

# **PRACTICAL APPLICATION OF SULFATE REDUCTION MEMBRANE TECHNOLOGY IN OILFIELD OPERATIONS**

**Kevin A. Juniell, NATCO Group – Houston, TX**  
**Robert Weston, NATCO Group – Camberley, England**  
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## **I. Introduction**

In oilfield offshore operations, there are instances when it is necessary to mix formation water with seawater. As an example, many offshore water injection operations use seawater as the injection fluid. As seawater is injected into the producing formation, it mixes with the formation water and is eventually co-produced with the oil and gas through the producing wellbores. In other cases, seawater is used as the wash water for offshore crude oil desalting operations, thereby mixing with the formation water inside the oil-treating vessel. Whenever seawater and formation waters are mixed, there is concern about chemical incompatibility effects, in particular the deposition of sulfate scales – calcium sulfate ( $\text{CaSO}_4$ ), barium sulfate ( $\text{BaSO}_4$ ) and strontium sulfate ( $\text{SrSO}_4$ ). Usually, seawater contains approximately 2,500-3,000 parts per million (PPM) of sulfate ion ( $\text{SO}_4$ ). If the formation water contains a significant level of calcium, barium or strontium, then the probability for sulfate scale to form increases. One effective way to minimize sulfate scale formation is to remove the sulfate ions from the seawater.

This paper describes the problems resulting from incompatibility of seawater and formation water. Then, a brief overview of the basic principles of membrane separation of sulfate ions is given. Also discussed are examples of offshore oilfield operations that require the mixing of seawater and formation water. The paper contains a discussion of the basics of process design of the sulfate removal membrane system as well as a discussion of certain important operational and maintenance considerations (pre-treatment, chemical treatment, chemical cleaning, operator training, data normalization, etc.). Finally, the paper has a summary of conclusions regarding proper design and operation of sulfate removal membrane plants in an oilfield environment. Please note that this paper is not meant to be an all-inclusive or exhaustive discussion of the topic. References are provided at the end of the paper for convenience in doing further research. It is hoped that enough information has been presented here to afford a good basic understanding of the relevant issues and thought processes involved with selecting and operating sulfate reduction membrane systems.

## **II. Mixing Seawater with Formation Water – What is the Danger?**

As previously mentioned, it is sometimes necessary to mix seawater with formation water in oilfield operations. When this becomes necessary, it is imperative that the operator be aware of the risk involved with mixing two waters that may be chemically incompatible. Formation water normally has a mixture of various dissolved solids in solution at a given temperature and pressure. Likewise, the amount and types of dissolved species in seawater is a function of the solubility of the specific ions at a certain pressure and temperature. The concentration of each ionic species in the formation water or in the

seawater is based on the solubility of a particular species at some equilibrium value for the given temperature and pressure. When two chemically different waters are mixed, the equilibrium of the separate waters is destabilized and the mixture will reach a new equilibrium state. As a result of the system re-establishing its equilibrium condition, it is possible that some of the dissolved ions will precipitate and form solid scales that deposit in the formation, tubing, piping or vessels. Of particular concern when seawater and formation waters are mixed are sulfate scales – calcium sulfate ( $\text{CaSO}_4$ ), barium sulfate ( $\text{BaSO}_4$ ) and strontium sulfate ( $\text{SrSO}_4$ ). Barium and strontium sulfate scales are problematic for two reasons. First of all, these scales are very resistant to chemical or mechanical removal. Once they deposit in tubing and surface piping, the only practical solution is to install new equipment. Secondly, these scales usually contain naturally occurring radioactive material (NORM) that may expose platform operators to harmful levels of radiation. Special handling and disposal methods must be employed at a relatively higher cost to the operation.

Seawater normally has approximately 2,500 PPM to 3,000 PPM of sulfate ions and formation water may have substantial amounts of calcium, barium or strontium ions. Therefore, prior to mixing seawater with produced water, it is recommended to perform a compatibility analysis that will determine the scaling tendency of the mixture.

This is not a new phenomenon, so operators of seawater injection plants have tried to address this issue in various ways, some of which are listed below:

- Re-inject produced water in lieu of seawater
- Periodic scale inhibitor squeezes
- Inject water from an aquifer with compatible water
- Inject low-sulfate seawater

Each of these options has inherent advantages and disadvantages. Therefore, the operating company must perform a cost-benefit analysis to determine the life-cycle costs associated with each option. Once the costs for each option have been “normalized”, then a selection process can be undertaken to choose the proper technology for the project.

### **III. Principles of Nanofiltration**

Nanofiltration is a membrane process that selectively reduces the concentration of sulfate ions to produce low-sulfate seawater (LSW). The process is similar to reverse osmosis in that a high-pressure water stream is fed to a semi-permeable membrane. The pressure must be high enough to force the water through the semi-permeable membrane while the membrane removes molecules that do not freely pass through. Nanofiltration differs from reverse osmosis because reverse osmosis tends to remove all ions in the water stream while nanofiltration targets the larger divalent ions, such as the sulfate ion. See Figure 1 and Table 1. Figure 1 shows a simplified schematic of the nanofiltration system. As indicated, the high-pressure feed stream enters the membrane and is split into two outlet streams – a low-sulfate seawater product stream (permeate) and a high-sulfate reject stream (concentrate). Table 1 shows a “typical” concentration of the primary

dissolved solids in seawater before and after passing through a nanofiltration membrane. As is clearly shown, the membrane rejects most of the sulfate ions and other larger ions, while most of the other dissolved components, notably chloride ions, freely pass through the membrane. This is the desired effect when using LSW for waterflood injection. Many injection formations contain clays that are sensitive to the salinity of the injected water, i.e. the clays may swell if the water is too "fresh". Hence, it is important to maintain the chloride content of the seawater during the sulfate reduction process.

Sulfate reducing nanofiltration membranes for use in oilfield applications are the result of a joint effort between the Marathon Oil Company and the DOW Chemical Company. Figure 2 shows a cutaway of the membrane product (named SR90-400), manufactured by DOW's Filmtec Division. The SR90-400 membrane is defined as a thin-film composite membrane consisting of three layers: a polyester support web, a microporous polysulfone interlayer, and an ultra-thin barrier layer on the top surface. The barrier layer has a negative surface charge that repels anions (negatively charged ions). Monovalent anion rejection (for example chloride ions) decreases with increasing salinity because the high concentration of positive cations starts to shield the negative surface charge of the membrane surface. Divalent ions (for example sulfate) have a high charge density such that the shielding effect is negated and the membrane barrier layer preferentially rejects these ions.

The membrane is constructed into a spiral wound element. This spiral wound configuration provides very high surface area per unit volume. By current DOW manufacturing standards, each element is 40" long and 8" outer diameter (OD). Six of these spiral wound elements are inserted into a pressure-containing vessel to create what is termed a separation unit. Elements are connected together in series within a pressure vessel. Pressurized water flows into the pressure vessel and through the channels between the spiral windings of the element. As the bulk seawater flows parallel to the membrane surface a portion of this stream passes through the membrane leaving behind the rejected particles (sulfate ions, suspended solids) in the concentrate stream. Since there is a continuous flow across the membrane surface, the rejected particles do not accumulate but instead are swept away by the concentrate stream. Thus, one feed stream is separated into two exit streams: the solution passing through the membrane surface (permeate) and the remaining reject stream (concentrate). See Figure 3. The reject stream becomes more and more concentrated with sulfate ions as it passes through each successive element prior to entering the final element in the sequence. The permeate stream of each element is collected in the common permeate tube installed in the center of each spiral wound element and flows to a permeate collection pipe outside the pressure vessel. The pressure of this stream has been reduced as a result of passing through the membrane.

In practice, the membrane vessels are normally arranged into a two-stage array with an approximate 2:1 staging ratio (i.e. twice the number of membrane vessels in Stage 1 as compared to Stage 2). See Figure 4. In this array, the seawater feed stream enters the Stage 1 bank of membranes for sulfate reduction. The permeate stream from Stage 1 is taken away as product. The concentrate stream from the Stage 1 bank of membranes is

routed to the inlet of the Stage 2 bank of membrane units for further sulfate reduction. Permeate from the Stage 2 bank is commingled with the permeate stream from the first bank and this combined stream is taken away as total LSW product. The concentrate stream from the Stage 2 bank is routed overboard for disposal. Normally, it is expected that 50% of the seawater by volume is converted to LSW per stage. Hence, in the 2:1 array, the total conversion of feed seawater to LSW is 75%, i.e. if 1,000 gallons per minute (GPM) of seawater is fed to the 2:1 array then 750 GPM will be converted to LSW and 250 GPM will be disposed of as concentrate (or waste). The 2:1 mode of operating provides a good balance between percent recovery of LSW and minimizing the required weight and size of the package.

#### **IV. Offshore Oilfield Applications**

The primary reason for installation of sulfate reducing membrane technology to date has been to minimize the formation of sulfate scales during seawater injection operations. In these instances, the incompatibility (or sulfate scaling tendency) of the seawater/formation water mixture is deemed severe enough to justify the use of sulfate reduction membranes. The logic is to remove the sulfate ion from the seawater as a preventive measure.

Another reason used to justify installing sulfate reduction membrane technology is to prepare seawater for use as wash water for crude oil desalting operations. The operator desired to use seawater to wash salt from oil, however, the sulfate scaling tendency between the seawater and the residual formation water contained in the oil was too severe. Hence, the operator elected to install a sulfate reduction membrane train to selectively reduce the sulfate content of the seawater thereby making it suitable for use as a salt wash fluid.

It is worth mentioning that there is a significant secondary benefit associated with reducing the sulfate content of seawater prior to injection for waterflood or pressure maintenance purposes. Sulfate reducing bacteria (SRB) feed on sulfate as a source of oxygen. In the process of digesting the sulfate ions, these microorganisms convert the sulfur component of the sulfate ion to hydrogen sulfide ( $H_2S$ ), a corrosive gas. This explains why some reservoirs began seawater injection operations free of  $H_2S$  but experienced increasing levels of  $H_2S$  as more and more seawater was injected over time. The SRB's thrived on the sulfate in the seawater and quickly multiplied in this food-rich environment. The more the population of SRB's grew, the more the production of  $H_2S$  grew. In operations with sulfate reducing membranes installed, reservoir souring has not occurred. In addition, the need for exotic corrosion-resistant materials and/or expensive chemicals has been eliminated.

As of the date of this paper, there have been a total of fourteen sulfate reduction plants awarded. Of that number, seven are currently operating, five are currently under construction or waiting commissioning and one plant has been decommissioned. The remaining plant was built but never started. Of the plants that are in operation, five are in the North Sea and the other two are offshore West Africa. Of the plants that are waiting

commissioning, two are offshore West Africa, two are offshore Brazil and the other one is in the North Sea. See Table 2.

Most of these plants have been installed to address concerns regarding the formation of sulfate scales. However, there is at least one installation that makes use of low-sulfate seawater for crude oil desalting.

#### **V. Process and Instrumentation Design Basics**

The major process information required to properly evaluate and design a sulfate reducing membrane system is as follows:

- Required product flow rate
- Water temperature at sulfate reduction system inlet (maximum/normal/minimum)
- Seawater source; depth of intake
- Oxygen content at sulfate reduction system inlet
- Existing pre-treatment – Yes/No
- Specify water quality needed after sulfate reduction treatment
- Complete water analysis including conductivity, pH, ionic analysis, total suspended solids, particle count/distribution, SDI, bacteria count, chlorine content
- Design pressure of upstream and downstream system
- Design temperature of upstream and downstream system
- Injection water profile
- Plot area available
- Height restrictions
- Access restrictions
- Lifting weight restrictions
- Minimum hazardous area rating

It will likely be necessary to collect representative formation water and seawater samples for physical/chemical analysis in order to provide the detailed information that is required. It is best if project specific information is used in the process design phase. However, in the event that accurate project-specific information cannot be provided, estimates based on best engineering judgment or previous similar experience can be used.

A very important aspect of the process design is establishing the optimal feed temperature. The feed temperature has a significant influence on the sulfate content in the product water; the lower the feed temperature, the lower the product seawater sulfate content and vice versa. However, lower temperatures also require the feed seawater to be pressurized to higher pressures in order to maintain the flow per unit area of membrane surface (flux rate). The feed temperature also affects the performance of the deoxygenation equipment; the higher the feed temperature the lower the oxygen content of the effluent seawater and vice versa. Hence, the process at lower temperatures will consume more oxygen scavenger chemical. There is thus a trade off between available power on an offshore installation, amount of pre-heat available, the operating cost of oxygen scavenger chemical and the desired sulfate content in the low-sulfate product.

The pre-treatment of the seawater prior to entering the nanofiltration membrane package is another aspect of the process design requiring attention. Proper pre-treatment of the seawater supplied to the nanofiltration membranes is necessary to maximize the efficiency and life of the membrane elements and to ensure trouble-free operation. Pre-treatment requirements include:

- Removal of fine suspended solids that can plug or block the membrane surface;
- Prevention of biological growth on the membrane surface;
- Prevention of scale formation on the membrane surface during concentration of the feed;
- Removal of any oxidizing biocides (e.g. chlorine) which can damage the membrane; and
- Pressurization as required for achieving nanofiltration separation.

The semi-permeable membrane will, in effect, act as a filter and plug with suspended solids that may be in the process stream. Upstream filtration to remove the majority of the suspended solids from the seawater will not only decrease downtime for cleaning, but also extend membrane life. The Silt Density Index (SDI) has been used as a measure for determining the degree of cleanliness required by the pre-treatment filtration system. SDI is a measure of the presence of all particles greater than 0.45 micron. There is no direct correlation between SDI and particle removal efficiency, except that, generally, the better the particle removal efficiency the lower the SDI. It has been found that SDI values in the order of 3 provide adequate water quality for proper operation of nanofiltration membranes. To achieve this, most nanofiltration pre-treatment systems comprise some form of coarse filtration (70-80 microns) followed by fine filtration to meet the requirement for SDI of 3. The majority of the existing systems in operation have opted to use multi-media filters for the fine filtration step. However, the use of cartridge filters is also a viable option. If multi-media filters are selected for this purpose, it should be noted that cationic polymeric filtration aids not be used, as they will severely foul the membrane surface if bleed through occurs. The need to save space and weight on some offshore facilities structures is driving nanofiltration membrane suppliers to develop other lighter more compact methods for accomplishing the removal of suspended solids.

Control of biological growth in the pre-treatment system is usually handled via injection of chlorine. However, chlorine is damaging to the membranes and is removed by injecting a de-chlorination chemical upstream of the membrane equipment (commonly bisulfite oxygen scavenger is used in offshore applications). Thereafter, periodic shock dosing with a non-oxidizing biocide chemical is done to protect the membranes from biofouling.

As seawater flows through the pressure vessel from the inlet to the outlet, the reject stream becomes more and more concentrated with sulfate ions. As a result of the increased concentration of sulfate ions in the reject stream, the tendency for calcium sulfate scale and/or magnesium sulfate scale to form on the surface of the membranes

also increases. To control the formation of scale deposits on the membrane surface, the membrane feed stream is dosed with an anti-scalant chemical.

The pressure required to force the seawater through to the permeate, or product, side of the membrane can be as high as 25-35 BARg or higher, depending on the water temperature. A booster pump is provided upstream of the membrane unit to provide sufficient pressure to feed the membranes as well as to satisfy the pressure requirements for any equipment downstream of the membranes (i.e. deaerator tower, injection pumps, etc.).

Before seawater is injected, the oxygen is removed to minimize the corrosive effects of oxygen on metallic components in the system. Traditionally, a mass-transfer tower (gas-strip or vacuum type) is used for this purpose. The location of the tower in relation to the sulfate membrane package must be given consideration by the operator. Table 3 presents a few considerations that should be taken into account when an operator is evaluating the process design. Based on the information contained in Table 3, an operator can make a value judgment as to which position the tower should be placed. It should be noted that among the existing operating units can be found instances of the tower positioned upstream of the membrane package as well as instances of the tower positioned downstream of the membrane package.

There are also process design issues related to system control and system performance monitoring that must be considered. As previously discussed, the sulfate rejection of the membranes is strongly dependent on the inlet temperature. In systems that involve seawater pre-heating, it becomes obvious that proper attention be given to the setting up of the temperature control loop so as to insure proper operation of the membrane system. Also, the membranes are sensitive to pressure surges or pulses; pressure surges can cause mechanical damage to the membranes. Hence, proper protection against pressure surges due to sudden flow stoppage, pump trips, or other such occurrences must be incorporated into the instrumentation and controls design.

The package control system is normally designed for automatic operation with limited operator intervention. The package is usually controlled locally via a unit control panel. This panel contains all the necessary instrumentation to enable safe running and control of the package.

The normal operating mode is for semi-automatic startup sequences, and in this mode the operator initiates a sequence that then proceeds automatically through a number of stages until the entire system is fully operational. Automatic shutdowns can be programmed to occur on fault.

The membrane system startup sequence includes a pre-flush followed by automatic startup of the membrane booster (feed) pump. Low sulfate product water is initially discharged overboard for a time period to allow for membrane performance to stabilize. Once the quality is proven then the LSW can be forwarded to the remainder of the downstream components.

Stop pushbuttons can be provided to enable planned shutdown as required. Each planned shutdown sequence includes a post-flush

Suitable plant instrumentation and safety interlocks should be provided to automatically shutdown the membrane trains and/or pre-treatment section on fault. A nanofiltration package can be automatically shutdown for a variety of reasons including: high differential pressure across a stage of membrane vessels, detection of chlorine, excessively high conversion, high feed pressure and low flow rates.

The trains can be stopped due to low product demand or for routine cleaning or for maintenance. The sequence is automatic and ensures that the membranes are flushed and that the system does not experience hydraulic shock.

Reduction-oxidation (redox) meters can be installed and positioned in an instrument sample line to detect the presence of chlorine. It is imperative that they are regularly cleaned to ensure correct operation, so the design makes provision for easy access and withdrawal.

At present, on-line instrumentation for the measurement of sulfate content is not sufficiently reliable for field use. On-line conductivity measurements can be included to provide a surrogate for sulfate. The conductivity of the feed water and product water from Stages 1 and 2 are continuously monitored. The actual sulfate content can be measured off-line via laboratory apparatus to test product quality as required. Of course, proper measurement devices to monitor basic process parameters as inlet and outlet flow rates, inlet and outlet pressures, inlet and outlet temperatures should be installed in the system. In addition, proper fluid sampling equipment should be installed to measure inlet, outlet and inter-stage sulfate concentrations.

## **VI. Operational and Maintenance Considerations**

Figures 5 and 6 depict an actual membrane skid packaged unit. From these figures, it can be seen that the skid package comprises mostly pressure vessels, membrane elements, small-diameter piping, instrumentation and valves. Hence, from a maintenance standpoint, the primary concern is maintenance of the membrane elements due to fouling. As stated earlier, it is for this reason that proper and careful consideration must be given to the design of the pre-treatment system, including the proper application of production performance chemicals. Also to be considered is the location of the seawater intake (lift pump suction). The seawater intake point should be located such that other streams will not contaminate it; e.g. sewage discharge, discharges from drilling operations, etc. It must be understood, however, that over a period of time, the performance of the membranes will be adversely affected by fouling agents. The primary fouling agents are suspended solids, bacterial microorganisms and scale deposits. The effects of such long-term fouling can be monitored so that periodic cleaning of the membranes can take place. Proper and timely performance of this major cleaning is vital to the longevity of the membranes, thereby reducing the operating costs (membrane replacement costs) of the plant.



Performance data (e.g. flow rate, temperature, pressure, conductivity) is communicated to the central control facility for data logging. This data set can then be processed, or normalized to detect fouling trends. The normalization process will show how the membrane plant is performing against the base line conditions, enabling the prediction of manual intervention or cleaning of the membranes. Figure 7 shows a graph containing the results of a set of normalized field data. The graph indicates the original "clean" condition of the membranes. It can also be noted that after each cleaning, the membrane performance does not completely return to the original condition. This offset in performance can be characterized by dividing the current performance condition by the original performance condition. The product of this division is termed the fouling factor. The fouling factor is monitored after each cleaning. The average life expectancy of a properly designed and maintained membrane element is 5-7 years.

The process of cleaning the membranes requires a system shutdown. Hence, it is critical that the time interval between cleaning cycles be as long as possible and that the cleaning process restore the membrane elements as close as possible to the "as-new" condition after each cleaning. Membrane cleaning is usually done on-site using tanks, pumps and piping installed specifically for this purpose. A critical part of the cleaning process is the assessment of the fouled membrane elements. This assessment should determine the required types of chemicals needed to maximize the effectiveness of the cleaning, thereby maximizing the time between cleaning cycles. It is best to use experienced and well-trained personnel to carry out this important analysis.

## **VII. Conclusions**

1. Injecting seawater into a producing formation requires knowledge of the potential incompatibilities between the seawater and the formation water. A major concern is the tendency of the seawater/formation water mix to precipitate barium sulfate or strontium sulfate scales.
2. Injection of low-sulfate seawater is a proven and effective method of treatment to reduce sulfate scale formation in seawater injection operations. Low-sulfate seawater is produced using the patented Marathon/DOW nanofiltration membrane.
3. Sulfate reduction performance of the membranes is largely a function of the feed temperature; the lower the feed temperature the more efficient the separation of sulfate from the seawater.
4. Proper pre-treatment to protect the membranes against rapid fouling by suspended solids, biological microorganisms and scale deposition is critical to the performance and longevity of the membranes.
5. During operation, the membranes will slowly become fouled to a point that performance declines below specification. At this point, the membranes must be taken out-of-service for a major cleaning. Performance data can be monitored in

order to detect fouling rates so that major cleaning of the membranes is properly scheduled.

6. Trained personnel who can properly assess performance data and fouling trends to make sound recommendations concerning cleaning cycles best handle maintenance of the membranes.
7. When properly maintained, membrane life is expected to be 5-7 years minimum.

## LIST OF FIGURES

- FIGURE 1. PROCESS SCHEMATIC
- FIGURE 2. SR90-400 MEMBRANE CUTAWAY
- FIGURE 3. CROSSFLOW CONCEPT
- FIGURE 4. TYPICAL 2:1 TWO-STAGE ARRAY
- FIGURE 5. SULFATE REDUCTION PACKAGE  
INSTALLATION
- FIGURE 6. END VIEW OF SMALL DIAMETER  
INTERCONNECTING PIPE
- FIGURE 7. NORMALIZED FOULING FACTOR GRAPH

FIGURE 1. PROCESS SCHEMATIC

SIMPLIFIED NANOFILTRATION DIAGRAM

Sulphate Removal

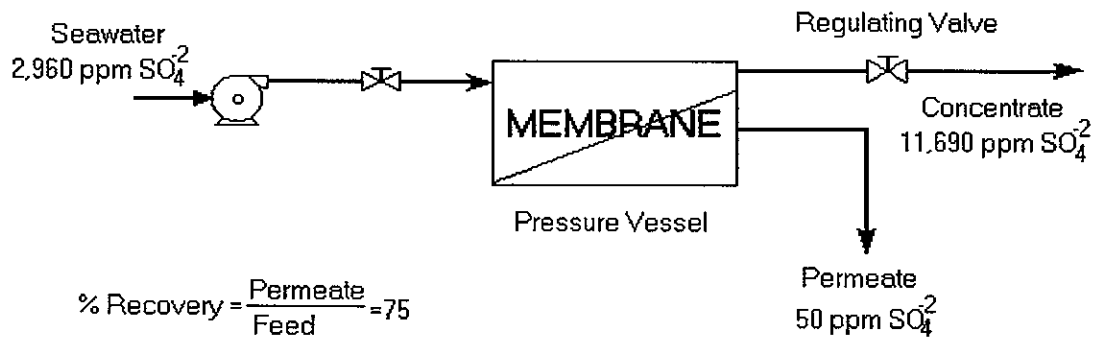


FIGURE 2. SR90-400 MEMBRANE CUTAWAY

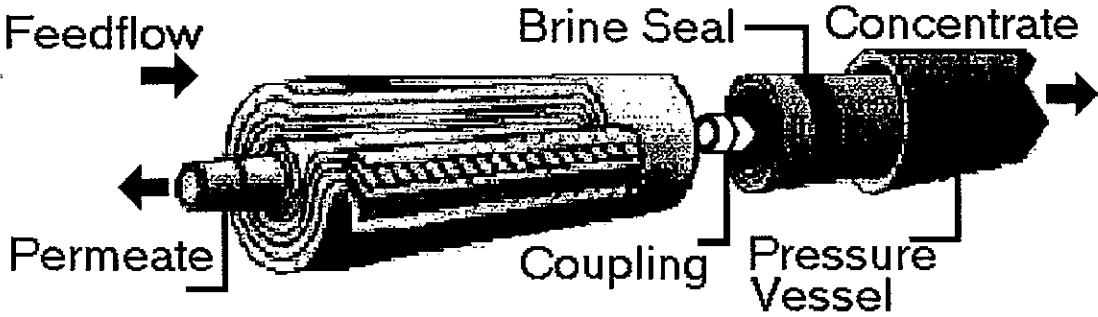
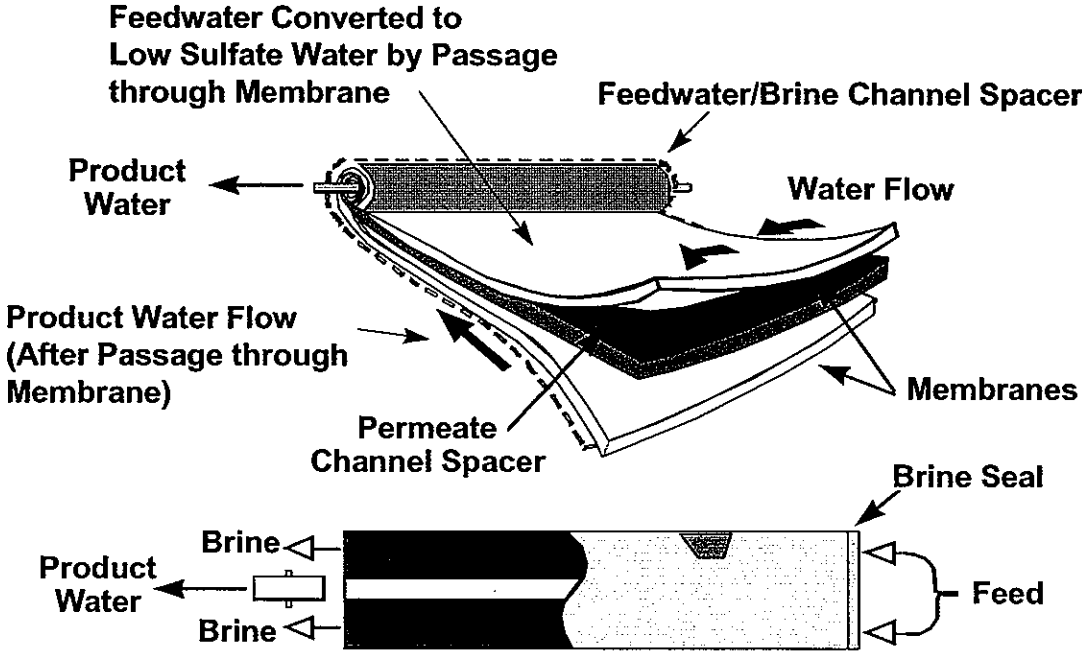


FIGURE 3. CROSSFLOW CONCEPT

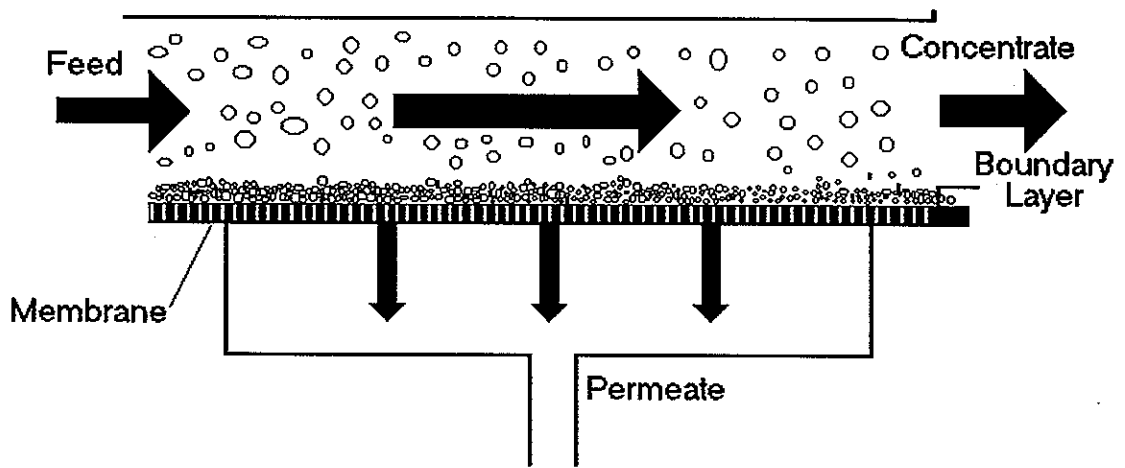


FIGURE 4. TYPICAL 2:1 TWO-STAGE ARRAY

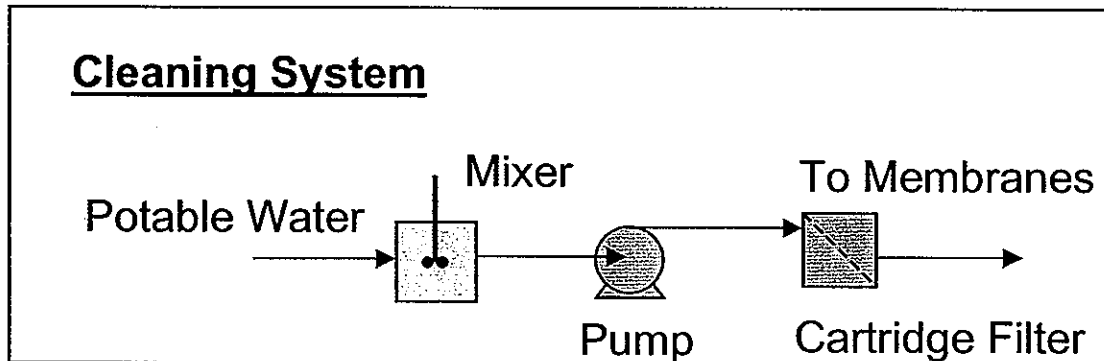
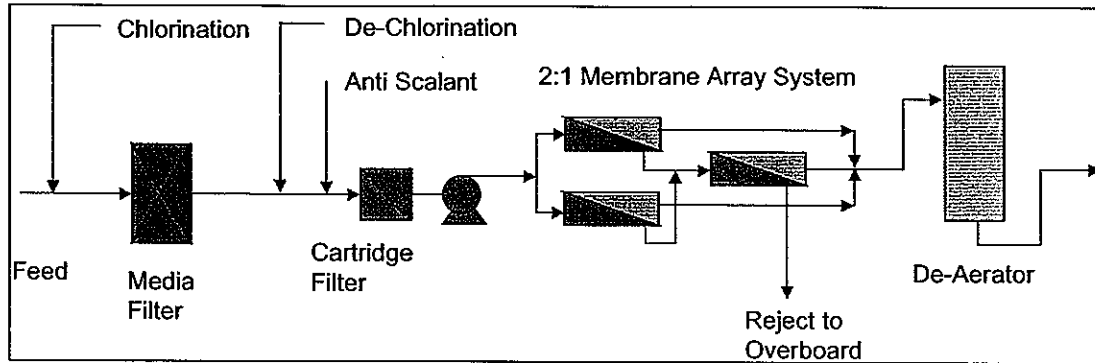


FIGURE 5. SULFATE REDUCTION PACKAGE  
INSTALLATION





FIGURE 6. END VIEW OF SMALL DIAMETER INTERCONNECTING PIPE

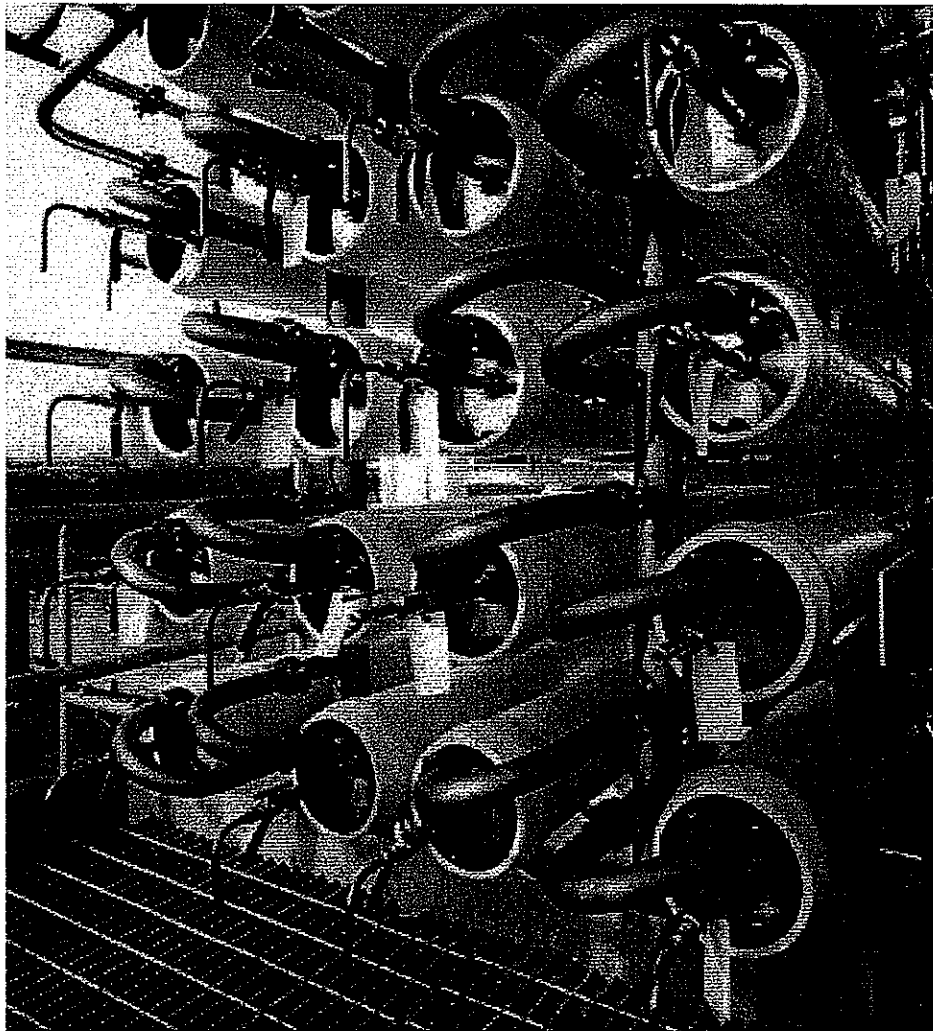
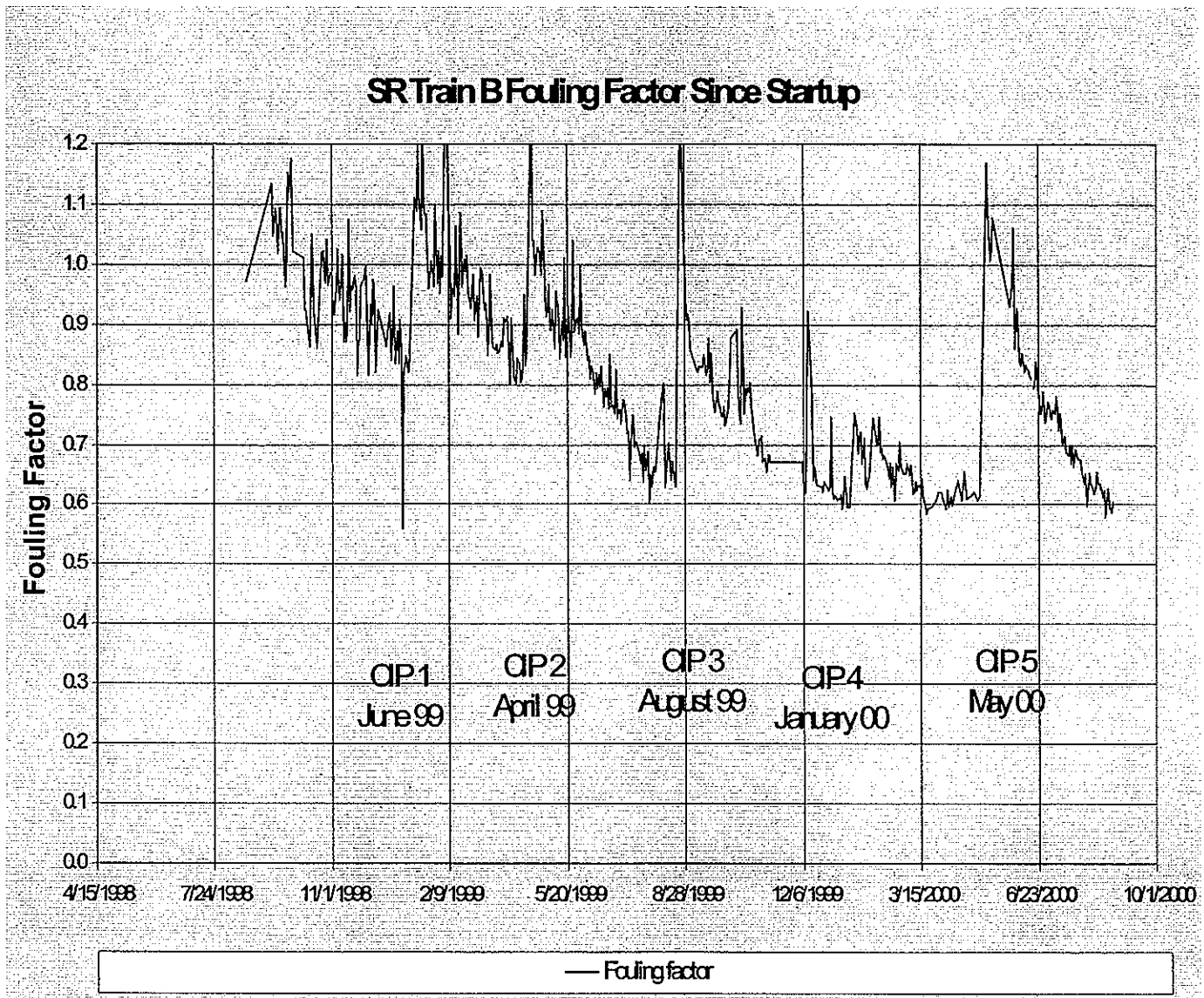


FIGURE 7. NORMALIZED FOULING FACTOR GRAPH



## LIST OF TABLES

- TABLE 1. CONCENTRATION OF IONS BEFORE AND AFTER MEMBRANE TREATMENT
- TABLE 2. LIST OF MEMBRANE INSTALLATIONS WORLDWIDE
- TABLE 3. CONSIDERATIONS FOR LOCATION OF THE MASS-TRANSFER COLUMN

TABLE 1. CONCENTRATION OF IONS BEFORE AND AFTER MEMBRANE TREATMENT

Constituent	Feed (mg/l)	Product (mg/l)
Sodium	11200	10690
Potassium	370	320
Calcium	400	330
Magnesium	1400	330
Chloride	19750	19000
Sulphate	2650	48
Bicarbonate	140	20
Total	35910	30736

**TABLE 2. LIST OF MEMBRANE INSTALLATIONS  
WORLDWIDE**

<b>Year Installed</b>	<b>Operator/Platform</b>	<b>Location</b>	<b>Design Flow Rate (BPD)</b>	<b>Current Status</b>
1990	Marathon Brae "A"	North Sea	120,000	Operating
1994	Agip Tiffany	North Sea	100,000	Operating
1994	Marathon Ewing Bank 873	Gulf of Mexico	20,000	Decommissioned
1998	BP ETAP	North Sea	20,300	Operating
1998	Texaco Petronius	Gulf of Mexico	72,000	Never Started
1998	Kerr-McGee Janice	North Sea	70,000	Operating
2000	Amerada Hess South Arne	North Sea	120,000	Operating
2002	Amerada Hess Ceiba	Equatorial Guinea	135,000	Operating
2002	TotalFinaElf Girassol	Angola	400,000	Operating
2003	Petrobras/SBM Roncador	Brazil	90,000	Awaiting Commissioning
2003	ExxonMobil Kizomba "A"	Angola	350,000	Awaiting Commissioning
2003	Statoil Heidrun	North Sea	200,000	Awaiting Commissioning
2004	ExxonMobil Kizomba "B"	Angola	350,000	Under Construction
2004 (?)	Petrobras Albacore Leste	Brazil	200,000	Under Construction

**TABLE 3. CONSIDERATIONS FOR LOCATION OF THE MASS-TRANSFER COLUMN**

<b>UPSTREAM DEAERATION</b>	<b>DOWNSTREAM DEAERATION</b>
sized for 100% flow	sized for 75% flow – size, weight and cost savings
Silicone based antifoams cannot be used so may need to design tower for lower irrigation rates to avoid use of antifoam, further increasing size of tower	Deaerator can be designed for maximum irrigation rates to keep size of tower to a minimum
Some mechanical dechlorination achieved in deaerator, reducing dechlorination chemical costs	Chemical dechlorination required upstream of sulfate reduction membranes
Good biological control as oxygen is only present upstream of the membrane system, when the water is still chlorinated	Potential for aerobic biological growth as aerated water is present in membrane system in absence of chlorine
316L SS high pressure piping may be used as the presence of deaerated seawater in membrane system will reduce potential for pitting corrosion	High grade SS high pressure piping required for aerated seawater
If deaerator feed is heated to obtain better tower performance, then this will adversely affect membrane performance	If heating is required for improved deaerator tower performance, this will not affect membrane performance
Extra pumps may be required to provide adequate NPSH for sulfate membrane feed pumps due to low NPSH available at deaerator outlet (for vacuum deaerator)	Can avoid NPSH problems for sulfate membrane feed pumps if deaerator located downstream

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