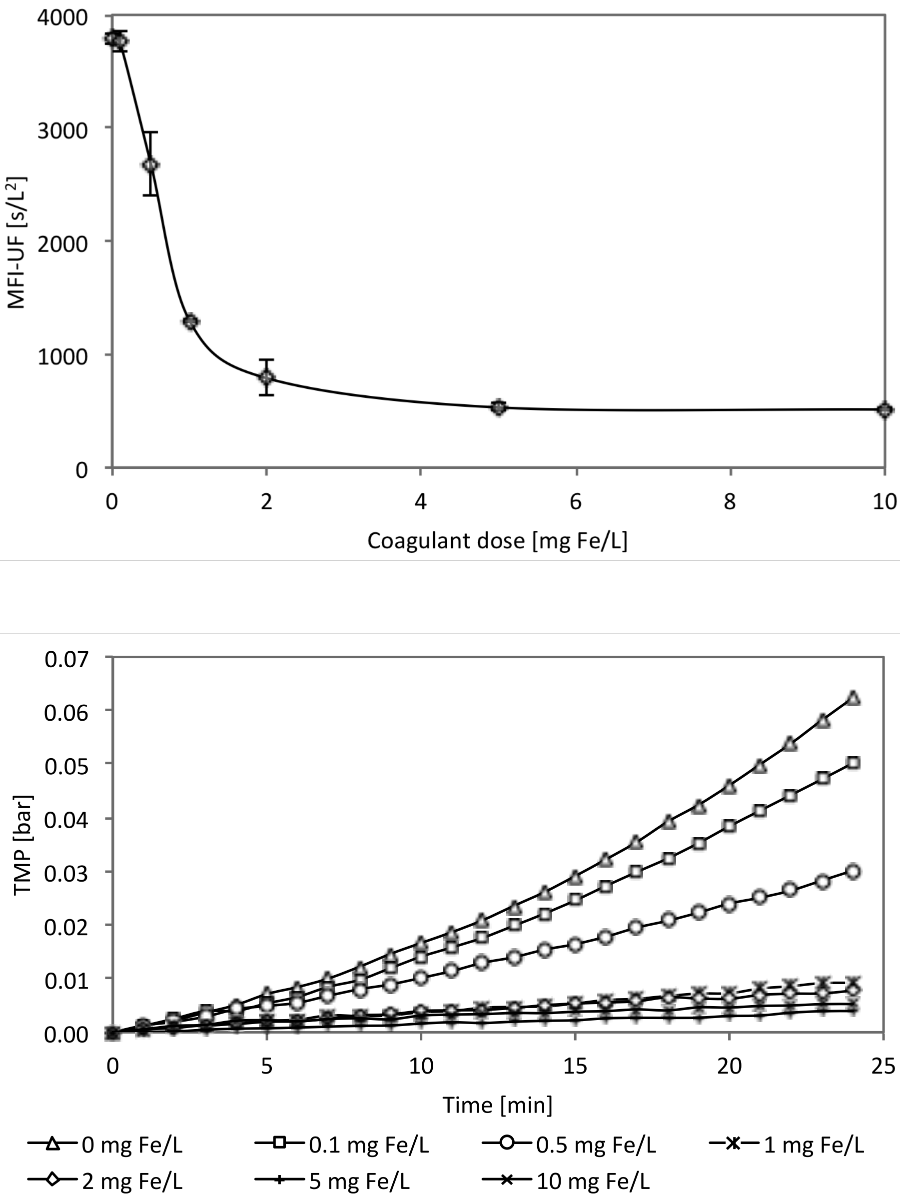


Figure 9.12. Effect of inline coagulation on fouling potential as measured by MFI-UF (top panel) and TMP development during filtration of algal laden seawater (0.5 mg C/L as biopolymers obtained from *Chaetoceros affinis*) through 150 kDa UF membranes at 100 L/m^{2}h. pH ranged from 8.0 to 7.1. (bottom panel). Figures: Tabatabai et al. 2014.

At coagulant doses < 0.5 mg Fe/L (pH = 8.0), the concentration of positively charged iron hydroxide flocs species is too small for any reaction to occur and hence AOM fouling potential and compressibility were not affected. At low coagulant dose (< 1 mg Fe/L) in natural seawater pH (~ 8), colloidal Fe-biopolymer complexes were formed and a slight reduction in fouling potential and compressibility were observed. At coagulant doses of 0.5-1 mg Fe/L (pH 7-8), AOM adsorption on iron hydroxide precipitates occurred, resulting in the formation of iron-biopolymer aggregates that were relatively large and less compressible. At higher coagulant doses, the cake/gel layer properties tended toward iron hydroxide flocs. Residual iron in all UF permeate samples was below the detection limit (20 μg Fe/L) (Tabatabai et al. 2014).

b) Fouling reversibility. Coagulation can reduce the extent of hydraulically irreversible fouling by AOM in PDI UF membranes. This was demonstrated at the Jacobahaven demonstration-scale on North Sea water during several successive bloom periods (Schurer et al., 2012, 2013). Ferric chloride dosed prior to the UF feed pump at an average dose of 0.5-1.5 mg Fe/L stabilized UF operation and reduced the frequency of chemically enhanced backwashing (CEB) at a nominal flux of 60 L/m^{2}h (Case Study 11.10). The efficiency of CEB in recovering membrane permeability was significantly reduced when coagulant was applied, indicating UF fouling by residual iron. Under such conditions, membrane permeability could only be restored by applying tailored CIP.

An alternative mode of coagulant application (i.e. coating) in seawater has shown promising results at laboratory-scale in terms of UF hydraulic performance at very low coagulant dose (0.5 mg Fe/L). In this process, a layer of preformed flocs of iron hydroxide (H_2FeO_3) is dosed at the start of each filtration cycle to create a protective barrier that prevents the attachment of sticky AOM (such as TEPs) to the membrane surface (Figure 9.13). The protective layer should be formed in a short amount of time at the start of the filtration cycle in order to prevent/minimize membrane-foulant interactions, and should not alter the intrinsic membrane permeability. Thereafter, seawater is filtered directly through the coated membranes. At the end of each filtration cycle, backwashing is applied whereby the coating layer containing AOM (including sticky TEPs) is lifted off the membrane surface and flushed out. Handling and treatment of spent coating material follows the same procedure as that of coagulated sludge. For the process to be successful, the coating layer should be highly permeable, so as not to reduce system efficiency, and easily backwashable.

Laboratory-scale experiments were conducted on feedwater containing AOM obtained from

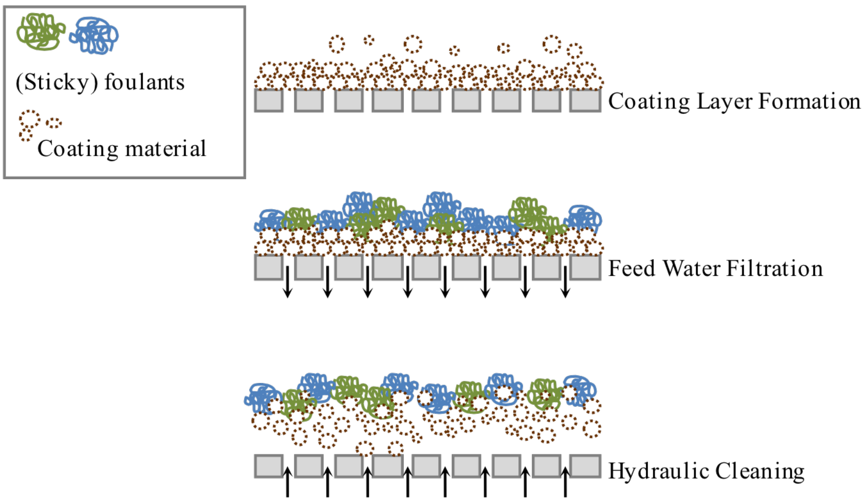


Figure 9.13. Simplified schematic presentation of UF coating by iron hydroxide particles to enhance hydraulic backwashing. Figure: Tabatabai 2014.

Chaetoceros affinis (as described previously) in synthetic seawater (total dissolved solids (TDS) = 35,000 ppm) to simulate bloom conditions in the North Sea. Coating suspensions with a range of particle size were created by precipitation of iron hydroxide and subsequent grinding at various intensities. Application of a coating layer prior to filtration of seawater with high AOM concentration (0.2 – 0.7 mg C/L) stabilized operation of PDI UF membranes with a nominal MWCO of 150 kDa, by significantly enhancing backwashability (Figure 9.14). Reducing particle size of the coating material to the submicron range (400-700 nm) allowed for a significant reduction in coating dose. Coating with nanoparticles of iron hydroxide, allowed for continuous stable operation at equivalent dose of 0.5 mg Fe/L. This is a significant improvement to inline coagulation in terms of required coagulant dose. coagulant dose. Furthermore, creating preformed flocs through precipitation and subsequent grinding may reduce the risk of UF fouling by residual iron (Tabatabai 2014).

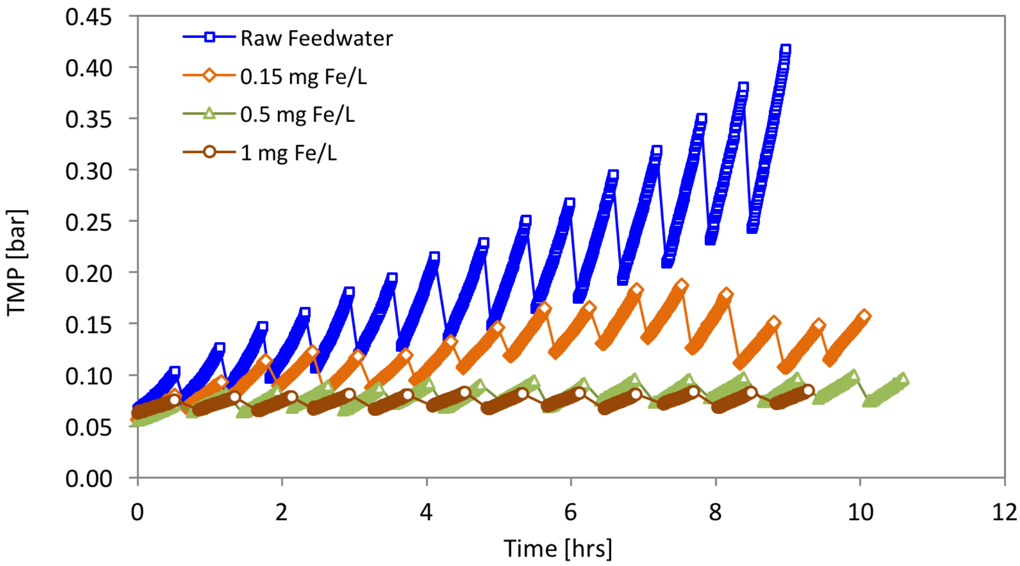


Figure 9.14. TMP development during filtration of algal laden seawater (0.5 mg C/L as biopolymers obtained from *Chaetoceros affinis*) through 150 kDa UF membranes coated with preformed iron hydroxide flocs at different equivalent dose, filtration flux 100 L/m^{2}h. Figure: Tabatabai et al. 2014.

At coagulant dose < 0.5 mg Fe/L (pH = 8.0), the concentration of positively charged iron hydroxide flocs species is too small for any reaction to occur and hence AOM fouling potential and compressibility were not affected. At low coagulant dose (< 1 mg Fe/L) and natural seawater pH (~ 8), colloidal iron-biopolymer complexes were formed and a slight reduction in fouling potential and compressibility was observed. At coagulant dose of 0.5-1 mg Fe/L (pH 7-8), AOM adsorption on iron hydroxide precipitates took place, resulting in the formation of iron-biopolymer aggregates that were relatively large and less compressible. At higher coagulant dose, the cake/gel layer properties tend toward the properties of iron hydroxide flocs. Residual iron in all UF permeate samples was below detection limit (20µg Fe/L) (Tabatabai et al. 2014).

c) Permeate quality. In SWRO plants, coagulant dosing prior to UF filtration can greatly reduce AOM flux through the UF and the seeding of biofouling in the RO (see Chapter 2). Commercially available PDI UF membranes based on polyethersulfone (PES) with nominal MWCO of 150 kDa can remove up to 45% of algal biopolymers compared to 25% when simulating conventional coagulation using 0.45 µm filtration (Figure 9.15). Inline coagulation at 0.5 mg Fe/L enhanced biopolymer removal by approximately 20%. At 5 and 10 mg Fe/L, biopolymer removal was further enhanced by 20%, resulting in a biopolymer removal of approximately 85%. Inline coagulation/UF exhibited superior biopolymer removal compared

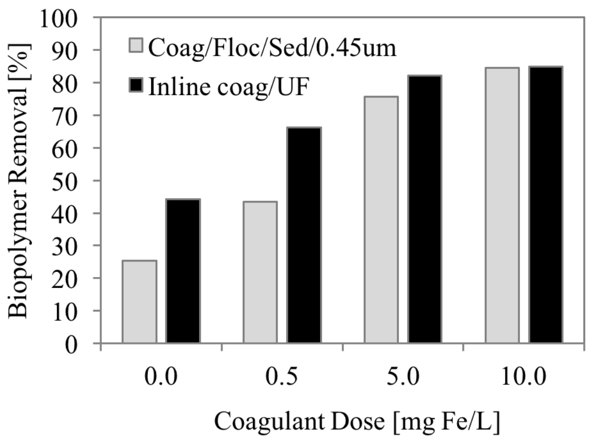


Figure 9.15. Removal of algal biopolymers as a function of coagulant dose for coagulation/flocculation/ sedimentation followed by 0.45 μm filtration and inline coagulation/UF.

to conventional coagulation/0.45 μm filtration at low coagulant dose (i.e., 0.5 mg Fe/L); however, this difference became marginal at higher coagulant dose, such that at 10 mg Fe/L no difference was observed in biopolymer con-centration between conventional pretreatment and UF pretreatment. Increase in dose shifts the predominant coagulation mechanism to sweep floc and inter-particle bridging, whereby removal is enhanced by adsorption to and/or enmeshment in precipitated iron hydroxide. Biopolymer removal was mainly through the removal of fractions larger than 100 kDa (Tabatabai et al. 2014).

Biopolymers were mainly composed of larger molecular weight fractions; approximately 80% of the total was larger than 100 kDa. As a consequence, reduction in biopolymer concentration was mainly due to the removal of compounds >100 kDa. Coagulation had a strong impact on the removal of biopolymers larger than 100 kDa for UF and conventional pretreatment simulated in laboratory scale experiments where removal was substantially higher at higher coagulant dose. In contrast, to GMF and DAF, MF/UF systems do not rely on coagulation to enhance permeate quality in terms of turbidity and SDI.

Higher MW biopolymer fractions of the AOM were well removed for inline coagulation with UF (150 kDa) and in experiments designed to simulate coagulation with flocculation coupled to different levels of pretreatment (e.g., sedimentation and sedimentation followed by 0.45 μm media filtration; see Figure 9.11 and Figure 2.2 in Chapter 2 for more details). Tabatabai (2014) showed optimum removal doses for ferric coagulant and pH to remove biopolymer, the fraction of biopolymers measured by $TEP_{0.4}$ (see Chapter 5 Section 5.3.1.2) and DOC, where feedwater concentrations were 0.55-0.63 mg/L, 0.27 mg/L and 1.7 mg/L respectively (Figure 9.9). The optimum dose from this experiment was 10 mg/L of ferric coagulant at pH 8, although pH was far less important for optimization than the ferric coagulant dose.

9.5 DAF PRETREATMENT FOR SWRO

9.5.1 Overview

DAF has been used in drinking water treatment since the 1960s for the removal of low density suspended solids and organics and for reducing turbidity. The performance of DAF is dependent on the preceding agglomeration (coagulation/flocculation) step. Unlike sedimentation however, where the aim is to generate large flocs to facilitate sedimentation, DAF does not require large flocs, as removal is achieved by floating floc-bubble aggregates. That coupled to technological developments has led to lower flocculation times, coagulant consumption, and sludge production.

From its earliest use, DAF was found to be highly effective in treating a variety of algal rich sources including freshwater and wastewater. For example, even as early as 1975, Hyde (1975) reported that waters containing 30,000,000-150,000,000 cells/L could be treated using flotation with flocculation periods of only 7-9 minutes. A consequence of these reduced flocculation times is that there is a saving not only in terms of space but also in the cost of the civil structure and mechanical and electrical equipment, as smaller flocculators can be used. This work was further expanded and reported by Valade et al. (1996) and Edzwald et al. (1999) confirming reduced flocculation times and, more significantly, that previously typical clarification loading rates of 10 m/h (referred to as conventional rate DAF) could be increased. Over the intervening years, further development leading to improved understanding in the application of ever higher loading rates has continued, achieving 50 m/h (referred to as high-rate DAF) with as little as 5 minutes of total flocculation time, as reported by Amato et al. (2012), albeit at pilot scale. Practically, DAF is still commonly designed with two stages of flocculation to optimize performance and prevent short circuiting during mild algal blooms or normal operations, which usually need longer retention times, especially if the source water contains relatively low turbidity.

In the 1990s, conventional-rate DAF was employed as part of the pretreatment scheme to treat algal blooms at a small scale SWRO plant at the Gas Atacama power station in Chile (see Case Study 11.7). High-rate DAF (Rictor /Aqua DAF) was later trialed at Taweelah in the Gulf in a 2002 pilot plant study prior to two stage GMF. This demonstrated that the required RO feedwater SDI_{15} could be obtained and that emulsified oil was removed in spiked tests (Rovel 2003). DAF was not pursued further at that time as algal cell counts remained < 100,000 cells/L. Subsequently, DAF and dissolved air flotation and filtration (DAFF) were employed at the El Coloso (Chile) and Tuas 1 (Singapore) SWRO desalination plants, respectively in the first large-scale applications of DAF/DAFF in SWRO pretreatment. Algal blooms were the main driver for including high rate DAF in the pretreatment scheme for the El Coloso plant. In the case of the Tuas 1 plant, DAFF (Figure 9.16) was primarily installed because of the potential presence of high solids (up to 60 mg/L) and oil (up to 10 mg/L), not for algal removal. Since, that time DAF is increasingly being incorporated in large-scale SWRO pretreatment schemes prior to GMF or UF for treating algal blooms, particularly in the Middle East, such as the Shuwaikh plant in Kuwait (see Case Study 11.5) with a pretreatment capacity of 350,000 m^3/d.

Figure 9.16. Dissolved air flotation and filtration (DAFF) installation at the Tuas, Singapore, SWRO desalination plant. Photo: PUB Singapore.

DAF is important in removing algal cells and reducing the suspended solids load for downstream pretreatment processes, as the DAF separation process is ‘gentle’ or low shear, thereby reducing cell lysis and release of fouling organics and algal toxins, mitigating the risk of AOM fouling in UF and RO. Removal of intact algae also assists in reducing taste and odor issues associated with algal blooms (see Chapter 10). The role of DAF in seawater RO systems and important parameters that can be optimized during an algal bloom are discussed in more detail below. Figure 9.17 shows a DAF float layer when treating a cyanobacterial (freshwater) bloom.

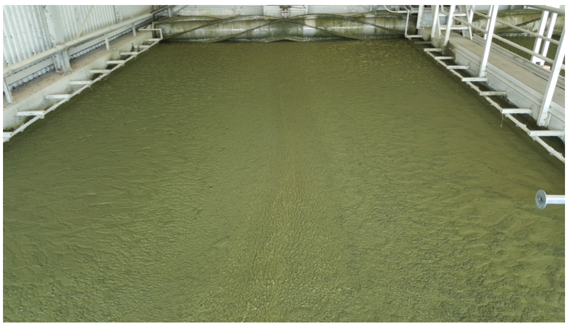


Figure 9.17. DAF float layer when treating a cyanobacterial bloom. South Australian Water Corporation Bolivar WWTP DAFF plant. Photo: Biomass lab, UNSW and SA Water.

9.5.2 Fundamental principles of DAF

The fundamental principle that has given rise to the development of DAF is that of enhancing the natural buoyancy of the particulates carried within a fluid by attaching them to micro-bubbles to encourage separation. The air used in the DAF process for freshwater applications is normally dissolved under a pressure of 400-600 kPa into a proportion of the previously clarified flow, termed the recycle (Edzwald 2010). The recycle rate applied will vary depending on the nature of the flow to be treated and in freshwater may range between 8-12%. The fundamental difference between fresh water and seawater (with all other conditions being equal), is the higher salinity of seawater, typically in the range of 35,000-45,000 mg/l total dissolved solids, which reduces the amount of air that can be dissolved. Henry's constant for the main gases that make up what is called air, (i.e. argon, nitrogen and oxygen) are all higher in seawater by approximately 30% in all cases and all temperatures, meaning they are all less soluble. This is sometimes referred to as the "salting out" affect and results in either a requirement for an increase in pressure of ~30% or the recycle rate having to be increased by ~20% (Haarhoff and Edzwald 2013). In practice, it is normal to increase these two variables together. For example, if for fresh water a 10% recycle is used then, on a seawater plant, the recycle rate would be increased to 12%. The pressurized recycle flow is then passed through various types of pressure reducing devices such as needle valves or fixed orifice nozzles resulting in the immediate release of a cloud of micro-bubbles within the contact zone of the flotation tank. It is within this contact zone that the bubbles and flocs to be removed are intermixed and floc-bubble aggregates are formed (Figure 9.18).

Historically there have been attempts to set feedwater quality limits (TSS, turbidity, algae cells, oil, and grease). Jansenns and Buekens (1993) suggested that the application of DAF be limited to the source water turbidity less than 100 NTU; however, the performance and application of DAF is dependent on the nature of the solids or pollutants to be removed and ensuring that the correct water chemistry conditions are applied during coagulation-flocculation at all times to maximize removal efficiencies.

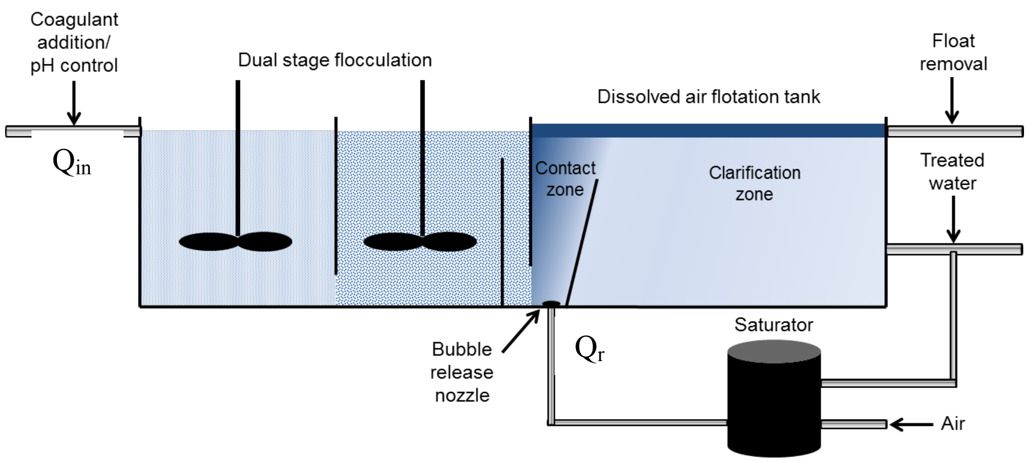


Figure 9.18. Schematic of a DAF system including dual stage flocculation.

DAF removal rates for algal cells are dependent on bloom cell concentrations - the higher the cell concentration, the higher the removal rate, similar to that observed for TSS and turbidity removal. For instance, DAF pilot trials treating waste stabilization pond effluent with algal cell concentrations averaging 3×10^8 cells/L showed that at least 99% of cells could be removed in the float when preceding coagulation-flocculation was optimized (Yap et al.

2012). In a seawater desalination context, however, pilot studies conducted using DAF have typically been undertaken at low algal concentrations (e.g. less than 100,000 cells/L (Bonnelye et al., 2004) and less than 3 μg/L chlorophyll *a* (Kim et al. 2011)) and therefore removal rates are expected to be lower. The type of algae will also impact removal efficiencies. For example, in drinking water applications the morphology (size and shape) of algal cells have been shown to impact treatability. Spherical cells < 5 μm or needle-shaped cells have been most difficult to remove by DAF, which is considered to be a result of the cell morphology making coagulation more difficult combined with the fact that they are likely to settle vertically, leaving only the narrow tip of the needle (< 5 μm) for collision by the bubble to (Henderson et al. 2008c; Konno 1993). Some motile species have been observed to swim out of flocs, meaning that flagellated species also tend to have a low removal rate. A small amount of pre-oxidation can inactivate motile species (Henderson et al., 2008c); however, great care is required to avoid excessive AOM release (see Section 9.1 on chlorination/dechlorination). Furthermore, AOM that is released by algal cells has been shown to impact treatability by DAF, as AOM concentration and character is species dependent (Henderson et al. 2010; Villacorte et al. 2015). For example, AOM has been observed to hinder coagulation by chelating coagulant, increasing the dose required, while other studies have shown that if its character is enriched in biopolymers, it can act as a bioflocculant and thus enhance flocculation via bridging mechanisms (Henderson et al. 2010; Pivokonsky et al. 2016). Further investigations are required for seawater applications to confirm that this also occurs for higher salinity feedwater.

9.5.3 Process design of DAF systems

The typical DAF tank is split into two primary sections: the "contact" and the "clarification" zones. The first, as the name suggests, is where the air is released from the air-saturated recycle flow through an arrangement of headers and nozzles or needle valves, forming a profusion of micro bubbles which, as they rise, intimately mix and attach to the floc carried through by the bulk flow. The total flow including micro bubbles and algal cell flocs normally exit the contact zone over a baffle, generally referred to as the "incline baffle", which in practice can actually be vertical. The design of this baffle forming the downstream boundary of the contact zone is critical to the design of the operation of the DAF tank as poor "contact" in this first zone will result in poor performance overall. The actual average retention time in the contact zone is typically ~60 seconds. The second zone in conventional designs is typically ~80% of the total tank volume and is where the air bubble agglomerates, with a density lower than water, formed via the contact zone are allowed time to rise to the surface where algal bloom agglomerates accumulate as floated sludge (Figure 9.18). The float is removed through a mechanical skimming unit (mechanical removal) or by solids overflow to the collection through (hydraulic removal). Mechanical removal results in a waste stream with solids concentration of 2 – 3%. The hydraulic removal produces wastewater with lower solids concentration in the rate of 0.5 – 1%. Typically, DAF tanks are covered to prevent disturbance of the float from wind and rain.

The excess air at this point (there should always be excess air) provides general buoyancy to the floated sludge and together with the hydrodynamic flows set up by the design, acts as a barrier preventing short circuiting and as a "filter". This filter layer is what is sometimes called the "whitewater" layer and comprises a range of bubbles sizes that are continually moving both vertically and horizontally. It is the management through proprietary designs of this whitewater layer that can impact the overall performance of the DAF system as the "whitewater" layer serves a number of functions and is not simply providing air to float flocs to the surface. It is for this reason that the use of the air:solids ratio to determine air dose (as used in sludge thickening applications) is not appropriate for seawater conditions, a TSS of several hundred is considered very high for SWRO pretreatment applications.

The size of bubbles regarded as ideal for freshwater DAF applications have traditionally been in the range of 10-100 µm with most having sizes of 40-80 µm (Edzwald 2010). This appears to be the case for desalination applications as Kim et al. (2011) also reported bubble sizes in the range of 10-100 µm when using seawater. It was reported that the mean bubble size in saline water was smaller than that in fresh water because of such factors as higher surface tension, higher ionic strength, and higher density of seawater (Besson and Guiraud 2012). The concentration of air bubbles (bubble density) will vary depending on a number of factors including sizes. These factors can include, but not be limited to, the recycle rate (that can vary typically in the range of 6-20% on a volumetric basis), recycle pressure, the saturator efficiency and temperature. However, for a typical system delivering the equivalent of ~8 g air/m^3 of throughput, the number of bubbles can be in the range of 1.8 – 2.5 x 10^5/mL. Smaller bubbles in seawater DAF systems may slightly offset the negative effects of salinity on the air demand. The smaller the bubble size (and the lower the water temperature), the slower the rise rate of the bubble, and thus, a larger flotation tank is required to allow bubbles to reach the surface (Gregory et al. 1999).

There are various DAF designs on the market and these range from what some may describe as horizontal, where the flow enters at one end (Figure 9.18) or from the center, and then flows along the tank length or radius to the outlet normally via an underflow baffle or a series of collector pipes. An alternative to this approach is combined dual media gravity filters with DAF which include the proprietary systems such as CoCo™ and Enflo-Filt™ or generic type called stack DAF, in-filter DAF or DAFF (Figure 9.19). These systems offer the end-user the advantage of space savings; however, the operation of the DAF in terms of loading rate is restricted by the limits placed on the filter and the physical property of an air bubble. Air bubbles with average diameters of 40-60 µm would have a rise rate in the range of 3-7 m/h, respectively (with large bubbles of 100 µm reaching rise rates of 20 m/h). This means that the higher net flotation rates now being utilized in high-rate DAF of 30-50 m/h cannot be used because of the problems associated with air being drawn into the filter bed causing air blinding. There is also the lack of available hydraulic driving head required for flow to pass through the filter to match the higher DAF rates. Moreover, the whole system (including the flotation cell) will need to be taken offline for 15 to 20 minutes during a filter backwash event.

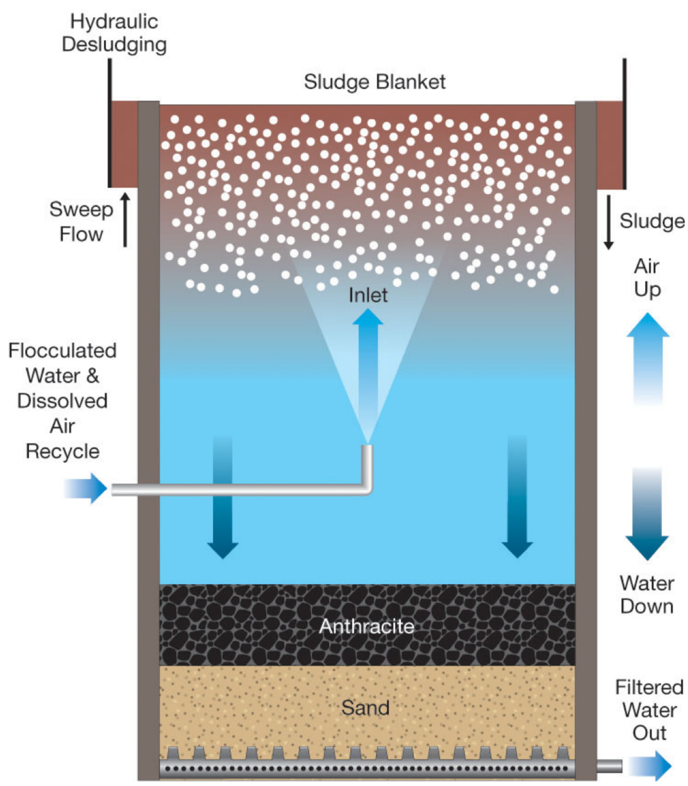


Figure 9.19. Combined DAF clarifier and granular media filter.

When no bloom, oil, or grease are present in the feedwater, some systems provide pipework to bypass the DAF. Some DAFF systems will operate in direct filter mode when blooms are not present.

The typical frequency of float removal tends to be site specific ranging from continuous to intermittent. This is true regardless of whether the sludge removal method is mechanical or hydraulic. The resulting sludge from the DAF tank can be dewatered and thickened separately but is generally mixed with the settled GMF or UF backwash water before thickening and dewatering by centrifuge or plate press. The clarified water following sludge treatment is typically returned to sea with the SWRO brine.

When designing a DAF system according to hydraulic loading, the influent flow is considered as the upstream flocculation system and excludes recycle flow. The tank area can then be designed according to the preferred hydraulic loading rate. The area calculated is that of the contact and clarification zones, whereby the velocity of the water flowing towards the base of the tank must be less than the rising velocity of floc-bubble aggregates to ensure separation (Edzwald et al., 2010). The area of the adjacent contact zone must be designed to ensure a residence time of 1-2.5 minutes. It is not always clear whether the hydraulic loading rate reported is for the clarification zone only.

9.5.4 DAF in SWRO pretreatment for removal of marine algae

A study by Haarhoff and Edzwald (2013), examining the differences in the application of DAF in seawater compared to freshwater, found that contact and separations zones were not significantly different. The largest difference was due to the lower solubility of air in seawater compared to freshwater. The dynamic viscosity, density and surface tension are all higher in seawater and though these differences are small at +8%, +3% and +1% respectively at a salinity of 35 g/kg and temperature of 20 °C (Haarhoff and Edzwald 2013) as viscosity cannot be ignored when designs are proposed where the operational load is high, i.e. typically ≥ 20 m/h. The reason why viscosity cannot be ignored at these elevated levels is because of the additional drag on the floc-bubble aggregates and the overall supporting whitewater layer as described by Amato et al. (2012). This ultimately impacts on the tank design, particularly in terms of its overall depth, which needs to be slightly deeper when considering the treatment of seawater. This will not always be necessary and will depend upon other factors such as the method used to draw off the subnatant and any water quality requirements imposed on the DAF product flow. At a minimum, an additional 30 cm of tank depth might be needed, as reported by Amato et al. (2012).

The percentage removal of algae in fresh water applications is typically 90-99% (Yap et al. 2012; Zhu et al. 2014) while practical experience shows that when used for seawater pretreatment, DAF systems usually yield significantly lower algal removal rates – 40 to 50% (Voutchkov 2013). Actual removal can be as low as 40%; however, as it is dependent on the cell concentration in the feed, the type and size of algal cells, and phase of growth (lag, exponential or stationary phase), due to differing amounts of AOM in solution (Henderson et al. 2008b). Oil and grease removal are difficult to quantify because of the three states in which it may be found, i.e. suspended (easy), emulsified, and dissolved (both more difficult). Therefore, in the case of oil, any removal rates need quantification by jar testing. Efficiency of turbidity reduction is a very poor parameter as it is not unusual to have higher turbidities post DAF than what may have been found in the raw. True turbidity removal efficiency can only be reported when the feed to the DAF is measured after flocculation.

9.5.5 Optimization of process parameters for marine HABs

While removal of algal cells via DAF is important during a bloom, operators must also consider the presence of associated AOM that may also deteriorate the water quality. This AOM may interfere with coagulation or reach downstream RO membranes and cause biofouling. It is therefore important that coagulation conditions are optimized, not just for cell removal, but also for simultaneous AOM removal. Application of coagulation-flocculation-DAF to remove AOM of larger molecular size (e.g., biopolymers and humic-type substances), in combination with biofiltration in GMF will maximize organic matter removal to protect the downstream SWRO membranes from accelerated biofouling (Shutova et al. 2016; Naidu et al. 2013). (This is discussed further in Chapter 2 and Section 9.4).

9.5.6 Summary

DAF is useful in removing algal cells and reducing the suspended solids load for downstream pretreatment, as the process is 'gentle' or low shear, preventing cell lysis and release of fouling organics and algal toxins, thereby reducing the risk of AOM fouling of UF and RO membranes. Practical full-scale experience to date with using DAF for pretreatment of seawater is very limited and have shown a wide range of algal removal efficiency; however, with optimization, DAF can be very effective as an SWRO pre-treatment, removing up to 99% of cells when preceding coagulation-flocculation is optimized. This removal efficiency depends on many factors – one of the key factors is the content of suspended solids in the source water. Designers of DAF for seawater plants should take care to consider salt water-specific parameters such as the "salting out" affect that results in either a requirement for an increase in pressure (~30%) or recycle rate (~20%). If the DAF is only operational periodically, operators could consider bringing the DAF online while cell counts are low so that the plant is fully operational when counts increase. This argues for plankton monitoring in the vicinity of the plant (Chapter 3) and within the plant (Chapter 5) so that effective actions can be taken sufficiently early to minimize clogging and fouling.

9.6 GRANULAR MEDIA FILTRATION

9.6.1 Overview

Conventional pretreatment using coagulation, flocculation, and filtration through granular media is the most commonly used source water pretreatment process for SWRO desalination plants today (other than cartridge filtration). This process includes filtration of the source seawater through one or more layers of granular media (e.g. anthracite coal, silica sand, garnet).

In the GMF process, suspended solids are removed through attachment to the filtration media particles and through blockage/capture by the filtration cake. Organics may be removed by biofiltration occurring in the filter (e.g., the utilization and breakdown of organic materials by microbes). The preferred process of filtration is capture of suspended solids with bed penetration (depth filtration) as opposed to surface filtration, since the latter results in a significantly faster increase of pressure loss and therefore shorter filter runs. During algal bloom events, coagulant overdosing may occur as described in Section 9.4.3 resulting in accumulation of large flocs at the filter surface, blinding the filter bed if the blooms are very intense.

Conventional filters used in SWRO pretreatment are typically rapid dual-media (anthracite and sand) filters (DMF) in a single-stage configuration; however, in some cases where the source water contains high levels of organics (total organic carbon (TOC) > 6 mg/L) and suspended solids (monthly average turbidity > 20 NTU/TSS > 30 mg/L), two-stage filtration systems are applied to achieve desired SDI levels. Under this configuration, the first filtration stage is mainly designed to remove macroalgae, solids, and organics that are present in suspended form. Often when a plant is subject to HABs, coagulation is employed in the first stage filtration. The second-stage filters are configured to retain fine solids (including HAB cells) and silt, and to remove a portion (20 to 50 %) of the soluble organics contained in the saline water by biofiltration.

Depending on the driving force for water filtration, GMF systems are classified as gravity or pressure filters. The main differences between the two 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 SWRO plants. Gravity pretreatment filters are used for both small and large SWRO desalination plants.

9.6.2 The filter operation cycle

GMF is a cyclical process, which incorporates two sequential modes of operation: 1) source water processing (filtration) mode; and 2) filter media backwash mode. During the filtration cycle the water moves in the direction of decreasing size gradation of the media and solids in the water are retained on and around the media grains.

As the feed water is filtered through the media, the content of solids and silt in this water decreases. Well-operating filters typically remove 90 to 99 % of the solids and silt in the source seawater (to an approximate size of 10 µm). Some of the marine microorganisms in the source water are also retained on the filter media forming biofilm around the filtration media granules. These microorganisms may consume a portion of the dissolved AOM from the source seawater such as biopolymers and TEP through biofiltration. The organic load removal efficiency of the filters is a function of four main factors: media depth, surface loading rate, coagulant concentration, and temperature. Removal of organics by the filters increases with depth and temperature and with the decrease of the filter-loading rate.

The solids retained in the pore volume between the filter grains reduce this volume over time and create hydraulic losses through the filter media (filter bed resistance). Most filters used in SWRO pretreatment operate at constant filtration rate, which means that the feed pressure of these filters increases over the filtration cycle to compensate for the head losses in the filter bed caused by accumulation of solids. Once the filter media head losses reach a certain preset maximum level, the filter is taken out of service and media backwash is activated. Deeper filters or larger surface area have larger capacity to retain solids and therefore, usually have longer filtration cycles.

Typical parameters used to monitor pretreatment GMF performance are SDI, turbidity, TOC and iron. Usually, turbidity of the feed and filtered seawater is measured continuously with online turbidity meters. For larger plants SDI_{15} of the filtered water may also be measured on line. The measurement frequency of TOC may be increased during an algal bloom event, especially when TOC exceeds 2 mg/L, chlorophyll *a* increases over 1 µg/L and/or algal counts exceed one or two million cells/L. Usually, if TOC, chlorophyll a or algal counts are below these levels, the algal bloom is not expected to have a major impact on pretreatment system operation and RO fouling. If these source water quality parameters exceed the above-mentioned thresholds, usually plant operators institute algal bloom mitigation strategies such as increasing the dose of coagulant fed to the source seawater, increasing acid addition in order to decrease pH, and thereby enhance algal removal, and/or decrease surface loading rate to enhance filter retention time and encourage biofiltration. (Note, however, that cell counts can be deceiving – one million cells of a small species will constitute a much lower cell volume or biomass than the same number of a larger species. Ideally, operators need to learn the species and cell concentrations of those species that cause problems, and the pretreatment strategies that were effective for those conditions. Simply stated - not all algal blooms are the same, so local experience needs to be documented and applied.

GMF are typically backwashed using filtered seawater or concentrate from the SWRO membrane system. Normally filter cell backwash frequency is once every 24 to 48 hours and spent (waste) backwash volume is 2 to 10 % of the intake seawater. During the severe 2008 algal bloom in the Arabian Gulf, backwash intervals at the Fujairah 1 plant in the UAE were dramatically reduced to 2 hours from 24 hours.

Use of SWRO concentrate instead of filtered effluent to backwash filter cells allows for a reduction in backwash volumes and a reduction in the energy needed to pump source water to the desalination plant; however, use of concentrate for filter backwash during algal blooms is not recommended because the concentrate will have an elevated content of AOM and biodegradable organics, which will not benefit the pretreatment process and may exacerbate RO membrane fouling. Additionally, osmotic shock from the higher salinity may cause cell lysis in some cases, releasing additional AOM.

During backwash of down-flow filters, the backwash water flows upwards through the filters, scours the filter grains, removes the solids accumulated on the filter grains, expands the filter bed, and transports the removed solids towards the backwash troughs. From experience, it is known that backwashing of filter media grains smaller than 0.8 mm with water only is inefficient. Therefore, a typical backwash regime currently includes a combination/sequence of air and water washing. Air creates greater turbulence and enhances particle scrubbing. The length of water and air backwashing cycles is a function of the solids content in the source water and the depth of media bed and typically is between 5 and 15 minutes.

GMF pretreatment efficiency is very dependent on the presence of a biofilm and the formation of a matrix of small particles around the granular filtration media that are needed to remove fine particles. Formation of such biofilm and filtration matrix is referred to as "filter cell maturation". Every time the filters are backwashed for removal of the residuals accumulated during the filtration process, a portion of the biofilm and solids matrix around the filtration media grains are removed and as a result, when the filter is put back in service after backwash, it usually does not produce pretreated water of quality compliant with the target SDI and turbidity values. It usually takes between 15 and 45 minutes after a filter cell is returned to service for the fine solids matrix to form to its previous level and for the backwashed filter to begin producing pretreated water of adequate quality. During this filter cell maturation period, the out-of-specification filtrate is usually discharged to the plant outfall.

Filter media type, uniformity, size, and depth are of key importance for the performance of pretreatment filters. Characteristics of media commonly used in SWRO desalination plants are presented in Table 9.1.

Single media filters (mono media) are not commonly used in SWRO pretreatment because of their limited ability to perform under varying source water conditions. In mono media filters, the fine size filtration media particles tend to aggregate at the top of the bed after a number of backwash runs. This reduces penetration of suspended solids and, therefore, mainly results in surface bed filtration. Typically, such filters could be used for desalination plants with subsurface intakes producing turbidity of < 2 NTU, TSS of < 5 mg/L and SDI_{15} < 5 (see the Sur plant in Oman, Chapter 6). A large-scale application of single media (0.7 mm sand) filters using an open onshore intake is at the Tampa Bay desalination plant in Florida (see Case Study 11.9).

A graduation of the filtration bed from coarse to fine particles can be achieved in dual media configuration by placing fine, high specific gravity filtration media as the lower filtration layer and coarse, low specific gravity filtration media as a top layer. Filtration media selection that provides coarse to fine filtration bed configuration includes 0.4 to 1 m anthracite or pumice as a top layer and 0.4 to 2.0 m of silica sand as a bottom filtration layer. DMF is the most commonly used GMF in SWRO plants worldwide. Deep DMF are often used if the desalination plant filtration system is designed to achieve enhanced removal of soluble organics from source water by biofiltration. In this case, the depth of the anthracite level is enhanced to between 1.5 and 1.8 m. In comparison to mono media filters, DMF systems operate at higher filtration rates, have longer filter run times, and produce less backwash water.

Tri-media filters are not commonly used in SWRO pretreatment and are primarily used for capturing small-size phytoplankton and fine silt that cannot be well retained by the top two layers in DMF. Tri-media filters typically comprise 0.45 to 0.6 m of anthracite as the top layer, which retains large-size algae (i.e. algae over 100 μm), 0.2 to 0.4 m of sand as a middle layer to remove medium-size algae (20 to 100 μm) and 0.10 to 0.15 m of garnet or limonite as the bottom layer. The third (garnet or limonite) layer of filtration is usually used only if the source water contains a large amount of very fine silt or the source water intake experiences algal blooms dominated by small algae (0.2 to 2 μm).

Since the cost of filter cells increases with depth, often instead of a deep, single tri-media gravity filter, a combination of coarser media (anthracite-sand) gravity filter followed by a pressure filter containing finer (sand and garnet) is used.

Most filters used in seawater pretreatment are down-flow 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 breakage 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 such as algal biopolymers may be released from the broken algal cells into the filtered water, which is undesirable as it can exacerbate biofouling of the downstream SWRO membranes.

Table 9.1. Typical media characteristics for GMF used in SWRO plants.

Media Type	Typical Effective Grain Size - mm	Specific Density tons/m^3
Pumice	0.8 – 2.0	1.2
Anthracite	0.8 – 2.0	1.4 – 1.7
Silica Sand	0.4 – 0.8	2.60 – 2.65
Garnet	0.2 – 0.6	3.50 – 4.30

9.6.3 Single and two-stage filtration

Two-stage filtration is typically used 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) or in river estuaries, which are exposed to an elevated turbidity levels occurring during the wet-season of the year.

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 while the second stage filter is configured as a DMF with design criteria described in the previous section. The first (coarse-media) filter typically removes 60 to 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 the 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 DMF and in a lower total amount of coagulant needed to achieve the same final filter effluent water quality, as compared to single-stage DMF.

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) if the second stage filters are designed as deep-bed (rather than shallow bed) filters they can achieve enhanced TOC and AOM removal by biofiltration. While deep, single-stage dual media filters can typically reduce 20 to 30 % of the TOC contained in the source seawater, the two-stage systems with deep second-stage filters can achieve 40 to 60 % TOC removal, mainly due to enhanced fine particle coagulation and biofiltration.

It should also be pointed out that if the filters are designed to achieve TOC removal by biofiltration, it would take at least four to six weeks for the filters to accumulate sustainable biofilm on the surface of the filter media to yield steady and consistent filter performance and TOC removal of 10 to 20%. If the source water temperature is relatively cold (i.e. below 15 ^{o}C), then the biofilm formation process may take several weeks longer.

9.6.4 Gravity filters

Typically, gravity filters are reinforced concrete structures that operate a water pressure drop through the media of between 1.8 and 3.0 m. The hydrostatic pressure over the filter bed provides the force needed to overcome the head loss in the media. Single-stage down-flow gravity DMF filters are the predominant type of filtration pretreatment technology used in desalination plants of capacity higher than 40,000 m^3/day. Table 9.2 provides examples of key design criteria for gravity filters at SWRO desalination plants of various size and water quality. Some of the largest SWRO desalination plants in the world in operation today such as the 325,000 m^3/day Ashkelon SWRO plant are gravity single-stage DMF.

Figure 9.20. Gravity filters protected with plastic covers for control of algal growth showing covers of gravity filters in Ashkelon, Israel.

Seawater always contains a measurable amount of algae, with the concentration usually increasing several times during the summer period and possibly increasing 10 times or more during periods of algal blooms (which may or may not exhibit themselves as HABs). There are thousands of algal species in the seawater, as discussed in Chapter 1, so generalizations like this should be viewed with caution.

Gravity filters (Figure 9.20) are typically covered with light plastic covers that protect the

Figure 9.21. Single-stage dual media gravity filters at the Gold Coast SWRO Desalination Plant

filter cells from direct sunlight to prevent algal growth or are installed in buildings (Figure 9.21).

An alternate method to aid control of SWRO biofouling is the installation of a granular-activated carbon media layer ("activated carbon cap") on the surface of the filters to enhance the removal of some of the polysaccharides and other organics in the source water. This approach has been tested in full-scale applications in the Middle East.

Table 9.2. Examples of large SWRO desalination plants with DMF gravity filters.

Desalination plant location and capacity	Pretreatment system configuration	Average and maximum filter loading rates	Notes
Ashkelon SWRO Plant, Israel – 325,000 m³/day	40 single-stage	10/12 m/h (avg./max)	Open intake – 1,000 m from shore
Sydney SWRO Plant, Australia – 250,000 m³/day	24 single-stage	8/12 m/h (avg./max)	Open intake – 300 m from shore
Fujairah 1 SWRO Plant, UAE – 170,000 m³/day	14 Single-stage	Filtration Rate - 8.5 m/h (avg.) 9.5 m/h (max)	Shallow offshore open intake. High bloom potential
Fujairah 2 SWRO Plant, UAE – 136,000 m³/day	16 DAFs, 12 - Single-stage	DAF rate – 21 m/h (avg.) 30 m/h (max)	Shallow offshore open intake. High bloom potential
		Filtration Rate – 10.5 m/h (avg.) 12.5 m/h (max)	DAF operated only if turbidity > 5 NTU
Gold Coast SWRO Plant, Australia – 125,000 m³/day	18 single-stage	8/10 m/h (avg./max)	Open intake – 1,500 m from shore

9.6.5 Pressure filters

Compared to gravity media filters that operate under a maximum water level over the filter bed of up to 3 m, pressure filters typically run at feed pressure equivalent to 15 to 30 m of water column, which has the potential to damage HAB cells. Therefore, pressure filters may have the disadvantage of causing accelerated biofouling when filtering source water with very high algal content. This effect is likely to manifest itself mainly 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_{15} < 4 and turbidity < 4 NTU). In addition, the Spanish desalination plant intakes are relatively deep and the algal content in the source water is commonly fairly low. Hence, the ingress of algae is lower and biofouling caused by breakage and decay of algal cells may not be as significant problem as it would be for shallow or near-shore open intakes (see Chapter 6).

Gravity media filters have a two- to three-times larger volume of filtration media and retention time than pressure filters for the same water production capacity. This is a benefit for plants exposed to algal blooms because source water turbidity and TSS could increase several times during algal bloom events. The higher solids retention capacity allows the gravity filters to handle such increases without decreasing the length of the filter cycle. Higher hydraulic retention capacity of the gravity filters allow these filters to develop a more robust biofiltration layer near the bottom of the sand media in the filters, which in turn results in more effective filtration.

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, HABs, shipping traffic, or ocean bottom dredging operations in the vicinity of the intake, seasonal change in underwater current direction, or spring upwelling of water from the bottom to the surface, then pressure filters will produce effluent with inferior effluent quality (SDI_{15} and turbidity) during such events and, therefore, their use would likely result in a more frequent RO cleaning.

Pressure filters have filter bed configurations similar to that of gravity filters, except that the filter media is contained in steel pressure vessels. They have found application mainly for small- and medium-sized seawater desalination plants – usually with production capacity of less than 20,000 m^3/d. There are, however, a number of installations worldwide where pressure filters are used for pretreatment of significantly larger volumes of water (see Table 9.3).

In most cases for good source water quality (SDI_{15} < 4 and turbidity < 4 NTU), pressure filters are designed as single stage, dual media (anthracite and sand) units. Some 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 (Figure 9.22) are customarily used in smaller plants and individual vessels have maximum diameter of 3 m. Horizontal pressure filters (Figure 9.23) are used more frequently in desalination plants and are more popular for medium and large-size plants. One example of a desalination plant using horizontal pressure GMF for seawater pretreatment is the 140,000 m^3/day Kwinana SWRO plant in Perth, Australia (Figure 9.23). Horizontal filters allow larger filtration area per filter vessel 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.

Table 9.3. Large seawater desalination plants with pressure granular media filters.

Desalination plant location and capacity	Pretreatment system configuration	Average and maximum filter loading rates	Notes
Al Dur SWRO Plant, Bahrain 218,000 m^3/day	DAF followed by horizontal pressure filters	DAF surface loading Rate – 25 to 30m^3/m^2.h Pressure filter rate – 18 to 24 m^3/m^2.h	Shallow offshore open intake in algal bloom prone area
Barcelona SWRO Plant 200,000 m^3/day	DAF followed by gravity filters and horizontal pressure filters	DAF surface Loading Rate – 25 to 30m^3/m^2.h Gravity filter rate – 8 to 10 m^3/m^2.h Pressure filter rate – 15 to 20 m^3/m^2.h	Deep offshore open intake in industrial port and near river estuary
Kwinana SWRO Plant, Perth, Australia 140,000 m^3/day	24 single-stage dual media pressure filters	14.0/18 m^3/m^2.h (avg./max)	Shallow open intake
Carboneras SWRO Plant, Spain 120,000 m^3/day	40 single-stage dual media pressure filters	12.0/15.0 m^3/m^2.h (avg./max)	Offshore open intake
El Coloso SWRO Plant, Chile 45,400 m^3/day	DAF followed by two-stage dual media horizontal pressure filters	DAF surface loading rate – 22 to 33m^3/m^2.h. Filter rate -25 m^3/m^2.h	Open intake in industrial port with frequent red tides

Figure 9.22. Vertical pressure pretreatment filters capacity. Photo: Voutchkov 2013.

Since pressure filters are completely enclosed, sunlight cannot reach the filter weirs, distribution system and media and induce green algal growth that would have negative impact on filter performance. The visual indications that can be observed for gravity filters are difficult or impossible to perform for pressure filters. Additionally, flow distribution into individual filters in a bank of filters may not be easily controlled compared with gravity filters, where equal flow splitting can be achieved through gravity.

Figure 9.23. Single-stage horizontal -pressure dual media filters at the Kwinana SWRO Plant. Photo: Water Corporation.

9.6.6 GMF filter performance

The purpose of the pretreatment filters for SWRO plants is not only to remove over 99% of all suspended solids in the source water, but also to reduce the content of the much finer silt particles by several orders of magnitude. Therefore, the design of these pretreatment facilities is usually governed by the filter effluent SDI_{15} target level rather than by target turbidity or pathogen removal rates. Full-scale experience at many GMF plants indicates that filters can consistently reduce source water turbidity to less than 0.1 NTU, and SDI_{15} less than 2-4 depending on source water quality.

The rate of algal removal by the filters will depend mainly upon the size of the algae and the size of the filter media. Most algae are typically retained on the surface of the top media (anthracite/pumice) (Bar-Zeev et al. 2012). Depending upon the size of media and size of the algae dominating in the source water, algal removal could typically vary between 20% and 90% or more. Based on recent research (Gustalli et al. 2013), UF membranes provide better removal of algae than granular dual media filters (99%+ vs 74%).

Results of a comprehensive study of the effect of bloom intensity on GMF algal removal efficiency indicates that the more intense the bloom, the lower the level of algal removal by the filters (Plantier at al. 2013). After seven hours of filtration, the overall efficiency of the tested GMF for a light algal bloom (30,000,000 algal cells/L) dominated by microalgae was 74%, while for a severe algal bloom (150,000,000 algal cells/L) this removal efficiency was reduced to 49%. Another important observation of this study was that the first 30 cm of the filter media retain more algae than the rest of the media, and that most of the algal retention occurs over the first three hours of the filtration cycle.

Desalination pretreatment GMF would typically provide 99% (2 logs) of removal of pathogens, but sometimes may have lower removal rates in terms of marine bacteria because these bacteria are typically of small size and can pass through the filters. It is interesting to note that while UF filters have two to three order of magnitude higher removal rates of marine bacteria, such removal is inadequate to prevent heavy biofouling of the downstream SWRO elements if sufficient amounts of bioavailable organics are present in the water.

Typical gravity and pressure dual media filters of conventional filter bed depth of 1.0 to 1.4 m have relatively low organic removal rates – 15 to 20% in mature filters. Algal biopolymer removal of 18% was reported in the Barcelona DAF-DMF pilot study (see Case Study 11.11) and TEP removals of 20 to 90% was reported in GMF (Bar-Zeev et al. 2012). The removal rate, however, increases significantly with depth and could reach 25 to 35% for filters with a total filter depth of 2.0 m or more. If a carbon cap is installed on the top of the filter media (above the layer of anthracite), TOC removal rate could be increased to 40 to 50%, although the performance of this removal requires monitoring over time as the removal capacity will become exhausted.

GMF with a total filter depth of 1.4 to 1.6 m typically removes 10 to 20% of the TOC, AOC and AOM in the source seawater (Voutchkov 2013). Deep gravity media filters (filter depth of 2.0 m or more) can remove over 40% percent of the TOC, AOC and AOM in the source water. Such filters develop a biofiltration zone at a depth of 1.6 to 2.0 m, which significantly enhances removal of dissolved organics contained in the source seawater.

9.7 MICROSCREENS FOR MEMBRANE PRETREATMENT

9.7.1 Overview

For SWRO desalination plants with membrane pretreatment (UF/MF), the seawater has to be prescreened to remove very fine (50 to 500 µm) sharp particles (e.g. broken shells) which could puncture the plastic membranes and compromise their integrity and performance. Generally, there are two types of microscreens; micro-strainers or disk filters. These devices can experience complete fouling of the screens that permanently builds differential pressure. Additionally, fouling due to macroalgae and jellyfish can be experienced during microscreening. The operation of microscreens during HABs is described in this section.

9.7.2 Types and configurations

Typically, microscreens, which could be the micro-strainer (Figure 9.24) or disk filter (Figure 9.25) type, are used in SWRO for large particle removal (50 to 500 µm). SWRO desalination plants with microscreens are also usually equipped with conventional coarse screens or a combination of coarse and fine screens, which retain debris >10mm upstream at the seawater intake.

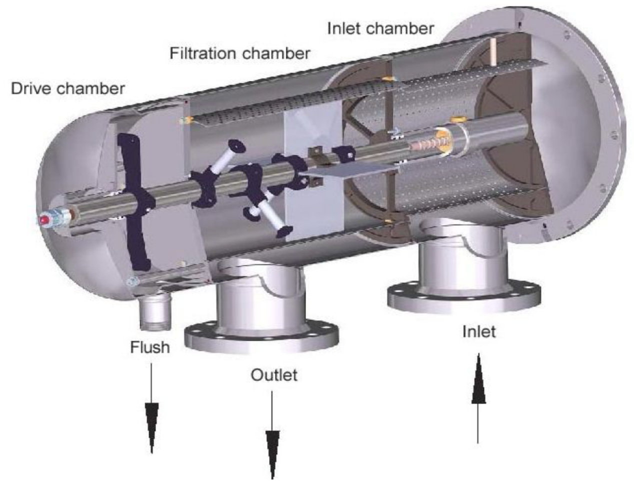


Figure 9.24. Self-cleaning micro-strainers. Photo: Voutchkov (2013).

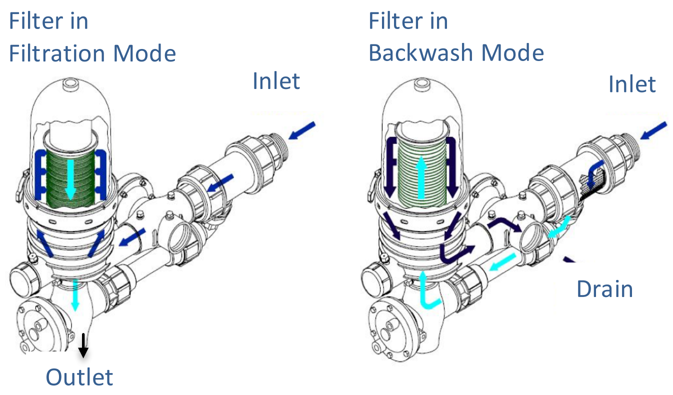


Figure 9.25. Disk filter – modes of operation. Source: Voutchkov (2013).

Most self-cleaning micro-strainers consist of screens with small openings located in a filtration chamber. The source water enters through 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 differential pressure between intake and filtered water overtime. When the differential pressure reaches a certain preset value, the deposited solids are removed by jet of backwash water. The self-cleaning process typically takes 30 to 40 seconds.

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

During the filtration process, the filtration disks 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 of the element. Source water debris and aquatic organisms (mainly phytoplankton) with a size smaller than the size of the microscreens (generally 80 to 120 µm) 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 headloss reaches a preset maximum level (typically 0.35 bar or less) the filters enter backwash mode. Considering the aperture of the screens, significant retention of algal cells could occur during a bloom. All debris (algal or otherwise) retained on the outer side of the filters is then flushed by tangential water jets of filtered seawater flow under 0.15 to 0.2 bar 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 through the plant ocean outfall – usually along with the concentrate. In some cases, the debris collected on the microscreens is disposed of through a separate pipe in the vicinity of the intake area. Hence during a bloom, algal cell wall debris and suspended solids could be re-entrained into the intake. Therefore, during a bloom, redirection of this waste stream could be considered to prevent downstream impacts.

Because of the relatively low differential pressure at which these filters operate, they are likely to minimize impingement of the marine organisms in the source water. Furthermore, 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 net entrainment.

One of the key issues associated with using membrane pretreatment is that the membrane fibers can be punctured by sharp objects contained in the source seawater, such as broken shells or sharp sand particles. In addition, seawater can contain barnacles, which in their embryonic phase of development are 130 to 150 µm in size and can pass through the screen openings unless these openings are 120 µm or smaller. Experience at the Southern Seawater Desalination Plant in Perth, Australia also shows that the source seawater could contain other marine organisms such as sponges, polychaetes, and diatoms which have diameter of only 3 to 10 µm and could be sharp-enough to puncture membrane fibers (Ransome et al. 2015).

If larval shellfish and barnacles pass the screens, they could attach to the walls of downstream pretreatment facilities, grow on these walls and ultimately 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 a very effective biocide for most other marine organisms. Therefore, the use of fine microscreens or disk filters (80 to 120- µm size) is essential for reliable operation of the entire seawater 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.

9.7.3 Algal bloom-related challenges

Full-scale experience shows that, depending upon the severity of the algal bloom and the size and type of microscreens, these devices face two operational challenges: 1) complete plugging of the screens which permanently builds differential pressure, triggering continuous backwash; and 2) rapid increase in algal mass accumulation on the screens which causes very frequent backwashes that in turn interrupt the normal operation of the pretreatment system and the desalination plant. Such challenges are very frequent in the case of near-shore or lagoon intakes of the Gulf (such as at Sohar) applying microscreens of size of 100 µm or smaller during conditions of algal cell abundance in the water of over 100,000 cells/L (see Case Study 11.3). In the case of the Jacobahaven demonstration plant, algal blooms resulted in a significant increase in the clogging rate of the smaller aperture 50 µm strainers, reducing the backwash interval to 5 minutes from 0.5 – 1.5 hour for non-bloom conditions at identical turbidity (see Case Study 11.10). Similar impacts on microscreens are observed during conditions of jellyfish outbreaks, strong winds or currents carrying sea grass or seaweeds (macroalgae) in the vicinity of the intake.

One solution has been to select a larger size microscreen (i.e. screen with pore sizes of 300 µm or higher). While this solution could address the problems during certain algal blooms and/or jellyfish outbreaks it creates potential problems with sharp particles entering the membrane pretreatment system and creating micro-punctures on the membranes over time. When adopting this solution, permanent MF/UF membrane integrity loss has usually been observed after plant operation of 6 months or more.

Another solution is to use conventional relatively coarse granular gravity media filters instead of microscreens in order to retain both the elevated content of algal mass and fine sharp particles during blooms. This approach is being considered for the West Basin Desalination Project in California, USA. As demonstrated by a comprehensive pilot study at the West Basin Demonstration Plant (Figure 9.26), long-term side-by-side testing of disk microscreens and GMF indicated comparable capability of these two pre-filters to remove harmful shell fragments; however, the deep-bed high-rate GMF provided higher quality filtrate during periods of poor raw ocean water quality (storm and algal bloom events). Testing showed that the GMF allowed more sustainable MF permeability and affected an increased MF cleaning interval compared to disc filtration during an algal bloom event (SPI Engineering 2010).

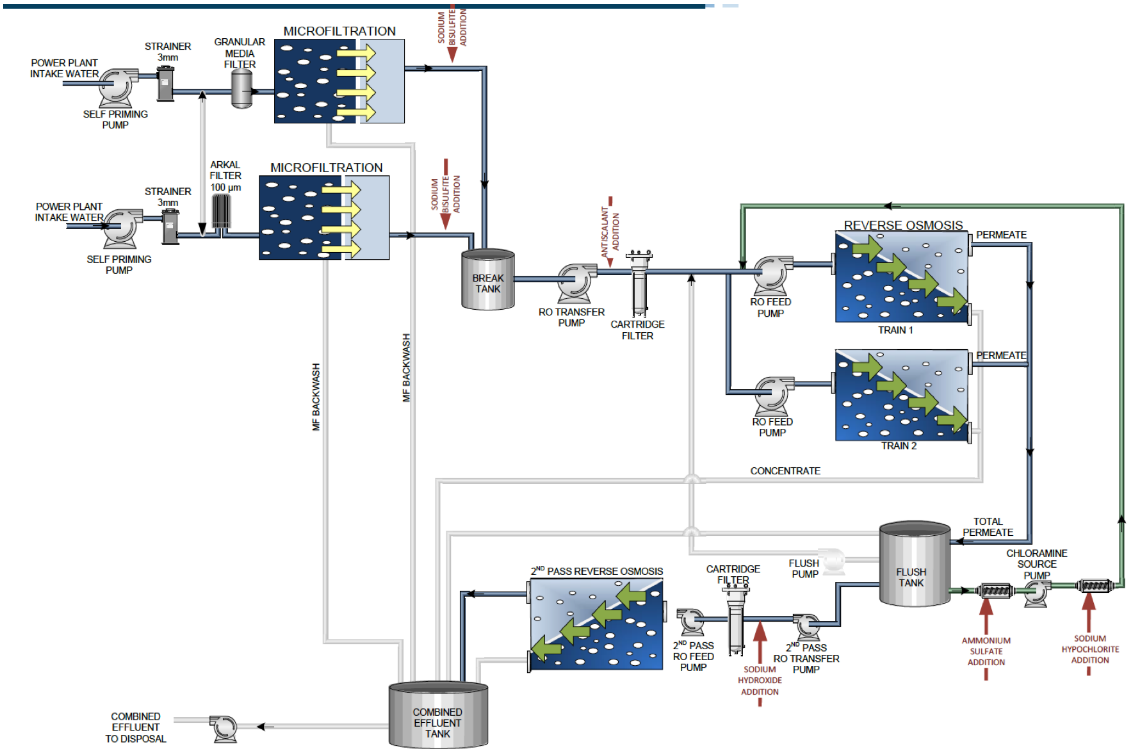


Figure 9.26. Configuration of West Basin SWRO Demonstration Plant. Figure: SPI Engineering (2010).

In order for microscreens to successfully handle conditions of high algal cell concentrations, jellyfish outbreaks or sea grass plugging, they need to have an effective focused auto-backwash system that maximizes the backwashing velocity across the strainer face with moving nozzles that operate over the strainer surface (Figure 9.24) or a variable area mesh that opens up on backwashing. For example, at the Southern Seawater Desalination Plant in Perth, Australia the original conventional microscreens had to be replaced by screens with a focused auto-backwash system to address operational challenges of intense screen plugging due to large quantity of macroalgal sea grass fragments carried to the intake via underwater currents (Ransome et al. 2015).

A microsreening technology that has potential to handle severe algal bloom conditions is microfiber filtration (Eshel et al. 2013). Microfiber filters consist of cartridges with multilayer textile threads, usually with pore sizes of 2 to 20 μm. While their ability to retain algae, organic matter and mineral solids is comparable to cartridge filters, they have the benefit that they can be backwashed automatically, similar to conventional microscreens. Full-scale test experiments of this technology on freshwater during algal bloom conditions in a lake have showed significant reductions in TEP (mean 47±21%), chlorophyll (mean 90±6%), total suspended solids (67±7%), turbidity (89±5%), and particles larger in size than 3 μm (93±4%). Such systems have not yet found full-scale implementation for seawater screening.

9.7.4 Summary

Fouling of microscreens due to HABs cause two main operational challenges: 1) complete plugging of the microscreens which permanently builds differential pressure triggering continuous backwash; and 2) rapid increase in algal mass accumulation on the screens which causes frequent backwashes that in turn interrupt the normal operation of the pretreatment system and the desalination plant. Monitoring of the time between backwashes in microscreen systems is therefore an important consideration during bloom periods.

9.8 MICROFILTRATION/ ULTRAFILTRATION

9.8.1 Overview

MF/UF membranes have been tested and applied as pretreatment for SWRO membranes for more than a decade (Wolf et al. 2005; Halpern et al. 2005). MF/UF membranes offer several advantages over GMF in SWRO pretreatment, namely, smaller footprint, consistently high permeate quality in terms of SDI, higher retention of organic macromolecules (including some AOM), lower overall chemical consumption (Wilf and Schierach 2001; Pearce 2010). Although, MF and UF are both applied in SWRO, UF is often employed due to its smaller pore size and better removal of particulate/colloidal organics, silt, and pathogens from seawater (Voutchkov 2009).

9.8.2 MF/UF filtration modes

In contrast to RO membranes that operate in cross flow, MF/UF membranes are typically operated in dead-end mode in two main configurations based on feed flow direction; PDI, and pressure driven outside-in (PDO). MF/UF membranes in outside-in configuration may also be operated in vacuum driven (submerged) mode, which are commonly polyvinyledene fluoride (PVDF). PDI membranes are generally based on polyethersulphone (PES) and blends thereof, although some PDI membranes based on cellulose triacetate and polysulphone are also available in the market (Pearce 2007). In PDI filtration, feedwater enters through the inside of the element and permeate is collected from the outside of the capillaries. Hydraulic cleaning (or backwash) is performed by reversing the flow, whereby product water flows through the membranes from the outside of the capillary, physically lifting the accumulated material from the membrane surface and flushing it to the waste stream. PDI membranes may be manufactured as single capillaries (hollow fibers) or multi-channel tubes. Multi-channel tubes combine several individual capillaries in a robust fiber, usually in a honeycomb arrangement. This configuration is considered to have higher stability and breaking resistance as compared to single capillaries.

PDI UF membranes based on PES have seen a substantial growth in installed capacity due to several large SWRO desalination projects, e.g., Ashdod (UF capacity 930,000 m^3/d), Shuwaikh (UF capacity 350,000 m^3/d). Among the manufacturers of PDI membranes applied in SWRO pretreatment are Pentair X-Flow, BASF inge, Hydranautics, Aquasource and 3M Membrana, while the main manufacturers of PDO membranes are Hyflux, Dow, Pall, Toray and GE Zenon. In general, PDO configuration can tolerate higher feed solids loadings (TSS > 300 mg/L; turbidity > 300 NTU) than PDI (TSS 100 mg/L; turbidity 100 NTU), and allows the use of air scour, which can enhance hydraulic cleaning. PDI configuration membranes tend to have higher permeability, due to the selection of PES rather than PVDF. However, comparative data on operation of PDI vs. PDO membranes on algal-laden seawater is not available. Table 9.4 summarizes the products of the main international suppliers in terms of membrane material and pore size/molecular weight cut-off (MWCO). AOM compounds – including particulate and colloidal algal biopolymers, TEPs, etc. – cover a wide size spectrum, ranging from a few nanometers to more than 1 millimeter (Figure 2.2 in Chapter 2). Based upon their MWCO (~ 80-150 kDa), most UF membranes are expected to remove only part of the higher molecular weight fraction of AOM, while MF membranes will remove an even smaller amount.

Table 9.4. Commercially available MF/UF membranes for SWRO pretreatment.

Pressure driven inside-out (PDI)				
Manufacturer	**Product**	**Material**	**Nominal pore size**	**Nominal MWCO**
Pentair X-Flow	Seaguard Seaflex	PES/PVP	0.02 μm	150 kDa
BASF inge	Multibore®	PES	0.02 μm	100-150 kDa
Hydranautics	Hydracap®	PES	0.02 μm	150 kDa
Aquasource	ALTEON™	Hydrophilic PS	0.02 μm	150 kDa
3M Membrana	UltraPES™	PES	n.a.	80 kDa
Pressure driven outside-in (PDO)				
Manufacturer	**Product**	**Material**	**Nominal pore size**	**Nominal MWCO**
Hyflux	Kristal®	PES	n.a.	120 kDa
Dow	Integraflux™	PVDF	0.02 μm	n.a.
Pall	Aria™	PVDF	0.1 μm	n.a.
Toray*	TORAYFIL®	PVDF	n.a.	150 kDa
GE Zenon*	Zeeweed®	PVDF	0.02 μm	n.a.

* Also available in vacuum driven (submerged) mode

MF/UF membranes are designed based on a certain feedwater quality composition. For instance, Hydranautics operation guidelines for their PDI membranes are turbidity < 200 NTU, algae counts < 1,500,000 cells/L, SUVA < 4, while Dow specifies turbidity < 50 NTU, TOC < 10 mg C/L, and TSS < 50 mg/L to reduce the extent of fouling and ensure longevity of their PDO membrane elements (Hydranautics 2015; Dow Water and Process Solutions 2015).

9.8.3 Fouling of MF/UF during algal blooms

PDI and PDO membranes have reportedly exhibited some degree of fouling during bloom periods due to high concentrations of algal cells and associated organics. The accumulation of AOM and algal particulates on the surface and/or within the pores of MF/UF membranes leads to a loss in membrane performance. As discussed in Chapter 2, high molecular weight AOM such as biopolymers, particularly the very sticky TEP, have been identified as the main cause of membrane fouling rather than the algal cells themselves.

Most MF/UF membrane plants are operated in constant flux mode to a flux set point to meet RO feedwater requirements. In constant flux filtration, fouling is observed as an increase in TMP through time, resulting in higher pumping requirements to maintain constant filtration flow. This may impact the efficiency of membrane cleaning. In general, MF/UF membrane fouling phenomena are a function of feedwater quality, membrane properties and operational conditions (e.g., filtration flux) and may be classified as:

- Hydraulically reversible or back-washable fouling (permeability is restored with hydraulic cleaning with air and water);
- Hydraulically irreversible or non-backwashable fouling (permeability cannot be restored with hydraulic cleaning alone and CEB may be required); and
- Chemically irreversible fouling (permeability cannot be restored with CEB and/or chemical cleaning in place (CIP))

AOM accumulation can cause a rapid increase in TMP and/or can increase non-backwashable fouling in MF/UF. The impact of AOM on MF/UF membrane permeability and backwashability can be explained using classic blocking and cake mechanisms as described in Chapter 2 (see Section 2.4.4.2).

Control of fouling that may occur during a HAB can be achieved by optimizing operational conditions and cleaning procedures (Murrer and Rosberg 1998; Brehant et al. 2002; Zhang et al. 2006; Ma et al. 2007; Bu-Rashid and Czolkoss 2007; Di Profio et al. 2011; Schurer et al. 2012) or by feedwater conditioning. Operating MF/UF membranes at a lower flux can reduce the extent of backwashable and non-backwashable fouling. Filtration flux affects the characteristics of the AOM cake/gel layer formed on the membrane surface; layers formed at low flux exhibit lower resistance to filtration and are less compressible than those formed at higher flux values (Salinas Rodriguez et al. 2012; Tabatabai et al. 2014). The downside of lowering filtration flux is that proportionally higher membrane surface area is required to meet production capacity, resulting in a larger plant footprint and higher CAPEX. This can be balanced in part by an increase in operational flexibility to allow backwash while maintaining capacity.

During HAB events, cleaning procedures can be tailored to control fouling in membrane pretreatment. Optimizing the frequency, duration and intensity of hydraulic cleaning regimes, the type and concentration of cleaning chemicals, the sequence in which they are applied, and the duration of soaking and rinsing steps can enhance permeability recovery and fouling control. Optimum conditions for cleaning are site and event specific and membrane manufacturer guidelines on cleaning regimes during HAB events are currently not available.

Alternatively, feedwater conditioning (e.g. with coagulation) can be undertaken to reduce the extent of back-washable and non-back-washable fouling in MF/UF systems. Experience in PDI UF operation in SWRO pretreatment has shown that other than for algal bloom events, coagulation is generally not required to stabilize UF hydraulic performance (Schurer et al. 2013), even during storms with turbidity peaks as high as 50 NTU (see Case Study 11.10). Most importantly, operating at lowered filtration flux of approximately 60 L/m^2h on North

Sea water allowed for longer filtration cycles (as compared to operation at 90 L/m^{2}h) and resulted in enhanced permeability recovery by backwashing and improved overall productivity during bloom conditions.

Coagulation may also further enhance UF operation during algal blooms when HAB cell numbers and TSS concentrations are high by partially acting on the following mechanisms (Figure 9.27):

- Reducing TMP development during filtration in PDI UF membranes;
- Reducing the extent of back-washable and non-back-washable fouling;
- Enhancing permeability recovery by backwashing;
- Improving permeate quality of PDI UF membranes.

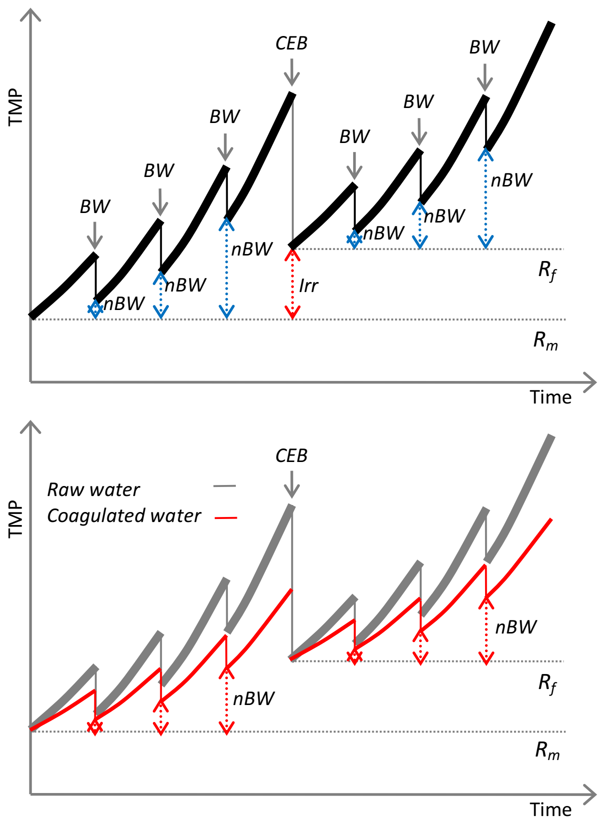


Figure 9.27. Idealized schematic presentation of TMP development and fouling in dead-end MF/UF systems for raw feedwater with algae (top panel) and coagulated feedwater (bottom panel). BW is backwash; nBW is non-back-washable fouling, Irr is irreversible fouling, Rm is clean membrane resistance (where resistance is TMP normalized for temperature variation) and Rf is fouled membrane resistance. Figures: Tabatabai (2014).

9.8.4 Removal of HAB cells using MF/UF

Removal of HAB cells typically exceeds 99% given UF/MF filters do not have any gross integrity breaches. Some shear of the HAB cells may be experienced during pumping and impingement against the membrane surface due to pressure, breaking a small number of cells (~1-2%) under normal UF conditions. During this process, a small amount of AOM may be released due to stress on the cells. Stress on the HAB cells without breakage may release more AOM than if the cells were completely broken. Coagulation prior to the UF may aid the process by encapsulating cells in floc and preventing some breakage (Chow et al. 1997; Dixon et al. 2011a,b). While these experiments were at a laboratory scale using cyanobacteria (both cultured and sampled from the field), the principles for marine species remain the same, although breakage percentages may be higher for marine HAB cells that may be less robust than cyanobacteria. Cell lysis may also be dependent on the species of HAB; however, this phenomenon requires further research.

The type of HAB cell may bear less importance on UF/MF performance than the cell count, as shown by Castaing et al. (2011) during a laboratory-scale trial for a variety of marine species. When comparing MF performance (0.2 µm, polysulfone, submerged outside-in) challenging the membrane with the marine HAB species *Heterocapsa triquetra, Alexandrium minutum* and *Prorocentrum lima* (Figure 9.28) showed that while small differences in the loss of relative permeate flux were identified between cell types, this was not as significant as filtration of 1,000,000 cells/L vs 30,000,000 cells/L using *A. minutum* (Figure 9.29). The small differences between cell type are most likely due to differences in the production of AOM from each cell type. The difference made by cell numbers in this experiment are not surprising given the stark difference between the cell numbers in each experiment.

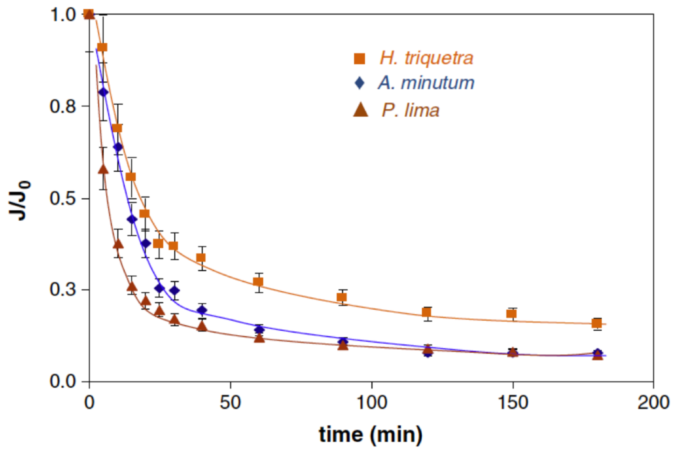


Figure 9.28. Relative permeate flux at 20 °C vs time for each tested microfiltration in presence of the different micro-algae studied.

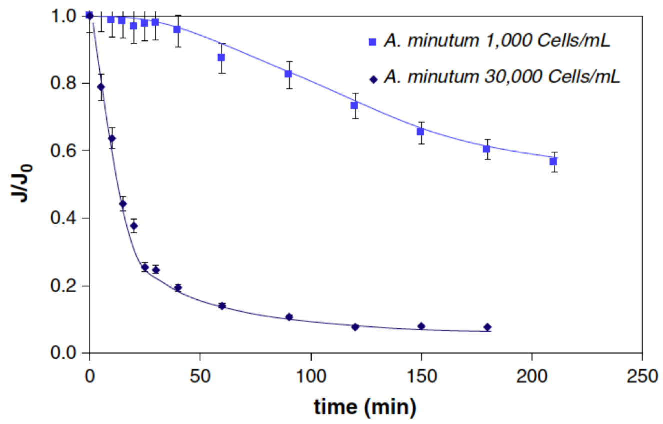


Figure 9.29. Relative permeate flow at 20°C versus time for *A. minutum* microfiltration.

Villacorte (2014) compared sixteen common species representing different major groups of algae, showing potential differences in MF/UF performance due to size and shape of HAB cells. These algal species cover a wide range of shapes and sizes and are presented in Table 9.5. Typical cells are spherical or ellipsoidal, but some are cylindrical (e.g. most diatoms) or elongated (e.g. *Pseudo-nitzschia* spp.). To normalize for such shape heterogeneities, the equivalent spherical diameter was calculated based the estimated cell volume of the cells. The equivalent cell diameters of the sixteen species range from 4 µm (*Microcystis* spp.) to 400µm (*Noctiluca scintillans*) and a median size of 12 µm. The maximum recorded concentrations of each species are also shown in Table 9.5. The species with the highest and lowest recorded concentration are *Microcystis* spp. (14,800,000,000 cells/L 1.4×10^{10}) and *Noctiluca scintillans* (1,900,000 cells/L), respectively. The calculated fouling potential (MFI-UF) of algal cells considering the maximum recorded cell concentrations ranged from 0.4 to 70 s/L^2.

When theoretically considered as spherical or partially spherical objects, the fouling potential of algal cells themselves appear to be low (< 100 s/L^2) even during severe bloom conditions compared to what is reported for ambient surface water, which is typically higher than 1000 s/L^2 (Salinas-Rodriguez 2011). The remarkable difference may be attributed to: 1) presence of particles much smaller than algae (e.g. TEP–like colloidal or particulate materials); and/or 2) algae that are surrounded by TEP–like materials that are blocking the gaps (interstitial voids) between deposited algal cells.

Table 9.5. Cell characteristics, recorded severe bloom concentrations, and calculated membrane fouling potentials of 16 species of common bloom-forming algae (Villacorte 2014).

Bloom-forming algae	**Cell shape (µm)**	**Eq. diam. (µm)** [(a)]	**Sphericity (-)**	**Severe bloom situation**	
				Cells/L [(b)]	**MFI-UF (s/L^3)**
Dinoflagellates					
Alexandrium tamarense	RE	32	0.995	10,000,000	0.38
Cochlodinium polykrikoides	RE	33	0.985	27,000,000	1.07
Karenia brevis	RE	36	0.984	37,000,000	1.60
Noctiluca scintillans	Sp	400	1.000	1,900,000	0.88
Prorocentrum micans	FE	44	0.914	50,000,000	3.06
Diatoms					
Chaetoceros affinis	OC	15	0.968	900,000,000	16.76
Pseudo-nitzschia spp.	0.8*PP	7	0.391	19,000,000	1.01
Skeletonema costatum	Cy	5	0.808	88,000,000	0.78
Thalassiosira spp.	Cy	12	0.867	100,000,000	1.86
Cyanobacteria					
Nodularia spp.	Cy	21	0.543	605,200,000	50.77
Anabaena spp.*	Sp	6	1.000	10,000,000,000	69.80
Microcytis spp.*	Sp	4	1.000	14,800,000,000	68.87
Haptophytes					
Emiliania huxleyi	Sp	5	1.000	115,000,000	0.67
Phaneocystis globosa	0.9*Sp	6	0.933	52,000,000	0.42
Raphidophytes					
Chattonella spp.	Co+0.5*Sp	15	0.665	10,000,000	0.39
Heterosigma askashiwo	Sp	20	1.000	32,000,000	1.68

*Non-marine species of algae; (a) Equivalent diameter of sphere with similar volume as the cell; (b) Maximum recorded concentrations reported in various literatures. RE = rotational ellipsoid; Sp = sphere; FE = flattened ellipsoid; OC = oval cylinder; PP = parallelepiped; Cy = cylinder; Co = cone.

Plugging of PDI fibers by HAB cells and associated AOM may also cause issues during a HAB bloom. To illustrate the potential effect of plugging, Villacorte (2014) calculated the expected loss of active membrane area and the localized flux increase for different species of algae at severe algal bloom situations based on cell size and abundance (Table 9.6). Based on the calculations, some bloom-forming species may cause severe plugging problems in capillary UF. Severe blooms caused by three species of algae (*Noctiluca scintillans, Prorocentrum micans, Nodularia* spp.) may cause complete plugging of capillaries within 30 minutes of filtration (at a flux of 80 L/m^2h). Two other species (*Chaetoceros affinis, Anabaena* spp.) caused more than 50% loss in active membrane area and 2-5 times increase in average flux of the remaining active membrane area. These findings indicate that large algae can cause blooms that may have severe implications to the operation of PDI UF due to plugging, as can high cell numbers.

These theoretical calculations show that plugging of capillaries by HAB cells during severe blooms may substantially increase membrane fouling. Plugging can also cause/enhance non-

Table 9.6. Calculated loss in effective membrane area and localized increase in average flux due to capillary plugging by 16 species of algae during severe bloom situations (Villacorte 2014).

Bloom-forming algae	Diameter (μm)	Cell conc. (cells/L)[(a)]	Active membrane area lost (%)[(b)]	Ave. flux after plugging ($L/m^2.h$)[(c)]
Alexandrium tamarense	32	10,000,000	9	87.5
Cochlodinium polykrikoides	33	27,000,000	25	107.2
Karenia brevis	36	37,000,000	45	146.0
Noctiluca scintillans	400	1,900,000	-----completely plugged------	
Prorocentrum micans	44	50,000,000	-----completely plugged------	
Chaetoceros affinis	15	900,000,000	80	390.7
Pseudo-nitzschia spp.	7	19,000,000	0.2	80.1
Skeletonema costatum	5	88,000,000	0.3	80.2
Thalassiosira spp.	12	100,000,000	5	83.8
Nodularia spp.	21	605,200,000	-----completely plugged------	
Anabaena spp.	6	10,000,000,000	57	184.1
Microcystis spp.	4	14,800,000,000	25	106.4
Emailiana huxleyi	5	115,000,000	0.4	80.3
Phaeocystis globosa	6	52,000,000	0.3	80.2
Chattonella spp.	15	10,000,000	1	80.7
Heterosigma akashiwo	20	32,000,000	7	85.8

(a) Maximum recorded concentrations reported in various literature; (b) Membrane area lost as a percentage of the initial clean membrane area after 30 seconds of filtration at flux 80 $L/m^2.h$; and (c) is the average flux of the remaining active membrane area after 30 minutes of filtration and keeping the permeate flow constant.

back-washable fouling if the accumulated HAB cells are not effectively removed from the feed channel by hydraulic cleanings. Backwashing and chemical cleaning may not be 100% effective in removing foulants in plugged capillaries and the plugged section of the capillaries will continue to increase in the succeeding filtration cycles and cause severe non-back-washable fouling. To minimize plugging problems in PDI UF membranes, the following could be applied:

a) shortening the filtration cycle e.g., from 30 min to 15 mins;
b) reducing the average flux (Note: various PDI UF plants are operating at a more conservative flux (50-60 L/m^2h) in areas where HAB blooms are common).

9.8.5 Summary

In summary, during algal bloom events MF/UF operation can be controlled by lowering filtration flux, applying amended backwash regimes, or through the addition of coagulant. Under optimum process conditions, UF operation can be stabilized by coagulation at doses as low as 0.5-1.5 mg Fe/L to yield less compressible AOM-iron hydroxide floc at this range. An alternative mode of coagulant application (i.e. coating UF membranes with pre-formed flocs of iron hydroxide) has proved efficient in controlling UF operation at very low dose (0.5 mg Fe/L) during severe algal bloom conditions in laboratory studies. To remove algal biopolymers from UF permeate by over 80%, doses as high as 5-10 mg Fe/L are required. Coagulation at such high dose in MF/UF systems results in increased cost and complexities associated with handling and treatment of coagulant-rich sludge and may also be detrimental to the hydraulic performance (non-back-washable fouling and membrane plugging by iron hydroxide flocs) of UF membranes. HAB cells are removed very well, greater than 99%, although a small amount of damage to the cells through shear stress and pressure may cause a release of AOM. Further, stress on live cells during the process may cause them to produce more AOM. Some difference between removal of different cell types has been apparent, particularly for larger cells such as *Noctiluca*, that can cause plugging of PDI UF fibers.

9.9 CARTRIDGE FILTERS FOR REVERSE OSMOSIS PRETREATMENT

9.9.1 Overview

Cartridge filters are a protection measure for SWRO membranes, pumps, and energy recovery devices. The main purpose they serve is to capture particulates in the pretreated source water that may have passed through the upstream pretreatment systems. During a bloom, cellular material should have been removed prior to the cartridge filters so in most cases an increase in differential pressure should not be due to cells. More likely, AOM can build up on the cartridge filters and cause biofouling. Buildup of iron residuals can also occur on cartridge filters. This section describes operation of cartridge filters during a bloom and the subsequent challenges.

9.9.2 Types and configurations

Cartridge filters are fine micro-filters of nominal size of 1 to 25 µm made of thin plastic fibers (typically polypropylene) which are wound around a central tube to form standard size cartridges (Figure 9.30). Often they are the only screening device between the intake wells and the SWRO system with well intakes producing high quality source water.

Figure 9.30. Cartridge filters installed in horizontal vessel.

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 to 1,524 cm 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.

Cartridge filters are typically installed downstream of the pretreatment to capture fine sand, particles and silt that could be contained in the pretreated seawater following GMF. When the source seawater is of very high quality ($SDI_{15} < 2$) 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 on intake well pumps or due to equipment/piping failure.

The main function of cartridge filters is to protect the downstream SWRO membranes, high-pressure pumps and energy recovery devices from damage, not to provide removal of large amount of particulate foulants from the source seawater. A typical indication of 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 by over one unit, 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. The biofouling of the cartridge filters is usually lower if they are chlorinated periodically. Therefore, it is recommended the point of dechlorination of pretreated water to be downstream of the cartridge filters.

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 interim pumping, the pretreatment membranes are more likely to be exposed to pressure surges. If the pretreatment membrane fiber material is weak and it easily breaks under pressure surge conditions, such pretreatment systems are more likely to experience fiber breaks. Broken membrane fibers can release small amounts of particles (or algal cells) into the SWRO feed water, which can cause accelerated membrane fouling. Therefore, 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).

9.9.3 Algal-bloom related challenges

During moderate and heavy algal blooms, cartridge filters usually experience shortened useful life due to accelerated bacterial growth on the cartridge surface and core. Such growth is triggered as a result of the elevated content of easily biodegradable dissolved organic solids released from the algal biomass, such as TEP. The accelerated biogrowth on the cartridge filters peaks during the period of the algae decay, which usually occurs several weeks after the beginning of a typical algal bloom. In order to counter the negative impact on the accelerated biogrowth on the cartridge filters and their premature plugging, and subsequent replacement, source water is usually chlorinated more frequently and the cartridges are exposed to chlorine at a higher dose (1 to 2 mg/L vs. 0.2 to 0.5 mg/L) for a longer period of time (6 hours vs. 2 to 3 hours). Figure 9.31 depicts a heavily fouled cartridge filter during period of severe algal blooms with cell concentrations exceeding 40,000,000 cells/L.

Figure 9.31. Heavily fouled cartridge filter during severe algal bloom event from the Pacific Ocean. Photo: Voutchkov (2013).

Electroadsorptive cartridge filtration is a novel technology, which holds promise for pretreatment of seawater exposed to frequent algal blooms. The electroadsorptive filter media removes TEP through a strong positive charge generated by nanofibers of the mineral boehmite and the tortuous path created by the depth filter media itself. The filter media has a mean flow pore of about 0.7 μm and very high nanofiber surface area that produces a filter with low-pressure drop but a high filtration efficiency and high loading capacity for TEP removal (Komlenic et al. 2013). Test results on Mediterranean seawater indicate that the electoradsorptive cartridge filters can achieve TEP and chlorophyll-*a* removal efficiencies of 40 to 75% and 60 to 80%, respectively.

9.10 REVERSE OSMOSIS

9.10.1 Overview

RO, the core of seawater desalination plant design, is discussed in this section. SWRO is very susceptible to fouling and the risk of fouling is very high during algal blooms. If pretreatment is operated efficiently, the risk to the RO will be greatly reduced. In this section the water quality required for the SWRO process unit is described, as well as the potential for fouling the RO, the most common types of RO fouling associated with algal bloom events, and diagnosis and mitigation of fouling events.

9.10.2 Raw water quality and pretreatment requirements for SWRO

Various membrane manufacturers have developed recommended pretreatment configurations based on seawater intake and water quality parameters to characterize the particulate and organic load of the raw source seawater to be desalinated, such as turbidity, total suspended solids, silt density index and total organic carbon (Chapter 5). These parameters are commonly used to indicate water quality of the raw water and across pretreatment. Of these indicators, only turbidity can be measured continuously online. The others are conducted as discrete measurements on water samples taken periodically. The configuration of the pretreatment system has evolved around specific water sources and aims to produce feedwater of acceptable quality to allow long-term operation.

In areas prone to HABs, the pretreatment system for open intakes requires augmentation by additional treatment steps, as suggested in Table 9.7 for open intakes. Generally, plants with seawater beach wells should attenuate the high organic and particulate load associated with HABs, as the wells provide pretreatment prior to intake, removing a substantial amount of the bacteria, algae, and algal organic matter (Chapter 6).

The majority of membrane manufacturers specify an upper limit of 5 for feedwater SDI_{15} in their guarantees. Field results show that for stable, long-term performance of RO elements, the SDI_{15} of feedwater should be consistently below 4. The SDI limit is dependent upon the lead element flux. Consistently lower feedwater SDI_{15} allows for higher lead element flux. For example, in RO systems with UF pretreatment, the lead element flux may be as high as 35 L/m^2h, while for conventional pretreatment systems, this may only be as high as 32 L/m^2h.

Field results have demonstrated that in the majority of cases water from deep beach wells has very low SDI_{15}, usually less than 1. RO systems, operating with good quality well water feed, practically do not show any pressure drop increase across the membranes or flux decline. Surface seawater, after a conventional pretreatment, usually has SDI_{15} in the 2 – 4 range. A RO system processing feedwater with SDI_{15} in the 2 – 3 range will show stable membrane performance. Membrane cleaning frequency for such feedwater does not exceed once or twice per year. RO systems processing feedwater of higher SDI_{15}, in the 3 - 4 range, usually suffer from some degree of membrane fouling and somewhat higher membrane cleaning frequency may be required. Long-term operation of RO systems with feedwater having SDI_{15} above 4 is not recommended, particularly when average system flux is above 15 L/m^2h.

Many systems are being designed with these higher average and lead element fluxes and such systems will be a major risk during HABs.

Table 9.7. Recommended configuration of pretreatment system according to raw water quality (Hydranautics 2015).

Water source	**Water quality parameters**	**Configuration of pretreatment system**	**Comments**
Seawater beach well	Turbidity < 0.2 NTU TSS < 2 mg/L SDI_{15} < 1.0	Cartridge filtration	If seawater is under influence of brackish water, acidification, and scale inhibitor may be required
Seawater beach well	Turbidity > 0.2 NTU TSS > 2 mg/L SDI_{15} > 1.0	Sand filtration Cartridge filtration	If seawater is under influence of brackish water, acidification, and scale inhibitor may be required
Seawater open intake	Turbidity < 5 NTU TSS < 5 mg/L TOC < 2 mg/L	Acidification Coagulation + flocculation Single stage granular dual media filtration	Short excursion of turbidity up to 20 NTU is possible for few days in year
Seawater open intake	Turbidity < 5 NTU TSS < 5 mg/L TOC < 2 mg/L	Membrane filtration	Short excursion of turbidity up to 20 NTU is possible for few days in year
Seawater open intake	Turbidity 5 -20 NTU TSS > 5 mg/L TOC > 2 mg/L	Acidification Coagulation + flocculation Two stage granular dual media filtration	Short excursion of turbidity up to 30 NTU is possible for few days in year
Seawater open intake	Turbidity 5 -20 NTU TSS > 5 mg/L TOC > 2 mg/L	Acidification Coagulation + flocculation Membrane filtration	Short excursion of turbidity up to 30 NTU is possible for few days in year
Seawater open intake	Turbidity > 20 – 30 NTU TSS > 5 mg/L TOC > 2 mg/L	Settling clarification Coagulation + flocculation Single stage granular dual media filtration	Suspended solids mainly inorganic particles(silt)
Seawater open intake	Turbidity > 20 – 30 NTU TSS > 5 mg/L TOC > 2 mg/L	Settling clarification Coagulation + flocculation Membrane filtration	Suspended solids mainly inorganic particles(silt
Seawater open intake	Turbidity > 20 - 30 NTU TSS > 5 mg/L TOC > 2 mg/L	DAF Coagulation + flocculation Single stage granular dual media filtration	Suspended solids mainly organic and biological particles(algae)
Seawater open intake	Turbidity > 20 - 30 NTU TSS > 5 mg/L TOC > 2 mg/L	DAF Coagulation + flocculation Membrane filtration	Suspended solids mainly organic and biological particles(algae)

Media filtration and membrane filtration are well understood. The prevention of biofilms was misunderstood in the past as discussed in the chlorination-dechlorination section of this Chapter. Recently most of the SWRO plants either eliminated the chlorination – dechlorination process completely or apply shock chlorination.

Another alternative for AOM reduction is addition of activated carbon (Huang et al. 2015) to absorb organics, although this is seldom practiced. A PAC system exists at the Adelaide Desalination Plant in Australia, where PAC can be dosed before and after the disk filters and PAC is removed by the UF, but no serious algal blooms have occurred at that plant to warrant use of the PAC system. Its primary use is for hydrocarbon removal.

9.10.3 Effect of algal blooms on SWRO operation

Algal blooms may result in deterioration of raw water quality depending on the density of the bloom and associated organics (intracellular or extracellular) or indirectly through a reduction in oxygen as algae die off. This may be detected through raw water monitoring of SDI, turbidity, TOC, and AOM such as TEP (see Chapter 5). An additional indicator that may be specifically associated with an algal bloom event is an increase in the concentration of chlorophyll-*a*, thus alerting RO operators of the potential for a fouling event. Chlorophyll background levels are generally specific for different regions and might be in the range of 1-5 μg/L. During a HAB bloom chlorophyll may increase significantly, reaching as high as 120 μg/L (Desormeaux et al. 2009; Franks et al. 2006). Some blooms will produce other algal pigments and thus less chlorophyll-*a*, and thus may not be detected by chlorophyll sensors (see Chapters 3 and 5).

During the HAB period, seawater drawn into the pretreatment system contains much higher concentrations of suspended particulate matter, which significantly increases the solids load in the influent to the filtration system (either media or membrane filtration). The result is a need to increase frequency of backwashing, backwash volume, backwash time and air scour time, which in turn reduces the yield of the pretreatment system and thus the flow to the RO trains. Another problem can be an increased concentration of dissolved organic carbon (or algal organic matter, AOM), which can easily pass the pre-treatment process. This can contribute to an increase in biofouling of RO membrane elements, resulting in an increase in differential pressure.

Alternatively, impacts may manifest in the RO due to fouling if pretreatment does not mitigate the increases in these parameters. SWRO systems with poorly optimized pretreatment may experience a rapid rate of differential pressure increase, due to particulate fouling and biofouling during or following a bloom event (Franks et al. 2006; Unni et al. 2011).

9.10.4 Ferric coagulant SWRO fouling during algal bloom events

When coagulation/flocculation is employed as a part of a seawater feed pretreatment scheme, it is important to dose the coagulant at a rate that will not result in filter breakthrough. While aluminum and ferric salts will both foul the RO process by a similar mechanism, only ferric is discussed as the most commonly used coagulant in SWRO (see Section 9.4.1). Ferric coagulant that will pass the filtration step and reach the membrane elements will result in iron hydroxide fouling of membrane elements (Figure 9.32). Ferric coagulant fouling can be diagnosed in the plant through a rapid increase in differential pressure, a rapid increase in normalized feed pressure and a rapid increase in normalized salt passage (normalization of RO plant data is discussed in ASTM method D4516 (ASTM, 2010) and requires diligent collection of plant data to be of use when diagnosing fouling). This trend is well described in Hydranautics Technical Bulletin, TSB-107, Table 1 (Hydranuatics, 2015).

Figure 9.32. Surface of membrane in the element that was operated in a SWRO system that was exposed to an excessive dose of ferric coagulant.

To demonstrate the localized effect of ferric coagulant foulant at the membrane level, a system performance evaluation was undertaken on a full pressure vessel of elements removed from a full-scale plant (Hydranautics 2015). Each element was individually wet tested and differential pressure analyzed, according to element position in the vessel (Table 9.8).

At standard conditions, the expected differential pressure from a single RO element should be close to 0.2 bar. This differential pressure is observed in tail elements seven and eight. The differential pressure on lead element one is approximately three times the nominal value. Differential pressure decreased gradually along the pressure vessel. Observing the distribution, it is evident that the case of increased pressure is result of poor quality of feedwater. In this particular SWRO system, the problem was partially resolved by optimization of the coagulation process and an additional filtration step, downstream of the media filters. During an algal bloom, selection of the correct dose of coagulant is complicated, as the concentration of HAB cells in the feedwater changes daily, and during exponential growth, even more frequently. The associated loading of AOM will also change with the cell count. It is thus difficult to both properly coagulate cells and avoid overdosing coagulant, unless regular jar tests are performed during a bloom and dosing information is used to make on site changes to minimize any potential overdose. Every SWRO system utilizing ferric coagulation will experience some degree of iron deposit in RO membrane elements.

Table 9.8. Differential pressure of individual elements according to their position in pressure vessel from a SWRO system experiencing excessive ferric coagulant dosing. Wet test for element performance was conducted at standard test conditions, but with 10% recovery rate rather than 8% standard.

Position	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**
Differential Pressure, bar	0.65	0.34	0.31	0.28	0.26	0.24	0.21	0.22

9.10.5 SWRO Biofouling during and following algal bloom events

All raw waters contain microorganisms such as bacteria and algae. The typical size of bacteria is about 1 μm. Most microorganisms will be removed during pretreatment, but those that remain can rapidly reproduce and form a biofilm under favorable conditions (Dow Water and Process Solutions 2015). Microorganisms entering an SWRO system find a large membrane surface where dissolved nutrients from the water are enriched due to concentration polarization, thus creating an ideal environment for the formation of a biofilm. Biological fouling of the membranes will seriously affect SWRO performance. Severe symptoms arising from well-developed biofouling results in blockage of RO element feed channels as shown in Figure 9.33 or even telescoping of elements in the direction of the tail end of the pressure vessel, causing permanent mechanical damage to the elements. The blockage of feed channels is expressed as an increase of differential pressure along the RO trains. An example of differential pressure increase variations in a SWRO plant that experienced severe biofouling, due to continuous chlorination – dechlorination is illustrated on Figure 9.34.

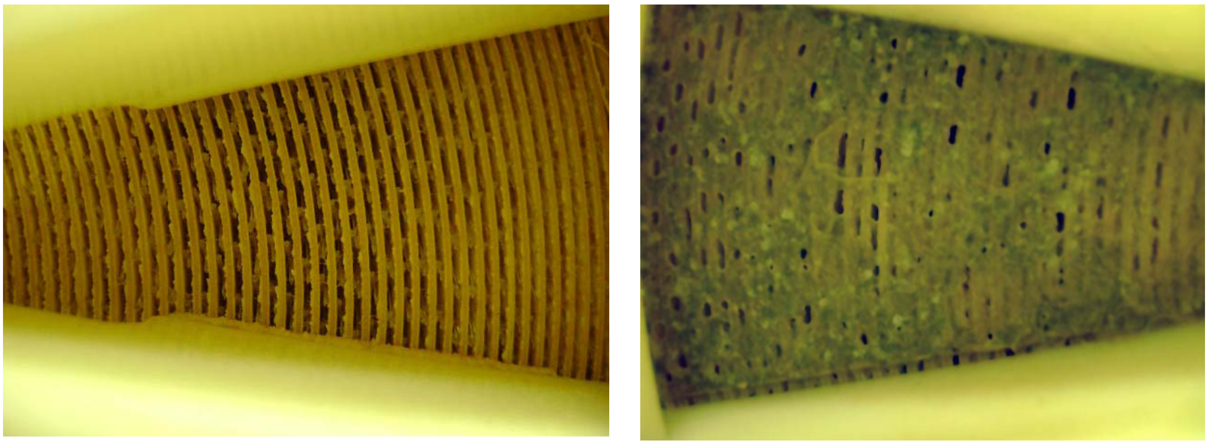


Figure 9.33. Left: Feed channels in clean RO membrane element; Right: Feed channels in an RO membrane element with significant biofouling. Photos: D.H. Paul Training.

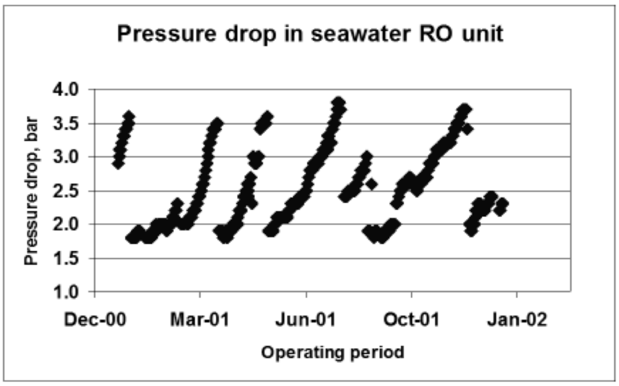


Figure 9.34. Differential pressure increase (shown here as pressure drop) in a SWRO plant as a result of biofouling Figure: Wilf et al. 2007.

The potential for biological fouling should be assessed during the design or pilot phase (where a pilot phase exists) so that the system can be designed accordingly. Warm waters, such as in the Gulf, generally have a higher biofouling potential than cold well waters.

When fouling of SWRO membranes occurs during an algal bloom, this is commonly biofouling. Biofouling will present itself to the operator as a marked increase in differential pressure in the first pass, first stage of a SWRO plant. An increase will develop over 1-2 weeks during or immediately after a HAB. To further diagnose such an issue, the normalized feed pressure will also experience a marked increase during this time and the normalized salt passage will remain normal or be slightly increased. The change in salt passage will distinguish biofouling from ferric coagulant fouling, as ferric coagulant fouling will have a much more rapid increase in normalized salt passage. This trend is well described in Hydranautics Technical Bulletin, TSB-107, Table 1 (Hydranautics 2015).

The regular assessment of the microbiological activity of the feed water should also be part of the operating discipline of a plant so that any increase in microbiological activity can be responded to at an early stage. AOC is one such method for undertaking this monitoring along with TEP and LC-OCD (see Chapter 5). The frequency of sampling and analysis depends upon the risk of biofouling.

Prevention of biofouling is an important consideration for operators as biofouling is very difficult to remove and can permanently damage RO membranes. Once a RO membrane is biofouled, subsequent biofouling will develop more rapidly. Figure 9.35 shows that during a laboratory-scale membrane fouling simulator (MFS) RO experiment, the addition of further AOM to the RO feedwater will cause an increase in differential pressure. Further to this, once a biofilm has 'pre-fouled' a membrane during previous fouling events, the differential pressure (feed channel pressure drop) will rise more rapidly than a brand new membrane despite cleaning of the membrane (Villacorte 2014).

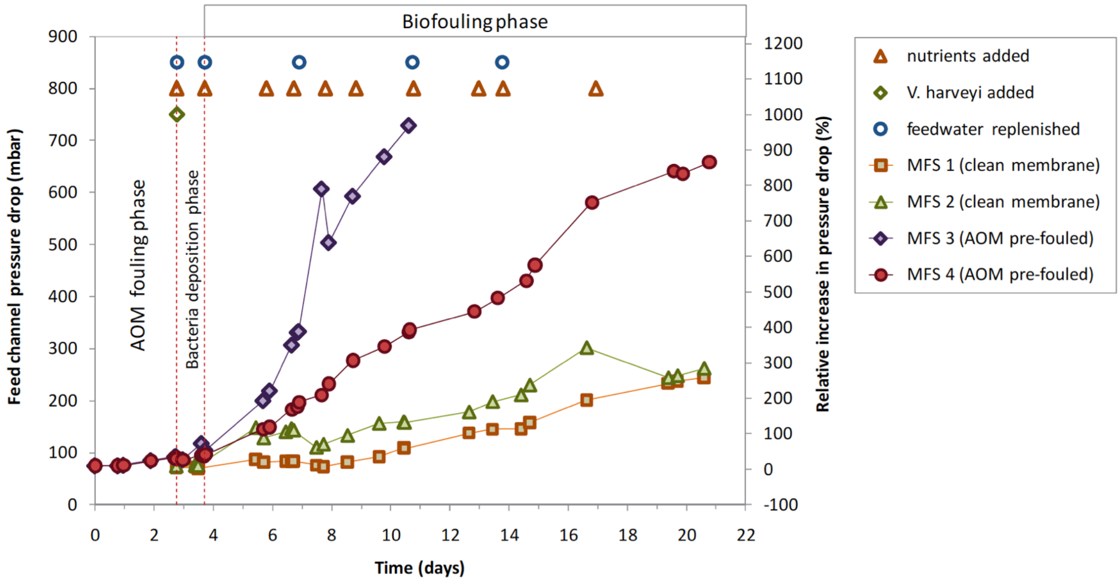


Figure 9.35. Differential pressure (feed channel pressure drop) across the feed channel of MFS cells with clean membrane (MFS 1 & 2) and pre-fouled with AOM (3 & 4). Source: Villacorte 2014.

The most successful approach to minimize biofouling is the limitation or removal of nutrients for microorganisms from the water in order to limit biological growth. This can be achieved through the careful operation of the pretreatment as discussed previously (Section 9.4-9.9). Coagulation is critically important in the pretreatment to remove biofouling. Without coagulation, a high percentage of AOM from the feedwater will pass filtration/flotation steps during blooms. If allowed to pass to the RO membranes, AOM will accelerate biofilm formation. Effective coagulation will enable consistent RO operation at the designed output capacity.

Additionally, the continuous addition of oxidation chemicals such as chlorine may increase the nutrient level because organic substances may be broken down to smaller biodegradable fragments (as discussed in Section 9.2). Dosing chemicals such as antiscalants or acids must be carefully selected because they may also serve as nutrients. Preventive treatments are much more effective than corrective treatments because single attached bacteria are easier to deactivate and remove than a thick, aged biofilm (Vrouwenvelder 2009). Other physical methods are targeted to remove microorganisms in the feedwater with microfiltration or ultrafiltration to deactivate them with UV radiation.

One method for minimizing the attachment of bacteria to a membrane surface and their growth is surface modification of the membrane (Dow Water and Process Solutions 2015).

This concept is available with fouling resistant elements. While popular for application in brackish water applications, they have been less popular for seawater RO elements to date.

Other methods for biofouling reduction are based on chemicals that have a biocidal effect on microorganisms. These sanitization chemicals are applied during the normal operation of the plant either as a continuous dose to the feedwater stream or preferably as a discontinuous (intermittent) dose in certain intervals. Typical treatment intervals are one to four per month (Dow Water and Process Solutions 2015). Non-oxidizing biocides are used as chlorine is oxidizing and cannot be used due to rapid damage of the polyamide layer in the RO.

One biocide chemical commonly used is sodium bisulfite that can be added into the feed stream as a shock treatment during normal plant operation. In a typical application, 500–1,000 mg/L sodium bisulfite is dosed for 30 minutes. Sodium bisulfite should be dosed as sodium metabisulfite (food-grade) that is free of impurities and not cobalt-activated. The treatment can be carried out periodically or when biofouling is suspected. The permeate produced during dose will contain some bisulfite, depending on the feed concentration, the membrane type and the operating conditions. Depending on the permeate quality requirements, permeate can be used or discarded during shock treatment (Dow Water and Process Solutions 2015).

As an alternative to sodium bisulfite, the non-oxidizing biocide DBNPA (2,2, dibromo-3-nitrilo-proprionamide) can also be used to prevent biofouling. DBNPA has been found to be cost effective, has acceptable transportation, storage, stability and handling characteristics, has broad spectrum control (e.g. planktonic and sessile organisms) and is quickly biodegraded in the environment. The standard method to apply DBNPA is shock (intermittent) dosing. The amount of DBNPA used depends on the severity of the biological fouling. With water that has reduced biofouling potential, dosing 10–30 mg/L of the active ingredient for 30 minutes to 3 hours every 5 days can be effective. Because DBNPA is deactivated by reducing agents (such as sodium bisulfite used for chlorine removal), a higher concentration of DBNPA will be required if there is residual reducing agent in the feed water. The concentration of DBNPA should be increased by 1 ppm of active ingredient for every ppm of residual reducing agent in the RO feed water. To remove the dead biofilm, an alkaline cleaning is also prudent. Biocides, their degradation products, and other ingredients in their formulations are not always completely rejected by RO membranes. For this reason, during shock dosing, it may be necessary to discharge the permeate because it may contain slightly elevated levels of organics. Note that although DBNPA is non-oxidizing, it does produce an oxidation/reduction potential (ORP) response in approximately the 400 mV range at concentrations between 0.5 and 3 mg/L. For comparison, chlorine and bromine give a response in the 700 mV range at 1 mg/L, which increases with increasing concentration. This increase in ORP is normal when adding DBNPA and it is recommended the ORP set-point is by-passed during DBNPA addition (Dow Water and Process Solutions 2015).

The optimal frequency for dosing sodium bisulfite or DBNPA will be site-specific and must be determined by the operating characteristics of the RO system. In RO systems operating with biologically active feed water, a biofilm can appear within 3–5 days after inoculation with viable organisms (Villacorte 2014). Consequently, the most common frequency of sanitization is every 3–5 days during peak biological activity (summer) and about every 7 days during low biological activity (winter). Dosing should be considered during a HAB bloom to prevent the onset of biofouling (Dow Water and Process Solutions 2015).

Care should be taken when disposing of biocide-affected waste streams as these can be detrimental in the environment and local permit conditions may prevent operators from disposing of these in the outfall.

9.10.6 SWRO operational strategies during a bloom

At low algal concentrations, the RO system will most likely continue to operate, with no system changes. Sometimes at moderate algae concentrations, the RO will operate at reduced capacity, due to lower output of the feedwater from the pretreatment system. At severe algal concentrations, operators may consider shutting down the RO to avoid severe (and somewhat irreversible) fouling of membrane elements.

Optimization of pretreatment may be the best strategy for maximizing SWRO efficiency during a bloom. Media filtration and membrane filtration optimization are discussed earlier in this Chapter (Section 9.6 and 9.8). Coagulation should be considered, as without coagulation, a high percentage of AOM from the feedwater will pass the pretreatment during the HAB blooms (Section 9.5). If allowed to pass to the RO membranes, AOM will accelerate biofilm formation. Acceleration of biofouling is also attributed to chlorination-dechlorination, as discussed in the chlorination-dechlorination section of this Chapter (9.2 and 9.3). Most SWRO plants either eliminate the chlorination – dechlorination process completely or apply shock chlorination. Shock chlorination should be reconsidered or modified during a bloom. Non-oxidizing biocide dosing to the RO could also be considered during a bloom period (e.g., DBNPA).

While RO operating conditions can be slightly changed during a bloom, the design envelope to produce the required amount of permeate at an acceptable quality is usually set precisely during the design process. When flux is lowered in an SWRO train, the permeate salinity increases, therefore lowering flux too much may result in permeate water that is out of TDS specification. In places where a boron specification for the permeate quality is low (e.g. 0.5mg/L), lowering the permeate flux may quickly produce an unacceptable permeate boron concentration. While some plants may be able to deal with this issue through a second pass design, others may not have this flexibility. Lowering recovery may slightly improve permeate quality and at the same time decrease the lead element flux to alleviate some fouling trends. This will result in higher pumping pressures and thus more energy to produce the same amount of water; however, the minor impact on lead element flux experienced when lowering recovery may not be sufficient to significantly alleviate the effects of fouling. While flux and recovery could be lowered to avoid some fouling during an algal bloom, a better result may be obtained by optimizing the pretreatment to yield better removal of AOM.

9.10.7 Membrane cleaning

Maintenance chemical cleaning for iron removal, applied periodically, will restore membrane performance to clean membrane conditions. The common cleaning procedure for iron removal (in the absence or organics) is recirculation of citric acid solution at a concentration of about 2%.

The cleaning applied for biofouling consists of high pH flushing combined with prolonged soaking. RO cleaning is discussed in greater detail in Appendix 5. A biofilm is difficult to remove because it protects its microorganisms against the action of shear forces and biocide chemicals. In addition, if not completely removed, remaining parts of a biofilm (in the form of AOM) lead to accelerated biofouling (Villacorte 2014). Once an element has biofouled once, it can be difficult to prevent further occurrences. Biological fouling prevention is therefore a major objective of the pretreatment process. The common cleaning procedure for biofouling removal is recirculation of an alkaline cleaning solution at pH 12. In extreme cases, a biocide can be used (Appendix 5).

When iron fouling is combined with biofouling, the removal of foulant from the membrane is difficult and restoration of performance much less effective. In such cases, the cleaning procedure consists of alternate steps of high pH (organics removal) and low pH (iron removal) cleaning. In extreme cases, a biocide can be used between the high pH and low pH step. To adequately restore RO system performance, in the case of mixed iron – biofouling, the cause of biofouling has to be addressed and eliminated (Appendix 5).

9.10.8 Summary

SWRO is very susceptible to fouling, particularly from overdosing ferric coagulants and through biofouling. During HABs, this risk is very high as it is difficult to predict the precise dose of ferric coagulant on any given day of a bloom, and AOM released from HAB cells can pass the pretreatment process and promote biofouling; however, if pre-treatment is operated efficiently, the risk to the RO will be greatly reduced in all but the most severe HABs. Cleaning of the RO after ferric coagulant fouling is not difficult, but cleaning of biofouled elements can be a challenge owing to the sticky nature of the biofilm. Other than optimization of the pre-treatment, only a minor amount of adjustment can be made to the lower flux and lower recovery of a SWRO process unit to prevent fouling from HAB AOM due to tight design parameters to achieve acceptable permeate quality.

9.11 SUMMARY OF BIOMASS REMOVAL IN SWRO

This chapter presents the most common treatment methods used in SWRO systems and discussed the impacts of a non-toxic HAB on plant operations. A summary of a broad set of industry knowledge on the matter is summarized, giving guidance to designers and operators alike. The following major concepts are discussed:

- Avoiding chlorination-dechlorination during a bloom will help to prevent downstream fouling. While chlorination will destroy HAB cells, broken cells will release AOM, causing downstream fouling (Section 9.2 and 9.3). The AOM is more important in terms of negative impacts than the intact cells, which are more easily removed during pretreatment.
- DAF is a good choice for cell removal in areas likely to experience heavy algal blooms, as it lifts and removes cells from the seawater in a relatively gentle manner (Section 9.5).
- GMF (Section 9.6) and UF (Section 9.8) can deal with cell removal in lighter blooms, but using a DAF upstream for heavy blooms is prudent.
- Coagulation assists all three pretreatments (DAF, GMF, and UF) and acts to remove AOM more effectively than the pretreatments alone. Less AOM in the pretreated water is also an important objective to alleviate SWRO fouling (Section 9.4).
- Microstrainers can foul with algal material and may have shorter cycle times during a bloom (Section 9.7).
- Cartridge filters are installed to protect the SWRO. These will most likely foul during or following a bloom and require regular replacement during longer more intense blooms (Section 9.9).
- There is little an operator can do to adjust the SWRO flux and recovery. A better strategy may be to focus on getting the pretreatment working efficiently and, consequently, the SWRO will operate more effectively. SWRO fouling from excess ferric coagulant and biofouling may still be an issue during intense blooms and may require focused cleaning after a bloom (Section 9.10).

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