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Techno-economic Evaluation of Conventional and Membrane-based Pretreatment before Reverse Osmosis

Shamma Ahmed Rashid Al-Malek

United Arab Emirates University
Deanship of Graduate Studies
M.Sc. Program in Water Resources

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By

Shamma Ahmed Rashid Al-Malek
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Author's Name: **Shamma Ahmed Al-Malek**

Main Supervisor: **Dr. Mohamed Abdulkarim**

Department of Chemical & Petroleum Engineering

College of Engineering, UAE University

Al Ain, UAE

Co-supervisor: **Dr. S.A. Agashichev**

Water and Electricity Research Center

Abu Dhabi Water and Electricity Authority

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Examination Committee

Dr. Mohamed Abdulkarim, Chair
Chemical & Petroleum Engineering Department
College of Engineering, UAE University
Al Ain, UAE

Dr. Sergei Agashichev
Senior Researcher
Research Center
Abu Dhabi Water & Electricity Authority
Abu Dhabi, UAE

Dr. Corrado Sommariva, External Examiner
President of European Desalination Society
Mott MacDonald Limited
Head of Desalination Team
Contract Professor at L' Aquila University

Dr. Samir Abu-Eishah, Internal Examiner
Associate Professor
Chemical & Petroleum Engineering Department
UAE University

United Arab Emirates University
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To my Father and Mother,

Brothers, Sisters and Friends

With Love ...

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ABSTRACT

In this thesis, a technological and economical comparison was conducted on both conventional and membrane based pre-treatment before reveres osmosis desalination. In order to produce high quality water at the design recovery ratio, reverse osmosis membranes should receive high quality feed water in terms of turbidity, suspended solids, and biological matters. Conventional pre-treatment schemes of media and cartridge filters after coagulation chambers have been used for years. Due to the poor performance and decline in reveres osmosis recovery, membrane based pre-treatment has emerged and has been under research and study in the recent years. Few conventional pre-treatment reverse osmosis plants have been replaced by a membrane based pre-treatment.

The following study is an attempt into that direction to state and verify the ability of membrane based pre-treatment scheme to produce the high quality feed water to the reveres osmosis membrane.

The methodology used for analysis of technological schemes includes three groups of technological and economic indicators, these are: (A) water quality data; (B) technological characteristics of equipment, and (C) economic characteristics of the processes.

The study is based on set of experimental projections of water quality data after pretreatment such as turbidity, the SDI index and total suspended solid. These data were received from several pilot system and full- scale plants. These plants are: (1) conventional pretreatment of the RO pilot plant installed by Ondeo Ltd. located at Al –Taweellah site, (2) conventional pretreatment installed at Al-Fujairah hybrid desalination plant, (3) MF pretreatment based on the "Zenon" system located on Al–Taweellah site, (4) UF pretreatment based on the "Aquasource" system located at Al–Taweellah site, (5) hybrid type of pretreatment proposed by GrahamTech Pte Ltd (Singapore) located in Bainouna power station, and (6) pretreatment scheme of RO desalination plant in Addur (Bahrain).

The average SDI_{15} index of filtrate over three months of study provided by the "Aquasource-UF and "Zenon-MF" systems were 1.5 and 2.9 respectively. Required level for reverse osmosis should not exceed an SDI_{15} of 3.0. Hybrid type pretreatment proposed by GrahamTech Pte Ltd demonstrated satisfactory level of performance characteristics. The SDI_{15} index observed to be less than 1.0 and daily degree of deterioration of normalized permeability was 0.27%. Comprehensive analysis for these case studies is explained in chapter 5.

Economic assessment of pretreatment schemes in design of the sea water RO was conducted on a demo plant. Design capacity of the demo plant is 250 m^3(of permeate)/day. The design of the demo plant was prepared by Japan Cooperation Center for the Middle East. The project is part of the "Advanced Hybrid Desalination System in Abu Dhabi".

The following economic forecasts were done in this study: total cost of filtrate (including investment and O&M of pretreatment) is estimated to be 0.48 and 0.57 \$/m^3(filtrate) for conventional and membrane pretreatment respectively. Cost of installed equipment is estimated to be 520.05 \$/m^3/d and 613.6 \$/m^3/d for conventional and membrane pretreatment respectively.

This study concludes that membrane-based pretreatment is a competitive technological alternative to conventional one. Membrane-based pretreatment can successfully coexist with conventional pretreatment rather than the process that should replace it. Some hybrid configuration of pretreatment schemes including conventional processes along with MF and UF can be recommended for further research.

Key words: techno-economic analysis; pretreatment; ultrafiltration, microfiltration

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Abbreviations

CF: Cartridge Filter

CP: Concentration Polarization

DFC: Direct Fixed Capital Cost

DMF: Dual Media Filter

ED: Electrodialysis

ICC: Indirect Capital Cost

IMS: Integrated Membrane System

MED: Multi Effect Distillation

MF: Microfiltration

MSF: Multi Stage Flash

MIGD: Million Imperial Gallon per Day

MVC: Mechanical Vapor Compression

NF: Nanofiltration

NOC: Natural Organic Compounds

NTU: Nephelometric Turbidity Unit

O & M: Operation and Maintenance

PC: Purchase Cost

PR: Performance Ratio

RO: Reverse Osmosis

SD: Standard Deviation

SDI: Silt Density Index

SEE: Single Effect Evaporation

SMF: Single Media Filter

SWRO: Seawater Reverse Osmosis

TBT: Top Brine Temperature

TDS: Total Dissolved Solids

TMPD: Transmembrane Pressure Difference

TPHC: Total Petroleum Hydrocarbons Components

TSS: Total Suspended Solids

TVC: Thermal Vapour Compression

UF: Ultrafiltration

Nomenclature

$\overline{X_i}$: Value estimated by linear approximation

A_t : Membrane permeability at operating temperature

$A_{t=25}$: Normalized membrane permeability at reference temperature

d: grain size diameter, m

K_1 : Cost of reference case

K_2 : Cost of required equipment

n: Number of readings.

n: Scale factor

P_{brine}: Pressure at brine side, bar

P_{feed}: Pressure at feed side, bar

Pi: The average permeability in the very beginning of the test period, m^3/m^2.bar.s

Pn: The average permeability at the end of the test period, m^3/m^2.bar.s

P_{perm}: Pressure at permeate side, bar

Q_{feed} : the feed flowrate of the liquid, m^3/h

Q_{filt} : the filtrate flowrate of the liquid, m^3/h

Q: Volumetric flowrate, m^3/h

Q_1 : Capacity of reference case

Q_2: Capacity of required equipment

Re: Renold number

t_1: The time required to filter 500 ml measured at the start of the test, minutes

t_2: The filtration time required for the same volume of the sample at the end of test, minutes

u: linear velocity of the feed, m/s

Xi: Experimental readings

Xin: The turbidity of the inlet stream, NTU

X_{OUT}: The turbidity of the outlet stream, NTU

Y_{IN}: The TSS for the inlet stream, mg/l

Y_{OUT} : The TSS for the outlet stream, mg/l

Z_{IN}: The concentration of iron in raw water, µg/l

Z_{OUT}: The concentration of iron in downstream after the Dual media filter, µg/l

ε: The energy, kW

ϕ : Particle shape factor, dimensionless

ε : Porosity of the layer, dimensionless

f : The friction factor

ρ : Density of the feed, kg/m^3

ΔP: Pressure drop, bar

μ_t: Viscosity at operating temperature, kg/m. s

μ_t: Viscosity at reference temperature, kg/m. s

τ: The duration of the SDI test, minutes

1 Introduction

Growth of water demand and depletion of recourses dictate necessity of development of new generation of desalination technologies. Recently there was a growth of scientific, engineering, and commercial interest to processes of membrane- based desalination. Being low resource consuming and ecologically friendly, membrane-based desalination is getting an attractive technological alternative to conventional desalination. Analysis and critical evaluation of membrane-based processes was carried out by different groups of experts [1-4]. These processes were characterized by the following advantages: No energy-consuming phase changes or potentially expensive reagents are needed; low capital and operating costs, less cumbersome maintenance (due to modular nature of process), short construction period and environmental benefits. These advantages were discussed by many authors [5-8]. According to data published by Wangnik [1-2], the cost of desalinated water has dropped considerably, but the cost of water produced by so called "conventional" treatment plants has risen, due to the over-exploitation of aquifers, intrusion of saline water, and also to increasing contamination of ground water. Analysis done by Wangnik [1-2] showed that the decline of MSF began in 1981; today this process plays a significant role only for very large capacities and for so-called dual-purpose plants. Market analysts have suggested that the RO market for modules and equipment, being estimated at US $ 914 million (in 1999) will grow by 8% a year [9]. The desalination market for the 2005-2015 period generates expenditure about $95 billion; of which around $48 billion will be derived from new capacities [10].

The world desalination market will be doubled to more then $70 billion during the next twenty years [10].

Reverse Osmosis, as a sustainable technological alternative for the UAE reconfiguration of technological policy and implementation of environmentally sound technologies, is getting an unavoidable trend of modern development. Recent documents and guidelines namely the Kyoto Protocol, Treaties of Maastricht and Amsterdam [11-19] put the foundation of comprehensive and environmentally sound polices of development. Recent inauguration of the Abu Dhabi Declaration [20] instituted sustainable technological policy of regional development that dictates some shift towards technological systems characterized by low level of CO_2 emission and decreased specific resource consumption. In this regard, the reverse osmosis (RO) technology can be considered as the one of potentially promising technological option. Some authors state that the RO can successfully coexist with multistage flash (MSF) rather than a process that should replace it. New generation of co-generative technologies including power generation along with MSF and RO desalination is becoming an attractive alternative from the standpoint of resource consumption and emissions. The UAE is expected to invest US$ 46 billion over the next decades in cogeneration projects for desalination [21]. In recent years the new generation of dual purpose technologies, namely triple hybrid including power generation, MSF and RO desalination is becoming an attractive alternative to conventional ones [22]. For example, the power-desalination complex in Fujairah has a capacity of 620 MW and 100 MIGD, where 62.5 MIGD (284,000 m^3/day) by multi-stage flash distillation (MSF) and 37.5 MIGD (170,000 m^3/day) by reverse osmosis (RO). Unlike conventional cogeneration processes, the triple hybrid includes RO process along with thermal desalination and power generation.

Published data [1-9, 23] indicates the growth of engineering and commercial interest to RO desalination. In particular, the research projects done within the framework of international research programs and carried out by regional research centers such as Middle East Desalination Research Center (MEDRC), Kuwait Institute for Scientific Research (KISR), etc. [24-26] confirms the growth of interest to membrane-based desalination. For example, 80% of the research projects carried out under the auspices of the MEDRC are focused on issues related to membrane-based desalination [25-26]. The USA and countries of the EU have put membrane research in the list of their priorities towards advanced technological programs [27].

Analysis of published data and regional experience confirms that pretreatment is the key issue in the way of practical implementation of reverse osmosis. Pretreatment focuses on prevention of membrane degradation and fouling. The main tasks of pretreatment are: (1) extension of membrane lifetime, (2) prevention of membrane fouling, (3) maintaining performance level. Different technological schemes can be used for pretreatment purposes. Majority of existing RO desalination plants are equipped with similar types of conventional pretreatment based on coagulation, flocculation and multimedia filtration. Since recently one can see the growth of new generation of pretreatment, namely, membrane-based pretreatment where micro- and ultra-filtrations are used instead of coagulation and multimedia filtration. Analysis of published data proved an increase of number of research projects related to different aspects of pretreatment before reverse osmosis [28-29]. Data published by European research teams regarding surface water treatment shows that membrane-based schemes have become commercially competitive and replaced the conventional processes on a vast scale. Some authors [30] outlined that the membrane pretreatment before RO is becoming a technologically competitive trend. Within the context of the problem, this study focuses on analysis of techno-economic aspects of conventional and membrane- based pretreatment before RO desalination.

1.1 Water resources in UAE

The quantity of water into supply has increased, despite that increase in water resources remain scarce in many locations of the United Arab Emirates. The scarcity of water may be caused by the following [31]:

- Rapid increase in population
- High per capita consumption
- Construction of villas and residential complexes
- Development of farms and forests

The United Arab Emirates rely on non-conventional water resources in addition to conventional resources, to meet the ever-increasing demands for water. The conventional water resources include seasonal floods, springs, falajes and groundwater. The non conventional resources are waste water treatment plants and desalination plants.

Because these resources do not meet the national demand for water, the non conventional resources are intensively used nowadays, especially desalinated water. Table 1-1 shows the water production desalinated water and ground water from 1997 to 2002 [32]

Table 1-1 : Water production growth of desalinated water compared to ground water from 1997 to 2002 in UAE (MIG/year), [32]

Source	1997	1998	1999	2000	2001	2002
Desalination	101168	113691	123723	134971	152804	177963
Groundwater	22362	23620	30566	30375	21715	24989

From the table it's obvious that water produced from the desalination plant increases annually, and the dependence on the groundwater source decreases due to the depletion of this source. This confirms that desalination is the central source of water in UAE now and in future, especially when the demand and consumption per capita increase sharply. The consumption per capita in 1997 was 126 gallon/day, this value increased to 157 gallon/day in 2002. The demand category is driven by the growth in population and per capita consumption; it can be classified to two categories; domestic demand and bulk demand. The formation of the domestic water demand is mainly residential, commercial establishments, hospitals, hotels, offices, and shops. The formation of bulk water demand is agriculture, landscaping, large industrial usage, palaces, airports, and other nondomestic bulk diversions. This demand category is driven by the increase in the number of farms and landscaping, and the development of industrial projects. The agriculture development to achieve self sufficiency in food supplies has higher share in water consumption; (about 80%); the estimated water demand, excluding water losses, for each farm of a normal size of 180 m X 150 m is 20,000 gal/day.

Water demand in the UAE is expected to double in the next ten years. From around 630 million gallon per day in 2000, it is expected to climb to 973 millions in 2005 and 1.24 billion in 2010 [33]. The second source of non-conventional sources is waste water treatment plants. The treated waste water in the UAE is used in the irrigation of public parks and beautifying streets and roundabouts of the major cities. It is worth mentioning that the sewage water after primary, secondary, and tertiary treatment can be used in agriculture irrigation especially if it is purged of toxic materials and bacteria. The four sewage treatment plants are located in Abu Dhabi, Al-Ain, Dubai, and Sharjah. Table 1-2 shows the capacity of each plant [34].

Table 1-2: UAE Waste water treatment plants capacities

City	Capacity (m^3/day)
Abu Dhabi	383,000
Al-Ain	85,000
Dubai	320,000
Sharjah	147,000

1.2 Desalination status in UAE

The total installed capacity of seawater desalination plants in GCC countries is estimated at about 7.92 m^3/day, which represents more than 58% of the total world capacity [35]. The share of each of the six GCC countries in the world capacity is shown in Figure 1-1. UAE is the second country after Saudi Arabia in water desalination industry, where 86% of the water supply comes from desalination of seawater or brackish water [32].

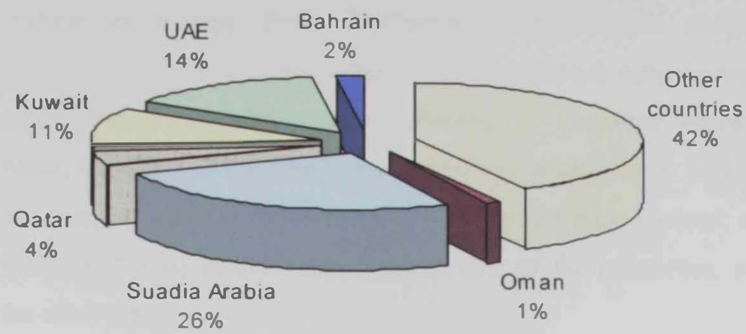

Figure 1-1: Percentage share of seawater desalination capacity of GCC countries [35]

UAE began desalinated water production in 1973 in Abu Dhabi at an annual production rate of 7 million m^3. In 2002 the production rate reached 479.82 million m^3, and since 1974 over sixty desalination plants have been commissioned in the UAE [36].

Today in UAE, the evaporation technique is dominant in desalination field, where 96% of desalinated water is produced by MSF and MED. The 4% remaining is produced by Reverse osmosis (Appendix A). Despite the obstacles that force this technique, especially in the Arabian Gulf, like water pretreatment and cleaning procedures required to control membrane fouling, which are the main important limitations of the technology to be generally applied [37], there is a rapid development in membrane materials, energy recovery systems and installation costs. These developments can play significant roles in encouragement of commercializing this technology. In 2003 the largest seawater reverse osmosis (SWRO) plant was commissioned in Al-Fujairah plant with capacity of 37.5 MIGD. In addition to Al-Fujairah plant, Abu Dhabi water and electricity Authority is in the stage of financing SWRO plant with larger capacity (50 MIGD) in Al-Taweelah complex.

1.3 Desalination technologies

Several different methods are available to desalinate seawater, of the three commercially proven processes: distillation, reverse osmosis and electrodialysis. Generally, distillation and reverse osmosis are used for seawater desalination, while reverse osmosis and electrodialysis are used for brackish water desalination. However, the selection and use of these processes depends mainly on site specifications and cost. The two most common methods used today are thermal desalination and membrane desalination. Thermal desalination uses a very simple and natural process to separate out solids: salt water is heated to produce water vapor that is in turn condensed to form fresh water. Some of the more specific desalination technologies that depend on heat to produce water vapor include multi-stage flash distillation, multiple-effect distillation and vapor compression. As shown in Figure 1-2, about half of the desalinated water is produced using some form of thermal distillation [2]. Membrane technology is the other major methods used to desalinate salt water, the share of the membrane technology worldwide is close to the MSF share. Like thermal technology, membrane desalination is based on a simple concept: salt water is forced across a membrane, producing potable water on one side of the membrane, and leaving behind briny water on the other side.

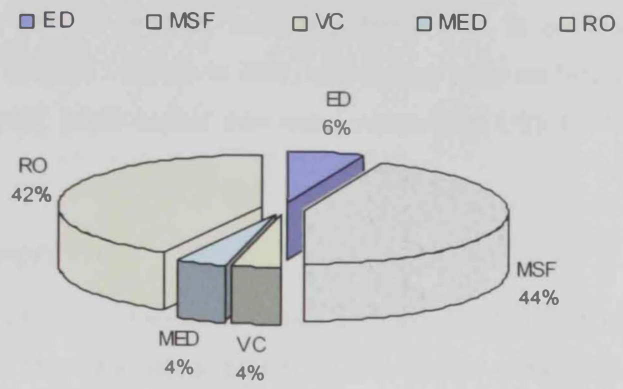

Figure 1-2: Worldwide percentage share of seawater desalination technologies [2]

1.3.1 Multi-Stage Flash Distillation

Multi-Stage Flash Distillation (MSF) distillation is currently the most common and simple technique in use, it has been operated commercially for more than 30 years [38]. MSF distiller can be designed for a range of performance ratio (the ratio between water production and energy consumption), with a practical limit of about 11:1. Capital cost increases with performance ratio due to the larger heat transfer surface areas needed and greater number of stages. The optimum value is usually in the range 7 to 9, depending on energy cost. A typical plant of an 8:1 performance ratio would have 16 to 18 heat recovery stages and three heat reject stages [5]. The main advantage of this technique is the ability to handle high production rate, Al-Shuwaihat power and water station in Abu Dhabi is designed to produce 16.7 MIGD for each distiller, where this rate was 12.5 MIGD in Al-Arabia Station. The capital costs of the MSF plants today vary from US$4.0 to US$7.0 per imperial gallon per day of installed capacity [39].

1.3.2 Multi-Effect Distillation

Multi-Effect Distillation (MED) is one of the most promising evaporation techniques existing today, it is predicted that the second generation in water desalination plants will be settled in MED as well as RO [40-41] where the energy consumption is less compared to MSF, (2 kWh/m^3 compared with 4 kWh/m^3 for MSF) [42]. Basically, this method can use low-temperature, low-pressure steam as the main energy source. Usually, 8 to 16 stages are common in such operations. This allows a good performance ratio which can go up to 15 [38]. The efficiency of the process is bound by high values of boiling point elevation at high concentrations. Unlike the MSF technique where water is produced mainly by turning sensible heat into latent heat of evaporation, the MED technique uses latent heat to produce secondary latent heat in each section. Layyah plant inaugurated two MED units of 5 MIGD in 2001, where these units are believed to be the largest of its type in the world [43]. MED capital cost today varies from US$ 4.5 to US$6.0 per imperial gallon per day of [39].

1.3.3 Vapor compression

Another distillation technology known as vapor compression (VC) is used for smaller scale desalination facilities. This process is based on the Carnot refrigeration cycle, in which a mechanical compressor rather than a heat source is used to compress the vapor from the evaporator to a higher pressure. As the compressed vapor condenses on one side of the tube heat transfer surface, seawater boils on the other side creating more vapor. This process uses electric energy rather than steam. The VC evaporator is more efficient than the previously described steam driven evaporators, but electric power is significantly more expensive than steam energy. VC units are commonly used for some small industries since they are more compact than other thermal processes and electric power is readily available. The number of VC units currently in operation is very small (4% worldwide) as compared to multi-stage flash systems, which are estimated at 44% worldwide [2].

The most important advancements in thermal desalination over the past 10 years have been increasing system efficiency and operational reliability. The operational enhancements have included scale control improvements, automation and controls, further operator training and better materials of construction. Additionally, increases in standard-unit sizes have increased the economies of scale for larger systems. However, these systems have very high-energy requirements and can be cost prohibitive unless low cost steam energy is available from a power plant.

The capacity of the thermal desalination processes varies over a wide range, from 500 m^3/d to 55,000 m^3/d. The average conventional sizes are 33,000 m^3/d for MSF, 12,000 m^3/d for MED and 3,000 m^3/d for VC [44].

1.3.4 Reverse osmosis

The RO membrane technique is considered the most promising for brackish and seawater desalination [45]. The RO uses dynamic pressure to overcome the osmotic pressure of the salt solution, hence causing water-selective permeation from the saline side of a membrane to the freshwater side. The RO process which employs membranes, has a simple layout, and is compact and modular. Existing units can be expanded to handle larger capacities. However, RO membranes are more sensitive to the conditions of the feed seawater, scaling, fouling and pH than thermal processes. Furthermore, unlike thermal processes, RO membranes do not provide high purity water. On the average, the permeate salinity varies over a range of 30–150 ppm [44]. The actual value depends on the process recovery, which is defined as the amount of product per unit mass of feedwater. Today the capital cost of the RO plant could vary from US$3.5-5.0 per GPD [39].

Darwish *et al* [46] summarized the main advantages of the RO system over the MSF system as follows:

- It consumes less energy, [mechanical energy delivered by motor(s)]
- It does not need to be combined to a power plant or to interfere with its operation. In fact, it can be operated only during non-peak power demand period.
- It has simple start/stop operation.
- It is delivered in modules, no need to shut off the whole plant for emergency or routine maintenance.

The fast growth in membrane technology shows that most significant improvements occurred in the following areas: (RO membrane, energy recovery system, and pretreatment scheme)

1.3.4.1 RO membrane

The RO membranes used are semi-permeable polymeric thin layers, adhering to a thick support layer. Membranes are usually made of cellulose acetates, polyamides, polyimides, and polysulfones. They differ as symmetric, asymmetric, and thin film composite membranes. Membranes are sensitive to changes in pH, small concentrations of oxidized substances like chlorine and chlorine oxides, a wide range of organic materials, and the presence of algae and bacteria. Therefore, careful pretreatment is needed in order to prevent membrane contamination and fouling which can occur as a result of suspended solids, bicarbonate ions, carbon dioxide; dissolved organic materials and chlorine compounds. Different antiscalants are used in order to prevent precipitation of dissolved salts due to increased concentration.

A number of module designs are possible and all are based on two types of membrane configuration: flat and tubular. Plate and frame and spiral-wound modules involve flat membranes whereas tubular, capillary and hollow fiber modules are based on tubular membrane configurations. The application of reverse osmosis in seawater desalination is a story of process improvement, whereas the first reverse osmosis units needed pressures of 120 bar, the turning to thin film composite membranes allowed systems to operate at significantly lower pressure-down to 60 bar [47]. In addition to that, membranes are more efficient now and can operate at higher temperatures, and have higher salt rejection. Membranes now have higher flux rates (flow rate per unit area), lower fouling potential, lower costs and longer lives than ever before. This trend may continue with further membrane system advancements. In order to allow the best ratio of the membrane area to operation volumes, two most convenient designs are made to fit the pressure vessels: the spiral-wound and the hollow-fibers membranes.

1.3.4.2 Energy Recovery System

One of the important performance measures of any continuously operated seawater RO system, the specific energy consumption. In seawater RO plants, operating pressures may reach or even exceed 70 bar and product water recovery is in the range of 30 to 35%, depending on feed water temperature and salinity, leaving the brine waste stream with a very significant amount of hydraulic energy which may reach 40% of the original energy supplied to the high pressure pumps of the seawater RO system. Energy recovery devices are used to recover work from the concentrate stream. The addition of energy recovery equipment typically reduces net power consumption by 25 to 40% [48] .Commercially available energy recovery devices are the Pelton wheel turbine, reverse running centrifugal pump and the hydraulic turbocharger [49]. In addition to theses, the pressure exchanger, this has recovery of up to 60% of pumping energy [50]. The efficiencies used for pumps and turbine are conservative values. Pump efficiency is in the range of 80–86 %, and turbine is in the range 84--88%, usually Pelton wheel turbines [46].

1.3.4.3 Pretreatment Scheme

The key component in success of RO desalination is pretreatment of water prior to RO stage. To prevent RO membrane fouling, all the organic, colloidal and biological matter needs to be removed from the feedwater. This can be achieved by various unit operations. Usually pretreatment process involves disinfection, coagulation, flocculation, filtration and adjustment of the solubility parameters to avoid precipitation of sparingly soluble salts in the membrane [51].Membrane-based pretreatment is one of the innovative approaches in treatment area as an alternative to conventional pretreatment techniques. More recently, a double membrane barrier concept has emerged as a possible alternative. This is accomplished by installing Microfiltration (MF)/Ultra filtration (UF) membrane upstream of the RO membrane in an Integrated Membrane System (IMS) to filter off bigger suspended solids, turbidity, bacteria, colloids, parasites and viruses for clarification and disinfection purposes [52]. This technology has now been optimized and is becoming competitive as compared to conventional processes for larger scale plant capacities. Projects using UF membrane with capacities greater than 100,000 m^3/d (21 MIGD) are being implemented [53]. This study will emphasize on this scheme due the rapid development of this new technology in both operational and economical aspects like capital and O&M costs.

2 Literature Review

2.1 Tasks of feed water pretreatment and performance decline factors

The main tasks of pretreatment are: (A) extending the lifetime of membrane; (B) preventing fouling of membrane; (C) maintaining performance characteristics such as rejection and recovery at the required level. According to data published by ESCWA [54], capital and operating cost associated with pretreatment subsystem accounts up to 60% of the total production cost. According to the data submitted by Durham and Wilton [55], conventional pretreatment system at Doha RO plant in Kuwait represents 23% of the total product water cost (4546 m^3/day). The type of pretreatment that must be designed specifically for individual application depend on several factors, namely: (1) designed configuration of the RO system and, (2) type of feed water. According to Tasaka *et al.* [56], the pretreatment methods can be subdivided into two broad categories: (A) pretreatment to prevent irreversible membrane degradation and (B) pretreatment to prevent reversible decline of membrane performance (or membrane fouling). Main performance decline factors are given in Table 2-1.

Table 2-1: Performance decline factors (degradation and fouling factors) [56]

	A. Membrane degradation (irreversible decline of membrane performance)		B. Membrane fouling (reversible decline of membrane performance)
A1	Physical deformation (compaction, drying)	B1	Deposition
A2	chemical degradation (hydrolysis, oxidation)	B2	Clogging of pores
A3	Biological degradation	B3	Cake (colloids)
		B4	Gel (organics)
		B5	Scale (insoluble inorganic salts)
		B6	Surface Sorption (specific organics such as surfactants)

Sheikholeslami [57] outlined the following approaches to fouling combating, namely: (1) fouling control, (2) pretreatment technologies, and (3) anti-fouling membrane modules. The first group (fouling control) comprises modifying technological parameters and operating conditions namely: (A) critical flux, (B) critical conversion which is a function of critical flux and (C) fouling control chemicals (they can act as scale inhibitors, scale crystal modifier or sequestrates (for Fe, Mg, etc.)

The next group, according to Sheikholeslami [57] is focused on innovative pretreatment technologies namely UF and MF based-pretreatment before RO. The third group covers Van der *et al.* [58] new generation of low-fouling composite membranes being more hydrophilic with reduced affinity of the surface to organics, In particular, Van der *et al.*[58] proposed to exclude using of antiscaling agents and to apply hydrochloric acid instead of sulfuric acid.

2.1.1 Pretreatment to Prevent Degradation of Membranes

Deterioration of membrane in itself is referred to as "Degradation"; it results in irreversible performance decline that, in turn, can be caused by chemical, physical, and biological factors. The concentration of chlorine and pH are the main two factors that can result in chemical damage of membranes. In particular, polyamide membranes are damaged even by low concentration of chlorine, thus the feed water must be dechlorinated before it enters the membrane system. Dechlorination can be done using (1) sodium bisulfite ($NaHSO_3$), (2) carbon filtration, and (3) treatment with gaseous sulfur dioxide (SO_2) [59].

Unlike polyamide membranes, cellulose acetate membranes are more vulnerable to hydrolysis. These membranes undergo rapid hydrolysis below pH=4 and pH>7, thus control of pH is particularly important for them, while dechlorination is vital for most polyamide membranes. "Degradation" can be a result of chemical oxidation catalyzed by iron compounds [56]. The same phenomenon caused by some transition metals is also reported for cellulose acetate membranes [56].

2.1.2 Main Fouling Factors and Methods of Elimination

Fouling is defined as the build up of deposits on the membrane surface that leads to performance decline [56]. Fouling can be considered as reversible change of membrane performance. It can be caused by different factors. According to Williams *et al.* [59] fouling factors can be divided into six categories: (1) suspended solids; (2) colloids; (3) scale forming salts; (4) metal oxides; (5) biological foulants and (6) organic foulants.

(1) Suspended solids: Coarse screening and hydrocyclones are used to remove large particles.

(2) Colloids are usually charged particles smaller than 1 μm in diameter. They are common in feed water and drastically reduce the productivity of the membrane. Several techniques can be used to remove colloids, the most common of them is coagulation-flocculation followed by conventional filtration. The typical coagulants used are alum $Al_2(SO_4)_3$, ferric chloride $FeCl_3$, and polymer or polyelectrolyte materials [59].

(3) Scale forming salts: Most common salt compounds are calcium carbonate, calcium fluoride, calcium sulphate, salts of barium, strontium, and silica [59]. To minimize or to eliminate the formation of scale deposits the following methods can be used: (1) Acidification by acid injections. Injected acid converts bicarbonate alkalinity, thus eliminating the formation of $CaCO_3$ scale. (2) Water softening using lime or lime soda. In this process, hydrated lime or soda ash is added to soften the water. Calcium and magnesium hydroxides are then removed as precipitates. This process can also remove some of the silica. (3) The addition of antiscaling agents or so called "threshold" agents. These compounds reduce the rate at which scale forms, allowing the system to operate with concentrations above the solubility limit. One of the most common "threshold" agents used to control calcium sulfate formation is sodium hexametaphosphate (SHMP).

(4) Metal oxide: deposits can be cleaned from membrane surface using acids.

(5) Biological foulants: To prevent bio-slime formation, the feed water is desinfected before it enters the RO system. Chlorination to 0.5 ppm by injection of chlorine gas or addition of hypochlorite is the most common method used. However, as discussed above, many RO membranes is damaged by chlorine. Therefore, the feed must then be dechlorinated, usually with sodium bisulfite, before it enters the system. Other disinfectants that can be used include ozone, ultraviolet light, formaldehyde, concentrated sodium bisulfite and copper sulphate [59].

(6) Organic foulants: (or fouling by natural organic matter (NOM)) such as humic acid fouling. Growth of organic film on the membrane surface can be caused by humic acids and their derivatives. In seawater it has been observed that this is the most widespread organic foulant present in coast water [60]. Humic substances are products of the incomplete chemical and biological degradation of plants and animal residues. They are complex, heterogeneous refractory organic compounds with reported molar masses ranging from several hundred to tens or hundred of thousands grams per mole. They include lignin, carbohydrates and protein. (Polyphenolic aromatic complexes such as humic acids, lignin and tannin are decay products of wood tissues of plants. They can occur in the form of a colloid as well). According to Karabelas and Yiantsios [61], natural organic compounds (NOC) are divided into humic substances (or polyhydroxyaromatics), and non-humic such as proteins, polysaccharides and aminosugars [61]. To prevent organic fouling the following methods can be used: coagulation and filtration, carbon adsorption and chemical oxidation [59].

2.2 Pretreatment before RO based on conventional schemes

Different unit operations can be used on the stage of pretreatment. The majority of the conventional schemes are based on coagulation and multimedia filtration. They include the following main stages: disinfection; coagulation; and filtration. Majority of existing RO desalination plants are equipped with similar types of conventional pretreatment. Some techno-economic aspects of pretreatment are considered by Ben Hamida [62] and Vrouwenvel *et al.* [63]. The scheme is shown in Figure 2-1 is a simplified flow- diagram of the conventional pretreatment.

Chlorinated water from the intake (1) passes through coagulation chamber (2), where ferric chloride ($FeCl_3$) is used as a coagulant. After coagulation chamber the seawater passes through the multimedia filters. (3) Then the filtered water passes through the cartridge filters (4) with further treatment by antiscalant and sodium bisulfite (SBS).

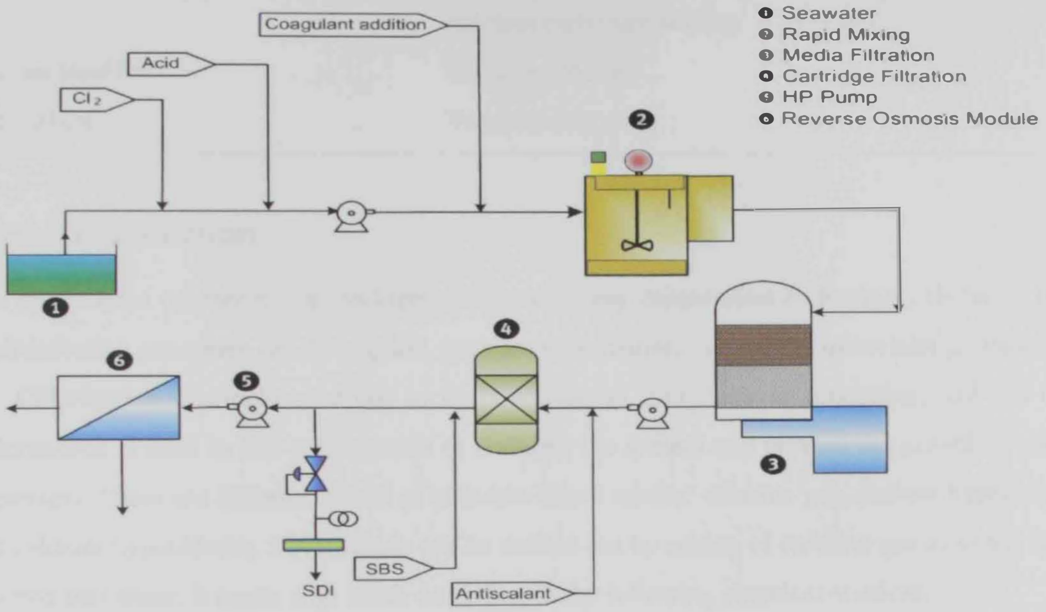

Figure 2-1: Conceptual flow diagram of conventional pretreatment before RO

Abubasher *et al.* [64] considered different unit operations that are used in the Middle East for SWRO feed water treatment. The study included results of pretreatment in terms of reduction in colony forming unit (CFU) per milliliter. The overall results of the study indicated that coagulation and dual media filtration (DMF) reduced bacteria concentration in the feedwater by 32-100%. In most cases coagulation and filtration effectively removed a large portion of total bacterial mass (82%) in the feed. Similar results were obtained by Al-Tisan *et al* [65] where the data on bacterial removal from feedwater were presented. The coagulation and media filtration were found to be efficient in removing of bacterial biomass (about 82%) from the feedwater.

The chemical dosing depends on the water characteristics and operation conditions of the plant. Table 2-2 shows the chemicals used in both conventional and membrane-based pretreatment. In addition, the membrane pretreatment system requires chemical cleaning by citric acid or sodium hypochlorite periodically to restore the flux.

Table 2-2: Chemicals used in pretreatment [64]

Chemicals	Purpose
Chlorine gas or sodium hypochlorite	Disinfecting
Ferric Chloride	Remove suspended matter and colloids
Cationic coagulant	Improve the coagulation process
Sulphuric Acid	To reduce the bicarbonate and avoid calcium carbonate scaling
Sodium bisulfite	Remove chlorine
Antiscalant	To prevent scaling

2.2.1 Disinfection

The disinfection operation is an indispensable part of any desalination technology. Different types of disinfection processes can be applied, such as: chlorination, ozonation, ultraviolet pretreatment, etc. Chlorination is widely used due to its effectiveness, simplicity in generation, and low cost. Chlorination is used in RO pretreatment to disinfect the system and prevent the growth of micro-organisms. There are different forms of chlorine–based agents: chlorine gas, sodium hypochlorite, and calcium hypochlorite. Chlorination can be carried out by adding of chlorine gas to water. Being injected into water, it reacts with water according to the following chemical reaction:

$$Cl_2 + H_2O \rightarrow HClO + HCl \qquad [2\text{-}1]$$

Hypochlorous acid and hydrochloric acid dissociate as follows:

$$HClO \rightarrow H^+ + ClO^- \qquad [2\text{-}2]$$

$$HCl \rightarrow H^+ + Cl^- \qquad [2\text{-}3]$$

The effectiveness of chlorination depends on pH, temperature, organic content of the water, contact time and concentration [62]. Chlorine can react with particles as well. The ratio of [HClO] to [ClO^-] is a function of pH. At low pH levels, the form [HClO] is becoming dominant, and as pH increases, the [ClO^-] anion goes up. Chlorine is more effective when the pH value ranges from 4.0 to 7.5. In spite of the effectiveness of disinfection at higher temperatures; the residual chlorine is quickly extinct at higher temperature. If the chlorine is added to water where there are different matter such as iron, manganese, nitrites, ammonia, and organic matters, it will first react with iron, manganese, and nitrites. As more chlorine is added, it will react with ammonia and organic substances to form a set of chloramines as shown by the following chemical reactions:

$$NH_3 + HClO \rightarrow NH_2Cl + H_2O \qquad [2\text{-}4]$$

[mono chloramines]

$$NH_2Cl + HClO \rightarrow NHCl_2 + H_2O \qquad [2\text{-}5]$$

[Dichlor amine]

$$NHCl_2 + HClO \rightarrow NCl_3\ H_2O \qquad [2\text{-}6]$$

[Trichlor amine]

Once the amount of chloramines reaches a minimum value at a point (called breakpoint chlorination) beyond which, the addition of chlorine will produce free residual chlorine required for effective disinfection.

2.2.2 Coagulation and coagulants

The coagulation is an indispensable stage of the scheme of conventional pretreatment, (while direct filtration proved to be ineffective in removing impurities such as bacteria, virus, soil particles and color) [66]. Coagulation is the process where the addition of a chemical reagent results in a reduction of the forces keeping dispersed particles apart. The process is referred to as coagulation and it can be achieved by destabilizing the electric charge of the particles, which are mostly negatively charged. A coagulant with positively charged ions is added to the water to neutralize the negative charges and hence promote coagulation. The coagulation is followed by flocculation whereby an agglomeration of the dispersed particles takes place. Mackenzie and David [66] outlined the following characteristics required for the coagulant: (1) insolubility within the natural pH range; the added coagulant must precipitate out of solution so that the ion is not left in the water, which will aid the colloid removal process, (2) a charge of a cation (a trivalent cation is the most efficient to neutralize the colloid charge), and (3) nontoxic behavior.

Aluminum and ferric ions are the most commonly used for these purposes. The pH of coagulation is an important characteristic and it has an effect on the efficiency of the coagulation process. Ferric salts work best in a pH range of 4.5-5.5, whereas aluminum salts are mostly effective around a pH range of 5.5-6.3 [67]. These pH values are set by adding an acid for lowering pH like sulphuric acid, or alkalis to raise the pH like lime or soda ash. A time factor is an important in flocculation process, the process must provide adequate time for the particles to come closer and form flocs, the required time for the coagulation process is 1 to 2 minutes, while for flocculation process is 10 to 30 minutes [68]. While selecting the coagulant the following factors have to be taken into consideration: type of the coagulant, concentration of the coagulant, proper mixing, residence time, pH, turbidity, alkalinity and temperature of the water. The optimum of the above factors must be determined from laboratory tests such as jar test in order to ensure a suitable coagulation process.

As mentioned above, ferric and aluminum salts meet the main requirements, that is why ferric chloride ($FeCl_3$) and aluminum sulfate ($Al_2(SO_4)_3$) are the most widely used as coagulants in water treatment. When a coagulant is added to the raw water, it first reacts with the alkalinity existing in water to form jelly-like flock-particles of ferric hydroxide, $Fe(OH)_3$ as per the following chemical reactions.

$$2FeCl_3 + 6HCO_3^- \rightarrow 2\,Fe(OH)_3 + 6Cl^- + 6CO_2 \qquad [2\text{-}7]$$

$$2Fe_2\,(SO_4)_3 + 6HCO_3^- \rightarrow 2\,Fe(OH)_3^- + 3SO_4^{2-} + 6CO_2 \qquad [2\text{-}8]$$

Once a coagulant is added to the raw water, the positively charged ions neutralize the negatively charged particles; which can be assisted by rapid mixing for a good coagulation. The neutralized particles begin to adhere to each other to form small particles called micro-flocks. These micro-flocks have positive charges from the coagulant added; continue to neutralize negatively charged particles until they become neutral particles. Finally, the micro-flock particles begin to agglomerate and stick together to form larger particles. This process is known as flocculation. To enhance coagulation, some additional electrolyte (referred to as coagulant aid) can be injected into water. Coagulant aid promotes create stronger and more settleable flocs and lead to reduction the amount of the required coagulant. Quantity of flocks formed with polyelectrolyte is low, but they are strong and not breakable and doesn't require pH control as in the case of aluminum coagulants. Polyelectrolyte is used in small dosages, (optimum dosage is usually 0.1 to 1.5 mg/L and for inorganic coagulants 2 to 8 mg/L) [62]. The most common coagulant aid used in seawater pretreatment is polyelectrolytes or polymers which can be classified as cationic, anionic, and nonionic polyelectrolytes. The most widely used coagulant aid is the cationic polyelectrolytes, whose chains have amines, imines, or quaternary ammonium groups [67]. When dissolved in water it produces positively charged ions which will neutralize the negatively charged colloids obtainable in water. According to Al Nuwaibit *et al.* [69], ferric ions at very low dosing rate (2 mg/l) coupled with cationic polymer (0.5 mg/l) were found to be the best economic chemicals for conventional treatment.

2.2.3 Filtration

Filtration is a process for the separation of suspended materials, mainly flocks formed in the coagulation/flocculation process and iron and manganese precipitates. The suspended materials are removed while passing through a porous media. The filter media is usually sand or a combination of sand, anthracite, garnet or similar substances manufactured for filtration. While passing through the filter media, the suspended particles adhered within the filter layer. To obtain a filter media of correct specifications, enough residence time should be allowed for proper screening to take place [51]. To avoid interpenetration of filter media (anthracite/sand), apart from having the correct uniformity coefficients and specific gravities, the size ranges should also be compatible. The coarser size fraction of the upper layer anthracite should not exceed more than five times the finer size fraction material of the lower layer sand [51].

The most widely used filters are gravity and pressure filters. They differ in the nature of the driving force and filtration rate. In gravity filters, water passes through the filter due to gravity, where the driving force is a gravity force and the filtration rate is 5-10 $m^3/m^2/h$. In pressure filters, the driving force is the difference in the applied pressure, and the filtration rate is 10-25 $m^3/m^2/h$ [62].

During filtration, the media grains become coated with flocs and the pores become clogged, the rate of clogging depends on the characteristics of the water, the more turbid the water, the faster the filter becomes clogged. The grain size of the filtering media has effect on the clogging rate, the finer the filtering material, the more quickly it becomes clogged. The filter should be cleaned in place by pumping water backward through its layers to restore the filter performance by removing the filtered particles from the bed. This operation called backwashing. After the backwashing operation, the different layers return to their original place, the largest particle settles first resulting in a fine sand layer on top and a course sand layer on the bottom. In the proper backwashing process the grains must be agitated and rubbed against each other to remove the sticky material to prevent mudballs formation which will clog the filter media and reduce the performance of the filtration process. The backwashing frequency process depends on many factors and can vary according to the operating conditions from one plant to another. Usually the filter operation run hours must not be more than 36 hours. According to Ben Hamida [62], the main conditions for starting the operation of backwashing are: (1) Head loss is higher than recommended by filter manufacturer and (2) an increase in nephelometric characteristics such as turbidity and the SDI index.

2.2.4 Energy consumption

According to Galloway *et al.* [70] the conventional and membrane pretreatments are characterized by similar energy consumptions, but dead-end regime requires less energy than the cross-flow one [71]. According to the estimates done by Galloway *et al.* [70] the processes with conventional and membrane-based pretreatment are characterized by the following values of specific energy consumption: 3.57 kWh/m^3 and 3.56 kWh/m^3 (of permeate), respectively.

2.3 Membrane-based pretreatment before RO

2.3.1 General aspects of membrane technology

The major pressure-driven membrane processes are: reverse osmosis (RO), nanofiltration (NF), ultrafiltration (UF) and microfiltration (MF). Classification based on the size of the particles to be separated as proposed by Ronald and Munir [67, 72] is shown in Table 2-3 and Figure 2-2.

Table 2-3 Characteristics of membrane processes [72]

Process	Retentate	Permeate
Microfiltration	Suspended particles, water	Dissolved solutes, water
Ultrafiltration	Large molecules, water	Small molecules, water
Nanofiltration	Small molecules, divalent ions, dissociate acids, water	Monovalent ions, undissociated acids, water

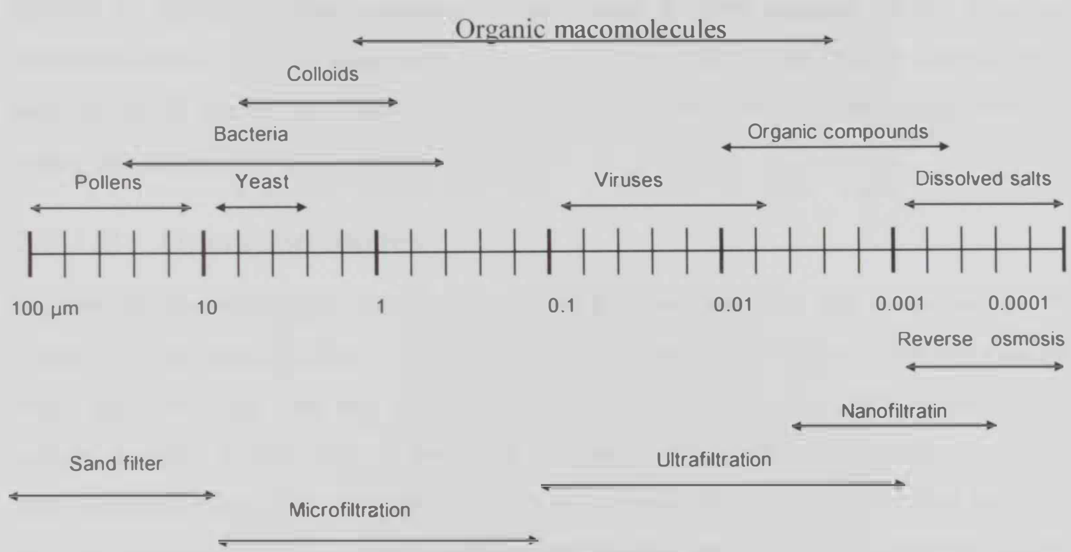

Figure 2-2: Range of separation of pressure driven processes [67]

There is a wide spectrum of configurations of membranes such as spiral wound flat sheet, hollow fiber, tubular and plate-and-frame. The prevalent configurations for pretreatment to RO are hollow fiber and spiral wound.

2.3.1.1 Membrane materials and morphological characteristics of membranes

MF and UF membranes can be developed from polymers, composite and inorganic. The polymeric membrane materials include cellulose acetate, polysulfone, polyamide, polypropylene, and other proprietary formulations. The composite membrane is also known as thin film composite. The major type of inorganic membrane is made from ceramic materials. The membrane material typically has a wide pH tolerance range to accommodate for low and high pH cleaning chemicals. The polysulfone, cellulose, ceramic and some of the proprietary materials have a free chlorine tolerance that allows for periodic or continuous sanitization, but polyamide membranes are more vulnerable to chlorine impact. The operating temperature is considered as an important property for membrane material, the operating temperatures for the polymeric membranes is much lower than these for ceramic membranes.

Membrane materials are characterized by the porous structure of the membrane material such as, the averaged pore size, pore size distributions, and the molecular weight cut-off (MWCO). The MWCO is defined as that molecular weight which is 90% rejected by the membrane. UF membranes have nominal molecular weight cut-off from 100,000 to 750,000 Dalton. The typical pore size of UF membranes ranges from 0.1 to 0.001 μm, while the MF membranes have pore sizes within the range of 0.1 to 10 μm [72].

2.3.1.2 Operating modes

MF and UF membranes are operated in two modes: dead-end flow and cross-flow, as shown in Figure 2-3. The dead-end flow or direct-flow mode of operation is similar to that of a cartridge filter where there are only feed and filtrate flows. The dead-end flow approach typically allows for optimal recovery of feed water in the 95 to 98% range, but is typically limited to feed streams of low suspended solids, less than 10 NTU. A typical cross-flow mode of operation has one part of entry as feed and two outlets one for retenate and another for filtrate. The cross-flow mode is used for feed waters with higher suspended solids; from 10 to 100 NTU. The cross-flow mode of operation typically results in 90 to 95 % recovery of the feed water [73].

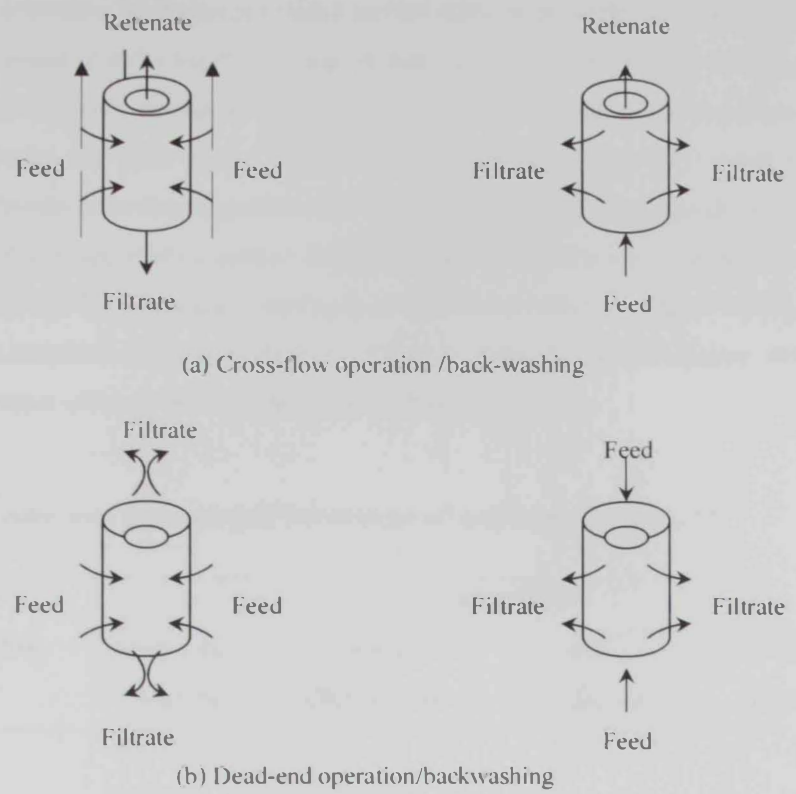

Figure 2-3: Schematic drawing of two basic module operations

2.3.1.3 Membrane regeneration

Gel polarization and pore blocking are the main factors that cause deterioration of performance characteristics. For the membrane performance to recover, there are two main methods of membrane regeneration, namely, hydraulic regeneration and chemical cleaning. The hydraulic methods include alternate pressurizing and depressurizing, back-flushing and changing the flow direction at a given frequency. The principle of backwashing is presented in Figure 2-3. After a certain period of time, the feed pressure is released and the direction of permeate reversed from the permeate side to the feed side in order to remove the fouling layer at the membrane surface. Periodic backwashing is used to minimize the need for chemical cleaning to once every 1 to 6 months. Typical flushing approach is based on short periodic backwashing, where filtrate water flow is reversed and pumped back into the filtrate-side of the MF/UF module and allowed to exit via the feed and concentrate ports. This reversal of normal service flow is designed to remove the foulant off the membrane surface and out of the feed channels. Typical frequency of the backwashing is 30- 60 seconds per every 15 to 60 minutes. To control biological fouling a disinfectant (chlorine, hydrogen peroxide) can be added to the backwash once every 1 or 4 hours. Cleaning is aimed at reducing the fouling problem. Chemical cleaning at an expected frequency of one to two months may be required to restore the flux. The frequency of membrane cleaning can be estimated during the pilot study. Typical cleaning chemicals are citric acid, hydrochloric acid, sodium hydroxide or sodium hypochlorite, with the selection being dependent on the foulants. The effects of various approaches against different types of fouling are summarized in Table 2-4. As indicated in Table 2-4, chemical cleaning is an effective control strategy for all types of membrane fouling. The emphasis is on how cleaning chemicals interact with the fouling materials and which chemical is more efficient in the cleaning procedures.

Table 2-4: Techniques preventing different types of membrane fouling [74]

Type of Fouling	Techniques			
	Hydraulic Cleaning	Feed Chlorination	Feed Acidification	Chemical Cleaning
Inorganic	-	-	++	++
Particulate	++	-	-	++
Microbial	+	++	+*	++
Organic	_	+	_	++

Note: - No effects or have negative effects. + Some positive effects, ++ Positive effect
*In conjunction of feed chlorination

2.3.2 Evolution of the concept and practical examples of membrane-based pretreatment before RO

Analysis of published data indicates an evolution of the principle of membrane- based pretreatment from the level of theoretical concept to practically commercialized technology. In recent years the membrane-based pretreatment has become widely considered as a viable alternative to conventional one. In particular, Cote *et al.* [75] showed that membrane-based schemes have become commercially competitive and replaced the conventional processes on a vast scale. These schemes are characterized by modular design, by smaller footprints for the same capacity, and other advantages that make them technologically attractive. Recent studies done by different researchers [73,76-77] outlined the following techno-economic advantages of the membrane-based pretreatment: (1) improvement of the nephelometric characteristics of water after pretreatment that makes them less vulnerable to seawater quality, (2) reduction of RO membranes fouling rate, (3) extension of the lifetime of the RO membranes, (4) decrease in the rate of chemical consumption, (5) decrease in the frequency of chemical cleaning,(6) reduction in the hydraulic resistance of the RO caused by fouling that, in turn, result in decreased energy consumption,(7) limited labor requirement, and (8) decreased manufacturing expenses in water production.

Drioli *at al.* [78] outlined the concept of integration of membrane operations for seawater desalination. It was pointed out that NF as a pretreatment allows increasing water recovery of the RO up to 50%. Pretreatment based on UF and MF leads to significant reduction in capital cost (from $ 2.00 - 4.00/gal to $ 1.75 -3.25/gal) that corresponds to 12-18%. Al-Sheikh [79] described SWRO pretreatment at the Jeddah plant in Saudi Arabia. Particular aspects of membrane–based pretreatment (installed on RO plant in Japan) are discussed by Taniguchi and Rosberg [80-81]. Rautenbach *et al.* [37] outlined the concept of seawater desalination based on UF pretreatment. Chakravorty and Layson [82] described the case of a membrane-based pretreatment where polypropylene membranes (with 0.2 micron) were used. An operating pressure was observed to be 100 kPa. The SDI_{15} index has been observed to be less then 3. Graeme *et al.* [83] presented data on UF-based pretreatment installed at Kindasa in Saudi Arabia. The performance of a UF membrane (at the trans-membrane pressure (TMP) < 0.2 bar) was observed to be stable at the level of 95-98 $L/m^2/h$, (No $FeCl_3$ dosing were done). The system is characterized by a decreased demand for chemical cleaning. (The RO followed by UF pretreatment doesn't require chemical cleaning regeneration over 6 months of the pilot test). No organic fouling of RO was observed to take place over the pilot test. Brehant *et al.* [76] described a UF pilot plant equipped with "Aquasource" system, based on hollow fiber polymeric membranes with molecular weight cut-off of 100,000 Dalton. Teuler *et al.* [71] described an "Aquasource" UF pilot system using cellulose derivative membrane and having an area of 7.2 m^2 and with molecular weight cut off of 100 kDalton. The pilot study demonstrated the following performance: (1) flux: 100 $L/h/m^2$ (at t=20 °C), (2) specific permeability (at t=20 C, ΔP=0.45 bar) was observed to be 240 $L/h/m^2/bar$; (3) specific energy consumption is 0.1 kWh/m^3(filtrate). Glueckstern *et al.* [84] submitted data on UF pretreatment with capillary membrane based on polyether sulfone polymer, where ID of capillary is 0.8 mm; membrane area is 25 m^2, and the MWCO is 150,000 Dalton.

Goto *et al.* [85] described a demonstration SWRO plant with a capacity of 200 m^3/day. The plant has a UF-based pretreatment with PVDF (Poly-Vinylidene Fluoride) membranes with nominal pore-size of 0.1μm and effective area of 50 m^2 per module. It can provide an SDI index <4 and resist high concentration of free chlorine. MF power consumption was 0.32 kWh/m^3.

According to expert's estimates the membrane-based pretreatment is characterized by smaller footprints for the same capacity. In particular, Galloway and Teng *et al.* [70, 77] stated that the membrane-based pretreatment requires less than 50 % of the area of conventional pretreatment. In addition it has a positive impact on RO characteristics due to the higher quality of pretreatment, namely, it will increase the flux rate of RO that, in turn, results in reduction of required RO size and capital cost [86].

New configuration of membrane system implying the concept of submerged or immersed membranes has become widely accepted in technology of wastewater recycling, membrane bioreactors, and surface water treatment plants. These systems consist of the shell-less hollow fibers being immersed into a tank that is open to the atmosphere. The driving force is generated by applying vacuum using a centrifugal pump. The applied vacuum can range from 0.14 to 0.63 bars. Permeate is removed by suction. For the shear stress at the membrane surface to increase, the air bubbling can be applied. Some authors referred to these processes as low pressure MF. Implication of this concept in technology of water recycling, in biotechnology, and in sewage treatment was considered by different authors, [75, 87-89], considered an application of the submerged membrane system produced by 'Zenon". The pilot tests conducted on seawater (at MPT of 25 kPa) gave a permeability of 25 $L/m^2.h$.

2.3.3 Water quality and technological aspects of membrane- based pretreatment

Data published by different researchers [28, 73, 76-77] pointed to advantage of membrane based pretreatment, such as decreased vulnerability of quality of filtrate to seawater quality. Quality of water after membrane- based pretreatment was observed to be capable of consistently reducing turbidity to less then 0.1 NTU disregarding turbidity level of influx. Taniguchi [80] indicated that the SDI value of the treated seawater with MF is kept below 4 disregarding the pollution of seawater. Glueckstern *et al.* [84] came to the similar conclusions, namely: quality of UF filtrate is less vulnerable to variation of seawater quality. Henthorne [86] compared the performance of membrane pretreatment and the conventional one. The study indicates that membrane pretreatment require lower chemical addition to achieve better water quality, in particular 0-1.5 ppm and 5-6 ppm ferric chloride, for membrane and conventional cases respectively. The membrane filtration provides sustainable water quality with the SDI index being from 0.6 to 2 regardless the quality of raw water. On the other hand, conventional pretreatment requires constant adjustment of rate of coagulant injection and has difficulty meeting values of the SDI index being less than 5 when the raw feedwater quality exceeds turbidity values greater than 10-15 NTU. Cleaning frequency was approximately every 42 days using conventional, compared to no cleaning requirements to date for the membrane pretreatment. Ebrahim [90] considered various types of pretreatment, such as conventional, microfiltration and beachwell technologies. Conventional pretreatment of surface water produces unsteady feedwater quality, since microfiltration is suitable for pretreatment of surface seawater feed for RO plants. Reduction in the COD and BOD were not always satisfactory in conventional pretreatment, where these characteristics were reduced substantially in MF filtrate. Similar results were achieved by Chua *et al.* [91] where the performance of different pretreatment system before RO was considered. The conventional pretreatment system was not efficient to remove oil and grease existing in seawater. The rejection of total suspended solids prior to the cartridge filter was observed to be 77.5%. In addition to the fluctuation of the SDI index of the filtrate, the chemical consumption and frequent replacements of cartridge filters were considered to be as a disadvantage of these systems. In the UF system, the sea water and filtrate were used as feed water respectively. In considered case the SDI of the UF filtrate was more stable than filtrate after sand filters. The MF pretreatment was tested and the result showed a moderate removal of colloidal silica and suspended solids. It was shown that the removal of colloidal silica and bacterial coliform could be improved using UF pretreatment. The observed degree of organics removal was found to be 16%, while no rejection in the case of oil and grease was found. The MF results showed that moderate removal of colloidal silica and suspended solids were possible through MF pretreatment. There was no rejection of reactive silica. Membrane pretreatment produces filtrate of a better quality, where the SDI index was found to be less than 3. The removal of fouling constitutes of seawater investigated by Brehant *et al.* [76] was observed to be more efficient using UF pretreatment than with conventional pretreatment. The UF reduced the SDI index from 13-25 to less than 0.8 whereas the DMF filtered water SDI remained between 2.7 and 3.4.

The comparison between sources of Gulf seawater was done by Bonnelye [92]. The study focuses on comparison of performance of pilot plants in the Gulf of Oman and Arabian Gulf. The pretreatment schemes were based on different technologies, conventional and membrane pretreatment. The conventional system applied to seawater from Gulf of Oman showed good quality of filtered water in terms of turbidity being at the level of 0.1 NTU, and the SDI index which decreased from 5 to 2.6. The efficiency of organics removal was found to be 25%. Similar result was obtained from the Arabian Gulf water, (where a minor increase in turbidity to 0.7 NTU and the SDI index 1.8-2.9). The membrane pretreatment gave good results in term of turbidity, algae and hydrocarbon removal, leading to a reliable SDI values being < 3. Another comparison study was done at the (Eilat site) and on the Mediterranean (Ashdod site), based on UF equipment with capillary backwashable elements operated in dead end flow mode. The conventional pretreatment unit included in-line flocculation followed by media filtration. During the test period both of them produced filtrate water of good quality, the SDI was in the range of 0.8-3.8 and turbidity ranges from 0.1 to 0.2 NTU. On the average, the UF pretreatment produced feed water with lower SDI and turbidity than the conventional pretreatment system. On the other hand according to Glueckstern *et al.* [84], the conventional pretreatment had a minor water loss during the backwashing resulting in higher recovery rate than the water recovery of the UF unit. Wilf and Schierach [28] proved that MF and UF can produce feedwater for RO of reliable quality. It is a cost competitive alternative to conventional technology and will result in improved economics of the SWRO through the reduction of chemical cost, required frequency of cleaning and costs of labor and membrane replacement. The total forecast reduction was expected to be approximately 10% of total water cost. Similar studies were done by Teng *et al.* [77] where the data on pilot study with different techniques such as ultrafiltration (UF) and microfiltration (MF) were presented. The results showed that membrane pretreatment produced a filtrate with SDI index less than 3. (The SDI of UF filtrate was less than that of MF).

UF pretreatment based on "Aquasource" was considered by Teuler *et al.* [70]. The specific flux of the UF membrane remained stable during the pilot test that seemed to be not sensitive to high salinity of the seawater and mechanical resistance. The test confirmed that the variation of seawater quality, transmembrane flux and transmembrane pressure have low impacts on the SDI index of filtrate.

Van Hoof *et al.* [93] studied UF for seawater pretreatment and reprocessing effluents of waste water treatment plant (WWTP). The UF system used in the pilot plant was developed by X-Flow in cooperation with the "NORIT membrane technology". The system was installed at Addur SWRO desalination plant. The SDI and turbidity were shown to be reduced while passing through UF (No preliminary pretreatment before the UF was applied). The SDI being 18-19 in feed water was reduced up to 1-1.5. Turbidity was reduced from 3-4 NTU to 0.3 NTU in filtrate.
Specific energy consumption is essential techno-economic characteristic it is dependent upon required driving force and hydraulic resistance. Data on energy consumption for the main and auxiliary equipment are given by Van Hoof *et al.* and Crespo and Boddeker [93-94] presented data on energy consumption for UF in dead- end mode with permeability 571 l/m^2-h, an energy consumption was observed to be 0.1 kWh/m^3. Goto *et al.* [75] provided data on energy consumption by MF pretreatment being installed on demonstration SWRO plant with capacity 200 m^3/day. MF power consumption is 0.32 kWh/m^3.

2.3.4 Economic aspects of the membrane-based pretreatment

According to data submitted by ESCWA [54], the cost of pretreatment in brackish and seawater desalination ranges from 35 to 60 % of the total water production cost. Data published by Durham and Walton [55] presented the pretreatment cost to be 22.97% of the total water cost, (data related to Doha Reverse Osmosis Plant with capacity 4546 m^3/day). Hafez and El-Menharawy [95] gave an itemized structure of the capital cost for SWRO desalination plants based on the IP project type where the cost of pretreatment varies from 9 to 12 % of the total cost.

According to research done by Glueckstern *et al.* [84] for the desalination plants on the Red Sea (Eilat site) and on the Mediterranean (Ashdod site), the cost of membrane pretreatment is higher than the cost of conventional pretreatment. The cost of UF equipment allocated to water cost is estimated to be \$ 0.048- \$ 0.057/m^3 while the cost of equipment for conventional scheme ranges from \$ 0.01 to \$ 0.02 /m^3. In contrast Ebrahim *et al.* [96] state that the total unit costs produced by conventional and microfiltration units are 0.093 \$/m3 (28.153 fils/m^3) and 0.04 \$/m^3 (12.264 fils/m^3), respectively. Roberto *et al.* [97] showed that specific production costs for MF and UF plants are about 5% lower than conventional water production, (but the dosing of activated carbon will increase the O&M costs to ~ 20% of the conventional schemes). A comparison of the capital costs reveals that membrane plants are 30-50% cheaper than the conventional treatment plants considered in the study.

Al-Malack [98], estimated the cost of microfiltration process and different schemes of conventional wastewater treatment. Capital cost analysis showed that the cost of slow sand filters was the lowest among all the systems being compared. More than 50% of reduction in the capital cost was obtained when cross flow microfiltration was replaced by slow sand filtration. The economic study showed that the capital cost of conventional tertiary treatment process is more than 65% higher than that of cross flow microfiltration, while the annual operation and maintenance expenses were 21% less.

Glueckstern *et al.* [84] gave the following economic data on membrane pretreatment: the specific investment, including site and utilities: 112.5-137.5 \$/m^3/day; where the cost of UF equipment 65.5 -87.5 \$/m^3/day; total UF filtrate cost (incl. investment+ operating cost): 0.048- 0.057 \$/m^3. Glueckstern *et al.* [84] outlined advantages of membrane based pretreatment such as decreased vulnerability of UF filtrate characteristics to seawater quality, but at the same time they pointed that the cost of membrane pretreatment is higher than the cost of conventional pretreatment. This process can be feasible for sites which require very expensive conventional pretreatment or where wide fluctuation of raw water quality can be expected.

Wilf and Schierach [28] expected that UF and MF to be an attractive pre-treatment technology. Since UF and MF have been commercialized, it enables a more advanced RO system design which should result in increased reliability and lower water cost. According their estimates, the total water cost should be reduced by about 10%.(Estimates are done for the cases where the flux rate in RO system has to be 13.6 L/m^2-h). Glueckstern *et al.* [84] gave techno economic data of membrane-based pretreatment. Coagulant was added to the UF feed at the rate 0.3 ppm. The flux rate of UF 60 120 l/m^2-h. UF membranes were backwashed with filtrate at the interval of 15 -30 min. (During the backwash a free chlorine (in the form of hypochloride) was added to the filtrate to the level of 20 ppm). Conventional system was more vulnerable to fluctuation of seawater quality.

2.3.4.1 Capital cost (Cost of installed equipment)

Configuration of the flow diagram and cost of installed equipment are site-specific and dependent upon different factors that make analysis of published data complicated. Majority of applied conventional schemes include the following unit operations and subsystems: (1) intake; (2) coagulation chamber; (3) multimedia and cartridge filters; (5) chlorination system, (6) chemical regeneration and backwashing systems. Membrane-based schemes available in literature have no unified configuration and in any particular case require special analysis. Glueckstern *et al.* [84] estimated the specific investment, including site and utilities as 112.5-137.5 \$/m^3/day, where the cost of UF equipment to be 65.5 -87.5 \$/m^3/day, while the cost of equipment for conventional pretreatment (including clarifier and two stage filtration) ranges from 40 to 45 \$/m^3/day. A study done by bureau of reclamation [99] reported results where the equipment cost for membrane and conventional treatment are 374.41 \$/m^3/day and 260.01 \$/m^3/day respectively. Galaway [70] concluded that cost of membrane treatment is higher than the cost of conventional treatment for SWRO. Hafez and El-Menharawi [95] estimated the specific cost for conventional pretreatment consisting of settling equipment, media and MF, to be 161 \$/m^3/day. Study done by Al-Malack [98] showed that capital cost for wastewater treatment using microfiltration expected to be 272.65 \$/m^3/day that is higher than the cost of conventional treatment 210.78 \$/m^3/day. (The conventional pretreatment consists of coagulation and sand filtration.) However, Ebrahim *et al.* [100] gave an estimate of the capital cost of microfiltration system to be 62.33 \$/m^3/day, while the capital cost for conventional treatment was estimated to be 229.87 \$/m^3/day. Gote and Liu [101] gave an estimate of specific cost of seawater pretreatment before RO based on submerged membranes as 238 \$/m^3/day. According to economic study done by Duranceau *et al.* [102], the total installed UF equipment cost is 0.68 \$/gal/day (180 \$/m^3/day). The data are relevant for the plant with capacity 1.67 MGD. Data presented by Wilf and Klinko [7] corresponds to the case where two stages of gravity filtration with 40% of permeate recovery were installed. The cost of pretreatment estimated to be 3.258 106 $ gives the specific cost of pretreatment 143.5 $/m^3/day. The operation cost was calculated for UF pretreatment.

2.3.4.2 Operation and maintenance cost (O&M cost)

According to data submitted by Van Hoof *et al.* [93] operating costs for UF pretreatment, are expected to be 0.07- 0.09 EUR/m3. O&M costs include the following items: (1) cost of energy; (2) cost of chemicals; (3) cost of membrane replacement; (4) labor cost.

Cost of energy is dependent upon specific energy consumption that is in turn is influenced by required driving force and resistance. Any desalination process is equipped by the following main and auxiliary equipment: intake pumps, chemicals dosing pumps and air scouring blowers. According to data published by Ebrahim and Abdel-Jawad [103], the cost of consumed energy is about 42.7% of the operating cost. The main energy consuming equipment is installed on the stage of RO desalination while the pretreatment is characterized by minor energy consumption. The power consumption cost given by Ebrahim *et al.* [100] is 0.023 $/ m^3 for membrane pretreatment and 0.032 $/ m^3 for conventional pretreatment when the power cost is 0.06 $/kWh. (Glueckstern *et al.* [84] found that power consumption of UF process is 0.0045$/ m^3 when the power cost is 0.05 $/kWh.

Table 2-5: Cost of energy based on available published data

	Energy Cost, $/kWh	Reference
1	0.04-0.09	Elttouney et al., 2002 [44]
2	0.05	Glueckstern et al., 2002 [84]
3	0.048	Hafez and El-Menharawy, 2002 [95]
4	0.059	Ebrahim et al., 2001 [100]

Cost of chemicals represents 5.5% of O&M costs for the whole process that suppose to be at the level of 0.05 $/m^3, [5,103]. According to their estimates the expenses for chemicals to be 5.7% of O&M cost that correspond to 0.047 $/m^3. Glueckstern *et al.* [84] and Ebrahim *et al.* [100] estimated the chemical cost from 13.7% to 26% of O&M costs (in case of conventional pretreatment). It was found the cost of chemicals in conventional pretreatment to be higher than in membrane pretreatment (See Table 2-6). The difference between these data can be explained by different water quality used in the studies.

Table 2-6: Cost of chemicals for conventional and membrane-based pretreatment [84,100]

	Pretreatment Scheme	Chemicals Cost, \$/m^3	
		Data by Glueckstern *et al.*, [84]	Data by Ebrahim *et al.*, [100]
1	Conventional system	0.048 (21.9%)*	0.0155 (26.2%)*
2	Membrane system	0.027 (13.7%)*	0.00545 (18.7%)*

*% of O&M Cost

Table 2-7: Cost of chemicals used for seawater and brackish water pretreatment [48]

		\$/kg	\$/m^3
1	Antifoam	2.54	0.003
2	Sulfuric acid	0.58	0.01
3	Antiscalant	4.38	0.01
4	Sodium hexametaphosphate	0.77	0.005
5	Caustic (NaOH)	0.51	0.005
6	Sodium sulfide	0.13	0.005
7	Chlorine	0.33	0.0031

2.3.4.3 Labor cost

The labor cost depends on various factors such as: technological level, structure of labor market, plant ownership etc. The cost of water while desalination is not labor intensive technology is not sensitive to labor coat. Data published by [100,104] stated that the labor cost ranges from 3- 5% of the total water cost.

2.4 Conclusions based on the analysis of published data

Analysis of published data indicates that membrane-based schemes have become commercially competitive. These schemes are characterized by the following techno-economic advantages: (1) smaller footprints for the same capacity, (2) improvement of nephelometric characteristics of water after pretreatment that makes them less vulnerable to seawater quality; (3) reduction of RO membranes fouling rate; (4) extension of lifetime of RO membranes; (5) decrease of rate of chemical consumption; (6) decrease frequency of chemical cleaning; (7) reduction in hydraulic resistance of RO caused by fouling that, un turn, result in decreased energy consumption; (8) limited labor requirement and decreased manufacturing expenses in water production.

3 Objectives and methodology of this study

This study focuses on comparison of conventional and membrane-based types of pretreatment. The study is based on experimental data received from pilot system and full-scale plants. Technological efficiency of pretreatment system can be characterized by the decrease of the fouling factors. This study focuses on technological and economic aspects of the technology that, in turn, is based on the estimation of the technological and economic indicators. Three groups of indicators were used for comparison of alternatives. The first group includes water quality data; the second one on technological characteristics of equipment, and the third group covers the economic characteristics of the process. Water quality after pretreatment (1st group) is characterized by the following main nephelometric indicators: turbidity, total suspended solid, SDI index and the degree of its rejection. The second group of indicators includes normalized permeability, transmembrane pressure difference, energy consumption, and deterioration of these characteristics during the test period. The third group includes cost of chemical and energy allocated to cost of produced water.

3.1 Objectives of the Study

Published data and regional experience, [24-30], confirm that the membrane pretreatment before RO is becoming a competitive technology. According to the published studies these schemes are characterized by many techno-economic advantages (see section 2.3.2). Within the context of the problem this study focuses on the analysis of techno-economic aspects of conventional and membrane-based pretreatment before RO desalination. It is aimed to the following objectives:

- Estimation of the **nephelometric characteristics of water** such as turbidity, total suspended solid, and the SDI index and the degree of its rejection.
- Estimation of the **technological characteristics of equipment** such as normalized permeability, transmembrane pressure, energy consumption and deterioration of these characteristics during the test period.
- Estimation of the **cost of chemical and energy** allocated to the cost of produced water.
- Estimation of the **cost of installed equipment** in the stage of pretreatment for the scheme of the demo plant proposed by Japan Cooperation Center for the Middle East. (The proposal is within the program "Advanced Hybrid Desalination System in Abu Dhabi). For the economic estimation of these schemes the "Super Pro-Designer software" **[www.intelligen.com]** is going to be used.

3.2 Methodology

3.2.1 Criteria for estimation of water quality

- **Silt Density Index (SDI_{15})**

It is a measure of the overall impurities that can cause membrane fouling (or blocking). It can be expressed as:

$$SDI_{\tau} = \frac{(1 - t_1 / t_2)}{\tau} 100\% \quad [3\text{-}1]$$

Where t_1 is the time required to filter 500 ml measured at the start of the test; t_2 is the filtration time required for the same volume of the sample at the end of test, and τ is the duration of the test (normally τ =15 minutes).

- **Turbidity**

It is a quantitative indicator of colloidal and heterogeneous matter and is measured in Nephelometric Turbidity Units (NTU). The rejection of turbidity through a pretreatment system can be expressed as follows:

$$R(X) = (X_{IN} - X_{OUT}) / X_{IN} \quad [3\text{-}2]$$

Where X_{IN} and X_{OUT} are the turbidity of the inlet and outlet streams respectively.

- **Total Suspended Solids (TSS)**

It is a quantitative indicator of suspended matter. Reduction of the TSS through pretreatment system can be expressed as:

$$R(Y) = (Y_{IN} - Y_{OUT}) / Y_{IN} \quad [3\text{-}3]$$

Where Y_{IN} and Y_{OUT} are the TSS in mg/l for the inlet and outlet streams, respectively.

- **Iron Reduction**

To avoid oxidation of ferrous ion (Fe^{+2}) into ferric oxide within a permeator, concentration of iron should be decreased before entering the RO unit. It can be expressed as:

$$R(Z) = (Z_{IN} - Z_{OUT}) / Z_{IN} \quad [3\text{-}4]$$

Where, Z_{IN} & Z_{OUT} are the concentration of iron, µg/l, in raw water and at the point of downstream after the Dual media filter, respectively.

3.2.2 Criteria for Estimation of Technological Characteristics of Equipment

- **Transmembrane Pressure Difference (TMPD)**

The transmembrane pressure difference is the driving force for the membrane separation which is defined as the difference in pressure between the filtrate side of the membrane and the permeate side of the membrane. Assuming linear distribution of the operating pressure in a high pressure channel, the TMPD can be expressed as:

$$TMPD = \frac{P_{FEED} + P_{BRINE}}{2} - P_{PERM} \qquad [3\text{-}5]$$

- **Specific Energy Consumption**

Specific energy consumption by pumps, kWh/m^3 (filtrate), is an essential criterion for techno-economic evaluation. It strongly depends on the recovery and can be expressed as follows:

$$\varepsilon = \frac{\Delta P * Q_{feed}}{Q_{filt}} \qquad [3\text{-}6]$$

Where ε the energy required, kW

ΔP: the differential pressure across the pump, bar

Q_{feed}: the feed flowrate of the liquid, m^3/h

Q_{filt}: the filtrate flowrate of the liquid, m^3/h

- **Specific Chemical Consumption**

The chemicals needed for the pretreatment stage are varying on the dosing regime and rate depending on the water quality specifications. Because this item can constitute a significant portion of the operation cost, it is worth to take it into consideration in the evaluation procedure.

- **Normalized Permeability of the Membranes**

The observed membrane permeability is a measure of the transmembrane flux. It can be expressed in m^3/m^2.s.bar. Being determined at the current operating temperature, it should be converted to its normalized value, at a reference temperature (t=25 °C). To account the effect of operating temperature, the observed permeability should be multiplied by a correction factor equals to ratio of viscosity at operating temperature to the viscosity at the reference temperature (t_0=25 °C).

$$A_{t=25} = A_t \frac{\mu_t}{\mu_{t=25}} \qquad [3\text{-}7]$$

- **Normalized Permeability Deterioration**

The decline of normalized permeability characterizes the deterioration of the membrane permeability during the test period; it can be expressed by the following relation:

$$Y = \frac{P_i - P_n}{P_i} * 100 \qquad [3\text{-}8]$$

Where Y is the deterioration in normalized permeability, %. P_i is the average permeability in the very beginning of the test period, P_n is the average permeability at the end of the test period. The previous criteria are used in the case studies evaluation which are described and analyzed in the next chapter.

- **Standard Deviation**

Many experimental readings are characterized by stochastic behaviour. For the degree of their fluctuation to be evaluated, the Standard Deviation [105] has to be evaluated. It can be expressed as:

$$SD = \sqrt{\frac{\Sigma(X_i - \overline{X}_i)^2}{n}} \qquad [3\text{-}9]$$

Where X_i : experimental reading; $\overline{X}_i$: value estimated by linear approximation, and n : number of readings.

3.2.3 Estimation of Economic Indicators of the Process (Capital and O&M Cost)

Capital and the O&M costs were considered as economic indicators in this study. The structure of the fixed capital investment (including direct and indirect cost segments) was adopted from Peters and Timmerhaus [106]. The fixed capital cost is equal to the sum of the direct (DC) and indirect costs (IC). The total direct cost (DC) includes the following items: equipment purchased cost; installation; piping & instrumentation; insulation; buildings; yard improvement; and auxiliary facilities. The total indirect cost (IC) includes payment for engineering and construction. The O&M cost includes the following of expenses: energy; membrane replacement; labor; etc.

4 Conventional and Membrane-Based Pretreatment before RO: Case Studies

The chapter focuses on analysis of technological schemes for conventional and membrane-based pretreatment. Evaluation of different schemes is based on the data gathered from the pilot systems and existing RO plants. The following pretreatment schemes were considered in the study: (1) Conventional pretreatment of the RO pilot plant installed by Ondeo Ltd. located at Al–Tweelah site; (2) Conventional pretreatment installed on Al-Fujairah hybrid desalination plant; (3) Membrane–based pretreatment based on the "Zenon" system located on Al–Taweelah site; (4) Membrane–based pretreatment based on the "Aquasource" system located at Al–Taweelah site, (5) the hybrid type of pretreatment proposed by Graham Tech Pte Ltd. (Singapore) located in Bainouna power station, and (6) RO desalination plant in Addur (Bahrain). The full description and analysis of each particular case is given below.

4.1 Pretreatment before RO based on conventional schemes

The majority of existing RO desalination plants are equipped with similar type of conventional pretreatment based on coagulation and multimedia filtration. The majority of the conventional schemes include (1) disinfection, (2) flocculation, and (3) filtration

4.1.1 Al–Fujairah hybrid plant

Al-Fujairah cogenerative plant is considered as the first one of its kind in the Middle East region and one of the biggest in the world that uses a combination of the two water desalination technologies. The production of power is 656 MW and of water 100 MIGD, respectively. The desalination plant produces 62.5 MIGD of water using the multi-stage flash (MSF) and 37.5 MIGD by reverse osmosis (RO). The desalted water is transported to the Northern Emirates and further to Abu Dhabi city through a 179-kilometer dual pipeline.

The quality of the seawater of the intake point of Al-Fujairah plant is characterized by more stable values of indicators. According to Bonnely *et al.* [92], the seawater of the Gulf of Oman is less vulnerable to seasonal variation and characterized by a turbidity around 0.2 NTU, and a SDI_5 index of less than 6. The seawater analysis at Gulf of Oman is given in Appendix B.

The water produced by the RO plant has a TDS of less than 180 ppm and a chloride content of less than 120 ppm. The RO plant consists of the following unit equipment: (1) Seawater intake, (2) Coagulation chamber, (3) Dual media filter, (4) Cartridge filter, (5) High pressure pumps with recovery turbines, and (6) Reverse osmosis systems.

A single stage RO would allow treated water with a TDS of about 500 ppm, therefore, a second pass stage, treating about 80% of the total capacity coming from the first pass is installed. The first pass is designed for a recovery rate of about 43% and consists of 18 RO trains running continuously in parallel. Each train includes a high pressure pump and a RO rack. Common headers are provided at the delivery of the high pressure pumps and at the outlet of the rejects. The second pass is designed for a recovery of 90%, which gives a total recovery rate of 41%. The second pass consists of 8 RO trains. The brine is returned to the turbine and excess energy is recovered before being discharged to the sea. The desalted water produced by the RO blocks in the first pass is divided into two streams; one stream is fed to the second pass RO blocks and the other, to the desalinated water tanks.

4.1.1.1 Conventional pretreatment scheme (Al-Fujairah hybrid plant)

The plant under study is equipped with conventional pretreatment that consists of the following unit operations: (1) disinfection, (2) flocculation, and (3) filtration. The scheme shown in Figure 4-l is a simplified flow-diagram of the pretreatment system. Chlorinated water from the intake (1) passes through the coagulation chamber (2), where ferric chloride ($FeCl_3$) is used as a coagulant, (3) After the coagulation chamber, the seawater is passed through the multimedia filters at a filtration rate of 8.66 m^3/ h/ m^2. The filtered water is collected in two storage tanks (3500 m^3), located under the filters. Then the filtered water passes through the cartridge filters (4) for further treatment by antiscalant and sodium bisulfite. The filtered water is then pumped to the RO plant through 18 horizontal pumps with fixed speed (one pump per RO block).

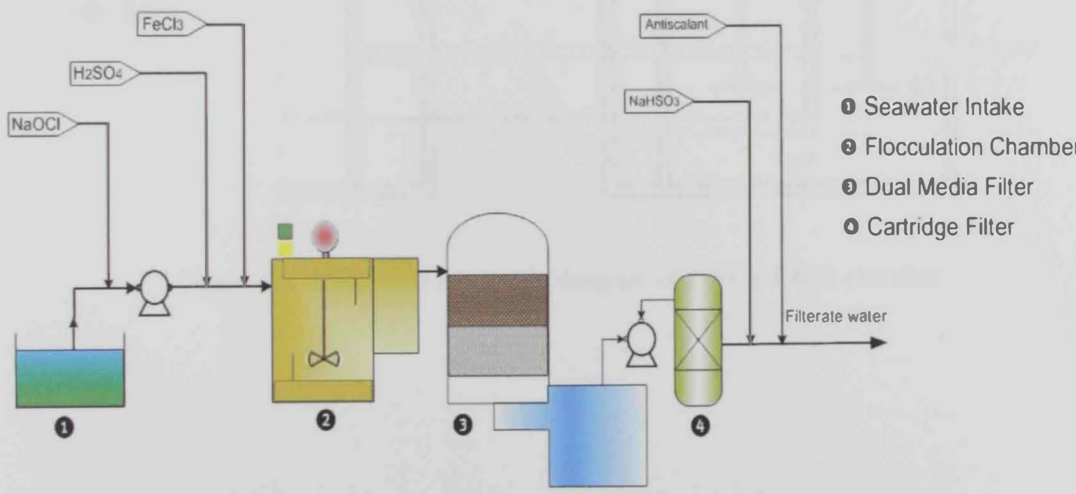

Figure 4-1: Schematic flow diagram of pretreatment before RO (Al-Fujairah hybrid plant)

The pretreatment under consideration is accompanied by injection of chemicals. The following chemicals were used: (1) sodium hypochlorite, NaOCl, (2) ferric chloride, $FeCl_3$; (3) cationic coagulant, (4) sulfuric acid, H_2SO_4; (5) antiscalant, and (6) sodium bisulfite. Data on some specific reagent consumptions are given below.

- **Disinfection**

To prevent biological growth, periodical chlorination in intake is applied. Shock injections of sodium hypochlorite (NaOCl) are used on a weekly basis (2-3 ppm for 2 hours once a week).

- **Coagulation (and flocculation)**

The process used to reduce the forces between particles is referred to as coagulation followed by flocculation. The purpose of flocculation is to increase the collisions of coagulated solids in order to agglomerate them for filterable (or settleable) solids. Flocculation is accomplished by agitation of coagulated particles in order to increase particles size (or density). Flocculation is carried out in two stages, each stage containing a chamber with two flocculators; each chamber has a 400 m^3 capacity. Injection rate of ferric chloride ($FeCl_3$) ranges from 3 to 5 ppm. To enhance the coagulation, (when the SDI > 20), a polyelectrolyte has to be used. Simplified schematic diagram is shown in Figure 4-2. Main characteristics of coagulation chamber are given in Table 4-1.

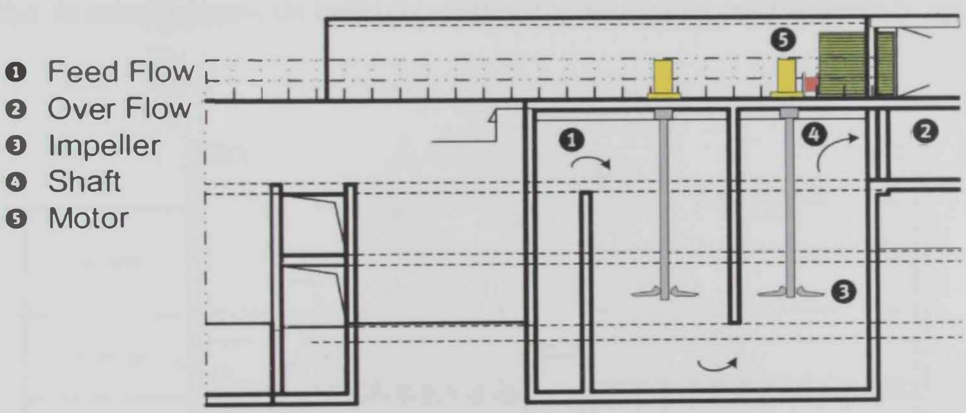

Figure 4-2: Simplified schematic diagram of a coagulation chamber

Table 4-1: Coagulation Chamber characteristics [107]

Number of chambers	2
Flow rate	21000 m^3/h
Dimensions of the tank	L = 5 m, w = 5 m, h = 9.4 m
Liquid depth	8.7 m
Nominal flow rate	9415 m^3/h (total inlet flow 18830 m^3/h)
Number of impellers	1
Number of blades per impeller per chamber	3
Blade diameter	2760 mm
Through speed	0.98 m/s

- **Multimedia filtration**

The filtration process is used to remove the suspended particles (whether these particles existed in the raw water or originated by a coagulation process). Each filter cell is made up of a rectangular concrete tank filled by three layers of filtering media: pumice, sand, and gravel. The upper layer of the granular filter media is characterized by large particles size and low density when the lower layer has a fine particles size and greater density if particles (Figure 4-3). The overall filtration area is 2170 m^2 (The system includes 14 units with a unit area of 155 m^2). The filtration rate is 8.66 $m^3/h/m^2$. The main characteristics of the dual media filter are shown in Appendix C. The suspended solids gradually block the porous space of the layer that, in turn, increases the hydraulic resistance of the filter. In order to remove the hydraulic resistance a backwashing process should be applied.

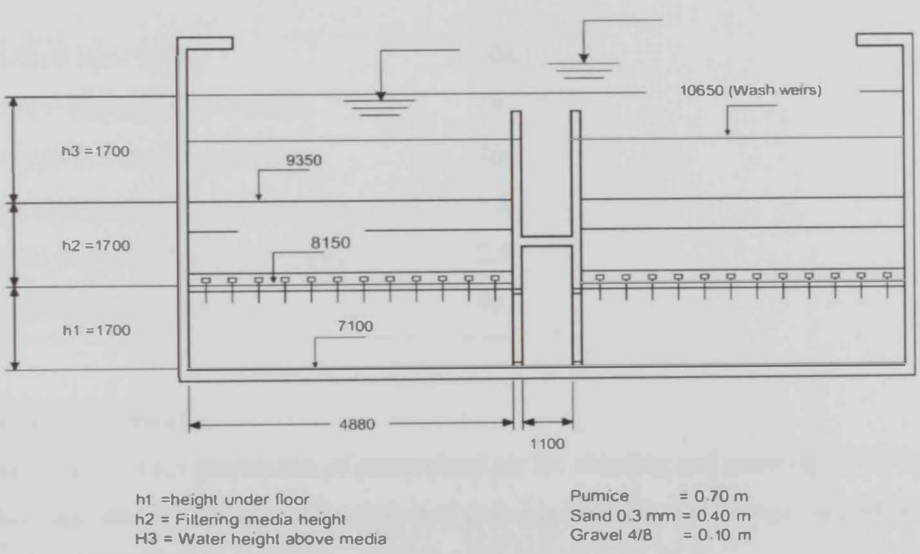

Figure 4-3: Schematic diagram of dual media filtration

There are two main steps in the filtration process: filtration and washing. The filtration step in which water first flows through the larger size material layer. This layer has an important retention capacity of suspended matters, with the larger particles removed, water flows through the lower layer. This layer enables the filtered water to be refined. Because the upstream level is kept constant, the outlet flow is equal to the incoming flow and clogging is compensated for until it reaches a maximum value that depends on the available head. In the washing step after partial drainage, agitation with air alone to detach the impurities for 5 minutes applied. Then intense washing with water alone with material fluidization also applied to evacuate the impurities and reclassification of the filtering medium. The lightest grains go upwards with ascending fluidization and the heaviest stays at low level, the water backwashing process takes 10 minutes. Filters should be washed when the head loss value in the filter reaches a preset level or when the filter run length reaches a preset time or when the filtered water turbidity or the SDI index exceed the preset value.

- **Cartridge filtration**

In order to prevent the RO system from accidental intrusion, a safety filter of cartridge type (with 5 microns nominal pore size) is installed, as shown in Figure 4-1. A safety filter is used. Differential pressure should not exceed 1.5 bars; otherwise the filter must be regenerated or replaced. Each cartridge filter is made up of a carbon steel vessel with a natural rubber inner layer. The support plates on which the filtering cartridge are mounted, are inside the vessel. The main characteristics of the cartridge filter are given in Table 4-2.

Table 4-2: Cartridge filter characteristics [UWEC]

Design flow rate (m^3/h)	1060
Number of cartridge filter vessels	18
Total number of cartridge filters	360
Design differential pressure (bar)	1.5
Maximum differential pressure (bar)	2.5
Total filteration area (m^2)	90

- **Auxiliary equipment**

Air blowers are used for production of pressurized air for cleaning and washing of the multi media filter. There are three blowers; two on duty and one stand by. The air blowers are working for 5-6 hours per day with a flowrate of 4250 m^3/h for each one.

Sodium hypochlorite (NaOCl) is generated on-site and requires only salt, water and electricity to produce the amount of sodium hypochlorite needed. The process of generating sodium hypochlorite involves passing seawater through an electrolytic cell where electrolysis takes place according to the following equation:

$$NaCl + H_2O + 2e \rightarrow NaOCl + H_2 \quad [4\text{-}1]$$

Products of electrolysis are sodium hypochlorite (NaOCl) solution and hydrogen (H_2) gas. Sodium hypochlorite solution containing hydrogen is transferred from the electrolyzer unit to storage tank and hydrogen gas disengages from the liquid phase in the upper part of the tank.

4.1.1.2 Water Quality Characteristics after Pretreatment (Al- Fujairah hybrid plant)

The overall efficiency of the pretreatment can be characterized by the decrease of fouling factors and can be quantified by the following characteristics: the silt density index (SDI_{15}), turbidity, total suspended solids, etc. Evaluation of the pretreatment is based on the quantification of these characteristics and estimation of the degree of their decrease. Analysis is based on data accumulated over the period from May to July of 2004. Data on sea water quality are given in Appendix B.

- **Silt Density Index (SDI_{15})**

The average value of the SDI_5 index before pretreatment is 15.5 and average value of the SDI_{15} index after pretreatment is 3.4. The experimental values of the SDI_{15} index (for seawater and water after pretreatment) are shown in Figure 4-4. The true standard deviation, for the seawater and the filtrate readings are equal to 1.07 and 0.3, respectively. The true standard deviation, *SD*, can be interpreted as a measure of sensitivity of filtrate to fluctuation of seawater quality. Comparison of the true standard deviation for seawater and for filtrate indicates that the filtrate readings are getting less randomized than that of seawater. Analysis of the behavior of the profile in Figure 4-4 indicates that the pretreatment system attained the required level of the SDI index. The SDI_{15} index of the filtered water ranges from 2 to 4.5 that meets the required level of filtered water quality.

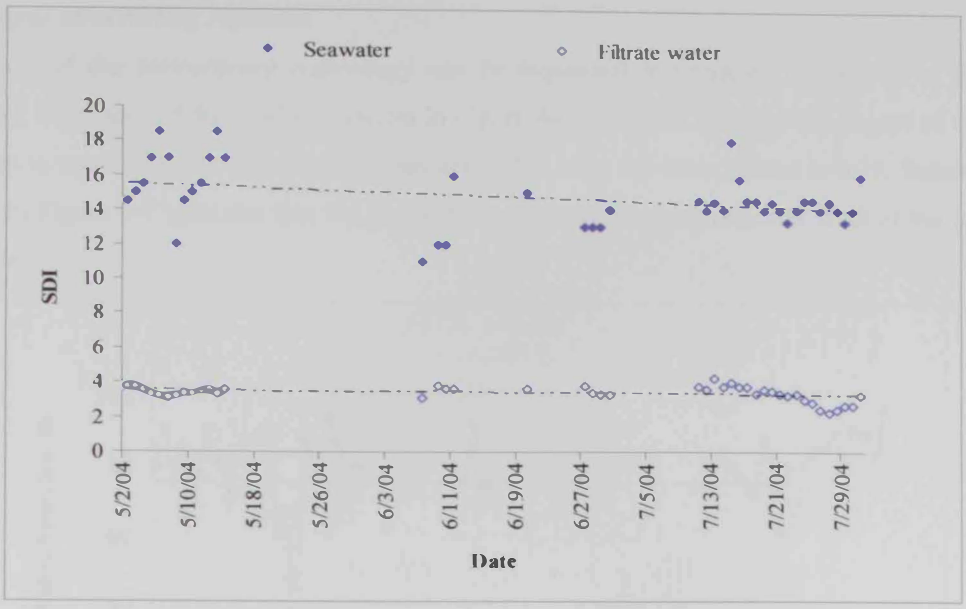

Figure 4-4: The SDI index of seawater and filtrate water (Al-Fujairah hybrid plant)

- **Turbidity**

The cartridge filter is referred to a safety filter, used to trap the accidental particles from the RO feedwater. Turbidity is a quantitative indicator of the colloidal and dispersed matters. Averaged value of turbidity after pretreatment is 0.157 NTU. Experimental values of turbidity (for seawater and water after cartridge filter) are shown in Figure 4-5. The standard deviation, for seawater and filtrate are equal 0.4 and 0.09, respectively. The profile in Figure 4-5 indicates that the pretreatment system provides the required low level of turbidity.

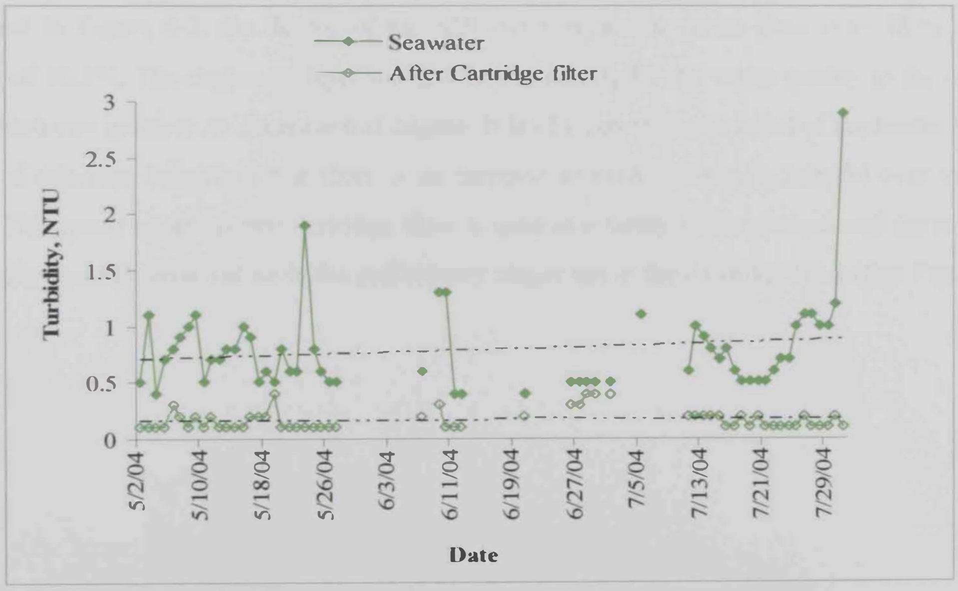

Figure 4-5: Turbidity of seawater and filtrate after cartridge filter (Al-Fujairah hybrid plant)

- **Degree of turbidity rejection**

Efficiency of the pretreatment technology can be expressed in terms of the degree of turbidity rejection; behaviour of the profile is shown in Figure 4-6.Averaged value of the degree of turbidity rejection is equal to 0.75. True standard deviation, *SD*, over the same period is 0.19. Behaviour of profile in Figure 4-7 indicates that the pretreatment system provides required level of the turbidity rejection.

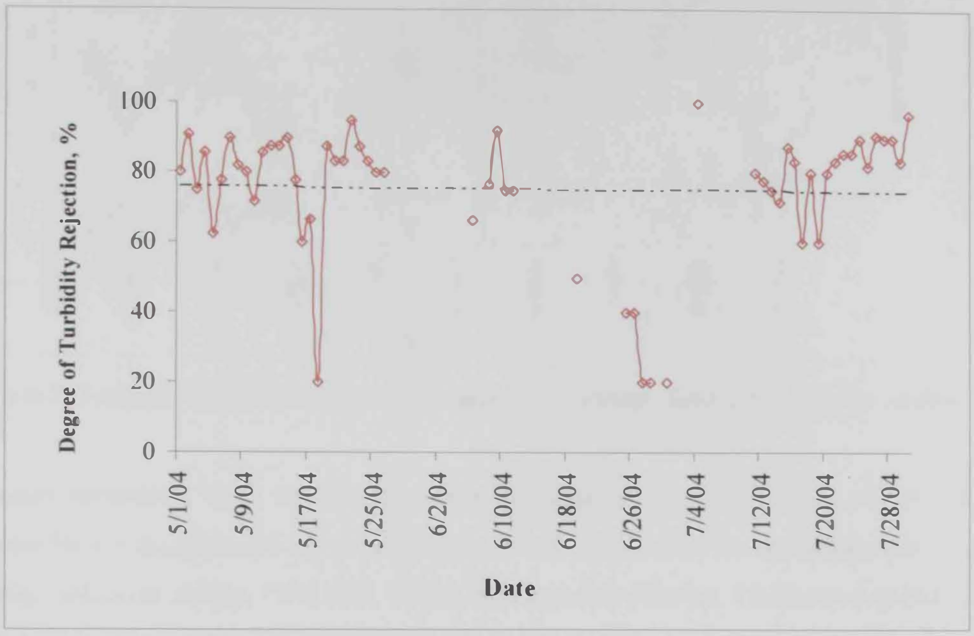

Figure 4-6: Degree of rejection of turbidity (Al-Fujairah hybrid plant)

As shown in Figure 4-7, the degree of the SDI index rejection varies from 0 to 38 % with an average of 10.3%. The degree of rejection is affected mainly by the water quality to the cartridge filter which can increase or decrease this degree. It is obvious from the trend of feedwater SDI and degree of rejection behaviors that there is an increase in each of them in parallel over the study period. This result confirms that cartridge filter is used as a safety filter no more and the main load of turbidity and SDI removal lie in the preliminary stages not in the cartridge filter (See Figure 4-7).

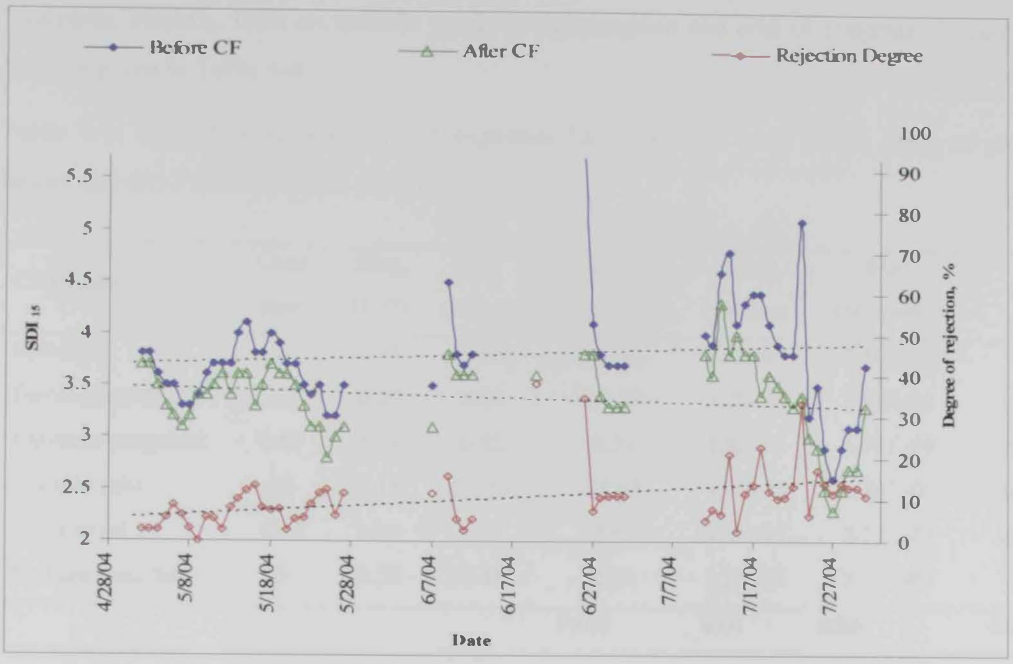

Figure 4-7: Turbidity of filtrate water before and after cartridge filter (Al-Fujairah hybrid plant)

The main parameters taken into consideration in mentoring the performance of the plant and as indicator for fouling potential (by measuring them before and after the treatments) are: temperature, turbidity, pH, conductivity, TDS, SDI, TPHC, iron and free chlorine. The measurements were taken regularly with a certain schedule. Table 4-3 presents the most important parameters for the pretreatment stage.

Table 4-3: Characteristics for estimation of efficiency of pretreatment (Data are averaged over the period from May 2004 to July 2004)

Characteristics	Before pretreatment	After pretreatment
Turbidity, NTU	0.368	0.157
SDI_{15} index	14.6	3.4
Iron, ppm	0.058	0.009
Total petroleum hydrocarbons, ppb	0.005	Not detected

4.1.1.3 Consumption of Chemicals on the Stage of Pretreatment before RO (Al-Fujairah hybrid plant)

The specific chemicals consumption is an essential indicator for the evaluation of the techno-economic characteristics of the process. The following reagents were used: (1) ferric chloride, $FeCl_3$; (2) sulfuric acid, H_2SO_4: (3) sodium hypochlorite, NaClO; (4) antiscalant; and (5) sodium bisulphite, Na_2SO_3. Data on specific reagents consumption and cost of reagents allocated to water cost are given in Table 4-4.

Table 4-4: Specific consumption and expenses for chemicals used in the stage of pretreatment before RO (Al-Fujairah hybrid plant)

Chemicals	Conc ppm	\$/kg, [107]	g/m³ (filtrate)	g/m³ (permeate)	\$/m³ (filtrate)	\$/m³ (permeate)	\$/day
Chlorine	3	0.55	0.034	0.08	1.8E-05	4.50E-05	0.35
Ferric chloride	3	0.27	8.16	20.58	2.2E-03	5.57E-03	959.40
Cationic coagulant	0.85	1.94	0.92	2.33	1.8E-03	4.53E-03	779.33
Sulfuric acid	25	0.18	27.19	68.63	4.9E-03	1.24E-02	2132.52
Antiscalant	1.05	1.94	1.05	2.65	2.0E-03	5.15E-03	886.67
Sodium bisulfite	6	0.50	0.2486	0.63	1.2E-04	3.14E-04	54.00
				Total	**0.01**	**0.03**	**4812.27**

Analysis of the published data indicates that the expenses for chemicals allocated to water cost ranges from 4.8-5.7% of O&M costs for the whole system. In particular Ebrahim and Abdel-Jawad [103], gave an average value of 5 % that corresponds to 0.026 \$/m³. Ray and McCray [12] however, gave a value of 5.8% of O &M (0.04 \$/m³). Thus the data obtained from the Al-Fujairah plant are in line with the published data.

4.1.1.4 Energy consumption in the Stage of Pretreatment before RO (Al-Fujairah hybrid plant)

This study considers energy consumption only in the pretreatment stage. (Energy consuming equipment located in the stage of the high pressure RO desalination is outside the scope of the study). The following energy consuming equipment were considered: (1) Intake pumps, (2) Flocculation chamber, (3) Chemical dosing pumps, (4) Air blower and (5) Cartridge filter pump. Data on energy consumption on the pretreatment before RO are consolidated in Table 4-5. The calculated energy consumption calculation for auxiliary equipment used is shown in Appendix D and E.

Table 4-5: Specific energy consumption* in the stage of pretreatment before RO (Al-Fujairah hybrid plant)

	Equipment	**Specific energy consumption kWh/m³ (permeate)**
1	Intake pump	0.248
2	Chemical dosing pumps	
	Antiscalant dosing pump	1.27E-06
	Sodium bisulphite dosing pump	1.74E-06
	Ferric chloride dosing pump	8.17E-06
	Sulfuric acid dosing pump	1.51E-05
	Cationic coagulant dosing pump	9.24E-07
3	Backwashing pump	0.069
4	Air blower	0.021
5	Flocculation mixers	2.9E-03
6	Cartridge filter pump	0.501
7	Auxiliary consumption	0.2
8	**Total specific energy consumption**	**1.04 kWh/m³ (permeate)** [0.416 kWh/m³ (filtrate)]

*All values in the table are expressed in terms of kWh per cubic meter of RO permeate. For the value to be converted to per cubic meter of filtrate after pretreatment, the table value should be multiplied by the degree of permeate recovery (40%). Thus, the specific energy consumption in terms of filtrate after pretreatment is equal to 0.416 kWh/m³(filtrate)

Analysis of the data shows that the chemical dosing pumps have insignificant share in the power consumption. The overall energy consumption by the RO desalination (including the high pressure desalination and post-treatment operations) is 5.15 kWh/m³ (permeate).The energy consumed in the stage of pretreatment (1.04 kWh/m³(permeate)) represents only 20.2% of the overall energy consumption.

The cost of energy allocated to water cost is estimated to be 0.03 \$/m³(permeate). This value is based on the assumption that the selling cost of electricity is 0.03 \$/kWh.

4.1.1.5 Summary of results (Al-Fujairah hybrid plant)

1. The SDI_{15} index of water after pretreatment ranges from 2 to 4.5, (while the SDI_5 for the feed water ranges from 12 to 19)
2. The turbidity of water after pretreatment ranges from 0.1 to 0.4 NTU. The installed pretreatment system provides 75% degree of rejection of turbidity.

3. The current study prove that 20 % (1.04 kWh/m^3(permeate)) of energy consumed by RO desalination is allocated in the stage of pretreatment. (Cost of energy allocated to water cost is estimated to be 0.03 $\$/m^3$(permeate).
4. The conventional pretreatment before RO is characterized by rather high level of chemicals consumption. (Cost of chemicals allocated to water cost are 0.03 $\$/m^3$(permeate).
5. The pretreatment studied can be characterized by the following techno-economic and operating drawbacks: (A) high rate of chemical consumption; (B) frequent backwashing and (C) difficulties in control and maintenance.

4.1.2 Conventional Pretreatment by Ondeo Ltd (Al -Tweelah pilot plant)

This section contains data on conventional pretreatment before RO pilot system proposed by the bidder "Degremont-Ondeo". The system was installed at Al–Tweelah site. Evaluation is based on data accumulated over the period from January 9, 2002 to October 30, 2002. It includes a set of the indicators specifying the water quality and the technological characteristics. The flow diagram of the pilot system by Ondeo-Degremont (capacity of 10 m^3/day) is shown in Figure 4-8. It contains the following main unit equipment: (1) Intake; (2) Settler; (3) Flotation Unit (Aqua-DAF™); (4) Dual media filters (two stages); (5) Cartridge filter.

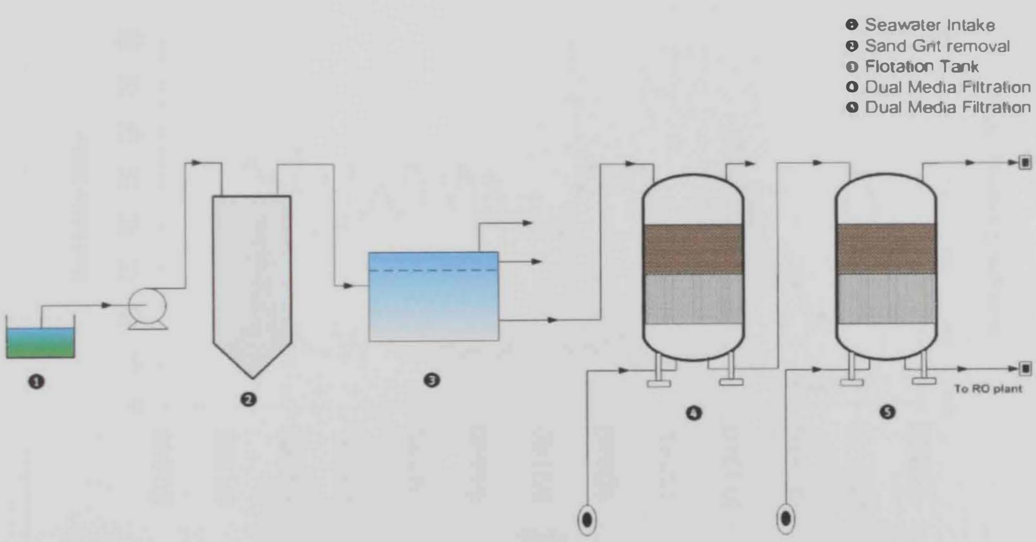

Figure 4-8: Schematic flow diagram of the Ondeo pilot plant

The flotation and the dual-media filtration systems are shown in Figure 4-9, and Figure 4-10, respectively. The indicators used for the evaluation of the pretreatment are the SDI index, turbidity, total suspended solids (TSS) and the degree of their rejection. As shown in Figure 4-11 the SDI index reading for the seawater ranges between 10 and 30 and the turbidity from 0.1 to 1 NTU.

Figure 4-9: Flotation system (Ondeo pilot plant)

Figure 4-10: Dual media filters (Ondeo pilot plant)

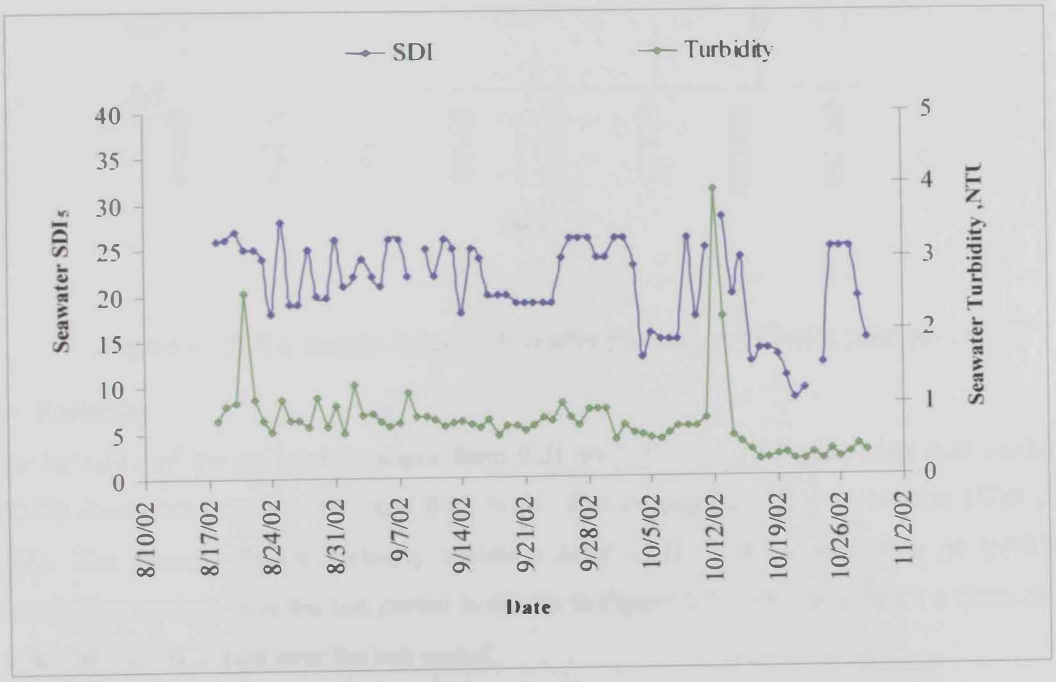

Figure 4-11: The SDI_{15} index and turbidity of seawater (Ondeo pilot plant)

- **Silt Density Index (SDI_{15})**

The average value of the SDI_{15} index after pretreatment is 2.34 as shown in Figure 4-12.The standard deviation 0.27. Experimental values of SDI_{15} after. The pretreatment varies from 1.5 to 3.1. This has been achieved by the efficient backwashing of the media filters. However, the data show high weekly variation. The small *SD* shows however a stable operation. True standard deviation, *SD*, for seawater and filtrate readings are equal to 4.9 and 0.27, respectively. Analysis of behavior of profile in Figure 4-12 indicates that the pretreatment system provides the required level of the SDI index. The SDI_{15} index after pretreatment is below 3.1 that meet required level of filtered water quality.

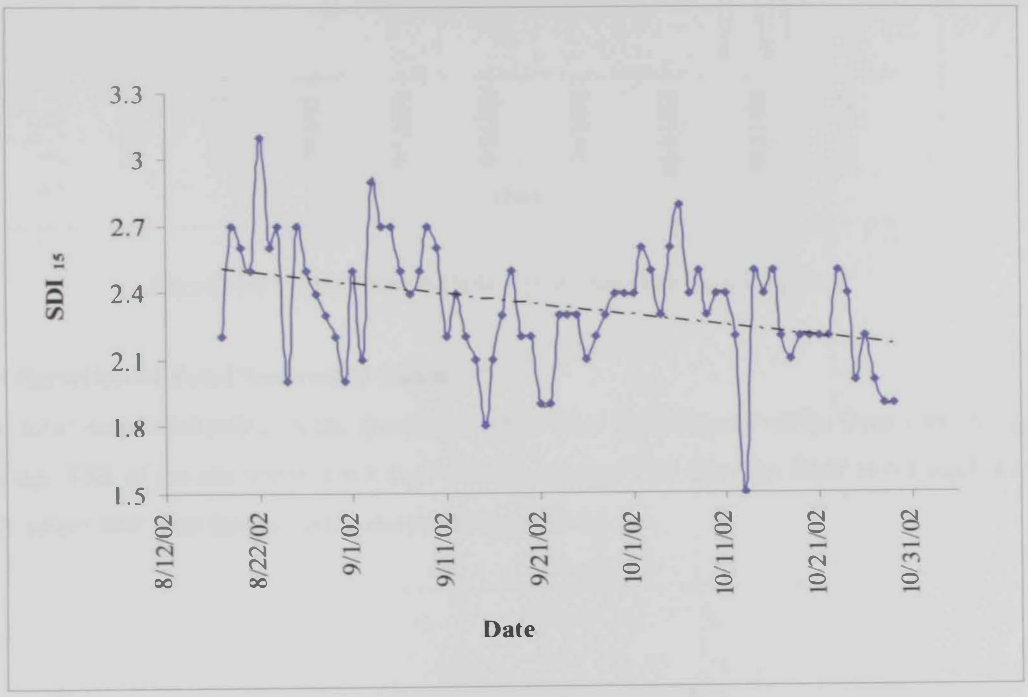

Figure 4-12: Silt density index (SDI_{15}) after pretreatment (Ondeo pilot plant)

- **Turbidity**

The turbidity of the raw water ranges from 0.21 to 3.92 NTU. Its value after dual media filter (DMF) fluctuates between 0.03 and 0.19 NTU. The average turbidity value after (DMF) is 0.1 NTU. The average degree turbidity rejection after DMF is 82%. Rejection of turbidity by pretreatment system over the test period is shown in Figure 4-13. The data shows a sharp decrease in the turbidity rejection over the test period.

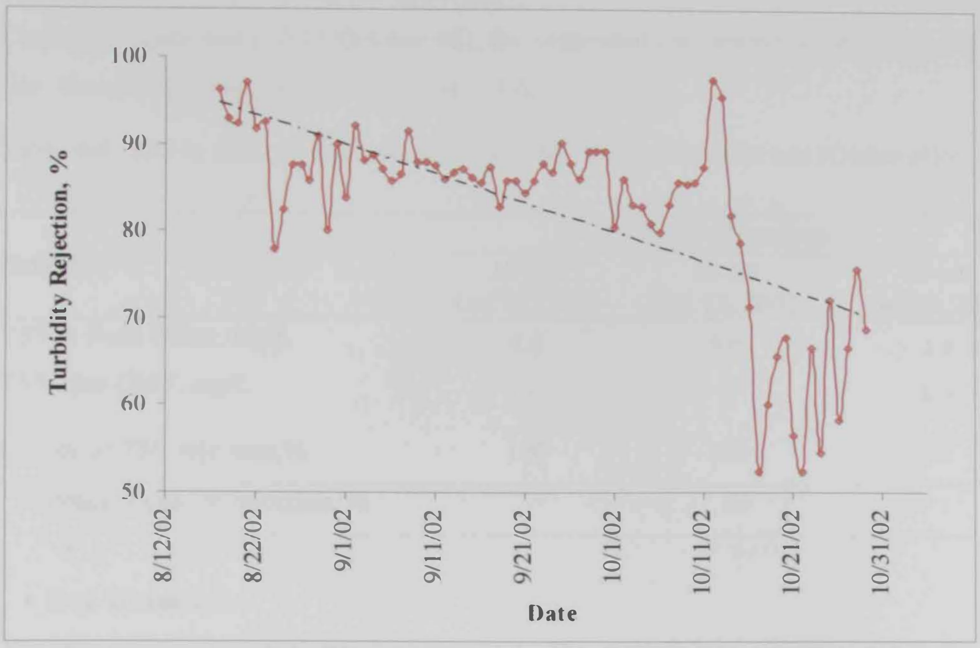

Figure 4-13: Degree of turbidity rejection (Ondeo pilot plant)

- **Rejection of Total Suspended Solids (TSS)**

The total suspended solids in the downflow water after pretreatment varies from 1 to 16 mg/l. The average TSS of the sea water is 9.3 mg/l and the average TSS after the DMF is 6.1 mg/l. Profile of TSS after DMF over test period is shown in Figure 4-14.

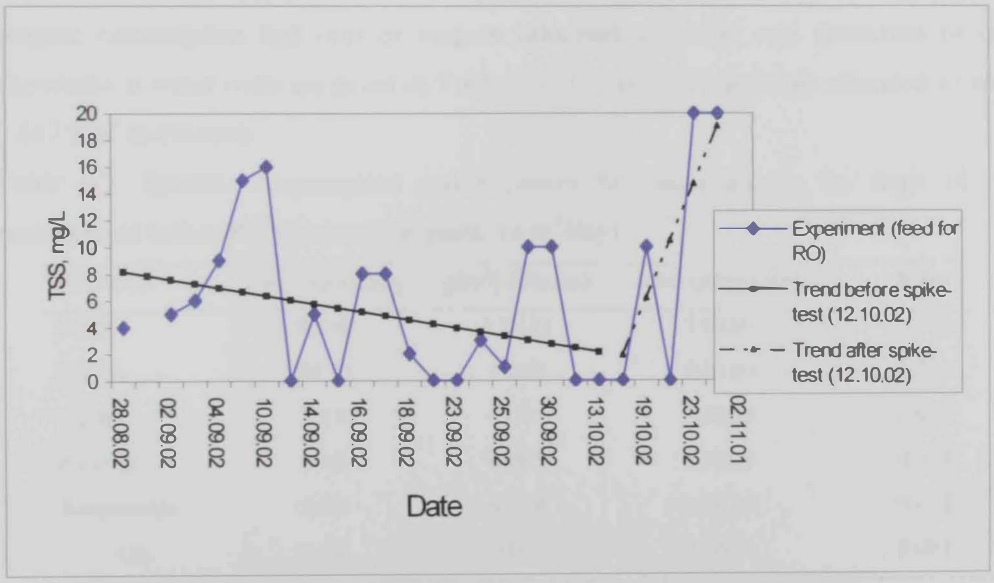

Figure 4-14: TSS after dual media filter (Ondeo pilot plant)

- **Rejection of TSS during the spike test**

During the spike test (12-13 October 02), the suspended contamination were injected into the feed line. Results of the test are shown in Table 4-6.

Table 4-6: TSS in the feed flow and after the DMF during the spike test (Ondeo pilot plant)

Indicator	Time of readings		
	10 am Oct 13, 2002	10 am Oct 13, 2002	10 am Oct 13, 2002
TSS in Feed Water, mg/L	8.0	9.0	1.9
TSS after DMF, mg/L	0	0	0.9
Degree of TSS rejection,%	100	100	52
Average degree of rejection, %		84	

- **Iron Reduction**

The iron analysis was performed once a week. The average iron concentration in the seawater was 55 μg/l and the average concentration in the filtered water was 35 μg/l.

4.1.2.1 Consumption of Chemicals in the Stage of Conventional Pretreatment before RO (Ondeo pilot plant)

The following chemicals were used: (1) ferric chloride, $FeCl_3$, (2) sulfuric acid, H_2SO_4, (3) sodium hypochlorite, NaClO, (4) antiscalant, and (5) sodium bisulphite, Na_2SO_3. The data on specific reagent consumption and cost of reagent allocated to water cost (structure of expenses for chemicals in water cost) are given in Table 4-7. The cost of chemicals allocated to water cost are 0.047 $\$/m^3$ (permeate).

Table 4-7: Specific consumption and expenses for chemicals in the stage of conventional pretreatment before RO (Ondeo pilot plant, 10 m^3/day)

Chemical	g/m^3(permeate)	g/m^3(Filtrate)	$\$/m^3$ (permeate)	\$/day
$FeCl_3$	96.58	1.0431	0.0260	0.314
H_2SO_4	98.33	1.062	0.0180	0.213
NaOH	2.11	0.023	0.0003	0.003
Polymer	0.45	0.005	0.0009	0.011
Antiscalant	0.77	0.008	0.0015	0.018
Cl_2	0.18	0.002	0.0001	0.001
		Total	**0.0470**	**0.560**

4.1.2.2 Summary of results (Ondeo pilot plant)

1. The SDI_{15} index of water after pretreatment ranged from 1.5 to 3.1 and the average SDI index was 2.34.
2. The turbidity of the raw water varies from 0.21 to 3.92 NTU, while the average value of turbidity after DMF is 0.1 NTU. The installed pretreatment system provides 82% degree of rejection of turbidity.
3. The average TSS of seawater was 9.3 mg/l, and after DMF became 6.1 mg/l. The degree of rejection of TSS during the spike test was found to be 84%.
4. The average concentration of iron in seawater was 55 μg/l and the average value in filtered water was 35 μg/l.
5. The conventional pretreatment before RO is characterized by high chemical consumption. (Cost of chemicals allocated to water cost was 0.047 $/m^3 (permeate)).

4.2 Membrane-based pretreatment before RO

This section focuses on the new generation of pre-treatment before reverse osmosis, namely membrane-based pretreatment where micro-or ultra-filtration are used instead of coagulation and multimedia filtration. The following pretreatment schemes were considered in the current study: (1) Membrane–based pretreatment based on the "Aquasource" system located at Al–Taweelah site (2) Membrane–based pretreatment based on the "Zenon" system located at Al–Taweelah site, (3) hybrid type of pretreatment proposed by GrahamTech Pte Ltd (Singapore) located at Bainouna site, and (4) membrane-based pretreatment implemented after reconstruction in RO desalination plant in Addur (Bahrain).

4.2.1 Membrane- based pretreatment by Aquasource Ltd. (Al -Taweelah Pilot Plant)

The Aquasource UF unit at Al-Taweelah pilot plantis comprised of two hollow fiber UF membrane modules mounted on a transportable skid. The skid is constructed of reinforced fiberglass and steel, and can be shipped by truck. The unit is self-contained, including all the components required for operation. It is connected to raw water, drain lines, backwash water and electrical power. The unit requires 2.8 m^2 of floor space. The Aquasource UF unit has two alternating operating regimes. Filtration and backwashing. During the backwash, the feed pump shuts down, valves are repositioned, and the backwash pump starts. The backwash pump draws treated water from the filtrate storage tank, chlorinates it, and forces the water under pressure in the reverse direction through the fibers. With the flow of water now from the outside to inside the fiber, the backwash water exits the inside of the fibers at the fiber ends, carrying with it particulate material accumulated during filtration. Table 4-8 presents the main characteristics of the Aquasource membranes.

Table 4-8: Characteristics of the Aquasource membrane

Driving force, bar	0.2-0.8
Flux, m^3/h	2.35
Operation mode	Crossflow
Number of fibers per module	2060
Flow direction	From outside to inside
Nominal membrane pore size, μm	0.01
Nominal molecular weight cutoff, dalton	~100,000
Membrane material	Cellulose acetate derivatives

Analysis of the system is based on data accumulated over the period from August 31 to October 31, 2002. It includes a set of indicators specifying water quality and technological characteristics such as SDI index, turbidity, transmembrane pressure drop, specific permeability, normalized permeability deterioration (NPD) and specific energy consumption.

- **Silt density index (SDI)**

The SDI_{15} index of the water after pretreatment ranges from 0.3 to 2.8. (Average value is 1.55). This value range satisfies the quality of feedwater required for RO, (For RO feed the $SDI_{15}<5$). The experimental readings of the SDI_{15} index after pretreatment are shown in Figure 4-15. The standard deviation between experimental readings and values estimated by linear approximation is 0.13. The increasing experimental trend of the SDI can be explained by the deterioration of the membrane characteristics by time.

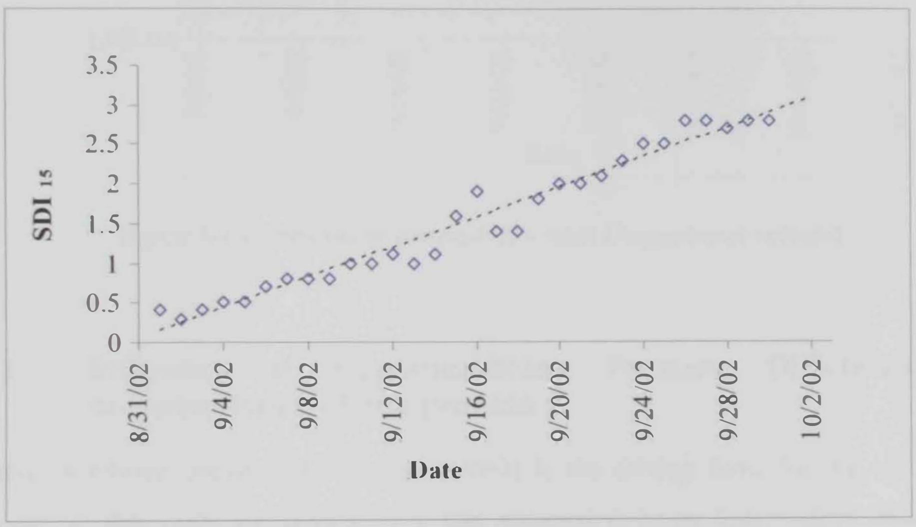

Figure 4-15: SDI_{15} index of filtrate water (Aquasource system)

- **Normalized permeability deterioration**

For the influence of viscosity on permeability of membrane to be excluded, the permeability observed at operating temperature has to be multiplied by the correction factor being equal to the ratio of viscosity at operating temperature to viscosity at the reference temperature (t_0= 25°C). The variation of permeability at operating temperature and its normalized values are shown in. The decline of permeability can be explained by the decrease of water viscosity or growth of membrane resistance or by both factors simultaneously (See Figure 4-16)

The permeability of the UF membranes declined by 41.7% over 42 days of the test period, while the normalized permeability (at t = 25°C) declined by 38.5%, that corresponds to 0.9 % per daily deterioration. According to Avlonitis *et al.* [108], the membrane should be cleaned whenever the normalized permeate flow drops to 10% or the differential pressure increased by 15% from the reference value recommended by the manufacturer.

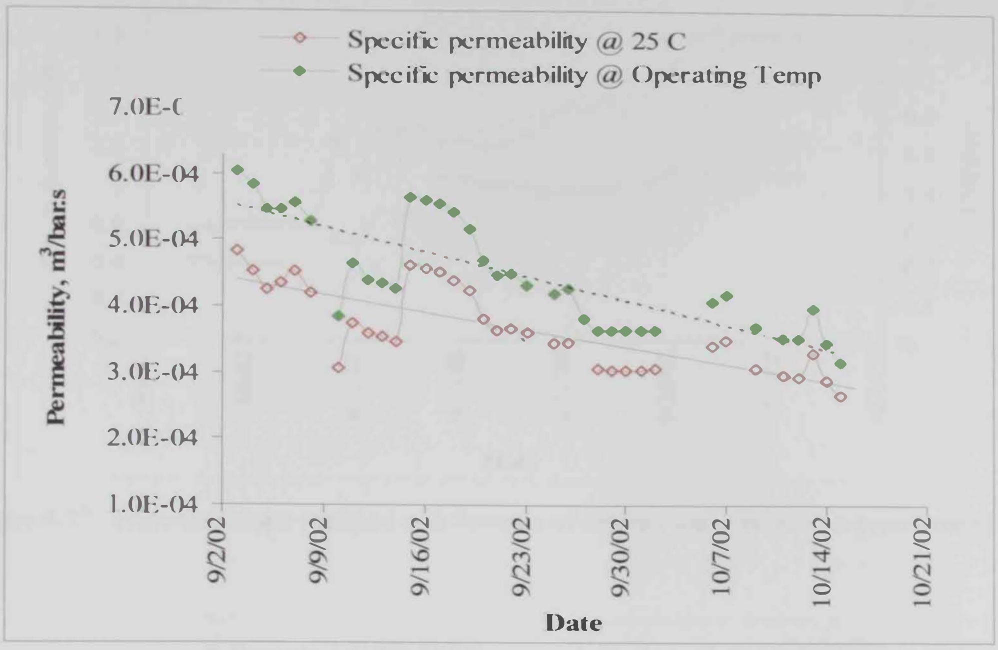

Figure 4-16: Membrane permeability rates (Aquasource system)

4.2.1.1 Influence of Transmembrane Pressure Difference on the characteristics of the process

The transmembrane pressure difference (TMPD) is the driving force for the ultra- and micro-filtrations. In this study the driving force was assumed to be an independent variable. The SDI index and permeability are influenced by this driving force (TMPD). To mathematically formulate the quantitative relation between them, the experimental profiles were expressed in terms of the driving force, namely, in functional forms such as: $A_t = f(\Delta P)$ and $SDI_{15} = f(\Delta P)$. Figure 4-17 presents the filtrate flowrate and transmembrane pressure difference during the test period. The experimental projections of driving forces and the SDI index vs. TMPD over two time periods are shown in Figure 4-18.

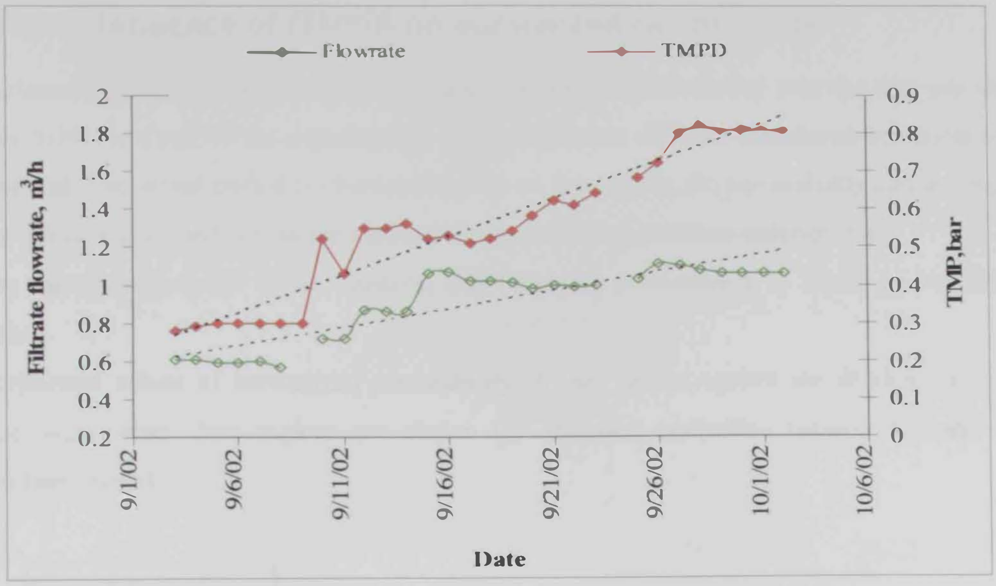

Figure 4-17: Transmembrane pressure and flowrate of filtered water profile (Aquasource system)

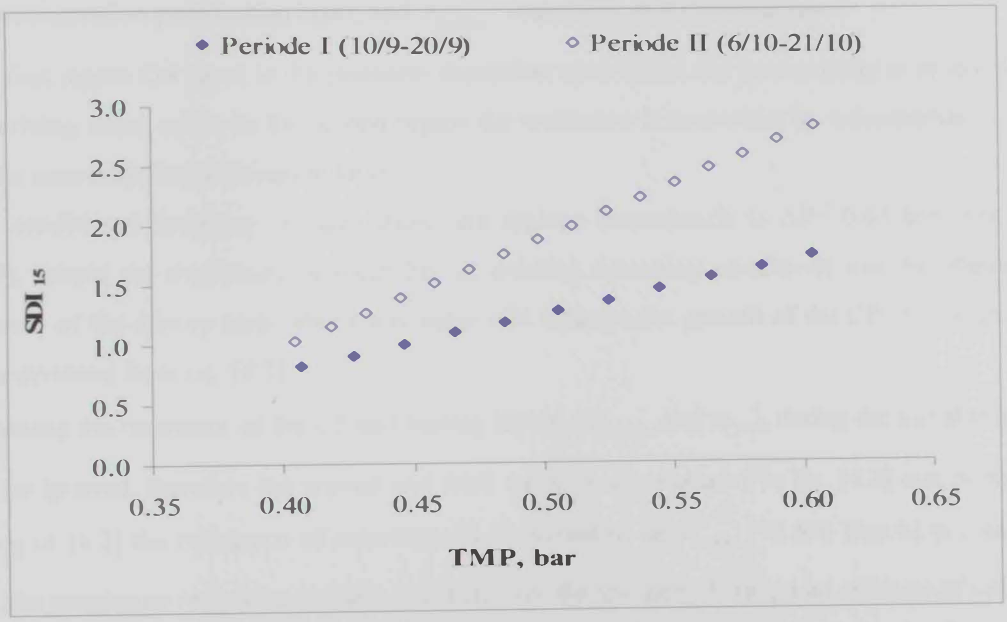

Figure 4-18: The SDI index of filtrate vs. driving force (Aquasource system)

4.2.1.2 Influence of (TMPD) on normalized permeability

Experimental profiles of the driving forces and normalized permeability over the time are shown in Figure 4-19. Analysis of the experimental profile indicates different functional behaviors over the test period. The initial period is characterized by an increase in the permeability and driving force, while during the second period the permeability is becoming pressure-independent.

Using the resistance- in- series concept, estimation of permeability is based on the following relation:

Experimental values of normalized permeability, A, are shown against the driving force, ΔP, in Figure 4-20, where two regions are shown (A: Pressure controlled region; B: mass transfer controlled region)

$$A_t = \frac{\Delta P}{\left(r_{membr} + r_{CP} + r_{fouling}\right)} \qquad [4\text{-}2]$$

Where A: normalized permeability; ΔP: driving force; resistance of membrane; r_{CP}: resistance of the concentration polarization layer, and $r_{fouling}$: resistance of the fouling layer.

The first region (24 days) is the pressure- dependent one, where the permeability is proportional to the driving force, while in the second region the resistance is controlled by mass-transfer, (namely by the concentration polarization layer).

The conditional boundary between these two regions corresponds to ΔP= 0.65 bar (See Figure 4-20), where the maximum permeability at existing operating conditions can be attained. An increase of the driving force above this value will enhance the growth of the CP resistance, as can be understood from Eq. [4.2]

Assuming the resistance of the CP and fouling layers, ($r_{fouling}$ and r_{CP}), during the initial test period can be ignored, therefore the second and third terms in denominator in Eq. [4.2] can be omitted). Using of [4.2] the resistance of membrane is estimated to be r_{membr} = 0.686 [bar-h]/m^3. Assuming that the membrane resistance remains constant over the test period we get an estimate of sum of the CP and fouling resistance. Inserting obtained value of the membrane resistance into Eq. [4.2], the CP and fouling resistances is estimated to be 0.236 [bar-h]/m^3, (that corresponds to 34% of hydraulic resistance of membrane matrix itself).

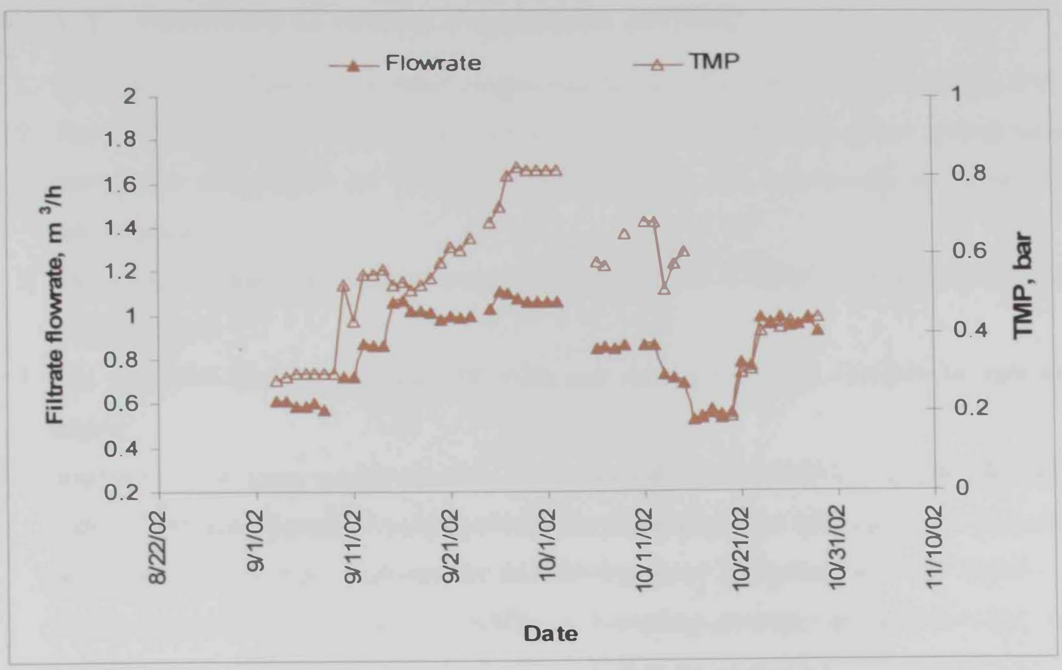

Figure 4-19: Filtrate flux and transmembrane pressure vs. time –Period I & II (Aquasource system)

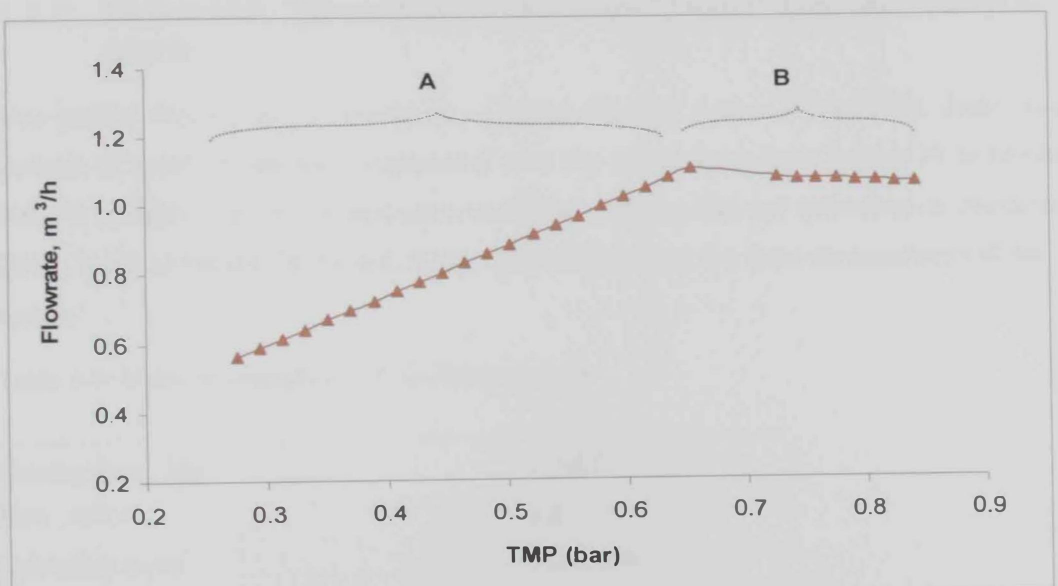

Figure 4-20: Normalized permeability vs. driving force (A- Pressure controlled region; B- mass transfer controlled region)

4.2.1.3 Summary of results (Aquasurce system)

1. The SDI index (of pretreated water) ranges from 0.3 to 2.8 and the averaged value is 1.55.
2. The permeability of UF membrane declined by 41.7% over 42 days of test period, while the normalized permeability (at 25°C) declined to 38.5%, that corresponds to 0.9 % of daily deterioration.
3. The energy consumption ranges between 0.006 to 0.025 kWh/m^3 with an average value of 0.016 kWh/m^3
4. The proportionality between the SDI index and driving force was found to be over the test period.
5. Analysis of the experimental profiles of the normalized permeability and the driving force indicate different functional behaviors over the test period. The initial period is characterized by an increase in both permeability and driving force (pressure-controlled region); while during the second period the permeability is becoming pressure-independent (mass transfer controlled regions). Transition from the first region to the second takes place at $\Delta P = 0.65$ bar.

4.2.2 Membrane- based Pretreatment by "Zenon" Ltd. (Al -Tawelah pilot plant)

This section focuses on the evaluation of micro filtration system proposed by Zenon Ltd. The analysis is based on the data accumulated over the period from from August 31 to October 31, 2002. It includes a set of the indicators specifying water quality and technological characteristics: SDI $_{15}$ index of filtrate TMPD and NDP. Table 4-9 presents the main characteristics of the Zenon system

Table 4-9: Main characteristics of the Zenon system

Driving force, bar	0.2-0.5
Flux , m^3/h	4.8
Operation mode	Dead-end
Number of fibers per module	4700
Flow Direction	From the outside
Nominal membrane Pore size, μm	0.035
Nominal Molecular Weight Cutoff, Dalton	~100,000
Membrane material	Proprietary Polymer

- **Silt Density Index (SDI_{15})**

The SDI index of filtered water produced by the Zenon system ranges from 2.3 to 3.4. The maximum reading (3.4) was higher than the required level. The averaged value of the SDI_{15} index after pretreatment was 2.9. Experimental values of the SDI_{15} index (for seawater and water after pretreatment) are shown in Figure 4-21. The true standard deviation, SD, for filtrate readings was equal to 0.31.

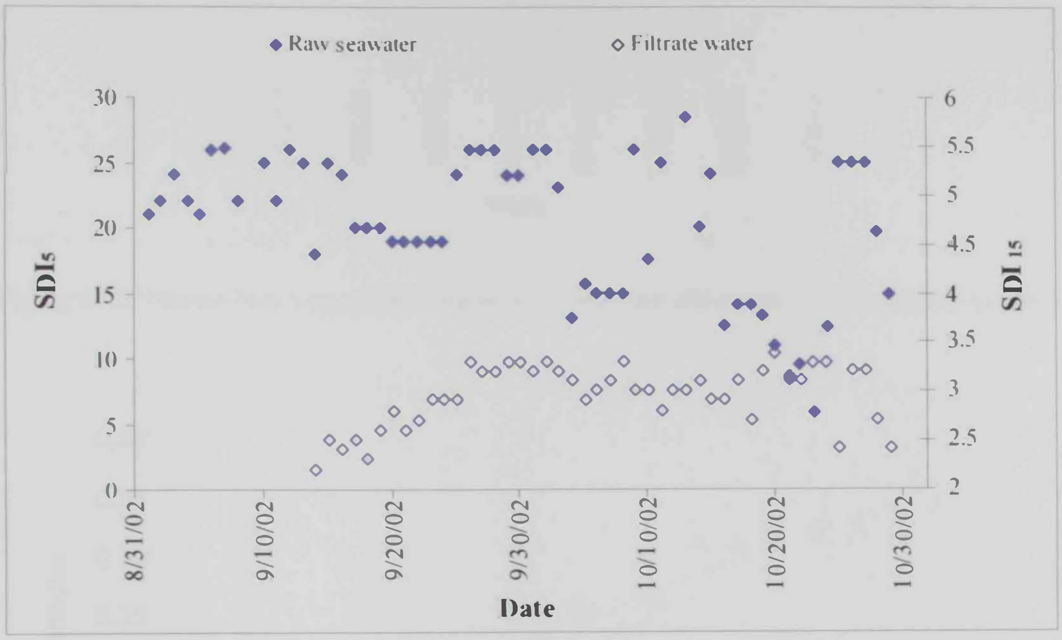

Figure 4-21: The SDI_5 index of water before and after pretreatment (Zenon system)

- **Transmembrane Pressure Difference (TMPD)**

The TMP and filtrate flowrate profile over the period (14/9- 31/10/2002) is shown in Figure 4-22. The profile of the driving force (TMPD) over the (13/9-3/10/2002) period of the pilot test is shown in Figure 4-23. True standard deviation, SD, for transmembrane pressure difference was 0.18.

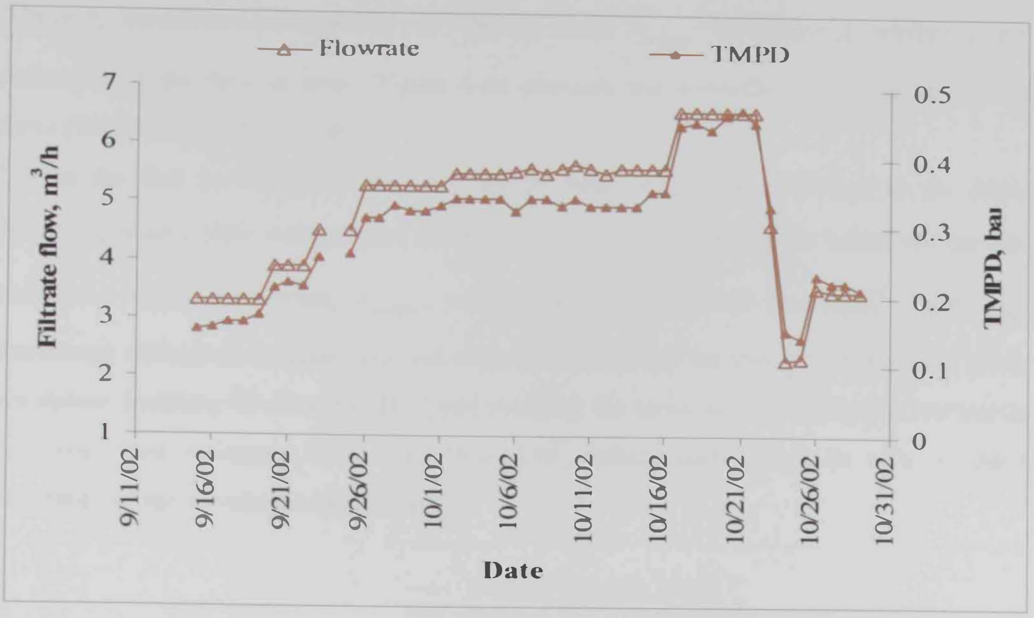

Figure 4-22: Filtrate Flowrate and transmembrane pressure difference profile (Zenon system)

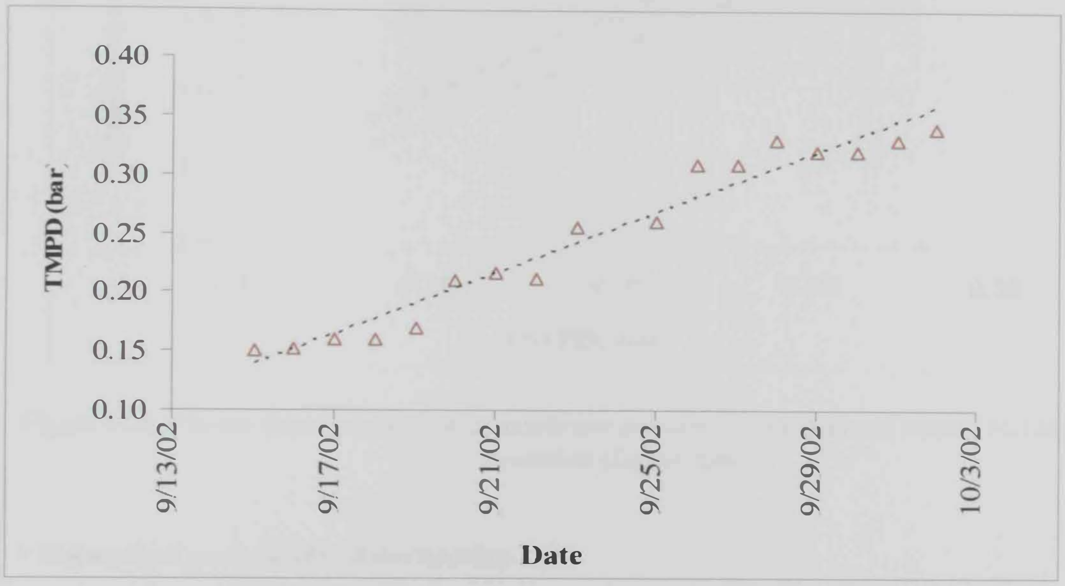

Figure 4-23: Transmembrane pressure drop over the period from 13/9 to 3/10 (Zenon system)

- **Influence of the TMPD on normalized permeability**

Using the resistance in series concept. Estimation of permeability is based on the following relation:

$$A_{t=25} = \frac{\Delta P}{\left(r_{membr} + r_{fouling}\right)} \qquad [4\text{-}3]$$

Where A: normalized permeability; ΔP: driving force; r_{membr} : resistance of membrane and $r_{fouling}$: resistance of the fouling layer. Figure 4-24 presents the normalized flowrate versus the driving force (TMPD) for both periods.

During the first period (16/09 to 1/10) the permeability was proportional to the driving force. Assuming a negligible resistance of the fouling layer, $r_{fouling}$, during the initial test period .then the resistance of the membrane, r_{membr}, was estimated to be 0.06 [bar-h]/m^3. Assuming that the membrane resistance remains constant over the test period we can get an estimate of the fouling resistance. Inserting this into Eq. [4.3] and applying the same assumption applied before the fouling resistance was estimated to be 0.01 [bar.h]/m^3, (which corresponds to 11% of the hydraulic resistance of the membrane matrix itself).

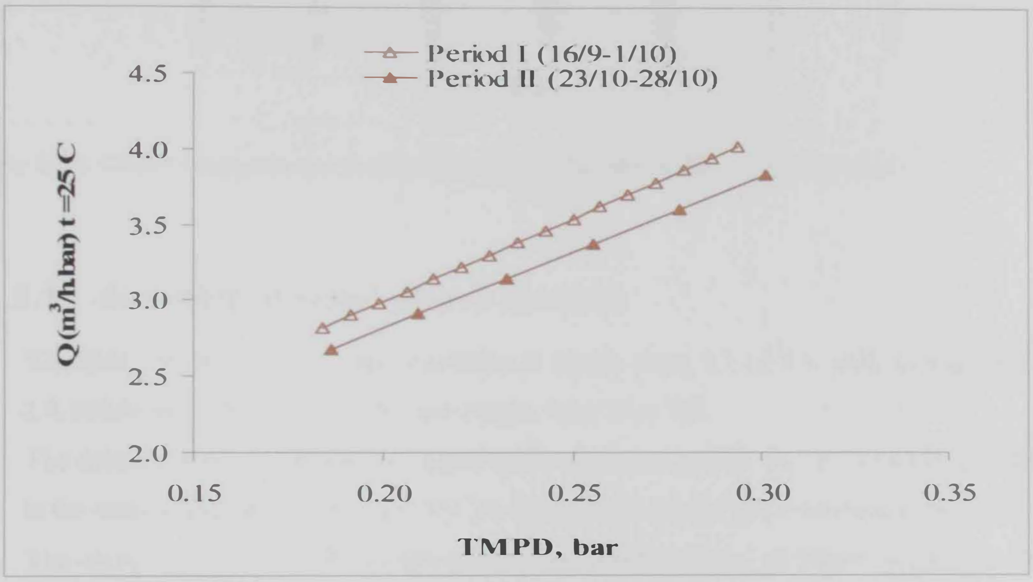

Figure 4-24: Filtrate flowrate versus transmembrane pressure during first and second periods of operation (Zenon system)

- **Normalized permeability deterioration**

The permeability of the UF membrane declined by 24% over the 44 days of the test period, while the normalized permeability (at t = 25°C) declined by 20%. This level of deterioration of transport characteristics indicates the necessity of regeneration of the UF membranes (Figure 4-25)

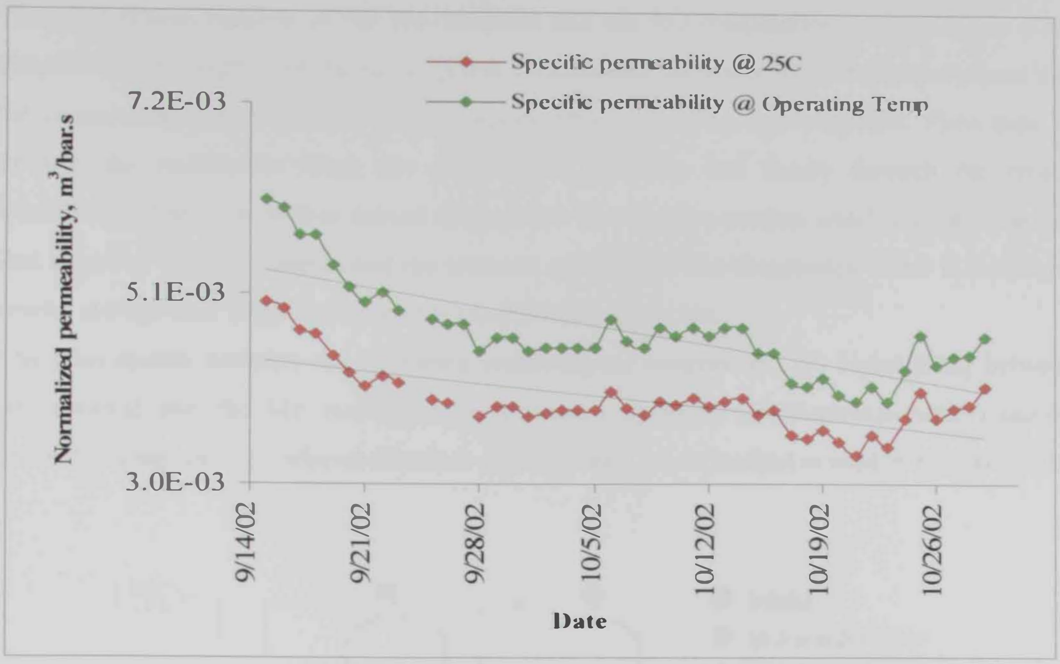

Figure 4-25: Observed and normalized values of specific permeability (Zenon system)

4.2.2.1 Summary of result (Zenon system)

1. The SDI_{15} index of water after pretreatment ranges from 2.3 to 3.4, with average value of 2.9, (while the SDI_5 for the feed water ranges from 20 to 30).
2. The deterioration of observed permeability over the test period was 24 % (while the decline in the normalized permeability (at 25°C) over the same period was estimated to be 20%).
3. The energy consumed by the Zenon systems was 0.008 kWh/m^3 of filtrate while it was 0.016 kWh/m^3 for the Aquasource system.
4. Function behaviour of permeability is driving force remains unchangeable over the test period (permeability is proportional by driving force)

4.2.3 Membrane-based (hybrid) pretreatment by "GrahamTek Pte (Singapore) Ltd."

This section covers the pre-treatment system for the reverse osmosis pilot systems proposed by "GrahamTek Pte Ltd" (Singapore). The system was located at Bainounah Power Company in Abu Dhabi and was operated during the period from May 10 to June 23, 2004 (about 1000 hours of continuous operation).

The pilot system consists of the pre-treatment and the RO desalination stages. Figure 4-26 is a simplified flow diagram of the pilot system. Chlorinated sea water from the intake passes through the coagulation chamber, where ferric chloride ($FeCl_3$) is used as a coagulant. Flow then passes through the multimedia filter, the polishing sand filter, and finally through the cross-flow microfilters. The flow is then passed through the desalination section which includes the 1st and 2nd stages of reverse osmosis and the pressure exchanger. The desalinated water is then sent to a nearby storage tank while the brine reject is drained back to sea.

The pilot system contains the following technological innovations: (1) Hybridizing between the conventional and the MF membrane pre-treatment systems, (2) Electromagnetic treatment to prevent scaling and (3) Reduced chemical consumption (no antiscalant is used in this process).

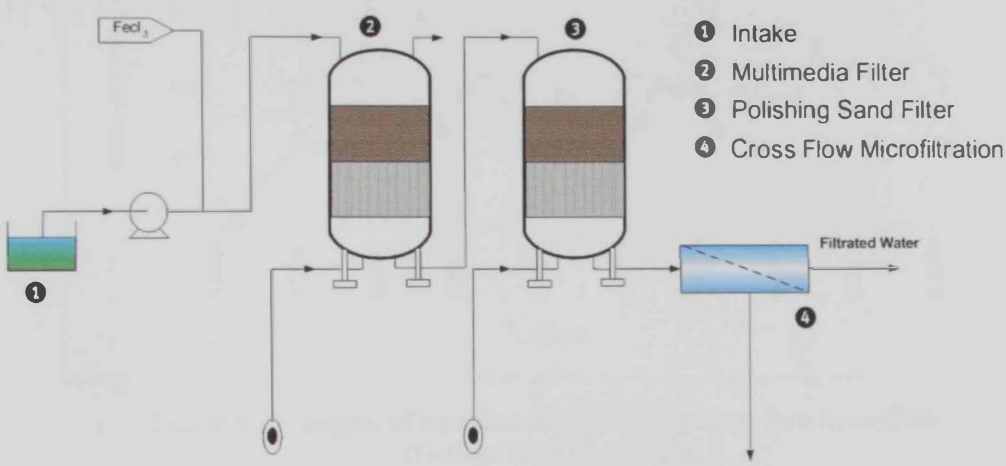

Figure 4-26: Schematic diagram of the GrahamTek Ltd pilot system

4.2.3.1 Performance Evaluation

This section focuses on the analysis of the last stage of the pre-treatment, namely, cross- flow microfiltration (MF), where membrane elements of spiral type with polysulfone membranes are used as the microfilter and media the total surface area is 418 m^2. According to the manufacturer requirements, the SDI_{15} index of raw water entering the microfilter should be less than 5. It is worth to emphasize that the SDI_{15} <5, could be attained only after passing through the coagulation chambers, multimedia and sand filters. The SDI_{15} index of the untreated seawater is site-specific and season-dependent; and has a much higher value and can reach 15-25. The efficiency of the pretreatment was characterized by the degree of variation of nephelometric characteristics such as turbidity and the SDI index.

- **Turbidity**

The average turbidity ranges from 0.344 NTU in raw water to 0.285 NTU in the down flow after the MF. (The maximum detected values in the MF feed and product were 1.6 and 0.4 NTU, respectively. The standard deviation, SD, for the seawater and filtrate readings is 0.05. The average degree of turbidity rejection was 19%. The profile of turbidity degree rejection was shown in Figure 4-27.

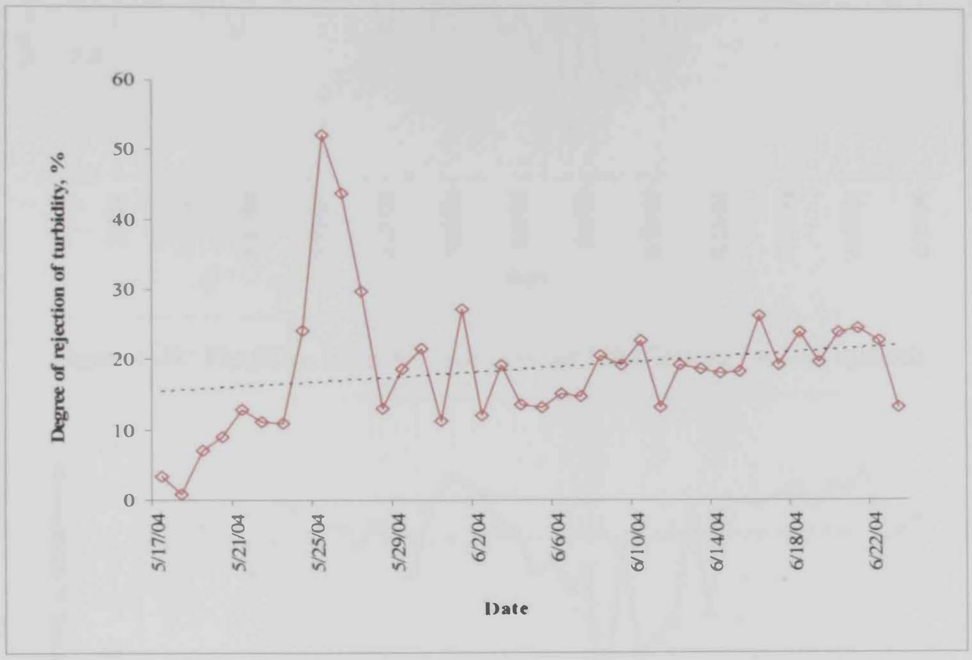

Figure 4-27: Degree of rejection of turbidity by cross flow microfilter (GrahamTek Ltd. system)

- **Silt Density Index**

The average SDI_{15} after microfilter (MF) was less than 1. The averaged SDI_{15} index after the polishing sand filter (in feed flow for MF) was 2.79. Experimental values of SDI_{15} after the polishing sand filter (feed flow for MF) are shown in Figure 4-28. The Degree of the SDI_{15} rejection is one of the efficiency indicators; it is shown in Figure 4-29.

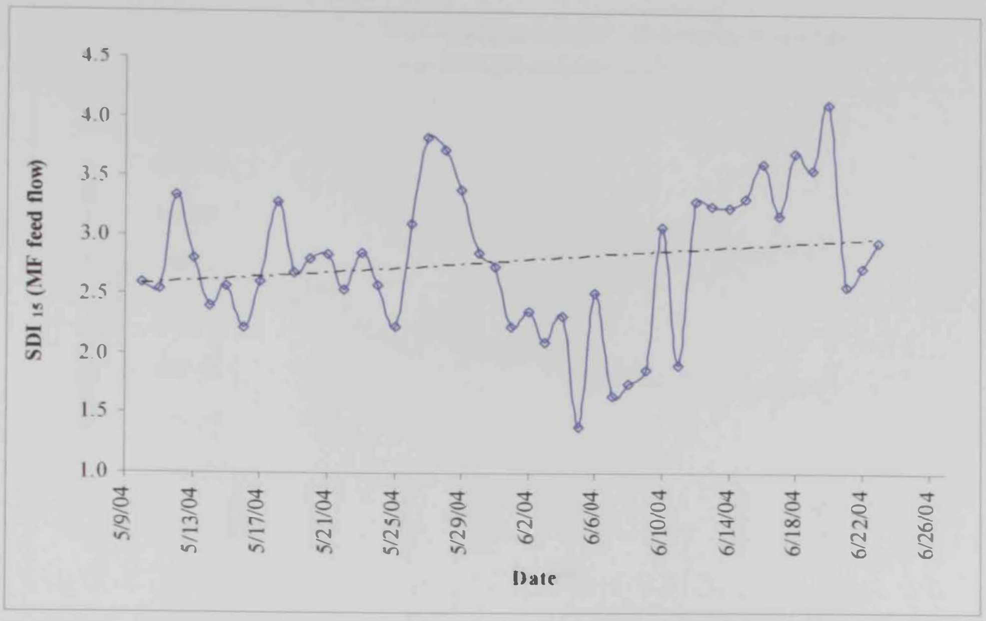

Figure 4-28: The SDI_{15} index for feed flow for MF (GrahamTek Ltd. system)

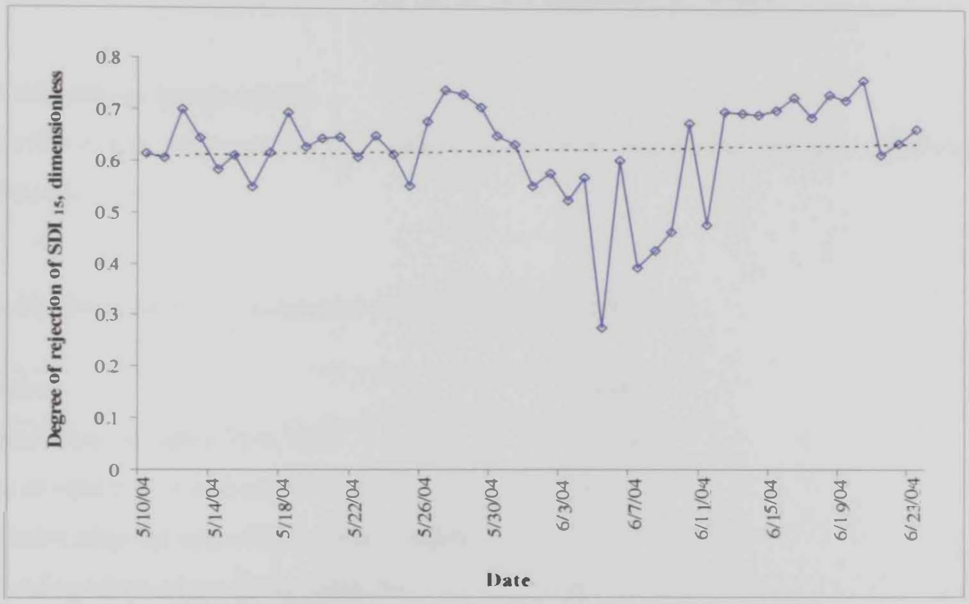

Figure 4-29: Degree of rejection of the SDI_{15} by cross-flow microfilter (GrahamTek Ltd. system)

- **Normalized permeability deterioration**

The pilot test was conducted without chemical cleaning of the MF membranes. (Air backwashing was applied for regeneration of permeability, namely, 90 seconds of air backwashing per one hour of microfiltration). At these conditions the specific permeability of the membranes declined by 9% (over the 45 days of test). The MF permeability normalized at 25°C indicated 12% decline over 45 days. See Figure 4-30. This level of deterioration of transport characteristics indicates the necessity of periodical chemical cleaning of the MF membranes.

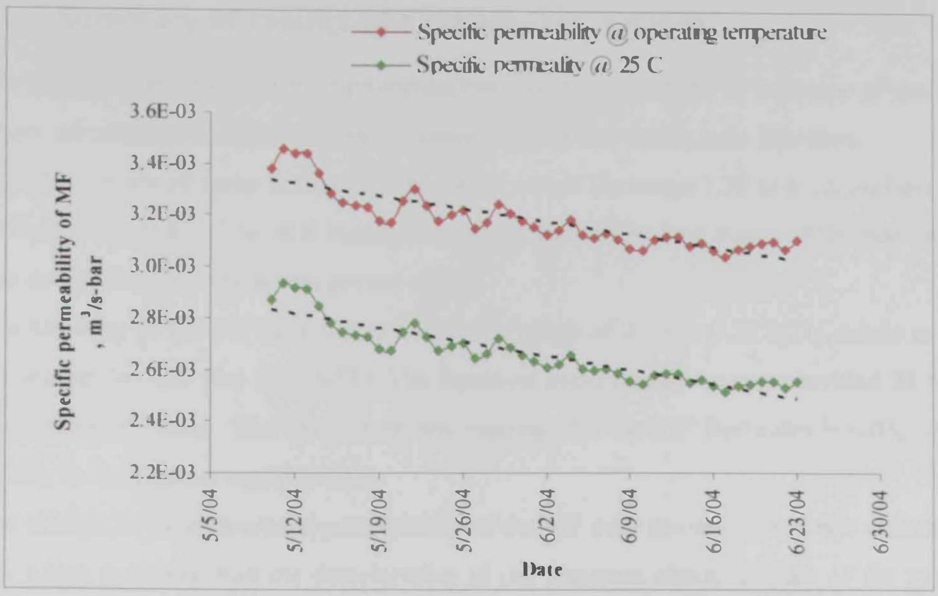

Figure 4-30: Observed and normalized values of specific permeability of cross-flow micro filtration pretreatment before the RO (GrahamTek Ltd. System)

- **Specific energy consumption**

The specific energy consumption for filtrated water produced by the MF was calculated to be 0.138. See Eq. [3-6]

Table 4-10: Indicators for Assessment of Crossflow Microfiltration

Indicator	Value*
Turbidity after the microfilter, NTU	0.1676
Degree of rejection of turbidity, %	28
SDI_{15}-index after the microfilter, dimensionless	<1
Degree of rejection of the SDI_{15} index, %	61
Specific permeability of microfiltration membranes, (atoperating temperature), m^3/m^2-s-bar	7.86E-06
Specific permeability of microfiltration membranes, (normalized to 25°C), m^3/m^2-s-bar	6.57E-06

*Values averaged over the test period

4.2.3.2 Summary of results of Graham-Tek system

1. The pilot system proposed by the GrahamTek Ltd. represents the hybrid type of pretreatment where the crossflow microfiltration is combined with the multimedia filtration.
2. The SDI_{15} index of water before micofiltration was in the range 1.38 to 4.12, and the standard deviation was 0.65. The SDI index after pretreatment was less than 1. SDI index of filtrate was acceptable over the whole period of test.
3. The turbidity of the MF feed water was in the range of 0.14 to 0.33 NTU, while its average value after the MF was 0.16 NTU. The installed pretreatment system provided 28 % degree of turbidity rejection. The SD of turbidity readings for the MF feedwater is 0.05, which was similar to the product reading value.
4. The decline in the normalized permeability of the MF over the test period was estimated to be 8% which indicates that the deterioration of the transport characteristics of the membranes and point out to the necessity of periodical regeneration using chemicals (No chemical cleaning was applied over the test period).
5. The average of specific energy consumption in the stage of the MF membrane was estimated to be 0.138 kWh/m^3 of filtrate water.

4.2.4 Addur Seawater Reverse osmosis desalination plant

This section focuses on the comparison between conventional and membrane-based pretreatment before the RO is applied in Addur seawater RO desalination plant (Bahrain). The Addur seawater RO desalination plant is one of four main desalination plants used in Kingdom of Bahrain. The other are Abu Jarjur (10 MIGD, brackish water), Sitra power and water station (25 MIGD,) and Hidd power and water plant (30 MIGD) [93]. The Addur reverse osmosis desalination plant was commissioned in 1990 and designed to produce 10 MIGD of desalinated water. The initial design of the pretreatment system was based on conventional scheme that include media and cartridge filters, (see Figure 4-31). The seawater is chlorinated by sodium hypochlorite solution, Sulfuric acid (25 mg/l) to reduce the pH, ferric chloride coagulant, (2.5 mg/l) and a coagulant aid (cationic polymer, 0.2 mg/l) are all injected into the seawater. The coagulated flocks are filtered off in the dual media filter (DMF) beds followed by filtration through micron guard filters and dechlorination by sodium bisulphate. The pretreated water is pressurized to 69 bar by a high pressure centrifugal pump and fed to the first pass of the RO train of membranes where the desalination takes place.

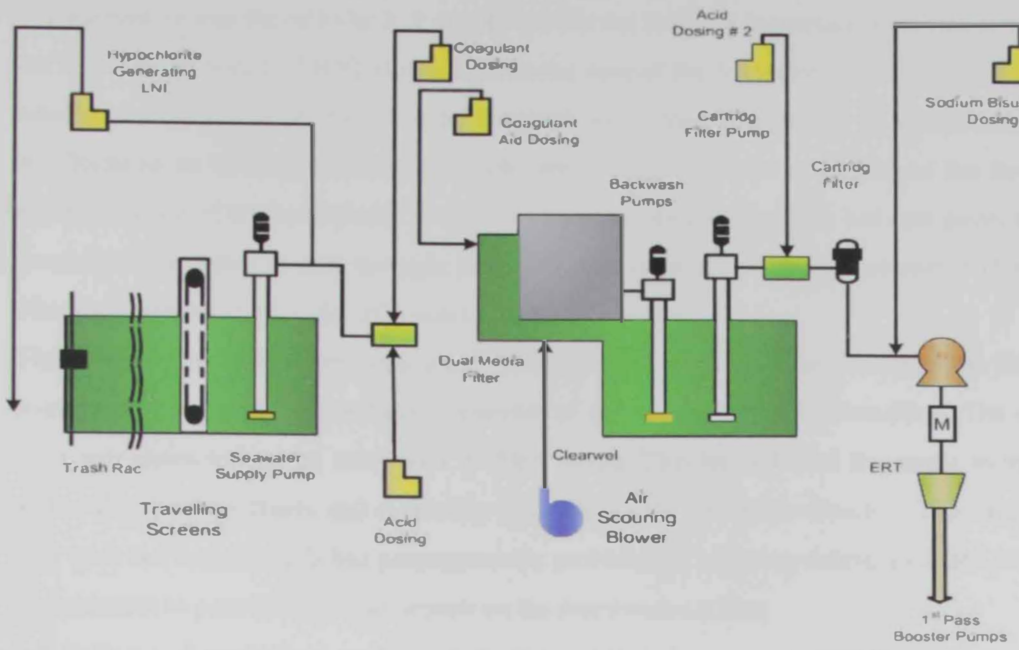

Figure 4-31: Simplified flow diagram of initial design of the Addur SWRO desalination plant [109]

By April 1992, the performance characteristics of the plant such as permeability, salt rejection and hydraulic resistance were observed deteriorating. The filtered water after the dual media filters did not achieve the required value of the SDI index (the design value of the SDI index after pretreatment was supposed to be ≤ 2.7). It was accompanied by side effects such as membrane degradation and secondary biological fouling caused by high cleaning frequency and dechlorination. According to Burashid and Hussain [109], the performance deterioration was caused by improper design and operating failures such as improper site selection; insufficient pretreatment, inefficient cleaning system, improper material selection. Some of these factors are considered below.

- **Site location**

The Addur SWRO plant is located in the southern part of Bahrain and the seawater intake point is located in the Addur Gulf of Addur. The area is characterized by high concentrations of organics and bioactive components. The intake is located 1.2 km offshore and at a depth of 3 m below water surface and downstream industrial and residential waste disposal. The elevated temperature of the water there makes favorable conditions for microorganisms to grow.

- **Insufficient pretreatment**

The pretreatment used was characterized by some regime failures such as the insufficient time for finalization of coagulation-flocculation operation. The time interval between injections of sulfuric acid, addition of ferric chloride coagulant, injection of coagulant aid and the cationic polyelectrolyte was found to be 2–3 minutes, while the required theoretical time was supposed to be not less then 10 minutes [109]. But in a particular case of the Addur system, it needs take at least 20 minutes for the flocks formation to be finalized. Since the experimental time was insufficient for the flocks to be formed, non-uniform flocks are formed. The non-uniformity of the flocks caused malfunctioning of the multimedia filter, namely, channeling in the filter bed that paves the way for the suspended matter to pass through, which, in turn, deteriorates the nephelometric characteristics after pretreatment such as the SDI index.

The design of the DMF filters with drain channels being constructed at the side of the filter bed had contributed greatly in the inefficient removal of the flocks during backwashing. The dual media filter was characterized by carry over of filter media. This had allowed the media to leak through and to the cartridge filters and eventually ended up in the membrane bundles. The cartridge filters also were not leak-proof. It has propagated the problems by allowing debris, sand and iron from the pretreatment to pass through and deposit on the membrane surface.

- **Failures of equipment or the system design due to improper material selection**

The initial design of the installed system was characterized by some failures. In particular, the distance between the RO trains and the point of dechlorination was 30 m while was considered to be unjustifiably long and used to create favorable conditions for microbiological growth. The next disadvantage the system was characterized by was improper selection and design of some pieces of equipment. In particular, characteristics and rated capacity of installed equipment were not in line with the specification requirements (Some of them others were characterized by lower capacity; some were overrated). The last point was improper material selection.

In 1993 the RO train feed header inspection revealed that the internal surfaces of the headers were coated with dark-brownish slime with a thickness of about 3-5 mm. Chemical analysis has shown that the fouling material was composed mainly of organics (70%) including iron and other minor inorganic constituents.

4.2.4.1 The pretreatment scheme of the Addur desalination plant after modifications of the initial design according to a rehabilitation program

In order to improve the performance characteristics of the plant, the pretreatment system was recommended to be modified, namely, the UF was suggested to be incorporated into the existing pretreatment system. In addition, the following changes in the technology were made according to recommendations: (1) The dual media filters were converted into single media filters with graded sand as the media (2) Sulfuric acid and SBS dosing systems were replaced (3) The membranes on two RO trains were replaced. In the beginning of 2005, when the full rehabilitation program has to be finalized, the Addur SWRO plant is expected to produce 5 MIGD of water with improved quality.

- **Design characteristics of the UF system**

A new enclosed building housing for the UF system was constructed at the open area south of the newly named single media filters. It consists of 9 UF trains each having 126 UF modules containing 3 UF elements. New RO feed and backwash pumps with RO feed well were installed with an independent membrane filter control room. The process flow diagram (after rehabilitation) is shown in Figure 4-33. Table 4-11 presents the main characteristics of UF system.

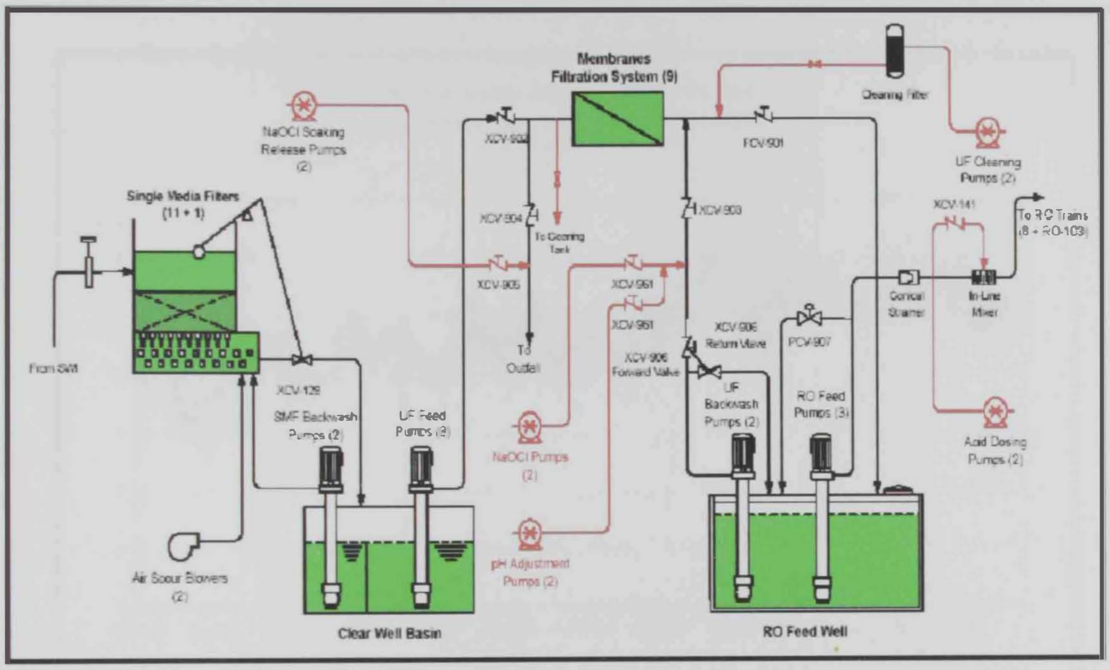

Figure 4-33: Modified flow-diagram of the process (after rehabilitation) [110]

Table 4-11: Main characteristics of the UF system [110]

Element	RS35/2000 Daltons cut off
Manufacturer	Nitto Dinko
Element configuration	Spiral wound, backwashable
Element material	Polysulfone
Membrane surface area per element	40 m^2
Total number of UF trains	9
Number of elements per train	378
Total number of UF membranes	3402

- **Performance characteristics of the UF system**

Few months after rehabilitation in July 2000 the performance of the RO membranes were deteriorated because of severe biological fouling. The experimental projections of the SDI index are shown in Figure 4-32. The performance characteristics such as the SDI index and the transmembrane pressure continued to deteriorate, which dictate that the frequency of chemical cleaning should increase (while according to the manufacturers recommendations the SDI index should not exceed 3). Figure 4-33 presents the measurements of the turbidity of the chlorinated raw seawater, single media system (SMS) outlet for four years of operation. The maximum reading of the filtrate turbidity was 1.16 NTU and 0.47 NTU as average value.

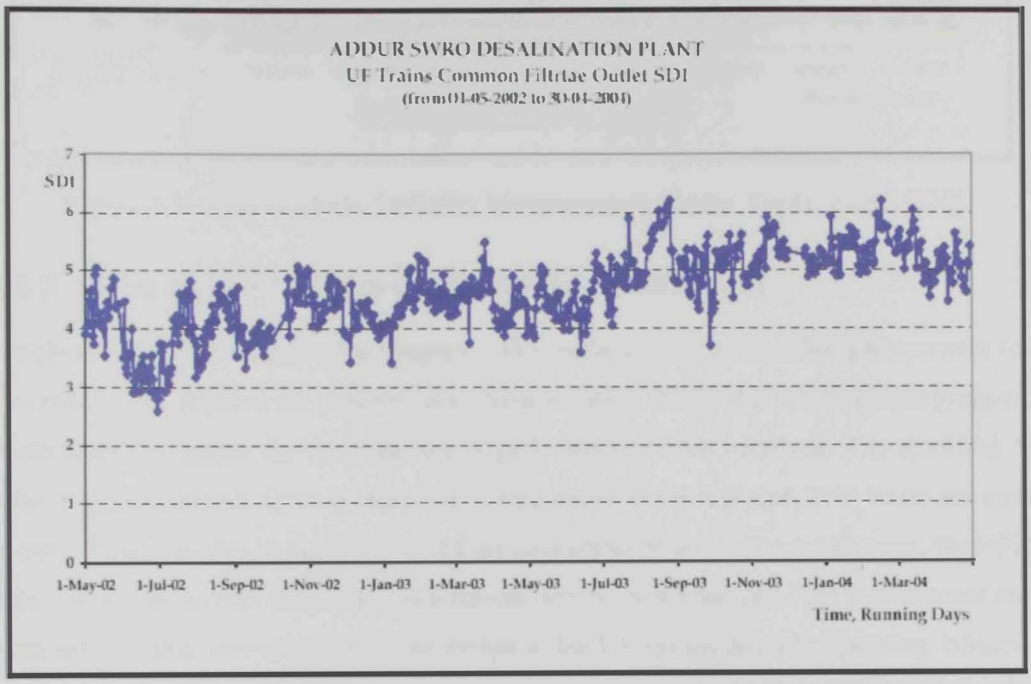

Figure 4-32: UF Trains Common Filtrate Outlet SDI (Addur SWRO plant) [110]

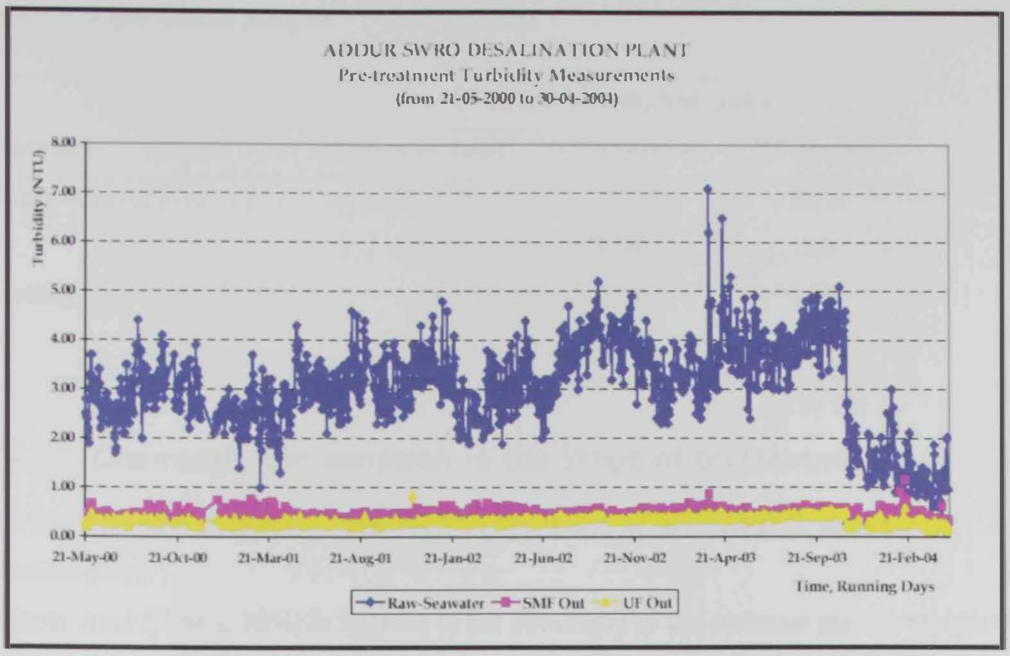

Figure 4-33: Pre-treatment Turbidity Measurements (Addur SWRO plant) [110]

4.2.4.2 The present status of the Addur SWRO plant

The implementation of rehabilitation program and installation of the modified pretreatment scheme did not manage to improve the performance characteristics. While the rehabilitation program was not implemented properly the many persistent problems remained unsolved. The modified design includes 5 trains, but only three of them are in operations at the moment. Two trains are normally taken out for cleaning due to rapid rise in differential pressure across the membranes. Burashid and Hussain [109], outlined the following main reasons for the performance of the pretreatment remains unimproved: (1) Improper selection of the design of the UF system and (2) Operating failures. The installed UF pretreatment was selected mainly on the basis of its ability to bring down the silt density index (SDI) from the range of 16--18 to a level of less than 3, disregarding the nature of the main fouling factors, which are natural organic materials and other colloidal particles in the Addur plant case. That is why the SDI index does not meet the RO feed water requirements. The SDI index of the UF filtrate ranges from 4 to 6. (The water quality analysis are given in Table 4-12. In addition, some operating failures and improper regime parameters such as hydrodynamic characteristics of the feed flow, type of cleaning solutions and pH operating values were found to take place as well.

Table 4-12: Water quality analysis

Parameter	Location of sampling point		
	Seawater Inlet	SMS Outlet	UF Outlet
Temperature (C°)	22.3	22.5	22.7
pH	8.14	8.08	6.6
Turbidity	1.85	0.29	0.19
SDI	18.97	16.17	≅ 5

4.2.4.3 Chemicals consumption in the stage of pretreatment

The following main chemicals were used in the stage of pretreatment at the Addur SWRO desalination plant:

- Sulfuric Acid (H_2SO_4, 98%) is injected in the streamline of the seawater passed through the UF prior to the entry of the RO. Rate of injection is 0.0132 liters per m^3 of seawater.
- NaClO solution is used for the sterilization/soaking of the UF trains. The NaClO solution is generated on site. The specific daily consumption of NaOCl is equal to 381 liters per one UF train per day. (Hence, the total daily consumption of NaOCl is supposed to be 381 L multiplied by the number of UF trains in operation).
- Citric acid is used for chemical cleaning of the UF trains, when ever the SDI of the UF train exceeds 3.

The rate of chemicals consumption and expenses are shown in Table 4-13. The total expenses of the chemicals is 0.21 $/m^3$ of permeate.

Table 4-13: Specific chemical consumption and expenses on the pretreatment stage

	$/kg	Flowrate, kg/m³	$/m³ filtrate	$/m³ permeate	$/day
Chlorine	0.55	0.043	0.007	0.024	206.98
Sulfuric acid	0.18	0.195	0.013	0.035	308.88
NaClO	0.55	0.261	0.052	0.143	1257.30
Citric Acid	2.75	0.00098	0.001	0.003	23.68
SBS	0.50	0.0249	0.004	0.012	109.50
Total			**0.07**	**0.21**	**1796.84**

4.2.4.4 Summary of Addur SWRO plant

1. Membrane-based pretreatment can successfully coexist with conventional pretreatment rather than a process that should replace it.
2. Different types of membrane operations, such as UF or MF, and various configurations of membrane elements can be recommended as alternatives cases for further study.
3. While selecting the pretreatment scheme, attention should be paid to the main fouling factors, The SDI index itself does not provide information regarding the physical nature of main fouling factors whether they are due to natural organic matter, colloidal or inorganic impurities.
4. The pretreatment implemented at Addur plant is characterized by elevated level of chemical consumption (chemicals cost allocated to cost of water is 0.21 $\$/m^3$ of permeate).

4.3 Result and Discussion

The analysis of the technological schemes is based on the methodology that includes three groups of technological and economic indicators: (A) the water quality data (B) The technological characteristics of equipment, and (C) The economic characteristics of the processes. The values of the main indicators are consolidated in Table 4-14.

The SDI index: The quality of water after pretreatment was based on the following the indicators: turbidity, the SDI index, and the total suspended solids. The SDI indexes measurements of the conventional pretreatment schemes are observed to be 2.34 and 3.38, for the systems installed (by Ondeo pilot plant) in Al-Taweelah RO pilot plant and Al-Fujairah site, respectively. (The experimental profile is shown in Figure 4-35).

The SDI_{15} index of the filtrate provided by membrane pretreatment by "Aquasource" UF and "Zenon" MF systems are observed to be 1.5 and 2.9 respectively. (The experimental profile is shown in Figure 4-36).

Turbidity: The degree of rejection of turbidity provided by conventional scheme based on Ondeo.Ltd system and hybrid pretreatment by GrahamTek, are observed to be 75 and 78%, respectively.

Table 4-14: Main characteristics of different pretreatment schemes

Characteristics	Case Studies					
	Conventional pretreatment		Membrane based pretreatment			
	Al Fujairah	*Ondeo-PP*	*Zenon*	*Aquasource*	*Graham-Tek*	*Al-ddur*
SDI $_5$ of feedwater	14.62	20.24	20.24	20.24	21.78	18.97
SDI $_{15}$ of filtrate	3.38	2.34	2.95	1.55	< 1	5.00
Turbidty of seawater, NTU	0.79	0.77	0.77	0.77		3.03
Turbidity of filtrate, NTU	0.16	0.10	<1	<1	0.19	0.47
TSS of filtrate, mg/l	-	6.10	-	-		
Deterioration of normalized specific permeability over the test period , %	-	-	20	36	12	
Daily deterioration of normalized permeability, % /day	-	-	0.45	0.86	0.27	
Specific energy consumption, kWh/m^3(filtrate)	0.413	-	0.01*	0.016*	0.125*	0.12*
Specific cost of energy, \$/m^3(filtrate)	0.0729	-	0.00028	0.00047	0.0037	0.0037
Specific cost of chemicals, \$/m^3(filtrate)	0.03	0.01	-	-	-	0.08

- Energy consumed by membrane system only

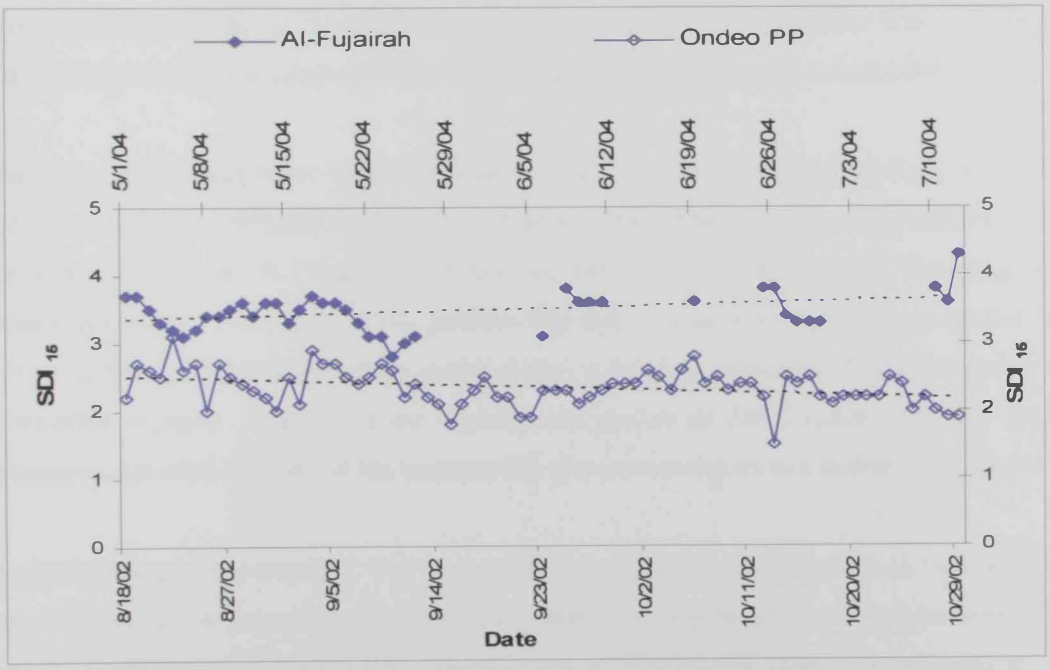

Figure 4-34 : Experimental profiles of the SDI_{15} index of filtrate produced by conventional pretreatment systems: (Al-Fujairah and Ondeo pilot plants)

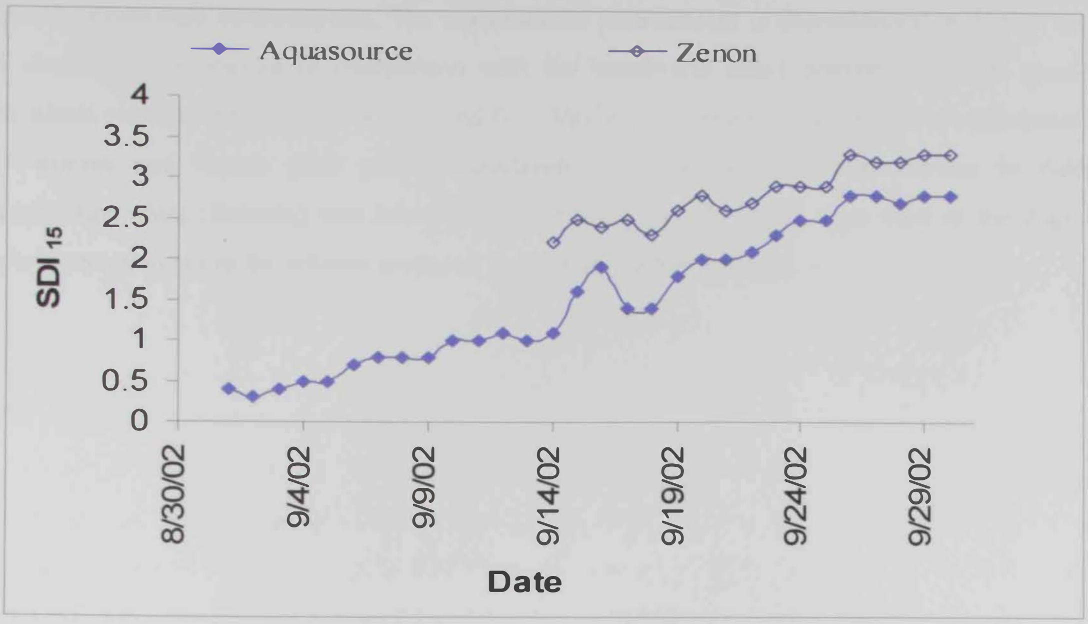

Figure 4-35: Experimental profiles of the SDI_{15} index of filtrate produced by different membrane pretreatment systems

Permeability: The rate of deterioration of permeability is an essential characteristic of performance. Daily deterioration of normalized specific permeability by the "Aquasource" UF and "Zenon" MF systems are observed to be 0.86 % and 0.45% respectively. The hybrid type of pretreatment proposed by GrahamTech system demonstrated a satisfactory level of performance, (daily degree of deterioration of normalized permeability was observed to be 0.27%).

Influence of driving force on permeability: Analysis of experimental profiles of UF and MF processes indicated different functional relation between driving force and permeability. Based on experimental data on UF ("Aquasource" system), two regions were observed. The initial period is characterized by an increase of the permeability and driving force (pressure-controlled region); while during the second period the permeability is became pressure-independent (mass transfer controlled regions). Analysis of the experimental profile of MF ("Zenon" system) confirms a pressure-controlled behavior of the permeability that corresponds to the models of classical MF.

Specific energy consumption. The conventional pretreatment is characterized by higher level of specific energy consumption in comparison with the membrane–based pretreatment. Specific energy consumption in conventional pretreatment in Al-Fujairah plant was estimated to be 1.04 kWh/m^3 and the MF-based pretreatment in Graham-Tek was 0.125 kWh/m^3.

Specific chemicals consumption. The conventional pretreatment is characterized by higher level of chemical consumption in comparison with the membrane based pretreatment. The specific chemicals cost was estimated to be 0.03 and 0.01 $\$/m^3$ of permeate for conventional pretreatment in Al-Fujairah and Ondeo pilot plant, respectively. Cost of chemical consumption in Addur desalination plant (Bahrain) was found to be 0.08 $\$/m^3$. No chemicals were used in the stage of hybrid pretreatment in the scheme proposed by GrahamTech system.

5 Modeling and Simulation of the pretreatment Schemes in the SWRO demo Plant

This chapter deals with the economic assessment of pretreatment schemes in the design of the SWRO demo plant. The design capacity of the SWRO demo plant is 250 m^3 (of permeate per day. The SWRO demo plant was designed by the Japan Cooperation Center for the Middle East. The project was intended to be within the program of the "Advanced Hybrid Desalination System in Abu Dhabi". The demonstration plant consists of two schemes of pretreatments; conventional and membrane based pretreatment. The productivities of the pretreatment processes were designed on the maximum SWRO feed quantities and those of BWRO are on the standard SWRO permeate quantities which are the sum of 2 SWRO units, namely, 138 m^3/day, as the standard design capacity. The production capacity of the BWRO will be 250 m^3/day with a target recovery rate of 90%.

This project is a joint study between Water Reuse Promotion Center in Japan and Abu Dhabi Water and Electricity Authority (ADWEA). The main objectives of this joint project are:

- To demonstrate the new technology for pretreatment stage before RO, especially for Arabian Gulf water.
- To investigate the potential of the membrane pretreatment to achieve the water quality requirement for the RO feed water.

5.1 The Pro-Designer Software

The pro-Designer software is a package that can perform steady state and batch process design and simulation. The Pro-Designer family of software currently includes BioPro Designer, BatchPro Designer, EnviroPro Designer and SuperPro Designer [www.intelligen.com]. The software is equipped with several features that will satisfy even the simulation of veteran when it comes to preliminary simulation and evaluation of process alternatives:

- Adjusting several key parameters affecting the performance of each operation
- Completely customizing the economic evaluation process
- Viewing utilization of resources (like raw materials, utilities, labor) in time for single or multiple overlapping batches.

5.1.1 Unit Operations and Equipment

The different units operations that are employed for water treatment in the SWRO process are:

A. Granular Media Filtration

The model used to simulate the behavior of a granular media filter assumes that the bed is composed of one or more layers of packing. Each layer is assumed to have a uniform distribution of particles with a given average binding capacity (over filtration time and length of that medium's bed depth) expressed in mg of solids per cm^2 of bed volume (including voids). Based on that binding capacity the effective binding capacity of the whole bed is estimated as the weighted average of each layer, with weight being the percent of total bed depth dedicated to each layer. The software uses Carmen-Kozeny model which can estimate the pressure drop across the clean bed and a description of the particulate properties required (as presented below):

$$\frac{\Delta P}{L} = \frac{f}{\phi}\frac{1-\varepsilon}{\varepsilon^3}\frac{\rho u^2}{d} \qquad [5\text{-}1]$$

$\Delta P/L$: pressure drop gradient, bar/ m

ρ : density of the feed, kg/m^3

u: linear velocity of the feed, m/s

d: grain size diameter, m

ε : porosity of the layer, dimensionless

ϕ : particle shape factor, dimensionless

f : is the friction factor, calculated as

$$f = 150\frac{1-\varepsilon}{\mathrm{Re}} + 1.75 \qquad [5\text{-}2]$$

$$\mathrm{Re} = \frac{\rho u d}{\mu} \qquad [5\text{-}3]$$

And μ is the viscosity of water, kg/m.s

The filtration stage is divided into two stages, dual and mono media filtrations. The first one consists of two layers, one for anthracite and sand, whereas the second filter consists of one layer only of sand.

B. Cartridge filter

Cartridge filter is a polishing step after media filtration. The filter media is mounted on a cartridge that is replaced periodically. The filter media pore size is 0.45 microns.

C. Pumping operation

The type of pumps used to transfer the water is centrifugal, while a diaphragm pump is used for chemicals dosing. In the Design Mode, the user specifies the desired pressure change (ΔP) and the model calculates the required power supply using the following equation:

$$\text{Power} = Q \, \Delta P / \eta \qquad [5\text{-}4]$$

Where,

Q : the volumetric flow rate, m^3/h

ΔP : the desired pressure change, bar

η : the pump efficiency

C. Microfiltration

Microfiltration is used to remove micron-size particles, applied in membrane based pretreatment scheme (B).

5.2 Scheme based on conventional pretreatment before RO

The seawater is pretreated with chlorine, coagulant and sulphuric acid and then fed to the media filters. The flocculated seawater is filtered through gravity dual- and mono media filters and collected in the storage tanks (45 m^3). The filtered water is passed through a safety filtration unit before reaching the RO plant to remove any suspended matter. The filtered water is then pumped up to the RO plant through high pressure pumps. Figure 5-1 presents a schematic diagram of the first line of pretreatment. The characteristic of different units of this scheme are shown in Table 5.1, 5.2, and 5.3.

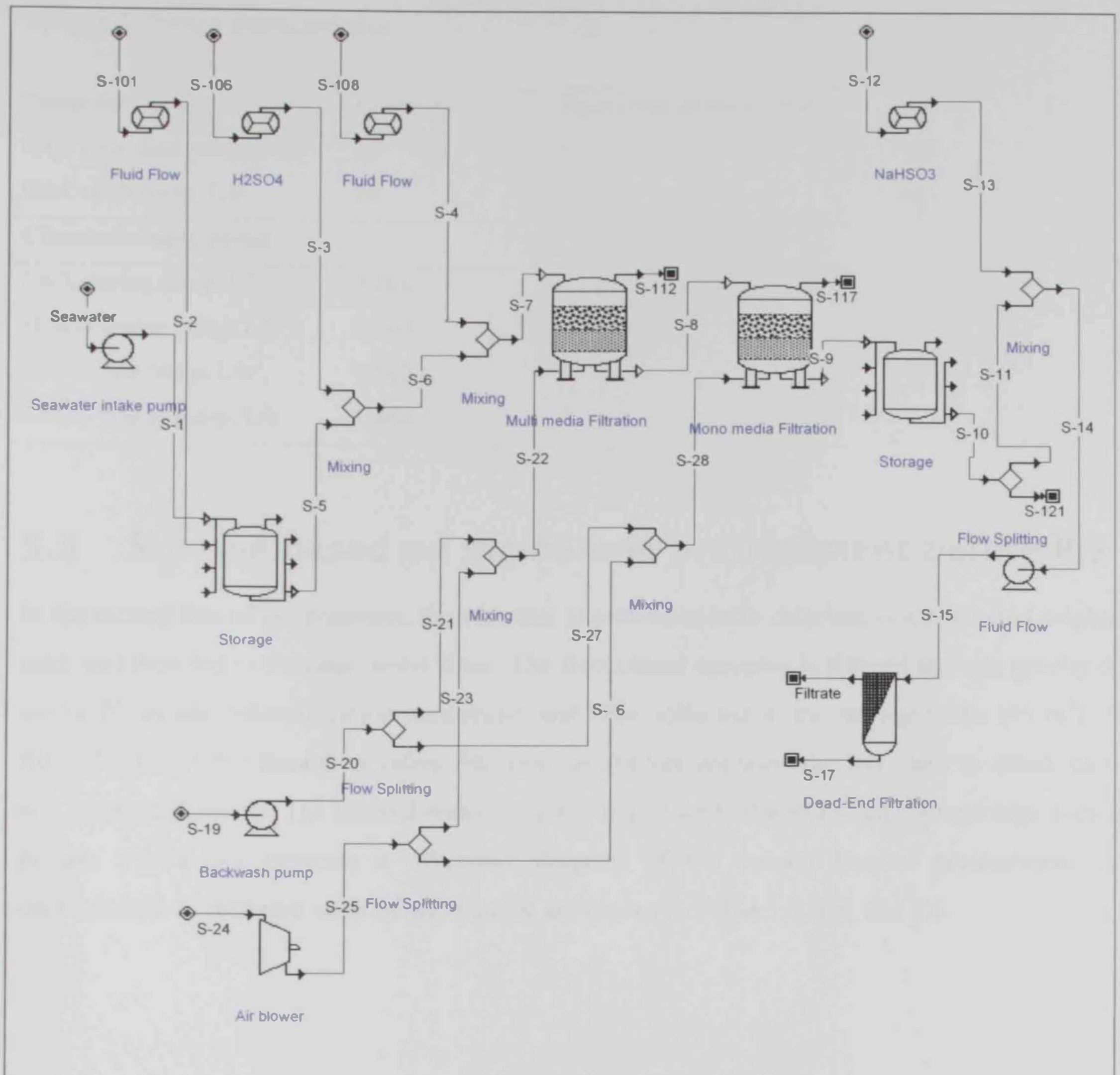

Figure 5-1: Schematic diagram based on conventional pretreatment (Scheme A)

Table 5-1: Characteristics of granule media filtration

Type	Length % of total	Density kg/m^3	Void fraction, %	Shape factor	Effective diameter, mm
Anthracite	63	1500	55	0.72	0.4
Sand	37	2600	42	0.82	0.6

Table 5-2: Cartridge filter Characteristics

Pore size (μm)	ΔP, bar	TSS rejection, %	Recovery, %	Filtrate flux, $L/m^2/h$
0.45	1	10	95	250

Table 5-3: Pumps characteristics

Pump function	Flowrate	Operating pressure, bar
Sand filter feed pump, m^3/h	18	5
Back wash pump, L/h	50	
Chemical dosing pump		
$FeCl_3$ dosing pump, L/h	0.204	5
H_2SO4 dosing pump, L/h	0.300	5
SBS dosing pump, L/h	0.402	5
NaClO dosing pump, L/h	1.860	5

5.3 Scheme based on membrane pretreatment before RO

In the second line of pretreatment, the seawater is pretreated with chlorine, coagulant and sulphuric acid, and then fed to the dual media filter. The flocculated seawater is filtered through gravity dual media filters and Microfiltration membrane, and then collected in the storage tanks (45 m^3). The filtered water is fed through a safety filtration unit before reaching the RO plant to avoid inlet of any suspended matter. The filtered water is then pumped up to the RO plant through high pressure pumps. Figure 5-2 presents a schematic diagram of the second line of pretreatment. The characteristic of different units of this scheme are shown in Table 5.4, 5.5, and 5.6.

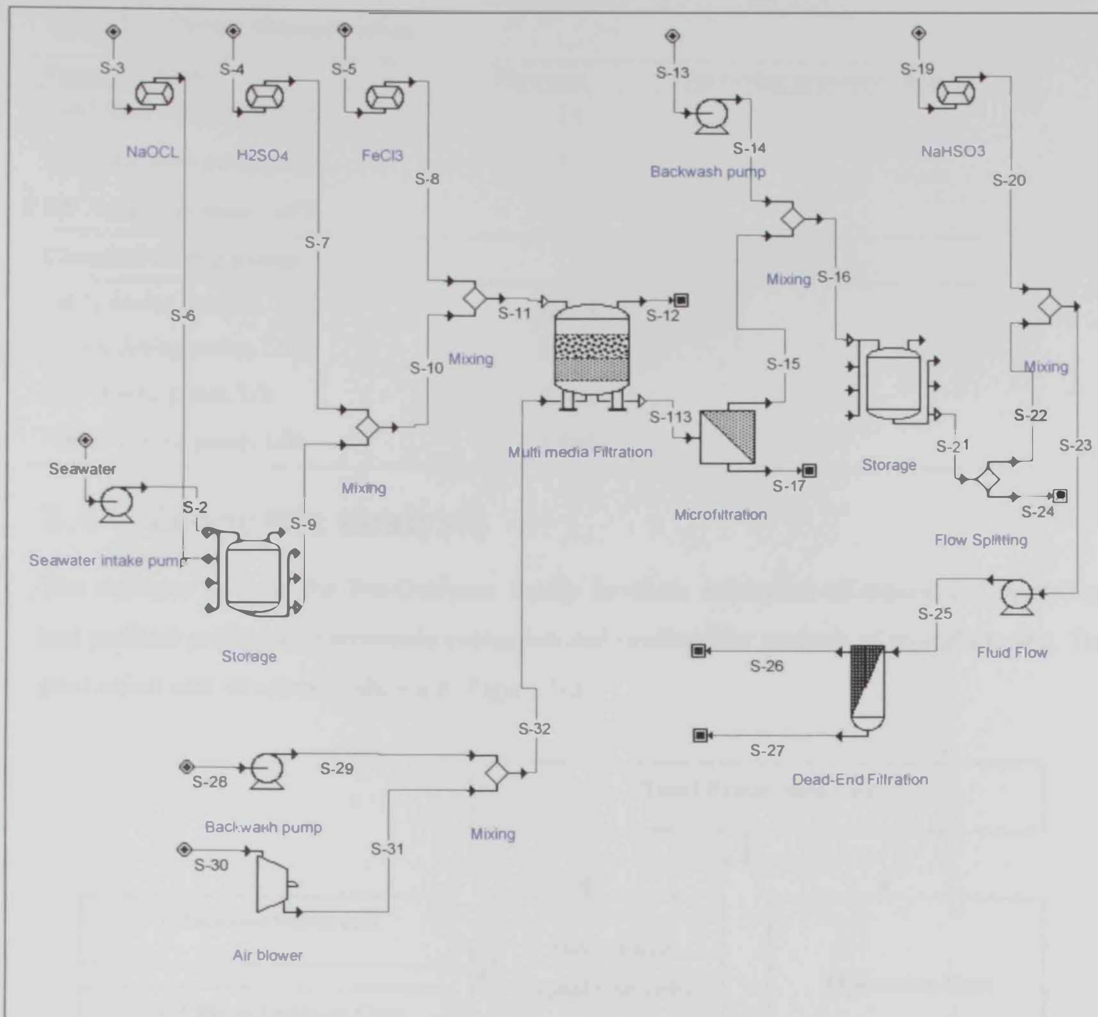

Figure 5-2: Schematic diagram based on membrane pretreatment (Scheme B)

Table 5-4: Characteristics of granule media (sand in membrane pretreatment)

Type	Length, % of total	Density, kg/m^3	Void fraction, %	Shape factor	Effective diameter, mm
Sand	100	2600	42	0.82	0.6

Table 5-5: Microfiltration membrane characteristics

Pore size, μm	ΔP, bar	TSS rejection, %	Recovery, %	Filtrate flux, L/m^2/h
0.1	1	30	90	94

Table 5-6: Pumps characteristics

Pump function	Flowrate	Operating pressure , bar
Sand filter feed pump, m^3/h	24	5
MF Back wash pump, m^3/h	54	2
MF backwash pump , m^3/h	120	
Chemical dosing pumps		
$FeCl_3$ dosing pump, L/h	0.204	5
H_2SO_4 dosing pump, L/h	0.300	5
SBS dosing pump, L/h	0.402	5
NaClO dosing pump, L/h	1.860	5

5.4 Economic analysis

The software tools of the Pro-Designer family facilitate estimation of capital and operating costs and perform preliminary economic evaluation and profitability analysis of manufacturing. The total production cost structure is shown in Figure 5-3.

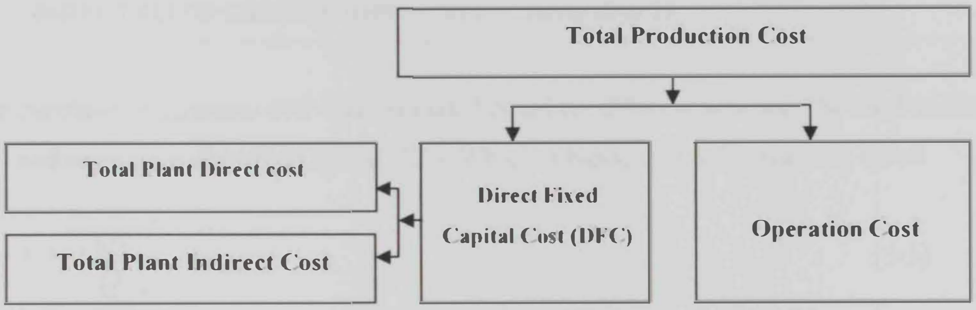

Figure 5-3: Structure of total production cost

5.4.1 Direct capital cost

The equipment purchased cost is the dominant category of the direct capital cost as shown in Table 5-7. Other important items are installation, process piping, etc. See Table 5-7.

Table 5-7: The structure of fixed capital investment, values for direct and indirect cost segments (Peters & Timmerhaus) [106]

A.	**TOTAL PLANT DIRECT COST (TPDC)**	
1	Equipment Purchase Cost (PC)	
2	Installation	(15% of PC)
3	Process Piping	(10% of PC)
4	Instrumentation	(6 % of PC)
5	Insulation	(5% of PC)
7	Buildings	(5% of PC)
8	Yard Improvement	(4% of PC)
9	Auxiliary Facilities	(5 % of PC)
B.	**TOTAL PLANT INDIRECT COST (TPIC)**	
10	Engineering	(5 % of TPDC)
11	Construction	(6 % of TPDC)
C.	TOTAL PLANT COST (TPC) = (TPDC + TPIC)	
12	Contingency	(6% of TPC)
13	Contractor's fee	(5% of TPC)
D.	**DIRECT FIXED CAPITAL (DFC) = TPC + Items 12 & 13**	

The purchased equipment cost was estimated based on different sources. The ship relation between cost and capacity is described by Eq. [5.5]. Which is based on a reference case value.

$$K_2 = K_1 * \left(\frac{Q_2}{Q_1}\right)^n \qquad [5\text{-}5]$$

K_1 = Cost of reference case

K_2 = Cost of required equipment

Q_1 = Capacity of reference case

Q_2 = Capacity of required equipment

n = Scale factor

The scale factor, n, is used to estimate the cost of a piece of equipment when no cost data are available for a reference size of an operational capacity. For membrane systems, the scale factor is normally larger, with n being in the range of 0.75 to 1 [48]. The major difference between the two schemes is that the second scheme has microfiltration membrane instead of polishing filter. The data for the reference case has been adopted from Wilf and Klinko [48], The cost of MF was calculated to be $ 36766, where the scale factor was assumed to be 0.9 .

The structure of the DC and the IDC was used based on [106]. The purchased cost of the different pieces of equipment were adopted from Super pro-designer software and published data [95, 106,111] (see Table 5-8)

Table 5-8: Cost of different pieces of equipment

Equipment	Purchased Cost	No. of paces	Total Cost , $	Reference
Feedwater Pump	4200	1	4200	111
Diaphragm Pumps (chemicals dosing)	2000	4	8000	Pro-designer
Granual Media Filter	23000	2	46000	95
Receiver Tank No.1	25000	1	23000	106
Filtrate Pump	4200	1	4200	111
Cartridge Filter	4600	1	4600	95
Centrifugal Compressor	6000	1	6000	106
Receiver Tank No.2	15000	1	17000	106
backwash pump	8000	1	8000	111
Total			**121000**	

*Characteristics of equipment are given in previous sections

The pro-designer software was used to calculate the cost for the pretreatment stage for both schemes. The structure of direct fixed capital cost (DFC) is shown in Table 5-7, where the percentage and components of fixed capital cost are presented. Each item is estimated by multiplying the total purchased equipment cost by a certain factor.

5.4.2 Indirect capital cost (ICC)

The indirect capital costs is the sum of the subsidiary costs associated with engineering and construction of the desalination plant and normally these items estimated by multiplying total equipment cost by a certain factor.

- **Operating and maintenance cost**

Costs of energy, labor, chemicals, spare parts, and major replacements or refurbishment required over the lifetime of the plant are included in the operational and maintenance costs.

A. Chemicals cost

The cost of chemicals used in feed treatment and cleaning operations, including sulphuric acid, caustic soda, ferric chloride, sodium bisulphite, antiscaling agents and chlorine is estimated based on the cost ($/kg) mentioned in Al-Fujairah plant section. The annual chemicals cost for first and second schemes are 2000 and 3137.5 \$/year, respectively. The total amount of chemicals used in scheme A is 9820 kg and in scheme B is 9757 kg.

B. Membrane replacement cost

Membrane replacement rates may range between 5 to 20 % per year. The lower limit applies to low-salinity brackish water when the system is supported by proper operational and pretreatment procedures. The upper limit tends to apply in situations where high-salinity seawater is used (as in the Gulf States) [54]. The membrane replacement modules (either MF or cartridge filters) are the key operating costs. The membrane replacement rate for microfiltration is 20% and for cartridge filter is 15%.

C. Labor cost

Labor is a key contributor on the operating expenses, according to Ray and McCray [48]. The expenses of labor for RO application is 0.4 \$/m^3 of product. Based on this estimation the labor cost for this plant is 28251 \$/year.

D. Energy cost

The energy requirement if the plant is 1.5 kwh/m^3, and for a capacity of m^3 /year and where the cost of energy is 0.03 \$/kWh, the total energy cost is 6145 \$/year.

Table 5-9 presents the summary of operating and maintenance expenses of the pretreatment stage in for both schemes.

Table 5-9: Summary of operating and maintenance cost per m^3 water produced

Annual operating cost items	\$/m^3, (%)	
	Scheme A	Scheme B
Chemicals	0.01 (4)	0.02 (6)
MF membrane replacement	-	0.05 (13)
Filter replacement	0.01 (2)	0.01(2)
Labor-Dependent	0.2 (61)	0.20 (52)
Energy	0.11 (32)	0.11(27)

5.5 Summary of results

The results of an economic evaluation of membrane pretreatment system designed to produce 430 m^3/d of filtrate. The total filtrate cost, including investment and operating cost, is calculated to be 0.48 and 0.57 \$/m^3 for conventional and membrane pretreatment, respectively. It is evident from Table 6 10 that membrane pretreatment is more expensive than the conventional pretreatment. This is due to the higher installed equipment cost for the membrane system. Installed equipment cost of conventional pretreatment, which includes two stages of filtration, is approximately 520.05 \$/m^3/d, whereas the cost for membrane pretreatment is 613.6 $/m^3$/d. It's worth mentioning to say that this cost exclude the intake and outfall expenses. According to published data for the cost of installed equipment ranges from 161 to 370 $/m^3$/day.

Table 5-10: Summary of cost expenses for conventional and pretreatment schemes *

	Cost category	unit	Scheme A	Scheme B
1	Total plant direct cost (TPDC)	$	181497	214146
		$/m^3$/day	422.09	498.01
	Total equipment purchased cost	$	120998	142764
		$/m^3$/day	520.05	613.6
2	Total plant indirect cost (TPIC)	$	19965	23556
	Engineering	$/m^3$/day	21.10	24.90
	Construction	$/m^3$/day	25.33	29.88
3	Total plant cost (TPC)			
	Contingency	$/m^3$/day	28.11	33.17
	Contractor's Fee	$/m^3$/day	23.43	27.64
4	Direct fixed capital cost (DFC)	$	223623	263849
5	Total operating cost	$/m^3$	0.33	0.39
6	Capital recovery cost	$/m^3$	0.16	0.19
7	Unit production cost (filtrate)	$/m^3$	**0.48**	**0.57**

* Life time= 10 years
* Plant capacity = 430 m^3/day

6 Conclusions and Recommendations

The proposed study focuses on comparing of conventional and membrane based schemes of pretreatment. The study is based on experimental data received from pilot plant and full scale plants. In this study the following pretreatment schemes were considered: (1) conventional pretreatment of the RO pilot plant installed by Ondeo Ltd. located at Al-Taweelah site, (2) conventional pretreatment installed on Al-Fujairah hybrid desalination plant, (3) MF pretreatment based on the "Zenon" system located on Al-Taweelah site, (4) UF pretreatment based on the "Aquasource" system located at Al-Taweelah site, (5) hybrid type of pretreatment proposed by GrahamTech pilot plant located in Bainouna power station, (6) pretreatment scheme of RO desalination plant in Addur (Bahrain), and (7) conventional and membrane scheme proposal for the SWRO demo plant. (The proposal was done by the Japan Cooperation Center for the Middle East within the program "Advanced Hybrid Desalination System in Abu Dhabi). Results and conclusions from this study can be summarized as follows:

The analysis of the technological schemes is based on the methodology proposed in this study, which includes three groups of technological and economic indicators such as: (A) water quality indicators, (B) technological characteristics of the equipment, and (C) economic characteristics of the processes.

Water quality indicators: turbidity, the SDI index and the total suspended solid. The SDI_{15} index of the filtrate provided by the "Aquasource" UF and the "Zenon" MF systems are 1.5 and 2.9, respectively.

The indicators of technological characteristics of the equipment include normalized permeability, transmembrane pressure, specific energy consumption and the deterioration of these characteristics during the test period. The daily deterioration of normalized specific permeability by the "Aquasource" UF and "Zenon" MF systems were 0.86 % and 0.45%, respectively.

The hybrid type of pretreatment proposed by GrahamTech pilot plant demonstrated a satisfactory level of performance characteristics. The observed SDI_{15} index was to be <1 and the daily degree of deterioration of normalized permeability was 0.27%.

The economic assessment of the pretreatment schemes in the design of the SWRO demo plant based on a design capacity of 250 m^3 (of permeate)/day. The total cost of filtrate (including investment and O&M) is estimated to be 0.48 and 0.57 $/$m^3$ (permeate) for the conventional and membrane pretreatments, respectively. The cost of installed equipment was estimated to be 520.05 $/$m^3$/d and 613.6 $/$m^3$/d for conventional and membrane pretreatment, respectively.

The analysis of the published statistics and experimental data confirms that the membrane-based pretreatment is a competitive technological alternative to the conventional one. The case of Addur plant is a good example to this. However, the membrane-based pretreatment can successfully coexist with conventional pretreatment rather than a process that should replace it. Some hybrid configuration of pretreatment schemes including conventional processes along with MF and UF can be recommended for further research.

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Appendices

Appendix A. Shares of Desalinated water from different entities in UAE based on the desalination technologies (MIGPY)

	ADWEA*	DEWA*	SEWA*	FEWA*	**Total**	
Thermal techniques	144605	55968	12733.63	4613	217919.63	95.24%
RO	4000	-	3361.69	3531.86	10893.55	4.76%
Total	148605	55968	16095.32	8144.86	228813.18	

*ADWEA: Abu Dhabi Water and Electricity Authority, DEWA: Dubai Water and Electricity Authority, SEWA: Sharjah Water and Electricity Authority, FEWA: Federal Water And Electricity Authority, 2003

Appendix B: Seawater analysis (Site: Al-Fujairah Desalination Plant)

Parameter	Unit	Results
pH	-	7.9-8.2
Turbidity	NTU	0.25
Conductivity	µs/cm	58000-66800
TDS	ppm	34180-45240
SDI_5	-	13-16
Total hardness as $CaCO_3$	ppm	6950-7177
Total alkalinity as CaCO3	ppm	83-116
Calcium	ppm	441-551
Iron	ppm	0.07
Chlorides	ppm	20766-24419
Sulphate	ppm	2000-2908
Silica	ppm	0.5
Copper	ppm	0.05
Total petroleum hydrocarbons (TPHC)	ppb	10-50

Appendix C: Dual media filter characteristic

Characteristic	Value
Total Inlet flow rate , m^3/hr	18,800
Number of filtering units	14
Compartments per filter unit	2
Compartment width , m	4.88
Compartment length, m	15.87
Filter unit area, m^2	155
Total filtration area , m^2	2,170
Nominal filtration flow rate per filter unit, m^3/hr	1,365
Nominal filtration rate, m/hr	8.65
Maximum filtration rate, m/hr	10.10
Support layer size/layer depth, mm/m	2-3/0.10
Sand effective size/layer depth, mm/m	0.3/0.4
Pumice stone effective size/layer depth, mm/m	1.65/0.7
Water height above filter media, m	1.75
Backwash rate, m^3/hr (m/hr)	6,200 (40)
Air scouring rate , m^3/hr (m/hr)	8,500 (55)

Appendix D: Pumps' energy consumption data

Operation characteristics	Intake pump	Chemical dosing pumps					Filtrated water pump	Backwash Pump
		Antiscalant	Sodium bisulphate	Ferric chloride	H_2SO_4	Cationic coagulant		
ΔP , bar	2.5	7	7	7	5	5	5	2
Flowrate, m^3/h	19200	0.016	0.023	0.105	0.273	0.017	4525	2070
Efficiency, η	0.75	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Overall energy consumption ε , (kW)	1777.8	0.009	0.013	0.059	0.108	0.007	3591.071	492.857
Specific energy consumption , kWh/m^3 (filtrate)	0.0982	5.03E-07	6.91E-07	3.24E-06	5.99E-06	3.66E-07	0.1984	0.0272
Specific energy consumption, kWh/m^3, (permeate)	0.248	1.27E-06	1.74E-06	8.17E-06	1.51E-05	9.24E-07	0.501	0.069

Appendix E: Flocculation chambers and air blowers' energy consumption data

Air blower	
ΔP, bar	0.5
Flowrate , m^3/h	4250
Efficiency, η	80
Overall energy consumption ε , (kW)	147.6
Specific energy consumption , kWh/m^3 (filtrate)	0.008
Specific energy consumption, kWh/m^3, (permeate)	0.021
Flucculation chamber	
Power input for two mixers, kW	20.5
Flowrate , m^3/h	19683
Specific energy consumption, kWh/m^3 (permeate)	2.9E-03
Specific energy consumption, kWh/m^3 (filtrate)	1.1E-03
Whole Pretreatment system	
Specific energy consumption, kWh/m^3 (filtrate)	0.41
Specific energy consumption, kWh/m^3 (permeate)	1.04

و613.6 باستخدام الطريقة الأولى والطريقة الثانية، على التوالي. أيضا كانت التكلفة الإجمالية لإنتاج المياه المصفى هي 0.48 و 0.57 $\$/m^3$ على التوالي.

وقد أوصت الدراسة بإجراء بحوث تجريبية أكبر وأكثر عمقا لتقنية التهجين بين الطريقة التقليدية و الأغشية سوءا MF أو UF.

ملخص الأطروحة

إن المعالجة الأولية لمياه التغذية الداخلة لأغشية التناضح العكسي مهمة وضرورية لضمان عمر أطول وأداء أفضل للأغشية، بالإضافة لتقليل كمية استهلاك المواد الكيميائية، وبالتالي تخفيض التكلفة الكلية للماء المنتج. يحتوي ماء البحر على نسبة عالية من الجزيئات العالقة والكائنات الدقيقة وهي حتما تتطلب معالجة أولية مكثفة قبل دخولها لأغشية التناضح العكسي.

تتميز المعالجة الأولية التقليدية باستخدام المواد الكيميائية والعمليات الفيزيائية كالتعقيم بالكلورين لمنع نمو البكتيريا، ثم تخثير الجزيئات العالقة وترشيحها وإزالة الأجزاء الكبيرة منها. تضاف بعد ذلك موانع التآكل لمنع الترسبات على سطح الأغشية. اتصفت لمعالجة التقليدية بارتفاع تكلفتها والإستهلاك الكبير للمواد الكيميائية ، بالإضافة لقصر فترة عمل أغشية الترشيح النهائية.

إن الحالات المختلفة المستخدمة في دراسة الطريقة التقليدية هي محطة الفجيرة و محطة التناضح العكسي التجريبية في محطة الطويلة من شركة Ondeo LTD، أما دراسة الطريقة الثانية وهي المعالجة الأولية بالأغشية فقد اشتملت على : Aquasource و Zenon وهي أنظمة تجريبية اختبرت في محطة الطويلة . بالإضافة للوحدة التجريبية التي وضعت تحت الإختبار في محطة بينونة لمدة ألف ساعة وهي مقدمة من Graham-Tek. تعتبر محطة الدور هي المحطة الوحيدة في الخليج العربي التي تدار بنظام الأغشية، ولذلك عمد الباحث لاتخاذها كدراسة حالة واقعية وليست تجريبية.

ولدراسة هذه الحالات المختلفة وضعت مجموعة من المعايير لتقييم أداء هذه الأنظمة من حيث جودة المياه المنتجة ومن الناحية التكنولوجية والإقتصادية. معايير جودة المياه تشمل بصورة رئيسية معدل كثافة الطمي والعكارة. أما الناحية التكنولوجية فقد اشتملت على معدل الطاقة المستهلكة و كمية الكيماويات المستهلكة باللإضافة لمعدل تدهور نفاذية الأغشية. تكلفة رأس المال وتكلفة التشغيل (وهي تشمل تكلفة المواد الكيماوية والطاقة و الأيدي العاملة) من العوامل التي تلعب دوراً مهماً في تقييم الكفاءة الإقتصادية لأنظمة التحلية المختلفة. وقد وجد من خلال الدراسة والدراسات المنشورة أن تهجين الطريقة التقليدية بنظام الأغشية يعطي نتائج أفضل من الناحية التشغيلية.

اشتملت الدراسة على تقييم اقتصاديّ لمحطة تجريبية مزمع إقامتها في الفترة المقبلة تعمل بالطريقتين التقليدية والأغشية الدقيقة ، أظهرت الدراسة أن تكلفة رأس المال 520.05

جامعة الإمارات العربية المتحدة

عمادة الدراسات العليا

عنوان الأطروحة: التقييم التقني و الاقتصادي لعملية المعالجة الأولية لمياه البحر باستعمال الأغشية في محطات التناضح العكسي

المؤلف: **شما أحمد راشد آل مالك**

المشرف الرئيسي: **د.محمد عبدالكريم**- أستاذ مساعد- قسم الهندسة الكيميائية جامعة الإمارات العربية المتحدة

المشرف المساعد: **أ.د. سيرجي أجاشيف** – باحث- مركز بحوث المياه والطاقة هيئة مياه وكهرباء أبوظبي

لجنة المناقشة: **د.محمد عبدالكريم**- أستاذ مساعد- قسم الهندسة الكيميائية

جامعة الإمارات العربية المتحدة- إ.ع.م

أ.د. كورادو سوماريفا- أستاذ- جامعة لا أكولا - إيطاليا

د.سمير أبوعيشة- أستاذ مشارك- قسم الهندسة الكيميائية

جامعة الإمارات العربية المتحدة- إ.ع.م

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أطروحة مقدمة من الطالبة

شما أحمد راشد آل مالك

بكالوريوس هندسة كيميائية

استكمالا لمتطلبات الحصول على
درجة الماجستير في علوم موارد المياه

يونيو 2005