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WSTA 4th Gulf Water Conference

February 13-17, 1999
State of Bahrain

Water in the Gulf, Challenges of the 21st Century

Ministry of
Electricity and
Water

Secretariat General
of the Gulf
Cooperation
Council (GCC)

Arabian
Gulf University
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Edited by
DR. WALEED K. AL-ZUBARI

CONFERENCE PROCEEDINGS Vol. 2

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The Fourth Gulf Water Conference
Water in the Gulf, Challenges of the 21st Century
State of Bahrain
27 Shawal-2 Dhu'l-qa'da, 1419 H
13-17 February, 1999

Conference Proceedings

Organized by
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Preface

The Countries of the Gulf Cooperation Council (GCC) are situated within an arid to semi-arid zones and are in general devoid of reliable surface water. Water requirements in these countries are met by groundwater (91%), desalinated water (7.5%), and treated wastewater (1.8%). In the past four decades, the GCC countries have experienced high population growth rates, along with expanded urbanization, increased economic activities, expansion of irrigated farming, and improvement of standards of living. The fast pace of development has resulted in a substantial increase in freshwater demands, which are met mainly by non-renewable groundwater abstraction. This heavy dependence on groundwater resources, particularly for agricultural development, has lead to their over-drafting beyond their natural renewal capacity, causing their depletion and quality deterioration, which has reduced their readiness as a source for direct use.

Despite the efforts made by the GCC countries in the provision of alternative water supplies, represented by the expensive desalinated water for the municipal and industrial sectors and the treated wastewater for the agricultural sector, there is a clear imbalance between water available resources and demands. A problem expected to continue in the future, as water is becoming the major constraint on agricultural, urban, and industrial sustainable development. These pressing conditions call for proper planning and management of the GCC countries' limited groundwater resources, a great challenge that faces their water authorities. Thus the theme of the present conference "Water in the Gulf, Challenges of the 21st Century" was chosen to emphasize this critical situation into the next century.

As at the previous WSTA conferences (Dubai, 1992; Bahrain, 1994; and Oman, 1997), the overall goals of the conference are to encourage scientific studies and research in the different fields of water resources, to create a forum of open discussion, and to exchange experiences among the Arabian Gulf States that the WSTA engendered throughout the three previous conferences.

The objectives of the convening conference are: 1) Identify priority issues and challenges facing the GCC countries to achieve sustainable water resources development in the 21st century; 2) Assess the current status of natural and alternative water resources in relation to present and future water demands in the GCC countries; 3) Review methods for conservation and efficient utilization of water in the various consuming sectors; 4) Review measures for water resources protection and maintaining water quality for various water uses; 5) Review local experiences for management, operation and maintenance of water projects in the GCC countries; 6) Review latest technologies and research in the assessment, development and management of water resources;

The Fourth Gulf Water Conference is organized by the Water Science and Technology Association (WSTA) in cooperation with the Ministry of Water and Electricity of the State of Bahrain, the Secretariat General of the Cooperation Council (GCC) for the Arab States of the Gulf, the Arabian Gulf University, and the Bahrain Center for Studies and Research. The Conference is co-organized and sponsored by the Arab Organization for Agricultural Development (AOAD), the International Atomic Energy Agency (IAEA), the International Hydrology Program (IHP), the UN Development Program (UNDP), the UN Economic & Social Commission for Western Asia (UNESCWA), the UN Environmental Program (UNEP/ROWA), the UNESCO Cairo office, the World Health Organization (WHO/EMRO), the European Desalination Society (EDS), and the International Desalination Association (IDA).

This conference proceedings contains 85 papers assembled into three volumes, one volume in Arabic (18 papers) and the other two in English (67 papers). The conference papers were selected by the Conference Scientific Committee from over 105 abstracts received from the call of papers. Many of these were modified to meet the standards of the Scientific Committee review. Eight papers are invited from the supporting organizations and renowned regional experts to give scientific presentations in respective technical sessions, and were supported by the WSTA, UNESCO (Cairo Office), and UNEP/ROWA. Conference sessions will be held on eight topics: Water Resources Planning and Management, Groundwater Resources, Desalinated Water, Wastewater Treatment, Surface Water, Agriculture Water & wastewater Reuse in Agriculture, Municipal Water Supply, and Drinking Water.

The scientific committee wishes to express its deep appreciation to the Governments of the GCC Countries and the GCC Secretariat General, and the sponsoring regional centers and organizations who kindly supported and endorsed this conference by providing keynote speakers.

Organization of these conferences requires considerable time and effort. As in the previous WSTA conferences, individuals from various sectors (industry, government and academia) have come forth and given generously their time. Special thanks are due to the members of the Organizing Committee, Scientific Committee, Information Committee, and Scientific Papers Coordinators and Reviewers.

Finally, we wish to acknowledge the immeasurable contributions made by the authors and their research associates who were not only willing to rework and modify their abstracts and manuscripts, but also had to meet an extremely tight time schedule. Without their efforts this document would not have been possible.

The Scientific Committee sincerely hopes that this conference will achieve its objectives and is both enjoyable and rewarding for you.

Dr. Waleed K Al-Zubari
Chairman, Conference Scientific Committee
Director, Desert and Arid Zones Sciences Program
School of Graduate Studies
Arabian Gulf University, Bahrain

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New Power and Water Co-Generation Concept with Application of Reverse Osmosis (RO) Desalination

T. Altmann

NEW POWER AND WATER CO-GENERATION CONCEPT WITH APPLICATION OF REVERSE OSMOSIS (RO) DESALINATION

T. Altmann
Salzgitter Anlagenbau GmbH (Preussag)

ABSTRACT

This article is the result of work carried out by Salzgitter Anlagenbau GmbH with the ultimate objective of defining the most suitable technique from an energy point of view for reliably producing desalinated water of the required specification with the lowest possible prime energy consumption. It describes how Salzgitter Anlagenbau GmbH has approached and achieved this objective. Furthermore, a new concept for a power/RO co-generation design is introduced. The merits of various desalting technologies will not be covered.

INTRODUCTION

Most large desalination plants are installed in dual-purpose power and water stations in order to make the best use of energy from fuel. Moreover, sea water intakes can be constructed to serve both facilities. In the past RO has been used mainly for single-purpose plants, while multi-stage-flash distillation (MSF) has traditionally been the technology most widely used for large dual-purpose plants, although multi-effect distillation (MED) has increased its market share. All of these distillation systems essentially work the same way, i.e. they use steam from a boiler to vaporize and then condense sea water. However, the evaporation and condensation phase changes required by the distillation processes consume more energy than the liquid separation at ambient temperature occurring in the reverse osmosis process. Energy remains a major factor in the operation of sea water desalting facilities and can be considered as the main single cost item of the total operating expenses excluding amortisation. Due to the increase in energy costs during the last years the reverse osmosis process is now emerging as the most energy- and cost-efficient desalting system as its specific prime energy consumption is significantly lower. Even greater energy efficiency is realised when RO system designs use energy recovery devices on the high pressure brine reject streams. However, there is still a potential for further considerable increases of specific water production in dual purpose plants.

SALZGITTER POWER/RO CO-GENERATION CONCEPT

Since the energy consumed in a power/RO co-generation plant for desalting sea water is process heat and electricity, there is an opportunity to make simultaneous use of the electricity and heat produced in all electricity generation facilities.

For this purpose RO provides a unique advantage to integrated power and water cogeneration plants since RO is the only commercial desalination process available on an industrial scale which can make use of both electricity and process heat as the main drive power for desalting sea water. This is achieved by driving the RO feed pumps either by electric motors or by condensing steam turbines. Presently, the most efficient way to cover the steam demand for desalination is through steam produced by a heat recovery steam generator at the exhaust of a combustion turbine power generation cycle.

The generated electrical power is used to operate electrically driven RO units (ROE), part of it being available for export to the grid. The exhaust heat from the combustion turbine is used to provide steam to steam-turbine driven RO units (ROS), and part of the steam could be used to generate additional electrical power in a steam-turbine generator, if required. Since the ratio of ROE and ROS units may vary according to the assignment of electricity and steam, this arrangement provides maximum flexibility to meet power and water demand.

When water demand is at a maximum, both ROE and ROS units operate at maximum. When the water demand decreases, the output from the ROE units is reduced and more power is exported to the electric grid. The ROS units continue to operate at their rated capacity. The maximum utilization of fuel energy can be achieved when the plant is designed for a 3:1 ratio of electrically and steam-turbine driven RO units. This plant configuration enables a total plant energy efficiency of more than 85%, resulting in overall water production per unit of prime energy being much higher than with integrated power/MSF plants.

However, contrary to existing power/MSF cogeneration plants where 8-10 megawatt of electrical power needs to be exported for each one million gallon of desalted water produced, there is no need for power export if the new RO process is applied since all the generated power and process heat can be used in the co-generation plant for the desalination process. Furthermore, in case of a shut-down of a power generation unit electricity from the grid can be used to secure continued water production.

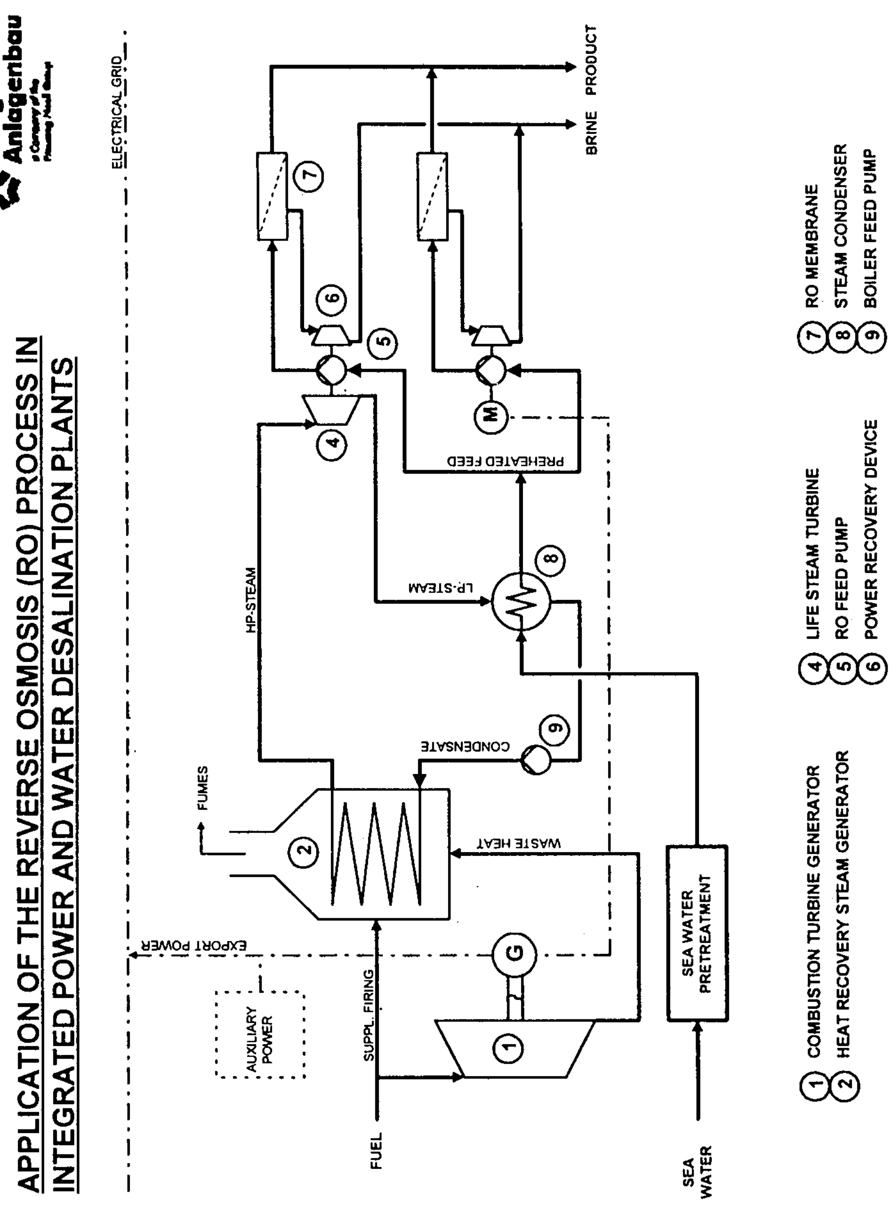

Figure 3: Power/RO co-generation concept

PROCESS OUTLINE (POWER/STEAM CYCLE)

Fuel is supplied to the co-generation plant and conditioned to the requirements of the selected combustion turbine. Power is generated by the generator which is driven by a gas turbine (1). The burned fuel gas from the turbine enters the heat recovery and steam generator (HRSG) (2), where condensate is turned into superheated steam and the fume gas leaves the HRSG (2) through a stack. Fluctuations in steam demand are compensated through supplementary firing.

Superheated steam is supplied to the life steam turbine (4). The exergy of the steam is converted into mechanical energy which is used to drive the RO feed pump (5) and consequently to elevate the pressure of the pre-treated sea water to the appropriate RO membrane (7) operating pressure. An impulse turbine (6) is mounted in line with the RO feed pump (5) in order to recover a considerable amount of the pumping energy. The steam exits the life steam turbine (4) at vacuum conditions and is condensed in the vacuum condenser (8) using pre-treated sea water. The condensate from condenser (8) is transferred to the deaerator. In the deaerator non-condensibles are stripped out to the atmosphere and the condensate is pumped to the heat recovery steam generator (2) by the boiler feed water pump (9).

Alternatively to a vacuum-condenser the exhaust steam out of the life steam turbine (3) can be used as energy source for operating a MED plant. This combination is called a "hybrid plant", with the MED units providing a supplementary source of lower TDS water for mixing with higher TDS water from the RO plant to yield a potable water product with a TDS of less than 450 ppm. Part of the MED distillate can also be used as boiler feed water supply. This concept can result in a lower cost RO system because its design can be less stringent regarding the TDS content of the RO product water.

It is obvious that there is some potential for the application of the "hybrid" system in the Gulf region where extensive preheating of already warm sea water is not desired and the high salinity of the sea water requires a high membrane area to yield potabale water at high water temperatures.

No separate sea water intake and outfall system is needed for the MED plant since the cooling water for the MED is used as feed-water for the RO plant and the pre-heated brine from the RO plant is used as feed water for the MED plant.

ADVANTAGES OF THE POWER/RO CO-GENERATION CONCEPT

Benefits in comparison to existing power/MSF plants:

- The RO process has the lowest sensitivity to energy cost since RO has the lowest prime energy consumption and the lowest portion of energy on total water costs. These savings become substantial as fuel costs increase.

- Since some of the major consumers of electrical power in the RO plant, i.e. the RO feed pump drives, have been replaced by steam-turbines, the specific electrical power consumption is less than the specific power consumption of MSF/MED plants because less feed water and brine needs to be pumped to and from the RO plant due to the much higher conversion ratio.

- The capitial cost of steam-driven RO units remains lower than comparable MSF/MED units due to considerable savings in the RO plant.

- The feed-to-product ratio of the RO plant is less than 2.5 : 1 whereas in distillation plants the ratio varies between 7-12 : 1 since a major part of the feed water is used as cooling water. Consequently, the required sea water intake and brine outfall structure for an RO plant is much smaller than for MSF/MED plants.

- The performance ratio[1)] of a steam-driven RO unit is approx. 50 (MSF/MED: 7-12). Such RO units produce 2-3 times as much water as MSF/MED plants using the same amount of steam energy.

- The power/RO plant allows a better match of power and water demand, eliminating the traditional MSF ratio of 8-10 MWe of export power for 1 MGD of product water since all the generated power and process heat can be used in the RO plant for the desalination process.

- Power from the grid can be imported to secure continuation of part of the water production in case of shutting down a power generation unit.

- The power/RO co-generation plant offers additional flexibility of operation by its increased ability to "follow" daily and seasonal power and water demands and higher efficiency operation can be obtained throughout the year.

[1)] Performance ratio = kg product water / kg life-steam

- The off-line time interval is shorter for RO than for distillation plants and load changes can also be made quicker by simply taking sections out of service without interrupting continous plant production. Start-ups and shut-downs of RO units can be accomplished in less than 30 minutes. The changes in steam flow required for the turbines can easily be controlled by means of supplementary firing.

- The construction of the power plant is less influenced by the design of the RO plant since the steam is taken from the heat recovery boiler and not extracted from the steam turbine in the power cycle. Electrically driven RO units are only connected by means of power transmission.

- Part of the water can be produced by steam-driven RO units at a joint location with the power plant and part of the water can be produced at an independent location close to the point-of-use to minimize water pipeline needs. Electrical power transmission provides an inexpensive and flexible means to connect the power generation and water desalination technologies.

- Preheated cooling water from the power plant can be used as feed water for the RO plant causing enhanced membrane performance.

- Product water post-treatment is simpler and less costly for the RO process since the RO product water is typically provided at ambient temperature, hence the product water does not need to be cooled. Furthermore, the RO product water usually contains a salt level which meets the WHO requirements for drinking water. Additional resalination, e.g. blending, of product water, a common practice with MED/MSF product water, is therefore not required to the same extent. Non-renewable ground-water resources are preserved.

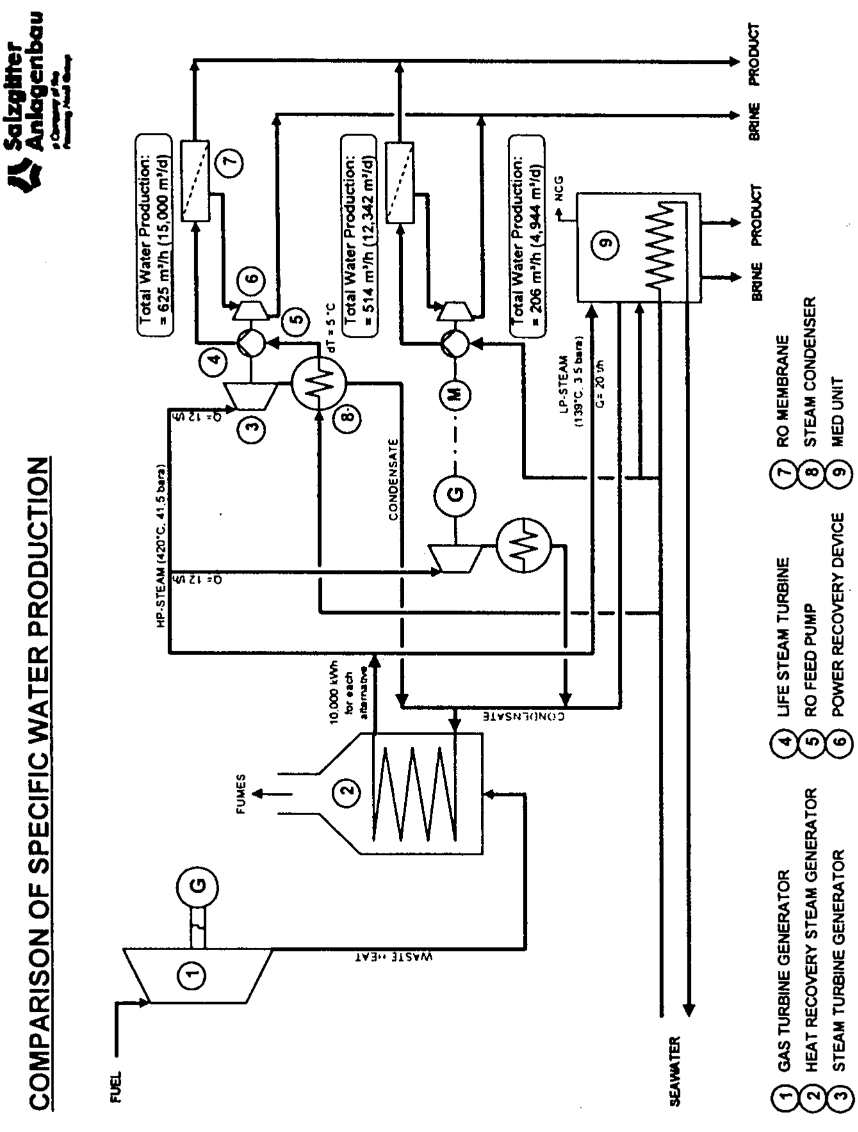

Figure 3: Comparison of specific water production

Improvement of the standard RO process design:

- RO feed water is used to condense the exhaust steam out of the turbines from the RO units by increasing the feed water temperature which depends on the ratio of ROE and ROS units in use. Preheating the feed water results in enhanced RO membrane performance. The productivity (flux) increases about 2-3 % for every degree Centigrade of temperature rise. The upper limit is set by the membrane tolerance at about 40-45°C. Increasing the productivity of the membranes will result in a reduction of membrane area needed to desalinate a given flow rate of feed water. This in turn will reduce the size, cost and pumping requirements of the overall RO facility.

- As an innovative energy-saving feature, speed-controlled steam turbines allow matching the pump discharge pressure to the required membrane pressure to avoid throttle losses in the feed control valve during operation at high temperatures where the feed pressure has to be reduced.

- In RO plants where steam-driven RO units are installed the RO feed pump can also be used for the membrane cleaning and flushing procedure. By operating the RO feed pumps at lower speed, the required flow and pressure can be adjusted to the specified cleaning and flushing requirements.

APPLICATION OF THE NEW POWER/RO CO-GENERATION TECHNOLOGY

- Design of large integrated power and water co-generation plants

- Conversion of existing fueled power stations to power/RO cogeneration plants

- Utilization of excess steam in existing dual purpose power/MSF plants to enhance water production

- Replacement of aged MSF plants in existing dual purpose plants

- Design of power and water co-generation plants for smaller consumers where large power plants are not economically justified

CONCLUSION

During the last years many reverse osmosis sea water desalination plants have been built worldwide. Plant sizes and cost effectiveness have increased each year and the technology has advanced rapidly. RO membranes are available now which can produce drinking water of satisfactory quality in a single stage, even from high salinity Gulf sea water. Developments in pretreatment and the use of energy recovery turbines have improved reliability and energy efficiency. Large capacity plants have proven reliable and continue to be installed. The RO technology has reached the point where the life-cylce cost are competitive with traditional MSF plants for single and dual purpose plants of any size.

The introduction of this new power/RO co-generation concept has positioned the RO process as the optimum technology for power and water co-generation plant installations since the RO process is the only desalination process which can make use of both electricity and process heat as the main drive power for the desalination of sea water. Consequently, using RO, large quantities of desalted water can be provided to regions where no further electrical power is needed.

For some applications RO/MED hybrid systems may offer further benefits if their respective technological strenghts are properly utilized in the hybrid plant.

A subtle but valuable advantage of the RO process is that the power-to-water-ratio is not fixed and therefore the power plant and the desalination plant can be sized and optimized independently by the end-user for the right balance of water need and electricity demand to achieve the maximum economical benefit.

An economic analysis done by Salzgitter Anlagenbau GmbH has shown that the product water cost from the power/RO cogeneration plant can be reduced by 15-20% as compared to power/MSF plants due to much better utilization of the specific fuel energy.

DEFINITIONS

Co-generation - "Co-generation" is an energy production process involving the simultaneous generation of thermal (e.g., process steam) and electrical energy by using a single primary heat source. It can be employed whenever there is a need for the two energy forms and whenever on-site electric power generation is justified or when thermal energy users are in close proximity to an electrical power generation facility.

Hybrid - The term "hybrid" refers to a combined distillation and RO plant, such combination being the "water portion" of a dual purpose power and water plant.

Reverse Osmosis (RO) - A pressure greater than the osmotic pressure of the feed water is applied to one side of a semi-permeable membrane, producing a flow of desalinated water through the membrane.

Multi-Stage Flash Distillation (MSF) - A thermal desalination process in which a brine stream flows through the bottom of several chambers (stages). The pressure in each chamber is maintained at a lower level than the saturation vapour pressure of the water and a portion of it "flashes" into steam and is then condensed.

Multi-Effect Distillation (MED) - A thermal desalination process in which evaporation takes place as a thin film of feed water passes over a heat transfer surface (outside of horizontal tubes - HTE or inside of vertical tubes - VTE). Vapour formed in the last effect is recompressed thermally (TVC) or mechanically (MVC).

Maintenance and Operation Experience of Co-Generation Plant

Saeed Mohammed Al-Shehabi

MAINTENANCE AND OPERATION EXPERINCE OF CO-GENERATION PLANT

Saeed Mohammed Al-Shehabi
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ABSTRACT

Since the early days of commissioning of phase No. II co-generation (Gas Turbine – Boilers – MSF Distillers) plant at Sitra Power and Water Station in 1984, the plant has experienced a series of recurrent failures seriously affecting the plant reliability. These failures and consequent shutdown at times, affected the plant availability especially during the peak load demands.

Recurrent problems occurred mainly in the areas of primary gas streams of gas reducing station; burner management system & control system of auxiliary boiler; economiser inlet header of heat recovery boiler and sea water pipes and structures of distiller.

KEYWORDS: Co-generation plant, Gas Turbine, Auxiliary & Stand alone boilers, Multi Stage Flash (MSF) Distiller, Problem identification, Failure analysis, Plant Reliability and availability.

INTRODUCTION

Unit-5 consisting of a *25 MW* Gas Turbine with a Waste Heat Recovery Boiler, an Auxiliary Boiler and a *5 MGPD* capacity MSF unit was commissioned in 1984.

Unit-5 is a self-contained plant with its separate gas supply system, sea water intake, outfall culvert and other auxiliary systems.

It initially experienced several trips and failures due to design deficiencies and component defects. Plant layout diagram shown in Figure No.1.

During years of experience in operation and maintenance of this plant, the system deficiencies were identified and effective maintenance techniques were applied. As a result of these efforts, improvement has been achieved not only in increased plant reliability but also in process control and plant operation.

Reliability improvement program requires management support and an appropriate organizational structure. The key strategies for improving reliability and overcoming system deficiencies are the systematic and careful analysis of whatever failure occur and failure avoidance via predictive modeling or system life cycle management or both.

This paper discusses **failure analysis strategy** and the maintenance solutions that were adopted at the plant to prevent recurrence of failures.

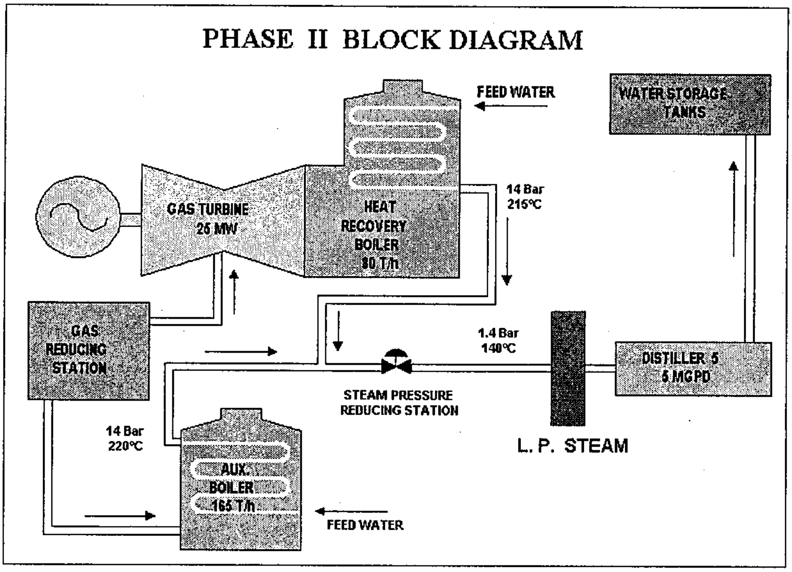

Fig. No. 1: Plant Layout Diagram

FAILURE ANALYSIS STRATEGY

To be effective, analysis must be based on a strategy that includes definite procedures, objectives and tools. Ideally, the analyses should be conducted by on interdisciplinary team whose members possess experience directly related to various aspects of the system being studied. The team studies the failures so that it can formulate courses of action to prevent recurrences. To do this, it must not only understand how the system was intended to function but also establish what combination of factors caused it go wrong.

No two failures are exactly alike. This is mostly because virtually all failures, particularly major ones, have multiple causes. Several conditions or events converge to precipitate the incident, sometimes suddenly. The table No.1 below suggests the range of causes commonly associated with major failures.

Table No.1 causes of major failures

Category	Cause
Human Factor	- Defective communications. - Role confusion - Improper attitude. - Perceptual flaws. - Misplaced priorities. - Ignorance and uncertainty. - Conceptual misunderstanding.
Environment	- Unforeseen conditions. - Unexpected deterioration or response. - Inadequate control or service conditions. - Use demands exceed system limitations. - Environmental in-compatibility.
Things (Equipment)	- Malfunction - Improper repair. - Design deficiency. - Inadequate maintenance. - Technical fault or defect. - Wrong or substandard material.

Steps of Failure Investigation

The steps to follow are not always carried out exactly in the sequence shown below, because there are many unavoidable interactive loops in the procedure.

- Broadly characterize the problem.
- Document the observations.
- Identify, collect and preserve evidence.
- Examine evidence.
- Reconstruct what may have happened.
- Determine why it happened.
- Formulate conclusion and recommendations.
- Report the results.

PHASE NO. II PLANT-PROBLEMS AND SOLUTIONS

GAS REDUCING STATION

The gas reducing station (figure 2) consisting of two full 100% capacity reduction streams, provides the required flow of clean dry gas at suitable pressure and temperature conditions to ensure safe, efficient and reliable operation of the gas turbine generator unit and Auxiliary Boiler. The system designed by Metano Impianti, Italy, and uses pressure regulators manufactured by Pietro Fiorentini. The system is designed for an inlet gas pressure of 70-35 bar (g) and temperature of 0°C - 50°C. Pressure reduction is done in two stages. The primary pressure reduction regulator reduces the gas pressure from 70 to 14 bar (g) supply to the gas turbine through fine filters, and one tee off line is taken to secondary streams for further reduction from 14 bar (g) to 4 bar(g) for supply to the Auxiliary Boiler by means of secondary regulator.

Each of primary and secondary streams is of 100% capacity.

The first reduction streams operate in service-standby mode. In the original scheme, the streams were designed to change over by means of pressure switch actuated block valves at both high and low pre-set pressures.

The streams also incorporate monitor regulators, which provide the back-up control in the event of failure of the working regulator. Two full flow safety relief valves are also provided at the 14 bar header for protection of the connected plant and equipment.

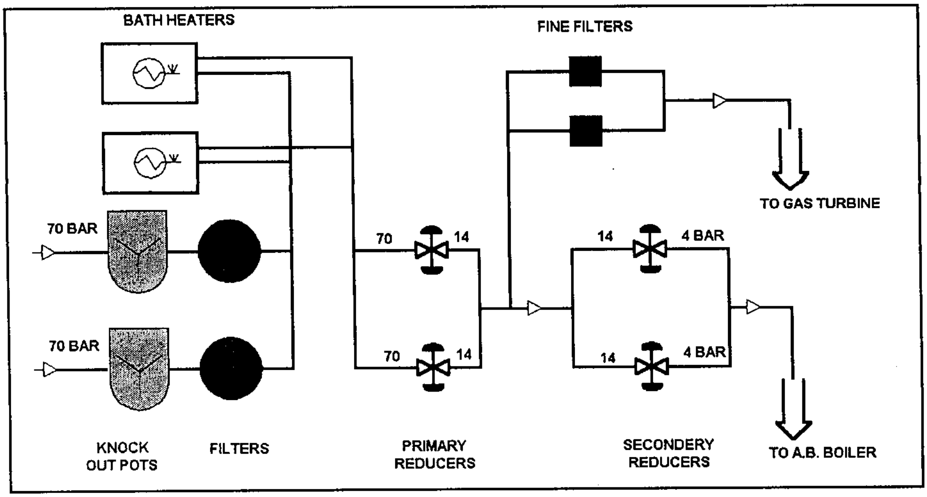

Fig. No. 2: Gas Reducing Station Schematic

NATURE OF FAILURE

The natural gas system had many failures since the commissioning in 1984. In most cases a regulator breakdown of the operating primary stream lead to chain of events causing complete shutoff of gas supply to both Gas Turbine and Auxiliary Boiler.

Factors leading to these situations were identified and the deficiencies were categories as under: -

1. Design defect (system / equipment).
2. Component failure.
3. Maintainability.
4. Human factor.

The gas system was divided into the following sub-systems for easy identification and analysis of various failures.

I. Primary streams.
II. Common vent system.
III. Design deficiencies.

Mode of failure depended on the operating conditions of the plant. Our study therefore took into consideration the following operating conditions for which the plant was designed to operate.

- Both Gas Turbine & Auxiliary Boiler in operation.
- Only Gas Turbine in operation.
- Only Auxiliary Boiler in operation.

I. PRIMARY STREAM FAILURES

Reliability of Regulators

Problem: Failures of internal parts such as valve seat stem, distributor and guide ring due to design deficiencies and poor quality of materials. Refer to Photo nos 1 & 2

Solution: Regulators were replaced with ones having stems of larger diameter. Half-yearly inspection and replacement of valve seat and guide ring and repair of the distributor.

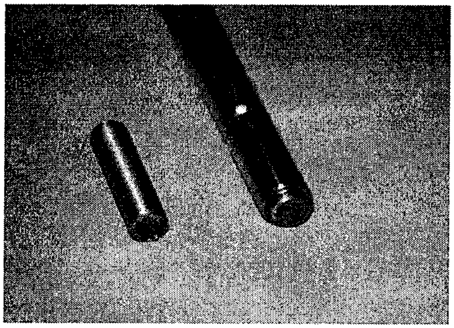

Photo No. 1: Regulator Broken Stem

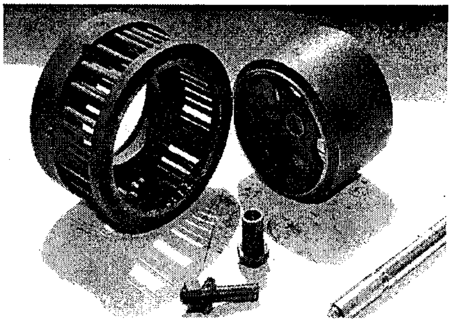

Photo No. 2: Damaged Regulator Distributor

RESPONSE TO LOAD CHANGES

Problem: Regulator does not respond quickly to large drops in load e.g. trip of Auxiliary Boiler or Gas Turbine causing pressure surge in the downstream header. This leads to automatic change over of stream in the event of pressure shooting over 16.5 bar.

Solution: Accelerators have been provided for fast takeover by the monitor and to prevent gas pressure reach high alarm level.

The pilot operated relief valve setting has been reduced to 15.7 bar from original 17 bar to limit the pressure surge.

Problem: Multiple failures of working stream controls cause an increase of header pressure to the trip value (19.0 bar g) which causes the trip of gas station and loss of gas supply to gas turbine and auxiliary boiler.

Solution: Readjustment of one of the relief valves pressure to 18-bar (g) to limit the header pressure below the trip value.

OPERATION PROCEDURES

Lack of detailed written procedures of the "as built" plant from the manufacturer relating to gas stream isolation, trial change-over and on-line testing of regulators has from time to time lead to plant operational mistakes.

Simplified procedures were written for testing & routine changeover of streams to check the integrity of regulators.

SYSTEM DESIGN DEFICIENCIES

Problem: Low capacity of the 14.0 bar header does not sustain the required gas pressure during a low pressure changeover operation of the streams, resulting in gas turbine trip on low gas pressure.

Solution: Change of operation mode to allow both downstream block valves to remain open for parallel operation of the two regulators in service / stand-by mode. Control pressure settings of working and monitor of the regulators were therefore modified to enable standby regulator to take over quickly in case of the service regulator failed close.

Problem: Oversized regulator operates at 10% opening for the full plant load requirement causing frequent failures of rubber seat due to excess stress.

Solution: Half-yearly inspection and replacement of the valve seat as required.

II. COMMON VENT LINE ARRANGMENT.

Problem: During the relief valve operation or manual venting (for regulator testing and adjustment) of 14 bar header, a back pressure is built up in the common vent header. This gas under pressure enters the block valve actuator via solenoid valve exhaust port and causes the block valve of the healthy stream to close. It also disables the function of the relief valves of the secondary stream. See Photo No. 3.

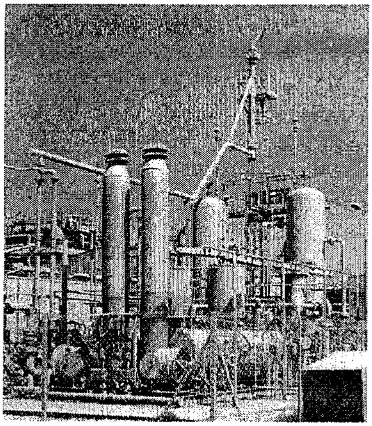

Photo No. 3: Gas Reducing Station Common Vent Line Arragement

Solution:

All primary streams regulators, secondary streams regulators, associated solenoid valves & control components vent lines have been separated by individual lines from the common vent header.

It has been proposed to have separate vent line for relief valves of high pressure and low pressure systems to prevent back pressure interference.

III. OTHER DESIGN DIFFICIENCIES

Problem: Secondary stream hunting due to 14.0 bar line pressure fluctuation.

Solution: Pilot regulator, which receives control gas from 14.0 bar line, is provided with a pre-regulator.

Problem: Bad quality push button switches causing unreliability.

Solution: All the local bush buttons switches have been replaced by quality ALLEN BRADLEY push button switches.

Problem: Due to compact design of the stream pipe-work, primary stream regulators monitoring section is not accessible for removal and inspection in situ.

Solution: Presently the defective regulator is removed with the help of a crane for repair & maintenance in the workshop. A support structure is required to be installed to use chain pulley block for regulator removal and fixing.

Problem: Damage of primary stream reduction regulator seats as a result of operation of the safety valves.

Solution: A 2- inch bypass line for the primary stream may be considered to facilitate testing of safety valves at the time of commissioning.

AUXILIARY BOILER

The Auxiliary Boiler provides steam required by the desalination plant in addition to steam supplied by the Heat Recovery Boiler or the total steam demand during gas turbine and / or heat recovery shutdown. This Boiler works generally in parallel with the Heat Recovery Boiler.

The Auxiliary Boiler consists of, Steam and water drums, Furnace, Superheater, Economizer, Desuperheater, Forced draft fan, Firing Equipment, Burners management system (BMS).

The Auxiliary Boiler is equipped with three dual fuel burners, designed for operation with natural gas or combination of gas and diesel. The combustion is maintained in safe condition and supervised by the "Burners Management System" (B.M.S)

In case of unsafe conditions B.M.S. ensures the prompt intervention of the safety devices to terminate dangerous operation.

FAILURES & DEFICIENCIES

- Tripping of the boiler as a result of disturbance caused by distiller trip.
- Tripping of the boiler when one of the three operating burners shuts off.
- Inadequate diagnostic system for trips & interlocks.

PROBLEMS AND SOLUTIONS

DRUM LEVEL SHRINK

When the distiller trips, there is a rapid increase in steam header pressure due to inadequate design of venting facility. The resulting increase of the drum pressure causes the drum level to shrink tripping the boiler on very low water level.

SOLUTION

A system modification is being considered to install a pilot operated relief valve on the steam header in addition to the existing vent control valve. It may be also required to provide interlock to switch the Auxiliary Boiler to minimum firing when the distiller trips.

TRIP ON HIGH GAS PRESSURE

When one of the three operating burners trips on any fault, burner gas pressure rises at the burner inlet causing the boiler to trip on very high pressure (700mbar).

Solution:

On manufacturer recommendation, the very high gas pressure trip switch setting has been increased to (1200mbar), and the pressure limiter system combustion control has been set to (500 mbar) instead of the original 650 mbar.

DIAGNOSTIC SYSTEM

In the event of the boiler trip, failure analysis can not be quickly and accurately done with existing chart recorders and conventional alarm system.

Solution: A data acquisition & logging system is being installed to records sequence of events leading to burner / boiler trips.

HEAT RECOVERY BOILER

The boiler is a waste heat recovery boiler manufactured by Macchi Milan, Italy, which uses exhaust gases of the gas turbine at the temperature 535°C.

The boiler is designed to work under forced circulation at feed inlet temperature 150°C @ 22 bar (g) to produce 80 Ton/hr. high pressure steam at 215°C @ 14 bar (g).

The heating surfaces of this boiler are located inside the duct. The exhaust gases enters the boiler from bottom passing across the superheater, evaporator and economizer and leaves through the stack.

The boiler consists of the following major elements.

- Boiler drum.
- Single stage finned tubes economizer.
- Double stage finned tubes evaporator.
- Superheater.
- Inlet & Outlet headers.

The inlet and outlet headers housed in individual steel boxes are designed in such a way that they remain outside the hot gas path and separated from the hot path by thermal refractory. The headers are insulated with Rockwell insulation covered by steel plate.

NATURE OF FAILURE

The economizer tubes are connected to the feed inlet header through "S" shaped welded bends (total number of 18 tubes) Refer Photo Nos. 3 & 4. Recurrent failures were experienced in the following areas: -

- Cracks at the "S" shaped bend welded joints to the header.
- Cracks at the "S" shaped bend welded joints to the tubes.
- Cracks at the header body itself at various locations.

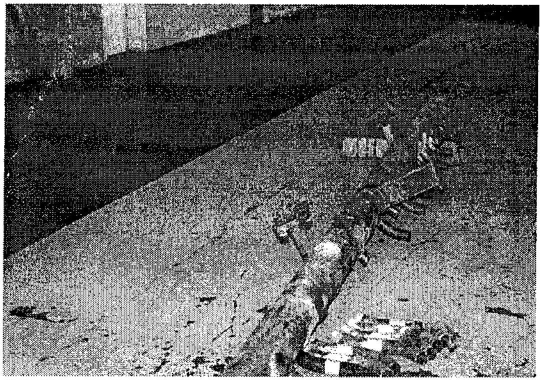

Photo No.4: Heat Recovery Boiler Eco. Feed inlet header

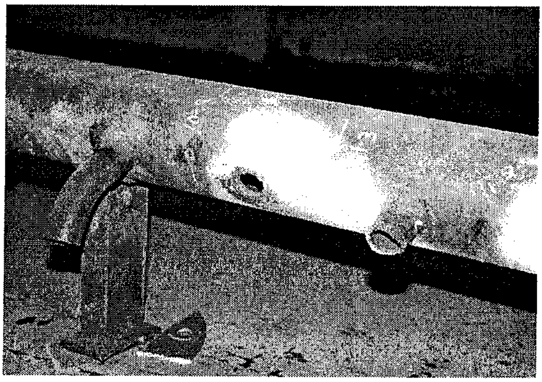

Photo No. 5: Heat Recovery Boiler Eco. Feed inlet header Cracks

FAILURE ANALYSIS

- The header support plate was found deformed leaving a gap of 4mm at the middle support guide causing uneven expansion of the header as well as the connected tubes, which lead to subsequent cracks.

- The 2mm clearance at the bottom as well as side gaps of the header expansion guide support, were found packed with hard scales of sulphur deposit preventing free expansion of the header and creating cracks at various areas of the header itself.

- Flue gas sulphur deposits and scales formation around the tubes fins specially at the area of the coil support beams near the U bends, prevented the free expansion of the coil, which lead to crack at the weaker portion of the "S" bend weld joints.

- Due to failure of the refractory at the welded joints of the tube "S" bends the sulphur deposit from the escaping flue gas caused pitting corrosion at the weld joints which ultimately lead to cracks.

CORRECTIVE MEASURES

- The header basement support plate was reinforced at the middle portion by welding a steel plate and the excess gap of 2mm at guide support was filled with a shim steel plate

- The header expansion guide supports were cleaned to maintain expansion clearances. Annual PPM was introduced to remove the scales during unit outage.

- Introduction of Non Destructive Testing (Magnaflux and Dye Penetrate Testing) for the header and welding joints during unit outage to detect any developing defects.

- Deposits accumulated between the coil support beam and the tube fins that were preventing the free expansion of the tubes were cleaned. Annual PPM was introduced to avoid scale accumulation.

- In order to avoid cracking of the refractory across the tubes, the "S" bends were provided with aluminum sleeves to allow relative expansion.

- However, in order to avoid catastrophic failure of the damaged header and as part of quality assurance and system reliability program the header along with the "S" bends was replaced with a new one. All the above preventive measures are being taken to prolong life of the new header.

DISTILLER PHASE NO. II

The desalination unit is multistage flash process (MSF) with cross-flow brine recirculation. The evaporator consists of 16 stages, 13 in the heat recovery section and 3 in the heat reject section, arranged in one block including the dearator.

The flash chambers are constructed of carbon steel. The three top temperature stages are protected against corrosion by copper nickel cladding, and the remaining chambers are coated with special epoxy coating.

The main distiller shell is fabricated from 20.0mm carbon steel plates except first two recovery stages, where the external thickness is 40.0mm.

NATURE OF THE PROBLEM

On the top of the distiller roof there are twenty-two (22) equi-spaced I beams longitudinally located across the width of the distiller. The structural importance of these beams is external stiffening of the stage roof top plate section, which in turn provide support for the tube bundles. The same beams also carry external pipe work and walkway gallery.

The I beams at the top of heat recovery stages were found in the advanced stage of corrosion to the extent that with the exception of part of the outer beams, total separation had occurred between the web and the lower flange. This was discovered during the 1994 outage cycle.

Also the distiller experienced several failures due to internal corrosion of pipes of sea water supply, brine and blowdown systems.

PROBLEM ANALYSIS

- Preliminary inspection, shows that the roof sections provided a catchment for rainwater which seeped through the insulation and stayed over the roof plate and supporting structures for long duration. Due to the lower temperature in the reject stages, this water with high chloride salts corroded these steel structures. These structural sections, subjected to downward force due to vacuum were found deformed especially at the central sections, see Photo Nos.6 & 7.

- The seawater and brine water pipes of wrong material quality were corroding and causing frequent leaks and pipes failures.

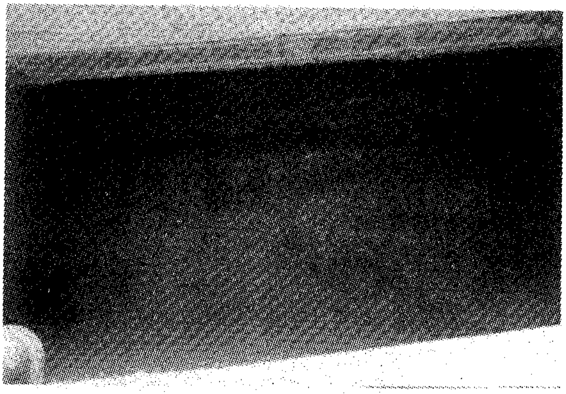

Photo No.6: Distiller Roof I Beams Corrosion

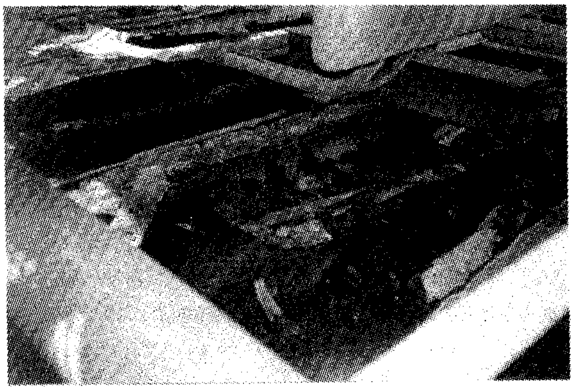

Photo No. 7: Distiller Roof Excessive Corroded Area

CORRECTIVE MEASURES

- Extensively damaged I beams were replaced during 1995 unit outage to avoid roof collapse.
- Removals of lagging material from the rooftop to prevent avoid water accumulation and limit the corrosion.
- Ultrasonic thickness measurement of roof plates for assessing the corrosion rate.
- Replacement of all affected I beams above the heat reject stages during 1998 unit outage.
- Sand blasting of all roof I beams, plates and painting with anticorrosive structural paint during 1998 unit outage.
- Future provision of drain lines from vacuum system to avoid water accumulation on the roof.
- Feasibility study to install sloped roof cladding to drain rainwater.
- Introduction of annual PPM to inspect the integrity of the structure.
- The existing sea and brine water carbon steel pipe lines which are less than ND 300 were replaced with S.S pipes and the pipes greater than ND 300 were internally cladded with 2MM S.S plates. The work was carried out in stages over a period of five years and completed in 1998.

LESSONS LEARNT FROM OUR EXPERIENCE

a) Before finalizing selection of critical equipment, other users should be consulted about the equipment performance & reliability.

b) Designing the system with adequate workspace should ensure equipment maintainability.

c) Detailed operation and maintenance service manuals of "as built" system will help in achieving reliable operation and quality maintenance.

d) Data acquisition system and sequence of events recorder should be provided for better understanding of plant problems.

e) Visual inspection and NDT of boiler headers should be carried out during annual outages to detect any sign of failure at an early stage.

f) The roof plates of the distiller evaporator should be kept dry. A proper drain system for the rainwater and other leaks to prevent seepage of water through the insulation cladding should be incorporated in the structural design.

g) Random inspection of sections of insulated steel structure should be annually done. Any damage of the protection painting will require more extensive inspection to plan for proper repair.

QUALITY IMPROVEMENT PROGRAM

One of the objectives of the management at Sitra Station to achieve zero defects has necessitated development and implementation of quality improvement programs that are designed to improve quality of maintenance and operation activities and to eliminate plant design deficiencies. The essential elements of such program are establishing management teams, analyzing equipment /system failures, tracing failures to their basic or root causes, eliminating the root causes of failure and continuing the program in an ongoing manner. These programs are briefly described below.

Quality circle has made major contributions towards quality improvement. It consists of shop floor workers who offer ideas and concepts for solutions of problems.

Reliability Committee consisting of trained and experienced engineers meet on a regular basis to analyze equipment /system failures, trace the failures to their basic or root causes and recommend actions to eliminate the root causes of failures.

Maintenance performance indicators, being used to measure the effectiveness of maintenance functions, are the numbers of reported defects over a period of time and the number of preventive maintenance jobs carried out over the same period. Other indicatores are forced outage rate, availability and reliability figures.

Corrosion is a major source of plant /equipment degradation. The system at present is being developed for monitoring and control of corrosion of structures and critical equipment.

Hidden defects can lead to catastrophic failures. During plant annual outages the inspection & NDT are carried out. The results of inspection with recommendation are submitted for short and long-term actions.

Condition based maintenance is now well introduced and essentially applied to rotating machines.

Auditing System requires that procedures, equipment/system manuals and failure records be reviewed at regular intervals to determine if failure to achieve a goal needs strict enforcement of existing procedures.

CONCLUSIONS

1. Application of quality control system for project stages of design, selection of material, installation, commissioning and continuous operation will lead to plant reliable performance.

2. Early application of the techniques of hazard analysis, failure mode and effect analysis, fault tree analysis is an effective tool to reduce equipment failure.

3. Development of an inspection procedure for plant integral parts to detect hidden deterioration and faults which may lead to enventual failures and costly maintenance at a later stage.

4. Quality improvement program should be integrated within plant management structure to fulfil reliability objectives.

REFRENCES

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2. S. Kalaf and Al-Alawi “Maintenance Philosophy and approach for dual purpose plant” Desalination, 1996 Elsevier Science B.V. Amsterdam.

Four Decades for the Multi-Stage Flash Desalination

Hisham El-Dessouky, Hisham M. Ettouney and
Faisal Al-Juwayhel

FOUR DECADES FOR THE MULTI-STAGE FLASH DESALINATION

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ABSTRACT

Worldwide, the multistage flash desalination (MSF) dominates the desalination industry with a market share of more than 60%. Since inception in the late 50's the MSF process have gone through several developments, which resulted in drastic cost reduction. The unit capacity has increased from 0.5 mgd to a current conventional capacity of 6-7.2 mgd. More recently, larger units with double the capacity (14.4 mgd) are commissioned in UAE. Process economics show that doubling the capacity may result in 20-24% reduction in unit product cost. Simultaneously, innovative design and optimization allowed for reduction in the specific unit volume, the specific heat transfer area, and the thickness of metal components. In addition, progress has been achieved in developments of chemical additives that prevents scale formation and allow for longer periods of operation. Accumulated experience and better understanding of system design and operation resulted in careful selection of various materials forming the system, i.e., tubes, tube plates, shells, valves, demisters. As result, corrosion and erosion rates are reduced to values that prevent system failure. Irrespective of these developments, the MSF process faces challenges from other competitive technologies, especially, reverse osmosis, mechanical vapor compression, and multiple effect evaporation. These systems are more energy efficient and have lower unit product cost. Therefore, achievement of new developments and innovations in the MSF process resulting in reduction in unit cost are essential to meet the market demands. In this regard, recent literature studies and conceptual designs include use of larger unit capacity (20-30 mgd), use of thermal vapor compression, adoption of hybrid configurations, or use of brine mixing MSF. Other attempts for cost reduction include use of inexpensive plastic tubing for the condensers, reinforced plastic for the stage walls, thin demister pads, and higher specific brine flow rate.

KEYWORDS: Seawater desalination, thermal processes, multistage flash desalination, desalination technology,

INTRODUCTION

Today the world production capacity stands at 22.8x10^6 m^3/d, with more than 12,500 operating desalination units, Wangnick, 1998. The world leaders in the desalination industry are the Kingdom of Saudi Arabia (KSA), USA, United Arab Emirates, and Kuwait with market shares of 27%, 15%, 11%, and 8%, respectively. The desalination in KSA, Kuwait, and UAE, is dominated by the multistage flash desalination (MSF) with percentages of 66%, 96%, and 90%, respectively, of their total production capacity. The design and operating features of the MSF process makes it capable of withstanding the harsh conditions found in the Gulf environment. The temperature of the Gulf water varies over a wide range from 35 to 12 °C, between the summer and winter seasons. Also, the air temperature reaches highs close to 50 °C during the long summer time, which extends from April to end of October. In addition, the Gulf seawater has high salinity, which may vary from lows of 42,000 ppm to highs that may reach 64,000 ppm. On the other the hand, the reverse osmosis process (RO) dominates the desalination industry in the USA. This is because the RO plants operate on low salinity water, which includes mostly brackish or river water sources. The features of the MSF process and the accumulated technical experience since its first installation in 1957 makes it the leading desalination process worldwide with a total market share of 60%. The second leading process is the reverse osmosis (RO) with a market share of 30% and the remaining 10% is divided among the multiple effect evaporation (MEE), mechanical vapor compression (MVC), and electro-dialysis (ED).

In 1957, the first flash desalination system was installed in Kuwait, Silver, 1980. During the same year, R. S. Silver made the first patented description of the MSF. The Kuwait installation was a once through system with 4 flashing stages and was constructed by Westinghouse. The unit had a large specific heat transfer area to provide a system performance ratio of 3.3. Subsequent success of the MSF process was mainly due to development of the system beyond the expensive form of the Westinghouse installation. This is achieved by optimizing the system design, which dictated increase of the number of stages to 10. Although, the optimized system had a performance ratio of 3.3, its cost savings came from reduction in the specific heat transfer area. Another factor that affected the production cost was the total unit capacity, which increased gradually over the years from 0.5 mgd to higher values of 6, 7.2, and 14.4 mgd. This increase was also associated with decrease in the specific heat transfer area and volume of flashing stage, Silver, 1980. Other gains were also achieved in system operation.

Temperley, 1995, showed that earlier units could not be operated for more than 600 hours and the heat transfer coefficient fell to less than 30% of the design value within 480 hours of operation. Since, improvements in antiscalent materials resulted in the increase of the plant to values close to 90%, during which time the production characteristics are close to the design conditions. Reports for the MSF plants in the Gulf countries show that availability varies from lows of 73% to highs of 86%. Plant shutdown occurs for scheduled overhauls and malfunctions. Commonly overhauls are performed during wintertime, when demand for water is lower and the availability is close to 80%. On the other hand during the summer time overhauls are restricted and the plant availability is around 88%. Developments in the additive materials and use of tube materials with low corrosion rates allowed for operation at top brine temperatures close 110 °C and use of acid cleaning for scale removal. Increase of the top brine temperature resulted in several operating advantages, which increases the system performance ratio, reduces the specific heat transfer area, the amount of makeup seawater, and consequently the associated pumping power. Evaluation of the desalination industry, Temperley, 1995, shows that the process has come a long way since the fifties. Today, the process has developed to a point, where life in the Gulf countries, many of the Caribbean and Mediterranean islands, and other arid regions around the world are fully supported by the desalination industry. Temperley, 1995, urged the desalination community to continue the research process and to develop comprehensive education and training programs for the desalination staff. This last point was further addressed by El-Dessouky et al., 1998a, in his evaluation for a comprehensive program for qualifying manpower for the desalination industry. The program by El-Dessouky et al., 1998a, includes elements for higher education, field and intensive training, and research.

To evaluate the role of research on advancement of the desalination processes, it is essential to consider achievements made by the research institutions in the USA. Between 1960 and 1980, a total $322.3 million dollars was spent on research or the equivalent of $1.7 billion in 1998 dollars, Hightower et al., 1994. These funds were used to realize the several achievements, which includes patenting and use of membranes for separation, development of the desalting membrane technology, development of the first MSF and multiple effect evaporation units, development and testing of construction materials, and investigation of several innovative concepts. Between, 1981 and 1994 research funds were reduced considerably, however, recently the US Congress authorized a new program for water desalination research and development. The program is funded for a six-year period beginning October 1997 with $3.7 million for fiscal year 1998.

This paper focuses on developments in the MSF system since the 1957 Kuwait installation. The discussion gives an account for current practice in various areas of the process. These include the process layout, production capacity, performance ratio, chemical treatment, product cost, materials of construction, and operating conditions. The last section is devoted to process innovations, which looks at future outlook for further development.

CONFIGURATION

The original MSF configuration was the once through arrangement, Figure 1. In this configuration the intake seawater is deaerated and chemically treated before flowing through the condenser tubes of the heat recovery section. The once through configuration has several operational draw backs; especially, lack of control on the intake seawater temperature, use of large amounts of chemicals to treat the large flow rate of the raw seawater, and the loss of large amount of energy in the large stream of brine reject. As a result the once through process could not hold against the brine circulation system and was limited to a small number of land base plants as well as low capacity and low performance ratio marine units, Tusel et al., 1997.

The main features of the brine circulation MSF, Figure 2, include

- Division of the flashing stages into a large number of heat recovery stages and a smaller number of heat rejection stages (commonly three).
- The excess energy added to the system by the heat steam is rejected to the cooling seawater in the heat rejection section.
- The remaining part is the makeup or feed seawater is mixed in the brine pool of the last stage in the heat rejection section.
- Part of the brine stream in the last stage of the heat reject section is circulated and flows through the condenser tubes in the heat recovery section.
- Part of the warm cooling seawater stream is used to control the temperature of the intake seawater, especially, during the wintertime. This feature is essential to maintain constant production capacity and stable operation.
- Treatment of the intake seawater stream is limited to simple screening and filtration.

The brine recycle stream is deaerated and chemically treated by addition of antiscalent, antifoaming, and other chemicals.

Although the number of stages and the production capacity of the MSF system are increased over the years, the brine circulation layout did not experience noticeable variations. Differences come in the tube arrangement, orifice design, demister and distillate tray configuration, water boxes layout, and stacking of the flashing stages. The condenser tubes are arranged in either cross or longitudinal direction to the brine flow. Neither arrangement has a distinguishable effect on the system performance. Each configuration requires different procedures for design, construction, cleaning, and maintenance. Irrespective of this, the majority of the existing MSF units have the cross tube arrangement. However, the large MSF configurations with capacities of 14.4 mgd utilized the long tube configuration, Al Radif, 1994.

PRODUCT UNIT COST

The product unit cost has decreased over the years as a result of developments in various technological aspects of the process. The breakdown of the product unit cost includes capital amortization, energy, labor, spares, chemical, and maintenance. More than 89% of the MSF capital, excluding the condenser and brine heater tubes, covers civil work, foundations, intake systems, flashing stages, instrumentation, controls, and monitoring station. The tubing cost may account for as little as 8% of the total capital, Hornburg et al, 1995.

Table 1 shows variations in the unit production cost from available literature data. As is shown a decreasing trend is observed with earlier values of $\$2/m^3$ and more recent values of $\$1/m^3$. As unit production cost is affected by the plant factor, interest rate, the energy cost, total production capacity, and system design. The increase in the plant factor is due to several reasons, which include long operational experience, better understanding of operating problems, development of more efficient chemical additives, and careful selection of construction materials. As for the last two factors, the new generation of chemical additives introduced in the late eighties allowed for continuous operation for long periods up to two years. However, during this period production may cease for reasons other than tube failure or complete scale blockage as it used to happen in early years of MSF. For example, units can be put in a standby mode as a result of unavailability of heating steam, loss of vacuum, trip of power source, increase in product conductivity, or need for on-line acid cleaning, Al-Gobaisi, 1994.

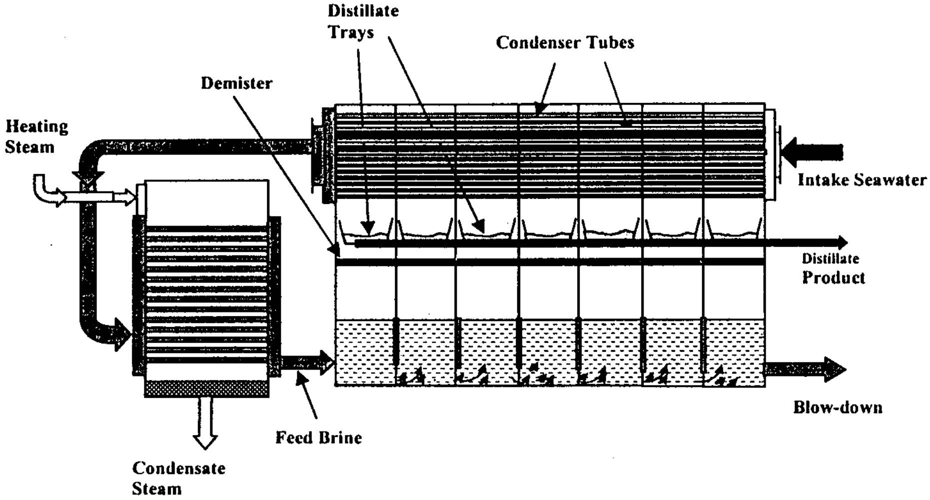

Figure 1: Once-Through Multi-Stage Flash Desalination Process

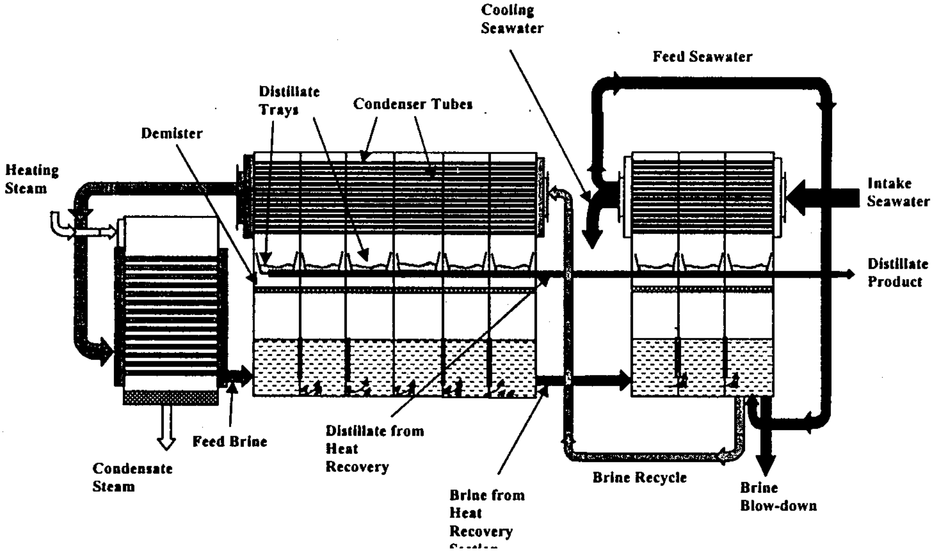

Figure 2: Multi-Stage Flash Desalination Process

Table 1: Variation in unit production cost for the MSF process

Water Cost ($/m^3)	Source
2.332	Darwish, 1989
1.86	Morin, 1993
1.854	Al-Mutaz, 1996
1.41	Sadhukhan and Misra, 1996
0.8	Bednarski and Minamide, 1997
0.55*	Wangnick et al., 1997
1.22	Leahy, 1998

* once through, 22 mgd unit capacity, conceptual design.

CAPACITY

The capacity of the MSF has increased from low values of 0.5 mgd in 1957 to current value of 6, 7.2 and 14.4 mgd, Table 2. Between 1957 and 1965, the capacity for the earlier designs of MSF varied between 0.5 and 1 mgd and the number of stages limited to 19. Subsequently, the capacity was doubled to 2 mgd in 1968 and to 4 mgd in 1970. The current conventional capacity of the MSF units of 6-7.2 mgd first appeared in 1985-1986. This capacity remains to dominate the MSF market, however, units with 14.4 mgd capacity appeared in 1996 and were installed in UAE.

Today, most of the MSF units are brine circulation, have a capacity of 6-7.2 mgd, and have 24 stages with 21 stages in the heat recovery section and 3 in the heat rejection section, Figure 2. This design has a control loop on the water intake to maintain its temperature within design limits during winter operation. This is necessary to maintain stable operation of the system and to prevent reduction of the system production capacity and performance ratio. As is shown in Figure 2, part of the reject cooling seawater is recycled and mixed with fresh intake seawater. Increase in the system capacity during the past four decades was restricted by technological limitations on manufacturing of larger size pumps and tubes of special materials and dimensions. For example, scale up of flashing units would require special orifice design, evaluation of the flow field pattern, release velocity of vapor, and entrainment rates of brine droplets. Pumping units should provide sufficient speed within the tube to maintain high heat transfer coefficient and to prevent settling of fouling materials. The larger vapor space in a double capacity unit may suffer from formation of dead zones, where non-condensable gases may accumulate and may cause increase in the unit pressure. Irrespective of this, doubling the system capacity has been and will always be an attractive means for cost reduction. New and old economic studies for performance evaluation of larger unit sizes show that doubling the capacity can lead to cost reduction of 20%, Wangnick et al., 1997.

Table 2: Developments in MSF capacity, number of stages, and performance ratio

Year	Number of stages	Capacity (mgd)	Performance Ratio
1957*	4	0.5	3.3
1960	19	1	5.7
1960	40	0.61	10
1962	40	1	10
1965	24	1	7-9
1968	24	2	7-9
1970	24	4	7-9
1978	24	5	7-9
1985	24	6	7-9
1985	24	7.2	7-9
1996+	24	14.4	7-9

* Once through, + long tube.

OPERATING CONDITIONS

The main operating parameter in the MSF system is the top brine temperature, which sets the system flashing range, the thermal performance ratio, the specific flow rate of brine circulation, and the specific power consumption. The top brine temperature varies over a range of 90-120 °C. Selection of a specific value depends on the operating range of the antiscalent additives. Operation at higher temperatures has the advantage of decreasing the required heat transfer area as a result of higher temperature differentials and heat transfer coefficient. Also, at higher top brine temperatures, lower flow rates of the intake seawater and brine circulation are used to generate the same amount of distillate product. As a result, lower amounts of chemical additives are used and the pumping power is reduced. The drawbacks of high temperature operation include the use of higher pressure heating steam, which comes at a higher cost and may lower the performance of the co-generation power plant. Also, corrosion, release rates of non-condensable gases, and thermal expansion problems become more evident at higher temperatures. Conventional MSF systems operate over a temperature range of 90-110 °C.

The shell load, which is defined as the flashing brine flow rate per unit shell width, is an important design and operating parameter, since it affects the residence time of the flashing brine in the flashing stages. The shell load varies over a range of $1.5\mathrm{x}10^5$-$1.5\mathrm{x}10^6$ (kg/h)/m. In large units with long tube configuration the shell load varies between $1.16\mathrm{x}10^6$ to $1.26\mathrm{x}10^6$ (kg/h)/m, while the cross tube configuration has lower values of $0.88\mathrm{x}10^6$ (kg/h)/m. Increase of the shell load reduces the flashing brine residence and as a result lower amounts of vapor are released. In other words, at low residence times the brine does not have sufficient time to equilibrate with the released vapor as a result of higher non-equilibrium losses. The residence time of the flashing brine also affects the vapor release velocity, or vapor disengagement velocity, from the brine surface. This velocity is limited to a maximum value of 9 m/s, in order to avoid high entrainment rates of brine droplets in the released vapor.

The vapor velocity in the space of the flashing chamber is an important design parameter, which is used together with the flashing rate, stage pressure, and stage width to determine the stage length. The vapor velocity is limited to a maximum value of 4 m/s, to avoid entrainment of brine droplets and reduction of the product quality. Also, higher vapor velocities would result in re-entrainment of the brine droplets which accumulates in the demister pad. In the heat rejection section, the low stage temperature reduces the pressure and increases the specific volume. As a result, the vapor velocity is the highest among all stages.

Other operating parameters that were evaluated during system development include effect of the pH and the amount of dissolved gases (oxygen and carbon dioxide), in the condensate and brine streams. The flashing process is associated with the release of the non-condensable dissolved gases. In the heat recovery section, improper venting would eventually lead to accumulation of these gases, which would reduce the heat transfer coefficient, the product condensation rate, and the net production capacity. More importantly, the carbon dioxide gas would dissolve in the condensate film, which in turn would lower its pH and promotes corrosion reactions. McGregor, 1992, reported uniform corrosion rates of more than 100 μm for Cu/Ni condenser tubes. Measurements show formation of carbonic acid and reduction of the pH of the distillate film to values below 5. This condition is resolved by proper venting procedures, dearation of the recycled brine, and chemical treatment for pH adjustment. Another approach is to use the more expensive titanium and high steel alloys, which increase in the tubing cost by more than 20%, Hornberg et al., 1995.

The use of high brine velocities inside the condenser tubes increases the heat transfer coefficient and reduces the fouling rate. Irrespective of this, high velocities are associated with increase in the pressure drop, tube erosion, and pumping power. Although the brine velocity for a clean system can be within safe limit, which do not cause erosion, fouling and marine growth can lead to reduction in the net flow area and subsequent increase in the brine velocity. The condenser tubes in the heat rejection section are prone to such conditions, since treatment of the intake seawater is limited to simple screening and filtration. Use of alloy materials limits the brine velocity to 4 m/s in 90/10 Cu/Ni and 5 m/s in 70/30 Cu/Ni, however, practical limits are within 1.5-2.3 m/s. Use of the more expensive titanium and high steel alloys would allow for increase in this limit to more than 20 m/s, without noticeable erosion effects, Hornburg et al, 1995.

PERFORMANCE RATIO

Developments in the MSF performance ratio over the past four decades are shown in Table 2. The early designs had performance ratios of 3.3. This value gradually to higher values and currently it stands at an average of 8. However, system design may result in performance ratios of 8.5 or even 10, in actual practice the average performance commonly reported is 8 and in some instances lower values close 7 are even reported. The MSF performance ratio is relatively low when compared to the performance ratios of multiple effect vapor compression systems, which varies over a range of 12-21, Ettouney et al., 1998. Also, conversion of the MSF performance ratio into specific power consumption shows that the MSF currently stands at a value of 20 kWh/m^3, Darwish et al., 1989. This value is 2-3 times more than the specific power consumption for reverse osmosis or mechanical vapor compression. Irrespective of this, the redeeming feature of the MSF system is its reliable operation, high plant factor, and huge field experience. Moreover, increase in the system capacity by more than 28 folds, i.e., from 0.5 mgd in 1957 to 14.4 mgd in 1996, has resulted in continuous reduction in the unit product cost (each doubling in unit capacity is associated with 20% savings in unit cost).

MATERIALS OF CONSTRUCTION

A list for common construction materials for the MSF system is shown in Table 3. Currently most of the MSF plants utilize Copper/Nickel tubes 90/10 and 70/30 for construction of the condenser tubes in the heat recovery, heat rejection, and brine heater. Occasionally, titanium tubes are used in the heat rejection section. The Copper nickel tubes have 0.9 mm wall thickness and its corrosion rates give a lifetime of 20-25 years. This result is reported in early and more recent studies, El-Dessouky and Khalifa, 1985, and Hornburg et al., 1995. However, recent environmental considerations dictate replacement of the copper based tubes, which are found in a large number of plants. In this case, the copper element contaminates the reject brine stream and causes a strong impact on the environment, Hornburg et al., 1995. The alternate is use high alloy stainless steel and titanium, which can withstand the harsh conditions and do not contaminate the product or the rejected brine. Common practices calls for use of combinations of various tubing materials to reduce the tubing cost. For example, the more expensive 70/30 Cu/Ni is used in the brine heater, the heat rejection section, and the first and second stage of the heat recovery section. This is necessary to withstand the effect of non-condensable dissolved gases, high temperature, and limited treatment of the intake seawater. In the remaining stages of the heat recovery section cheaper tubing materials can be used, i.e., aluminum brass. However, if acid treatment is applicable the tubes are made of 90/10 Cu/Ni. Today, in many installations, titanium tubes are used in the heat rejection section, to overcome all types of operational problems and in turn to keep continuous and steady operation. Other drawbacks for use of the copper tubes occurs during startup of new plants, since all the product stream is must be rejected until the copper content in the product water is reduced to safe levels, Al Sum et al., 1994.

Table 3: Common materials used in MSF systems

Part	Material
Flash Chambers operating above 85 °C	Carbon steel fully clad with 2 mm SS 316 L or 90/10 Cu/Ni
Flash Chambers below 85 °C	Carbon steel fully clad with 2 mm SS 316 L or 90/10 Cu/Ni
Internals of flash chambers (orifices, weirs, splash plates, inter-stage walls, demister supports)	Stainless steel, SS 316L
Distillate trays and tube support plates	Stainless steel, SS 316L
Heat recovery tubes, stages 1-3	Cu/Ni 70/30
Heat recovery tubes, other stages	Cu/Ni 90/10
Heat rejection tubes	Cu/Ni 70/30
Tube sheets	Cu/Ni 90/10
Water boxes	Carbon steel clad with Cu/Ni 90/10
Brine heater shell	Carbon steel
Brine heater distillate well	Cu/Ni 90/10
Brine heater tubes	Cu/Ni 70/30

CHEMICAL TREATMENT

Use of chemical additives came a long way since the early years of MSF operation, Temperley, 1995. Chemical additives are primarily used to prevent scale formation, foaming, and corrosion. Scales include alkaline compounds, $CaCO_3$ and $Mg(OH)_2$, and non-alkaline materials include $CaSO_4$, $CaSO_4.1/2H_2O$, and $CaSO_4.2H_2O$. For prevention of alkaline scale two methods are used; acid dosing and antiscaling additives. Sulfuric acid is used for acid dosing. Earlier developments of antiscalent materials focused on development of materials that would prolong the number of operating hours before scaling problems dictates system shutdown and cleaning. The first commercial antiscalent was the polyphosphate, which operates at top brine temperatures below 90 °C. This is followed by development of Belgard EVN in 1984, which can withstand higher top brine temperatures. Success of these materials turned the focus on development of less expensive antiscalent material with similar or superior performance. In 1988, a cheaper material and superior product is put in use, which is the Belgard 2000, Shams El Din, 1988, Finan et al, 1989. More recently, the commercial antiscalent known as Sokolant is put in use with further reduction in its unit cost versus the Belgard 2000. Success of the antiscalent materials is manifested in the data on plant shutdown causes reported by Al-Gobaisi, 1996. In his report, more than 50% of plant shutdown for installations before 1977 was due to scale and tube failure related problems. This percentage drops to 0% in installations after 1990.

STAGE CONFIGURATION

MSF flashing stages includes various elements; primarily condenser tubes, stage walls, demister, distillate trays, venting devices, brine orifice, ball cleaning devices, and instrumentation for measuring temperature, pressure, pH, and product conductivity. The stage walls are designed to withstand pressures up to 3 bars. The stage should provide enough area for vapor disengagement, with vapor velocity below 4 m/s. Also, a height of 2.13 m between the flashing brine surface and the demister is necessary to allow for settling of the brine droplets. The cross section area of the stage should provide enough area for the demister to keep the vapor velocity within the demister below 9 m/s to avoid entrainment of brine droplets, El-Dessouky, et al, 1999a. In addition, the stage should provide enough volume to house the condenser tubes and to allow for proper venting of the non-condensable gases. The stage width is determined by the brine orifice load, which is defined as the flow rate of the brine per unit width of the stage. On the other hand, the necessary demister area, the length of the tube bundle, and the vapor velocity, affects the stage length. In typical MSF plant in the Gulf area with a production capacity of 6 mgd, corrosion allowance for the chamber walls in the first and last flash chambers is 12.7 mm and 9.3 mm for other chambers. The wall thickness is 38.1 mm and the stage dimensions are 15.8 m wide, 3.61-3.72 m length in heat recovery, and 3.62-2.9 m length in the heat rejection. The stage height in the upper deck is 3.9 m and in the lower deck is 4.3 m.

The brine submerged orifice or the interstage device controls the flow of the flashing brine between stages and its residence time with the stage. Orifice shape can be a simple rectangular opening in the wall separating two adjacent stages. A splash plate or weir is used to increase the water depth in front of the orifice to insure that the orifice is fully submerged.

Demisters are an essential part of flashing stages and its operational or design problems can lead to high product salinity, contamination of the surfaces of the condenser tube, and increase in the pressure drop, El-Dessouky et al, 1998a. The first effect is a result of excessive entrainment of the brine droplets by the vapor stream. As this condition arises as a result of blockage of the demister pad, therefore, the vapor velocity increases within the demister pad and rentrainment of the brine accumulated in the demister. As a result, the product stream is discarded and a plant shut down is essential in order to clean or replace the demister pads. Increase in pressure drop also occurs due to poor demister design or demister blockage. This reduces the pressure of the vapor stream and consequently the vapor temperature. As a result, the temperature of the brine stream flowing inside the tubes does not increase to the desired levels and eventually higher amounts of steam must be used to reach the design value for the top brine temperature. The demister specific surface area varies between 0.5-1.5 $m^2/(kg/s)$, while the demister vapor velocity is limited to 9 m/s and vapor release velocity is limited to 4 m/s.

MODELING

Modeling of the MSF is developed in various areas, which primarily include design and rating models based on system material and energy balances as well as a set of well tested correlations for the heat transfer coefficient, thermodynamic losses, and physical properties. These models generate design data which includes the heat transfer areas in the condenser and the brine heater, temperature profiles, salinity profiles, and flow rates. The models may vary from the simple, which constitutes a set of closed form equation, to the very detailed, which contains nonlinear equation that require iterative and computer solution. Review of the developments in the MSF design and rating models include the following:

- Silver, 1970, developed a simple model to express dependence of the MSF performance ratio on the thermodynamic losses.
- El-Dessouky et al., 1985, showed the effect of the thermodynamic losses on the system parameters that affect its final cost, i.e., the performance ratio and the specific heat transfer area.
- Detailed mathematical models, which are based on stage to stage calculations and require computer iterative solution, are developed by Omar, 1983 and Khan, 1986.
- A tri-diagonal matrix solution to improve convergence of the computer iterative procedure and to reduce computation time is developed by Helal et al, 1986.
- Husain, 1988, showed that fitting of model results to the plant data should take into consideration heat losses in each stage, vapor leak, presence of non-condensable gases, and evaporation from the distillate trays.
- Darwish, 1991, performed thermal analysis of the MSF system to arrive at quantitative assessment of the effect of design and operating parameters on the system performance.

- El-Dessouky et al, 1995, developed a stage to stage model which takes into consideration the effect of non-condensable gases, losses to the surroundings, dependence of the thermophysical properties on stage properties. The model includes detailed account for evaluation of the heat transfer coefficients in the brine heater and the stage condensers. Results are found to compare favorably against industrial data.
- A fixed-point iteration scheme is developed by El-Dessouky and Bingulac, 1996, to simplify problem formulation and computer coding.
- El-Dessouky et al., 1998b presented a step by step mathematical development for the origin of the MSF process. The development starts at evaluation of the performance of a single stage flash unit and goes through evaluation of the once through, brine mixing, and the conventional MSF systems.

INNOVATIONS

Innovations in MSF address several important issues including the increase in the performance ratio beyond a value of 8, increase of the unit production capacity, use of new construction materials, and further development in the chemical treatment.

El-Dessouky and Ettouney, 1999, proposed increase in the thermal performance ratio through the use of the vapor compression technique. This system includes entrainment part of the vapor from each stage in the heat rejection, which is compressed in a steam jet ejector, Figure 3. This result in reduction of the energy rejected by the system into the cooling water stream and as a result increases the system the performance ratio. Another form for the system can be based on mechanical vapor compression, where the entire vapor formed in the rejection section is compressed mechanically. Heating of the feed seawater takes place by exchange of heat with the brine reject and product distillate streams. Practicality of this approach remains questionable since existing compressor technology can provide up to 10 °C increase in the temperature of the compressed vapor. This may require the use of a multi-stage compressor configuration, which may not be simple to operate and expensive.

Another innovative approach that always seemed to improve the MSF performance and reduce its product unit cost is the increase in its production capacity. The latest increase in the unit production capacity is the jump from 7.2 to 14.4 mgd. Irrespective of this increase in the unit capacity may not be the ultimate answer, since it would require installation of expensive standby units. Also, replacement of existing unit takes long period of time. At the present time, the majority of the MSF units are the 6-7.2 mgd units. Replacement of these units is subject to the success of the 14.4 mgd units, which have been commissioned in UAE in 1996.

Development of very larger scale MSF units, with capacities greater than 20 mgd, have been in focus the OSW studies in sixties to the more recent study Wangnick et al., 1997 and Borsani et al, 1995. The OSW studies included design of units with capacities between 50 and 250 mgd, though, these projects never materialized commercially. Constraints for designing large MSF plants are high capital, 1.5 to 2 billion, large operation maintenance team, a minimum of US$ 100, and a unit capacity of 100 mgd would required a power plant of 500 MW. An important operational problem in such large system is to maintain uniform temperatures in the tube bundle. In addition, manufacturing and transportation of long tubes, 25-28 m, may prove difficult to overcome.

El-Dessouky et al., 1999b, proposed a novel approach for enhancement of the performance of existing MSF systems. The proposed system eliminates the heat rejection section and makes use of a mixing tank for brine recycle and feed seawater, Figure 4. This configuration can be adopted in a simple manner in the once-through configuration. The main feature of this system is the control of the feed seawater temperature during winter operation and the reduction of the amount of brine reject.

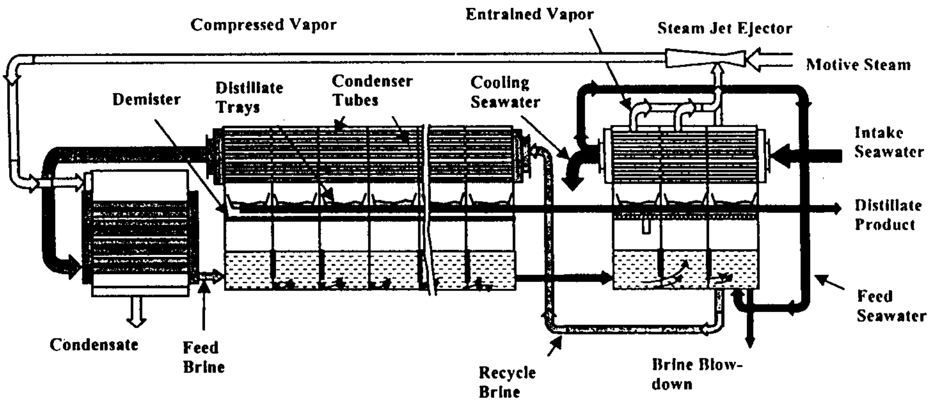

Figure 3: Multi-Stage Flash Desalination with Thermal Vapor Compression

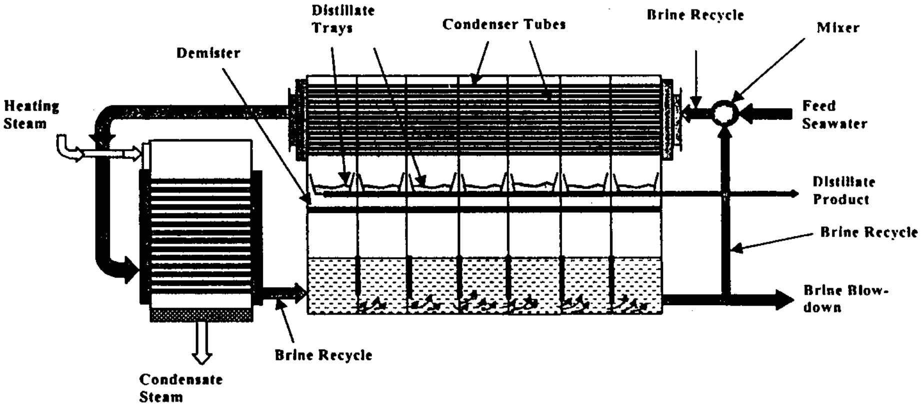

Figure 4: Brine Mixing MSF-M Desalination Process

CONCLUSIONS

The MSF process has gone through a large number of developments since its first operation in the late 50s. These developments are achieved in the unit capacity, specific power consumption, specific unit volume, and specific heat transfer area. As a result, drastic reduction in the unit production cost has occurred. The accumulated field experience in design, manufacturing, installation, operation, and maintenance made the MSF process dominating the desalination market, especially thermal desalination processes. Challengers of the MSF are lead by the RO, MVC, and MEE processes. These systems are more energy efficient and have a lower unit production cost than the MSF. Although, use of the RO is limited in most cases to treatment of river and brackish water as well as seawaters with lower salinity than Gulf seawater, progress in membrane development may result in overcoming operational difficulties in the harsh conditions of the Gulf. On the other hand, the MVC and MEE remain to have limited field experience with a small number of operating units. In this regard, recent market trends show that the number of new installations for either system is increasing steadily. Maintaining the market status of the MSF necessitates further research, innovations and field testing. Developments leading to unit cost reduction may consider design and operation of larger units, vapor compression, brine mixing, and redesign of the system internals.

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Desalination of Seawater and Brackish Water with Reverse Osmosis and the Disc Tube Module DT

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DESALINATION OF SEA WATER AND BRACKISH WATER WITH REVERSE OSMOSIS AND THE DISC TUBE MODULE DT

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ABSTRACT

The "open channel" Disc Tube Module type DT can operate effectively and economically at increased turbidity and Silt Density Index levels for reverse osmosis. This module is applied reliably on sea water which has been pretreated only with normal mechanical filters such as a sand filter and a cartridge filter, usually making dosing of desinfectants as well as acid and antiscalant pretreatment unnecessary. More than 1,600 plants with a permeate production in the range of 2 to 300 m^3/d have been installed since 1985 and are operating successfully to secure the water consumption on cruise liners and merchant ships as well as in the navy and on installations offshore.

The chemical pretreatment, necessary in conventional reverse osmosis practice in the field, is avoided using this module type. Therefore onshore applications are gaining importance because there is no negative impact on the environment, that has to be expected usually because of long term accumulation effects of the additives from conventional reverse osmosis plants that are going back with the concentrate to the sea or the raw water source. Due to the low pretreatment requirements of the module, it can be operated for a wide range of further desalination and purification applications.

Design details of the DT module and results from practical field work in sea water desalination and treatment of brackish water are discussed.

KEYWORDS: reverse osmosis, desalination, open channel membrane module energy recovery, pretreatment.

INTRODUCTION

The increasing world population, the rising living standard and the expanding industrialization are the causes of the ever increasing demand for potable water and for water with high quality for industrial applications. The areas affected are not only the arid world regions with their chronic water deficiency, but also in an increasing rate the urban agglomerations and industrial centers in which the capacity limits of natural supplies have almost been reached.

The phrase "WATER IS LIFE" is generally used nowadays to express very clear these problems of the lack of potable water. But this statement should be concretized: "NO LIFE WITHOUT CLEAN WATER IN SUFFICIENT AMOUNT" [1], as this helps to remind mankind that nearly every type of contamination at least can be a source for water pollution and thus a danger for the basis of life and, that the water produced artificially has to meet well defined quality standards.

In the near future it will not be possible anymore to satisfy the growing water demand by conventional methods of water procurement and processing. Therefore an increased utilization of separation techniques for the production of potable and industrial water from polluted and saline waters is called for [2].

By the application of separation processes based on membrane technology in the treatment of effluents of widely varying nature the burden can be removed and the recovered water returned to the consumer (recycling) or used otherwise. This would result in reduced water consumption, alleviating at the same time some of the wastewater problems with the various effects they entail.

WATER DESALINATION

For the recovery of fresh from saline waters, which is already used to a moderate extent, immeasurable quantities of brackish and ocean water are available. The different separation processes used in this application differ, however, widely with respect to their specific power and investment requirements. Increasingly, small or medium-sized plants are almost exclusively being built according to the RO technology [3].

From about 20.5 million m^3/d fresh water produced nowadays from saline water of different sources about 7.5 million m^3/d are generated using reverse osmosis [3] with increasing tendency for this process.

In Spain for example the desalting capacity for brackish water will be increased from 205.000 m³/d in 1998 to 225.300 m³/d in the year 2002. In the field of desalination of sea water the capacity of 215.000 m³/d in 1998 is expected to increase to 488.500 m³/d in 2002. One of these plants is the system which is being installed in the bay of Palma de Mallorca with 42.000 m³/d permeate production, where start up is expected in the fall of 1998 [4].

In sea water desalination beside a lowest possible pretreatment, features like high efficiency, easy handling, low energy consumption and optimized operation costs are expected by the potential customers, who at the same time want to use membrane technology under rough working conditions and as far as possible in an unattended operation mode. The DT open channel module, using selected membranes and a special plant design, can meet these requirements.

DISC TUBE MODULE SYSTEM

History

Based on the plate and frame module technology, in 1982 the company ROCHEM RO-Wasserbehandlung GmbH – since 1.01.1998 PALL ROCHEM Wassertechnik GmbH – started to produce reverse osmosis systems for water desalination. In the initial phase, the company's activities were focussed on systems for the production of potable water from sea water to be installed on ships and for the offshore industry. The standardized PLC (programmable logic control) controlled modular systems were designed for a fail-safe unattended operation and in order to ensure for the end user 24 hours per day reliability and high product water quality.

Utilizing the long term experience gained in the desalination of sea water and later in brackish water and tap water treatment a new module system was developed – the Disc – Tube Module. The successful design of this module was made possible by an integrated RO concept, since the developer posesses the rare capability of being both module and systems manufacturer.

Fundamentals

The patented DT module represents a successful concept for membrane applications in molecular separation, desalination and purification of liquids. It can operate effectively and economically at increased turbidity and Silt Density Index levels for reverse osmosis, nanofiltration and for high pressure ultrafiltration applications.

The DT-module is a fully modular design comprising the following components (Figure 1):
- membrane cushion
- hydraulic carrier disc with internal O ring gasket
- end flanges with lip seals
- tubular pressure vessel
- central rod with tailpiece and nut

The membrane cushions for the DT-Module are produced from flat membranes with internal fabric fleece (for drainage of the permeate). By contrast with other design concepts, the membrane cushion is octagonal and has a round opening in the center which is used to drain the permeate during operation. The outer edges of the membrane cushions are welded by an ultrasonic method.

The hydraulic discs of the DT-Module have at their outer edge a washer which extends equally on both sides. In the center are a ring gap formed by ribbing through which the raw water runs during operation, a gasket groove which takes the O ring seal between membrane cushion and hydraulic disc, and the opening for the tension rod with the drainage gaps for the permeate.

These hydraulic discs and membrane cushions are stacked alternately on the tension rod to form a membrane stack. Open flow channels are formed in this way on the raw water side between the hydraulic discs and the membrane cushions fitted in between.The completed membrane stack is bordered at both ends by end flanges with lip seals. The stack is compressed by tightening the nut on the tension rod. The reaction is provided by a joining flange, through which the permeate is drained during operation.

The preassembled stack of membrane elements is inserted into a pressure pipe vessel. This pressure vessel not only holds the stack of membrane elements but also acts as a hydraulic component, since a gap is formed between the inner wall of the pipe vessel and the gaskets at the edge of the hydraulic discs, which guides the incoming feed to the ring gap in the first hydraulic disc of the membrane element stack.

The principle of operation of the DT-Module assembled from the components described above is shown in Figure 1.

Operational Details

As a result of the special design features of the DT module, there are a number of aspects which have a favourable effect on the process and the operation of the module:

* The extremely short flow paths (6 cm) along the membrane, before renewed mixing of the feed by the deflection at the outside edge of the hydraulic disc or the ring gap, limit the concentration polarisation.

* As a result of this limiting, the system is unlikely to block, keeping the permeate flow at a high level.

* The delayed formation of a secondary membrane allows longer cleaning intervals and thereby extends the service life.

* The clearly defined sealing functions, i.e. the sealing off the feed-side high pressure area from the permeate-side low pressure area with O rings, and the sealing of the feed-side high pressure area against the ambient pressure by lip seals, avoid leaking problems in the high pressure area.

* The transfer of the operation pressure stress from the membrane element stack to a commercially available 8 inch pressure pipe vessel increases the operational reliability, simplifies the system and the assembly procedure and also helps to reduce the cost of the module.

* Since the sealing pressure of the edges of the hydraulic disc in the compressed state is limited deliberately, the optimised emergency flow characteristics enhance the operational reliability of the module. Even extreme pressure differences do not result in destruction of the module, but only in a limited local breakdown within the system.

* Due to the open ring gaps or flow paths between the membrane cushion and the hydraulic discs, even fluids with high colloid or solids content can be processed without problems.

* The open channels allow efficient cleaning of the module, since the interfering substances which have been removed from the membrane by the cleansing agent can be carried away freely.

* The relatively high packing density for an "open" module system of 200 m^2/m^3 allows the construction of compact modules and installations.

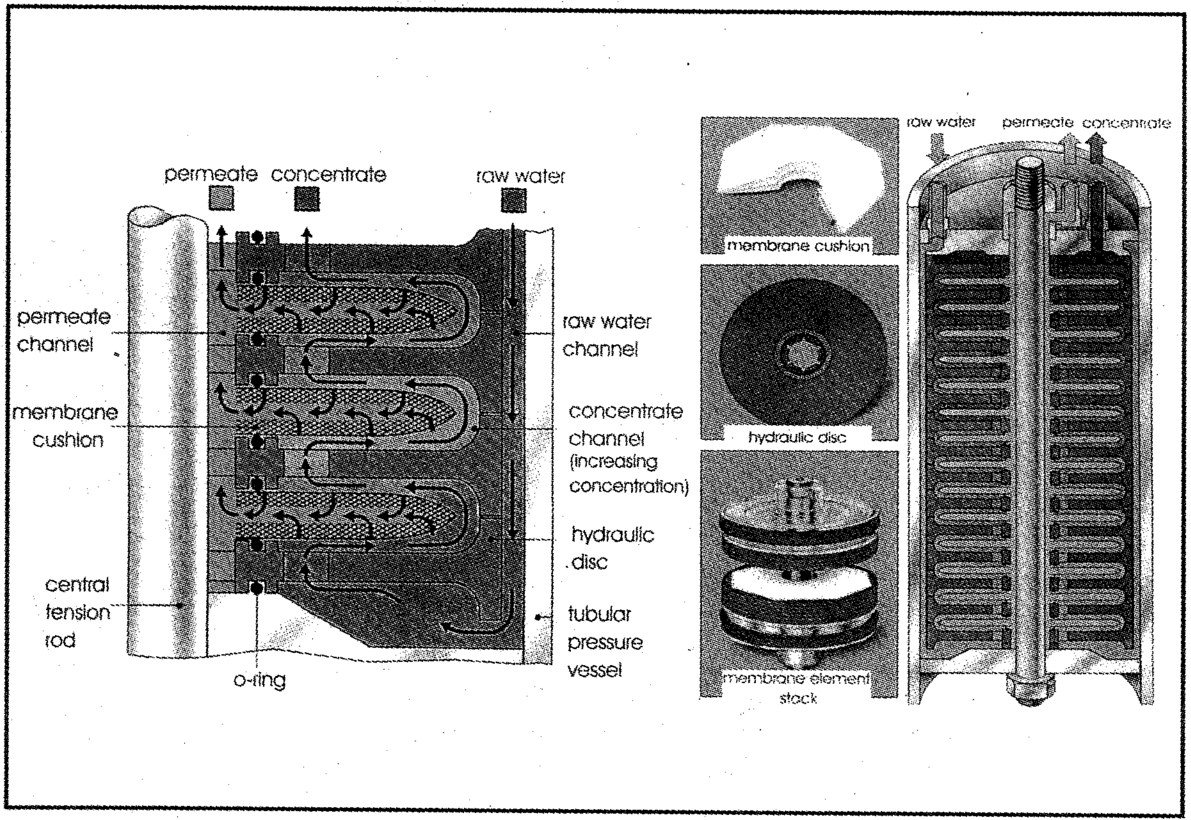

Figure 1: Components and principle of operation of the DT-module
Source: PALL ROCHEM

As further advantages one can mention:

- * Distortion of the membrane element stack internally or during insertion into the pipe vessel is avoided by the symmetrical fully modular design.
- * Flexibility of the application of the module, because the hydraulic discs as well as the membranes used with them can be chosen to suit the particular problem (membrane index).
- * Reduction of operating costs through quick and easy assembly and dismantling of the module for membrane replacement.
- * Cost effective operation, because the narrow gap technology only requires small fluid volumes and therefore low energy input to achieve sufficient flow across the membranes.

APPLICATION IN WATER DESALINATION

The open channel PALL ROCHEM RO DT module can operate without problems on filtered seawater with a modified Silt Density Index (SDI) as high as 15 or more. Thus simple sand filters and cartridge filters can easily achieve the SDI limits of the Rochem system. Other systems specify SDI limits as low as 3. These much lower limits required by other membrane systems can only be achieved by more complex filtration and chemical pretreatment.

This detail is one reason for a comparatively low energy demand in desalination:

- Recovery rate of 30% from seawater - energy demand average 9 to 10 kWh/m^3 of pure water (without energy recovery).
- Recovery rate up to 75% from brackish water (TDS = 5,000 ppm) - average 4 kWh/m^3 of pure water.
- Recovery rate up to 85% from tap water (TDS = 800 ppm) - average 2 kWh/m^3 of pure water.

These figures relate to standard systems. If brine energy recovery is also employed for seawater desalination energy costs can be reduced by as much as a further 30%.

Due to the module design, unlike other reverse osmosis systems the DT module technology means that no permanent feedwater pretreatment with desinfectants, acid and antiscaling is required at sea water desalination. Besides cost saving (less expenditure for additives and their handling as well as by avoiding components for storage and dosing), this means more safety in operation and less environmental burden.

RO SYSTEMS AND PLANTS

Reverse osmosis systems and plants equiped with the DT module for the production of fresh water from seawater are working on ships, submarines and in the offshore industry as well as on shore for the water supply to hotels and holiday camps or for (the desalination of brackish water) land based utilities. The permeate production ranges from less than 2 m^3/d up to 500 m^3/d for the largest single unit, in various standard systems. They could be of higher volumes on request.

The plants can be delivered as partially or completely automatic system. Due to the truly modular design, construction of compact plants, their transportation and installation are simplified. Also a flexibel use of limited floor space is possible. The system shown in Figure 2 is capable to produce between 0.5 and 2.0 m^3/h fresh water from seawater, depending on the number of modules installed.

Figure 3 shows a seawater system for the production of 250 m^3/d, built into a 40" container.

An impression of compact installation of the modules. Thanks to the design features they are easy to service and to handle. The detail of which is shown in Figure 4.

Energy recovery systems can also be incorporated into the design as shown in Figure 5.

It is thus worth reiterating, that energy is one of the most important features in sea water desalination, as energy saving helps to reduce not only fresh water production costs but also indirectly environmental related problems [5,6].

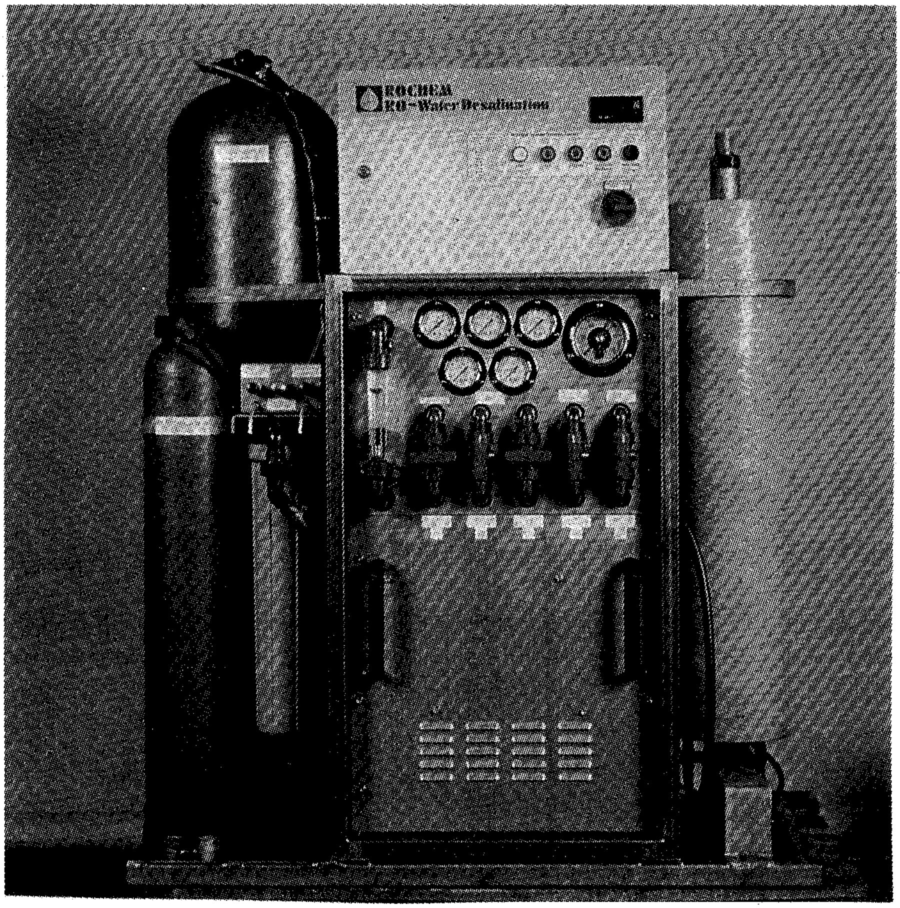

Figure 2: System RORO 0510 with DT-module for the production of 0.5 – 2.0 m³/h fresh water from sea water (depending on the amount of modules installed, i.e. the installed membrane area) source: PALL ROCHEM

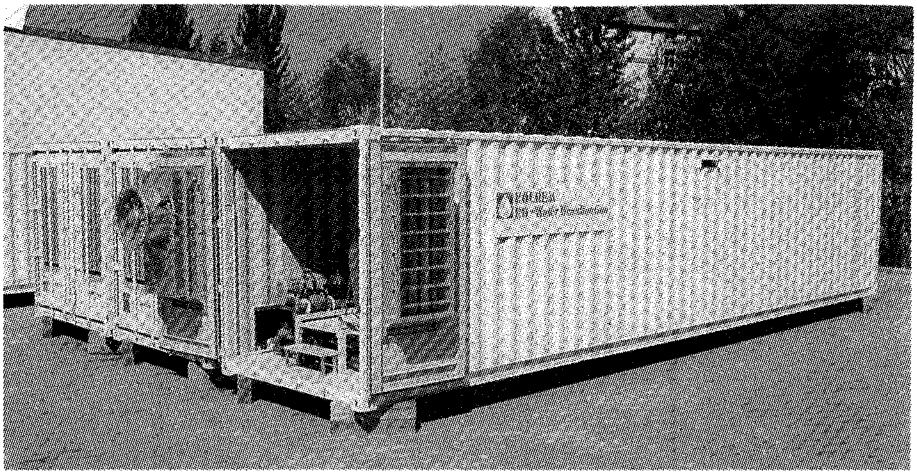

Figure 3: Compact system RORO 8000 DT 250 S for 250 m^3/d, installed in a 40" container, after final test in factory, ready for shipping

FUTURE

The combination of an open channel design with narrow gap technology and a successful membrane cushion concept in an advanced module system makes it possible to use reverse osmosis in applications like desalination, industrial waste water treatment or aqueous fluids processing, even if the water to be treated is of high Silt Density Index corespondingly high turbidity.

Together with a cost effective operation because of comparatively low energy and cleaning chemicals demand and a favourable price to performance ratio the application of this technique in new markets can be foreseen. Where either due to process constraints or economic reasons reverse osmosis were not be considered till now and where through the combination of the highly efficient DT module system along with other features, further improvement can be obtained. This includes the installation of reverse osmosis as stand alone units for onshore application with limited demand of potable water like hotels, holiday camps, villages and small towns, where fail safe automatic operation, easy handling, independancy from logistics and infrastructure regarding chemicals, high reliability and trouble free shut down behaviour as well as the lowest possible influence on the environment are expected.

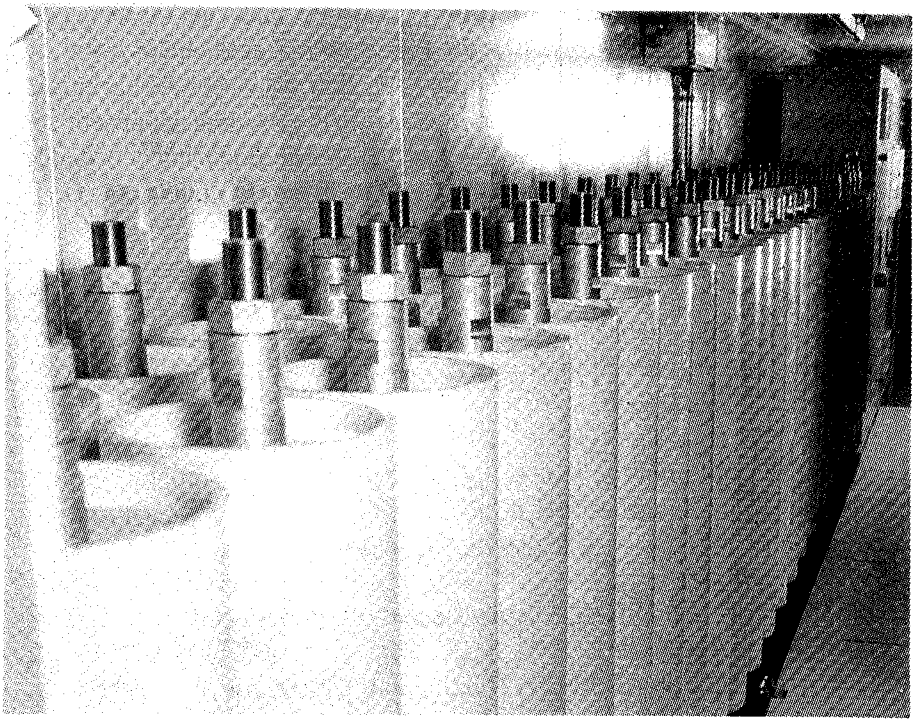

Figure 4: Detail of the system RORO 8000, configuration of the DT-modules

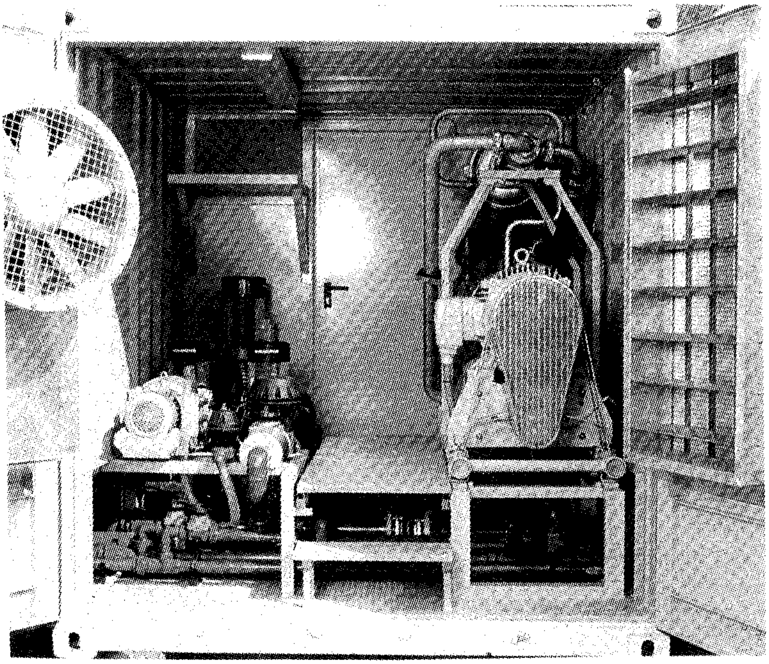

Figure 5: Pumps and energy recovery unit

The successful operation of this type of module and the fact that no permanent feed water pretreatment with desinfectants, acid and antiscaling chemicals are required when this module is used in sea water desalination. These can be viewed as factors that demonstrate the innovative potentials and the growing importance of reverse osmosis technology in accordance with guidelines for sustainable environmentally friendly development.

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Reverse Osmosis Process Chemistry Relevant to the Gulf

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REVERSE OSMOSIS PROCESS CHEMISTRY RELEVANT TO THE GULF

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ABSTRACT

The cost of seawater desalination by reverse osmosis technology (RO) continues to drop with time. This portends accelerating adoption of RO around the world in the 21st century. Since chemicals are used in the operation of RO systems as coagulants, antifoulants and cleaners continuing and increasing adoption of RO process in the Gulf necessitates conservation and efficient utilisation of these chemicals. This is possible through understanding and optimization of the RO process chemistry. In this paper, we survey the chemical factors that affect the operational characteristics of RO systems. The factors include the chemistry of the feedwater, membrane fouling mechanisms, composition of foulants found in membrane elements from many plants around the world and examples of antifoulants and their application. Current approaches towards the development of antifoulants to prevent fouling and membrane cleaners for plant maintenance is discussed in this context. The chemistry presented is part of a conceptual framework in which process optimization and chemical usage conservation can progress well into the next century.

KEYWORDS: Reverse Osmosis, Desalination, Chemistry, Process, Membrane Foulants, Antifoulants, Cleaners, Conceptual Framework, Optimization.

INTRODUCTION

In the Middle East, particularly around the Gulf, over 50% of the world's desalination plants exist. A large segment of the desalination installations employ reverse osmosis (RO) membrane technology. Recent bid prices of seawater desalinated with RO were offered for sale for as low as US$0.60/ M^3 in Florida (USA), which amounts to less than half of any reported costs for a plant currently in operation (1). This portends accelerating adoption of RO around the world in the 21st century. Since chemicals are used in the operation of RO systems as coagulants, antifoulants and cleaners, continuing and increasing adoption of RO process in the Gulf necessitates conservation and efficient utilization of these chemicals. This is possible through understanding and optimization of the RO process chemistry.

OVERVIEW

In this paper, we survey the chemical factors that affect the operational characteristics of RO systems. The factors include the chemistry of the feedwater, membrane fouling mechanisms, composition of foulants found in membrane elements from many plants around the world and examples of antifoulants and their application. Current approaches to the development of antifoulants to prevent fouling and membrane cleaners for plant maintenance is discussed in this context. The chemistry presented is part of a conceptual framework in which process optimization and chemical usage conservation can progress well into the next century.

FEEDWATER CHEMISTRY

The oceans are the most abundant sources of water, amounting to 99.37% of water on earth (2). The Gulf water can be differentiated from the typical seawater due to its relatively small volume, high evaporation rate, restricted circulation and mixing with fresh seawater and high utilization rate. These factors are known to result in significantly higher salinity (TDS 55,775 mg/l compared to 35,159 mg/l of standard seawater) (3). Sand (50 microns - 2mm), silt (5-10 microns), clay (<5 microns) and submicron colloidal particles are introduced by natural and industrial drainage and erosion. Just as active a process is the growth, excretion, death and decay of animals and microorganisms. The biotic life in the Gulf waters can be expected to be different from other localized areas of the seas and should be studied and considered in the control of reverse osmosis processes. The quantity of the different components in seawater is not proportional to the quantity of the components that enter the sea, but is inversely proportional to the facility with which the components in the seawater are made insoluble by chemical actions in the sea (4). When river water meets the highly saline seawater, much of the suspended colloidal matter coagulate and deposit as organic matter-rich silt in the delta regions. This is the same mechanism as the deposition of coagulated colloidal matter in the RO system with the increase in salinity of the brine. Complexation interactions and precipitation of particles is an active natural process in seawater. Despite the high salinity, biotic life abound. Some of the types of colloids that exist in the natural waters, especially in the sea, are listed in Table 1. Polyphenolic complexes like humic acid, lignin and tannin are decay products of woody tissues of plants are noticeable due to their tendency to foul RO membranes by themselves or coagulated with other foulants (5). Polysaccharides that constitute cell walls of microorganisms and plants and excreted variously asgums, slimes and biofilms are prominent players in membrane fouling. In the ocean, they are added to polysaccharides from sea organisms especially the plankton (6), which upon aging become kerogen. It is believed that a white material called "marine snow" that has been observed to fall in deep oceans are these aged and ubiquitous polysaccharides (7).

Table 1: Colloidal Matter in Natural Waters

Microorganisms
Biological Debris (plant and animal)
Polysaccharides (gums, slime, plankton, fibrils)
Lipoproteins (secretions)
Clay (hydrous aluminum and iron silicates)
Silt
Oils
Kerogen (aged polysaccharides, marine snow)
Humic Acid/Lignins/Tannins
Iron and manganese oxides
Calcium carbonate
Sulfür and sulfate

MEMBRANE FOULING MECHANISMS

Feedwaters to RO systems typically are concentrated by a factor of 2 - 4 (50%-75% recovery) during desalination. For simplicity in classification, three classes of fouling (8-14) can be said to occur.

First, the solubility limits of various dissolved salts in the brine stream may be exceeded, leading to deposition and growth of crystals in the flow channels and membrane surface of the RO elements. This type of fouling is referred to as scaling. Most common examples of scales are calcium carbonate, sulfates of calcium, strontium and barium, and calcium fluoride. In the examination of foulants (5) under the microscope, crystals have well defined shapes. Inhibitors injected into feedwaters to suppress crystallization are called antiscalants.

Secondly, particulate matter pre-existing in feedwater can agglomerate and adhere to the membrane and brine flow channels due to increased concentration, salinity, compaction, flocculation, surface interactions and other physical and chemical factors. The particles may be large enough to be removed with 1 to 5 micron RO prefilters, such as silt, carbon fines, coagula and microbial clusters. There may also be colloidal particles that escape filtration even with 0.45 micron filters used in Silt Density Index measurements. The colloids can be organic, inorganic or composits. Ferric, aluminium and manganese hydroxides and silica grow through increasing particle sizes by polymerization and cross-linking with organic and inorganic polymers to become gels and amorphous foulants commonly seen on membranes (5,15). Biotic debris such as polysaccharides and dead cellular matter contribute largely to this type of foulants. We have found that anticoagulants and antideposition agents show promise in inhibiting this fouling process. Antifoulants then can be a term applied specifically in this particulate fouling context. In a broader sense, the term antifoulant is used in the field to include antiscalants as well (9).

Thirdly, a prominent source of fouling, as is true of all water treatment or distribution systems, is the growth and anchoring of microorganisms. Moderate temperatures and minimal nutrient levels in RO waters can support at times explosive growths of microorganisms. Bacteria with cell division times of 20 minutes can grow from a normal count per unit volume of water to a too numerous count in the period of an 8 hour shift. Due to the tendency of bacteria to secrete polymers that anchor themselves to surfaces to facilitate growth as the biofilm, this fouling mechanism is unique and poses a serious threat to operation of RO systems (16). This threat is compounded by the great difficulty of completely removing biofilm from membrane surfaces.

COMPOSITION OF FOULANTS

Through analysis of the composition of foulants on seawater and brackish water RO membrane elements sent from plants across the US and some overseas locations, we have seen examples of simple scales and more frequently the more difficult to clean composites of foulants and biomass (5). Inorganic scales are relatively easy to identify by qualitative tests and quantitation by ion emission spectroscopy using Inductively Coupled Plasma (ICP) or by Scanning Electron Microscopy with Energy Diffusive X-Ray Analysis (SEM-EDX) methods. Our experience confirms the limited set of calcium carbonate, sulfate of calcium, strontium and barium, and calcium fluoride as the common scales round in RO systems. Known responses to specific cleaning formulations also allow us to confirm the identity of the scale.

Elemental gravimetric analyses of carbon, hydrogen, nitrogen along with quantitation of residual ash after combustion allowed us to examine the organic components of foulants (in the limited sense of coagulated particulate matter) and biomass. Correlation of the results of these analytical data on foulants with feedwater chemistry, plant design and operations data, we have made the following observations (5). These observations have been helpful in guiding process improvements to minimize fouling and improve cleaning strategies.

1. Organic matter contribute more than 10% of the mass of foulants in all recent samples, with the majority of the samples contamina 50-100% of organic matter.
2. Biomass constituted the major part of the organic contents of the foulants.
3. Synthetic polymers used in water pretreatment can often be differentiated from biomass in the foulants.
4. Anionic polymers used as antiscalants and dispersants can deposit on membranes and act to concentrate high valence metal ions (Fe, Al, Ca) on the membrane surface.
5. Colloidal silica and clays have complex and variable contents of iron, aluminum and calcium.
6. Elemental phosphorous is associated with biomass, phosphate and phosphonates.

These six aspects of fouling can each occur alone or frequently two or more aspects simultaneously. Discernment and differentiation of the contributing fouling mechanisms is a developing science. This science is vital to the development of antifouling chemicals and effective cleaning methods.

ANTIFOULANT DESIGN AND APPLICATION

The terni antifoulant used here is in its broadest meaning covenng scaling, particulate fouling and microbial fouling – the three classes of fouling mechanisms discussed above. We will briefly discuss strategies aimed at each of these three areas.

For scale control, the development and application of antiscalants is well known and reviewed in the field of boiling water and cooling water chemistry (17,18,19). and applied to boilers, evaporators, cooling towers and cooling systems. Anionic polymers, polyphosphates and organophosphorous compounds, sometimes referred to as threshold inhibitors and dispersants, are used in substoichiometric amounts, usually in the range of 1-5 mg/liter concentrations. By binding to surfaces of growing crystal nuclei, the rates of crystallization from supersaturated solutions are retarded, and crystal packing orders are modified. By this mechanism, crystallization rates are so retarded that although supersaturation of solutes in the water will eventually equilibrate through crystallization, within the residence time of the water in the system, there is little or no scale formation. The uniqueness of RO among water conditioning systems is that the residence time is very short (a few seconds), concentration of seed crystals is low, and temperature is constant. For this reason higher levels of supersaturation without crystallization are possible. On the other hand, the limits of saturation and rates of scaling are hard to model, measure and predict. Interferences come from other solutes in the water, organic or inorganic (19). Assumptions. of RO fouling limits very considerably among practitioners.

For controlling fouling by preexisting particulate matter, the task is much more challenging due to the variety of types of potential foulants and the complexity of their interactions (4, 5, 11, 13, 15), with each other in the same water and with the membrane. Stability and agglomeration of colloidal particles is a subject of major importance in natural waters (4) as well as.in the treatment of process waters (2). Drawing on the basic science of colloids and testing of model foulants suggested by our RO foulant analysis data (5), progress is made steadily with the development of anticoagulants and anti-deposition agents. Case studies from various plants will be presented in the near future.

Concerning prevention and management of biofouling in water treatment systems, the literature is extensive. Much of the art and science found useful generally, is applicable to RO systems as well. Several factors peculiar to the RO system can be mentioned. Chemicals used to sanitize and clean the system have to be chemically compatible with the thin, salt-rejecting, barrier membrane. Of prime concern is that accumulation and exponential growth of the microorganisms should not be allowed to occur within the system. Pretreatment of feedwater, adequate maintenance of upstream unit operations, continuous flow of water through the RO unit, good monitoring and sanitization program and use of preservatives during downtime (20) are important to this end. Normalized flow and differential pressure in the system are sensitive indicators of biofouling. To prevent irreversible fouling, trend charting and readiness to perform adequate cleaning are important aspects of the system operation. Immediate assistance should be sought where cleaning has been inadequate. Powerful new cleaners are now available, providing alternatives to membrane replacement in the worst cases of biofouling.

PRACTICAL IMPLICATIONS FOR GULF RO SYSTEMS

Given the above observations and in our opinion, an enlarged conceptual framework from what is commonly considered, we can look at some practical implications for RO systems in the Gulf region. Since scaling is not generally a problem for seawater and very high TDS well waters, we would like to address here, firstly, antifoulant for controlling coagulation of colloidal and polymeric organic matter. Petroleum sources, active biotic life in warm waters, low exchange rate in the Gulf with fresh seawater through tides, and high industrial utilization present unique conditions of high levels of organic matter prone to coagulation in RO systems, bringing down with them as foulants, inorganic colloidal particles. Secondly, as we have found in our extensive analysis of composition of foulants and correlations with responses to cleaning, the presence of biomass, and natural and synthetic organic polymers on membranes make cleanings challenging. We will briefly discuss improved cleaning with cleaning enhancers.

Exciting developments have occurred in the expansion of a line of antifoulants (21) that not only prevent coagulation in RO systems, but also act as dispersants and antideposition agents for partially coagulated matter. Small RO systems typically used in making bottled drinking water from poor quality surface waters with biomass and particulate contents are helped by injection of antifoulants without other pretreatment. For larger RO systems where pretreatment can be designed and controlled, the economy and convenience of treating colloidal particulate matter with antifoulant alone can be an attractive alternative to clarification and Silt Density Index reduction with coagulant addition and demanding filtration schemes. Case studies will be published elsewhere.

Difficult cleaning problems typically call for resolution by expert companies. Through thorough review of the water and pretreatment chemistry, analysis of the foulant composition and source, and selection of proprietary antifoulants and high performance cleaners, both fouling avoidance and reliable maintenance cleaning can be attained. Most RO systems are cleaned with generic chemicals recommended by membrane manufacturers such as shown in Table 2. While these generic chemicals often produce satisfactory results, many systems require better cleaners to achieve full recoveries from fouling. Some key ingredients from proprietary products that are missing in generic cleaner formulations (Table 2) are made into concentrated cleaning enhancers (21) to be used in v/v dilution ratios of upto 1:250, in conjunction with the generic formulations (see Table 2). A case study on the use of the cleaner enhancers (Better Clean^{TM}-A and D) in an off-line pilot cleaning of single elements from an industrial waste water reclammation RO is given in Appendix 1.

CONCLUSIONS

This paper provides the conceptual framework in which contributions to the RO process chemistry is being made. By reviewing the chemistry of seawater and the natural processes that lead to deposition of solids are instructive for the control of the same or similar processes within the RO system. Known membrane fouling mechanisms are reviewed. Analysis of the composition of foulants sampled from membranes from RO plants provided insight on the specific contributors to fouling. Such knowledge provides the basis for designing antifouling strategies and testing methodology for developing antifoulants. Reference is given to new antifoulants developed to control coagulation of organic foulants in the RO system, and to new cleaner enhancers to improve the cleaning effectiveness of generic cleaning chemicals. We look forward to the expansion of this conceptual framework and reporting on advances towards the day when RO plants can be uniformly sturdy in operation and predictable in performance. Optimal utilization of RO chemicals should be a continuing target into the next century, particularly important to the conservation of Gulf water quality, and economical production of water.

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Table 2: BetterClean™-Boosted Cleaning Solutions Formulated With Generic Chemicals in Concentrations Recommended By Major Membrane Manufacturers

BetterClean™-	A	B	C	D	S
Types of Foulants-	Acid Soluble Foulants	Biofilm, Bacterial Biological Matter	Oils and Organic Matter	Organic and Inorganic Colloids	Silica and Silicates
Fluid Systems	1% citric Acid	/<---------------- 1% STPP (TFC only) ---------------->/ 1% EDTA.Na 1% TSP 1% Borax (TFC only) 0.1% Triton X-100 (CA only)			
Hydranauties	2% Citric Acid 0.1% Triton X-100 or Tergitol 8 etc	2% STPP 0.26% SDDBS pH 7.5	0.5% Na Perborate 0.1% Triton X-100 or Tergitol 8 etc.	2% STPP 0.8% EDTA.Na 0.1% Triton X-100 or Tergitol 8 etc	—
Filmtec	0.2% HCl or 0.5% H_3PO_4 or 2.0% Citric Acid or 0.2% Sulfamic Acid	/<------- 1% EDTA.Na ------->/ 0.1%NaOH to pH=12 or following pH=12 solutions: 0.05% Na-DDS or 0.1% STPP+1% EDTA.Na, or 0.1% TSP+1% EDTA.Na		0.1% NaOH 0.05% Na-DDS pH=12	—
Desal	HCl pH=3 or Citric Acid pH=2	/<----------------	1% TSP 1% STPP 0.1% Na.DDS 1% EDTA.Na pH=8-11.5	---------------->/	—
Permasep	0.5% HCl (pH 2.3 min) or 0.5% H_3PO_4 (pH2.3 min) or 0.2% Sulfamic Acid or 2% Citric Acid, pH=4 or 2% Citric Acid 2% Na_2EDTA NH_4OH to pH=4 or 1% $Na_2S_2O_4$ or 1% NaHMP	1% Na_2EDTA 0.1% NaOH pH=11 max	0.3% NaPerborate 0.25% Na-DDBS pH=10 or 1% Na_2EDTA 1% STPP 1% TSP, pH=11 or 2% STPP 0.25% Na-DDBS or 1% NaHMP	0.3% NaPerborate 0.25% Na-DDBS pH=10 or 2% STP 0.25% Na-DDBS pH=10 or 1% NaHMP	0.5% NaOH, pH=11 or 1% Na_2 EDTA 0.1% NaOH, pH=11 or 0.3% NaPerborate 0.25% Na-DDBS pH=10 or 2% STPP 0.25% Na-DDBS pH=10

KEY : STPP = Sodium Tripolyphosphate
EDTA.Na = Sodium Salt of Ethylenediaminetetraacetic Acid
TSP = Trisodium Phosphate
Triton X-100 and Tergitol 8 = Non-ionic Surfactants
Na-DDS = Sodium Dodecylsulfate
Na-DDBS = Sodium Dodecylbenzenesulfonic Acid
Na_2EDTA = Disodium Salt of Ethylenediaminetetraacetic Acid
NaHMP = Sodium Hexametahosphate

APPENDIX 1
Case Study on Cleaning Enhancement of Reverse Osmosis Membranes with Better Clean

INTRODUCTION: To improve the effectiveness of routine maintenance cleaning of a reverse osmosis system used to recover water from an industrial waste water stream, the use of Better Clean products are investigated. A batch of thin-film composit membrane elements (TriSep ACMII 8.3"x40") were test-cleaned off site. The routine cleaning conditions using phosphoric acid at pH 3 followed by sodium hydroxide at pH 11.0- 11.5 were found helpful in the plant, but can be improved upon. This study resulted in a recommendation to use BetterClean A and BetterClean D in conjunction with the low pH and high pH generic cleaners respectively to attain two to five-fold improvement in permeate flow.

PROCEDURE: Elements were cleaned individually by recirculation of 20 gallons of cleaning solution at low pressure at 30-40 gallons per minute (GPM) cross flowrate and no permeation for 30 minutes. The temperatures throughout was approximately 77°F (25°C). Tests were performed at 112.5 psi (half of 225 psi manufacturer specification condition). Permeate flow rates were recorded as gallons per minute, before each cleaning, and after 10 or more minutes of rinsing with San Diego city water which is filtered through activated carbon bed to remove residual chlorine. To normalise to specification conditions of 225 psi, each increase of 0.10 GPM in the obseved permeate flow is equivalent to approximately 288 GPD under specification conditions.

RESULTS: The **increase in GPM** of permeate flow after each cleaning step is recorded and tabulated below (Table 1). The magnitude of the increase with and without BetterClean A and D are compared.

Table 1: Increase in Permeate Flow (GPM) After Each Cleaning Step

Element No.\Cleaners:	H_3PO_4 pH_3	H_3PO_4 pH_3 +BC-A	NaOH pH11-11.5	NaOH pH13 +BC-D	Total Improvement in Permeate Flow* (in GPM)
1. 80127	0.015	—	0.105	—	0.120
2. 80077	0.027	—	0.094#	—	0.121
3. 80198	0.039	—	0.070	—	0.109
4. 80152	—	0.036	—	0.317	0.353
5. 80178	—	0.023	—	0.334	0.357
6. 80175	—	0.090	—	0.418	0.508
7. 79298	—	0.025	—	0.549	0.574

* Salt rejections have improved or are unaffected.

Cleaning the element with pH13 NaOH after this step only increased flow by 0.080 GPM.

CONCLUSION

The use of BetterClean A with phosphoric acid and BetterClean D with sodium hydroxide at pH 13 would increase the overall effectiveness of the two step cleaning by three to five-fold.

Degradation Study of Cellulose Triacetate Hollow Fine Fiber SWRO Membranes

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DEGRADATION STUDY OF CELLULOSE TRIACETATE HOLLOW FINE FIBER SWRO MEMBRANES

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ABSTRACT

Cellulose acetate (CA) membranes are liable to chemical attack mainly by operation at improper pH or by chlorine and other oxidizing agents. The chemical attack could lead either to hydrolysis of the pendant acetyl group or to oxidation of polymer backbone leading to chain scission. Hydrolysis can be detected by determining the degree of polymer acetylation, whereas oxidation can be detected by changes in polymer molecular weight as measured directly, for example by the intrinsic viscosity measurement, and/or indirectly by determining polymer tensile strength. In the present study degree of acetylation, intrinsic viscosity and tensile strength measurements, were used for identifying the causes of the poor performance of two cellulose triacetate hollow fine fiber membranes obtained from a commercial SWRO plant after being in operation for five and two years, respectively. To establish causes of degradation in the commercial samples, analyses were also performed on fresh virgin fiber, on a second sample exposed to high concentration of chlorine, and a third sample exposed to high pH. By comparison to the virgin sample, it was established that fiber exposure to high concentration of chlorine and to high pH induced decrease in polymer chain molecular weight and in degree of acetylation, respectively. Both the commercial SWRO membranes were found to have undergone both hydrolysis as well as oxidation. However, polymer chain oxidation was established as the major cause for the degradation in performance of the commercial sample which was in operation for nearly five years, while hydrolysis of the pendant acetyl group was established as the major cause for the degradation in performance of the second commercial membrane sample which was in operation for two years. The paper describes the detail of the measurements made and results obtained.

KEYWORDS: cellulose acetate, RO membranes, chemical degradation, degree of acetylation, intrinsic viscosity and tensile strength.

INTRODUCTION

Cellulose acetate (CA) membranes are widely used in RO desalination industries because of the overall advantages of their properties, cost and ease of use. One of the most outstanding features of CA membrane is their claimed resistance to Cl_2. However, occasionally the CA membranes exhibited reversible and irreversible deterioration in their performance. Reversible decline, which is usually caused by fouling due to accumulation of scale, suspended solids or microbiological growth, can be restored to some extent by chemical cleaning. However, in case of irreversible decline, the performance could not be restored and in such cases membrane replacement was required. The irreversible decline is induced either by physical or by chemical membrane damage. Physical damage is caused either by compaction due to creeping, or by scratching or by mechanical causes such as vibration. The major chemical processes of CA membrane deterioration are hydrolysis and/or oxidation. Both hydrolysis and oxidation will lead to poor membrane performance.

CA membranes are most stable at pH 4.7 and the extreme pH, less or above this value, will accelerate the process of hydrolysis of acetyl group, pendant to the polymer chain. It is reported that deacetylation of cellulose acetate is caused by chlorine at pH 7, in presence of metal salts. In this study [Murphy, 1991] revealed that deacetylation was more effective in presence of cobalt and ruthenium, moderately in presence of iron, nickel, rhodium, palladium and silver, and slightly in presence of copper and gold salts. Hydrolysis leads to reduction in degree of acetylation which in tum leads to loss of salt rejection capacity of the RO membranes.

Another major cause for the chemical degradation of CA is oxidation. Oxidation is accelerated at high concentration of oxidizing agents such as chlorine (Cl_2), which is used in RO process as disinfectant [Motomura and Taniguchi, 1981]. Studies [Fujiwara et al., 1994; Kumano et al., 1994] indicated that chlorine in presence of heavy metals lead to oxidation as well as some times to hydrolysis of CA membranes. Cobalt and copper have a remarkable effect on the deterioration of the CA membranes while iron, manganese and nickel affect membrane performance but to a lesser extent. Also, temperature, pH and chlorine concentration have an influence on CA membrane performance. The effect of oxidation is mainly through the scission of membrane polymer chain, thus reducing both its molecular weight as well as mechanical strength, resulting in a decline in salt rejection and to an increase in permeate flux of RO membrane. The oxidation can be detected by either mechanical strength measurement and/or by molecular weight determination. Several methods are used in determination of molecular weight of polymer including intrinsic viscosity which is directly proportional to the polymer molecular weight.

In this study, because of its simplicity and availability the intrinsic viscosity, the tensile strength measurements and determination of acetyl content were employed, to establish the causes of the observed poor performance of two commercial SWRO cellulose triacetate (CTA) hollow fine fiber (HFF) membranes. For comparison reasons, the performance of a fresh virgin membrane, a known oxidized membrane and a hydrolyzed membrane, each of CTA HFF origin were also examined by the same three methods.

EXPERIMENTAL

Materials

Membranes

The two commercial SWRO CTA HFF membranes with poor performance are brine side membrane elements. After about 5 years in operation, the first one (Mem#1) had permeate conductivity of 13,480 µS/cm. The second one (Mem#2) after about 2 years in operation, it had permeate conductivity of 3,920 µS/cm. Virgin membrane fibers and oxidized membrane fibers were obtained from membrane manufacturer as well as a new membrane module which was used for the hydrolysis. Both the commercial membranes and the hydrolyzed membrane were autopsied and samples were collected from inner, middle and outer portions for detailed analysis.

Hydrolyzed CTA membrane was prepared by continuously operating a new SWRO CTA HFF membrane element (size 5" x 20") for about 375 hours in recirculation mode using non chlorinated raw seawater feed with SDI ≈ 4.5. The pH of the feed was maintained at about 9.5 by adding NaOH as and when required. The membrane performance (flow and conductivity) were continuously monitored until the performance of the membrane was remarkably reduced. At the end of the experiment, the membrane was removed and autopsied, and samples were collected for further analyses. Before carrying out any analysis all the membranes cut were thoroughly cleaned with distilled water for about 15 hours.

Degree of Acetylation

Acetylation degree of membranes, which is normally expressed as percent of acetyl group in the polymer, was determined using standard ASTM procedure with slight modification [ASTM, 1983]. In this method, a known amount of membrane sample (0.7 g) was first added to 70 ml of acetone in an Erlenmeyer flask and kept stirring using a magnetic stirrer, for 1 hour followed by addition of 5 ml of methanol. After 30 minutes of additional stirring, IN NaOH was added in excess (15 ml), to the highly swollen (not completely soluble) membrane sample solution and kept stirring for another 1 hour. After adding hot (60°C) distilled water for washing down the sides of the flask and continuing stirring for another 10 minutes, the unreacted excess amount of NaOH was titrated against standard sulfuric acid (0.5 N) using phenolphthalein indicator. When the pink color disappeared completely, an excess (0.2 - 0.3ml) of sulfuric acid was added and retitrated with 0.1 N NaOH. For each sample, a duplicate was also carried out as well as two blank analyses. The acetylation degree was then calculated as the percentage weight of combined acetic acid formed to the total weight of membrane polymer during hydrolysis by the excess NaOH.

Stress Strain Measurements

Instron Series IX automated Materials Testing System, Series 4200 Interface, was used to measure the tensile properties. Hollow fine fibers samples of length 7 to 8 cm were soaked in distilled water prior to the tensile measurement. To prevent the fibers from drying, distilled water was sprayed on to the fiber while running the test. The test conditions were as follows :

Number of test specimens	10
Sample length between the grips	2 cm
Cross head speed	2 cm/min.

Percentage elongation and tensile strength at break were automatically calculated by the computerized Instron for each of the fibers.

Intrinsic viscosity measurement

The intrinsic viscosity [η], which is related to the polymer molecular weight, was determined following the standard procedure [Avlonitis et al., 1992]. Ubbelohde type capillary viscometer was used at constant temperature of 30 ± 1°C. After determining the efflux time (t_o) for the solvent methylene chloride/ methanol (9/1) mixture, the same (t_i) was determined for the known concentration of the sample. For each concentration of sample, five measurements were made and average of three measurements were taken after discarding both the lowest and highest readings. The measurements were made at different concentrations for each sample by successive dilution using the above solvent. The intrinsic viscosity $[\eta]$ was then obtained as intercept of the plot of η_{sp}/c against c, where c is concentration (g/dl) of membrane sample and η_{sp} is specific viscosity. The specific viscosity can be obtained from relative viscosity (η_{rel}) as $\eta_{rel}-1$. The η_{rel} can be determined from the efflux time of solvent t_o and sample t_i as $\eta_{rel} = t_i / t_o$.

RESULTS & DISCUSSION

Degree of Acetylation

The degree of acetylation for all the tested membrane samples, i.e., virgin, oxidized, hydrolyzed, and two commercial samples with poor performance, are shown in Figure 1. For all the membrane samples analyzed, the difference between different portions (i.e., inner, middle and outer) of membrane was found to be not significant as can be seen from the figure. It is very clear from the figure that the hydrolyzed membrane is definitely having lower degree of acetylation (acetyl content) than that of the virgin or the oxidized membranes, where no hydrolysis is supposed to take place. The commercial membrane (Mem#1) which was in operation for nearly 5 years showed only a slight loss (about 1.4%) in degree of acetylation, whereas the Mem#2, which was in operation for about 2 years, showed a remarkable loss in degree of acetylation (about 5%), indicating a severe degree of the membrane polymer hydrolysis. So, it can be said that the hydrolysis was the main reason for the poor performance of Mem#2 and the observed slight hydrolysis of Mem#1 have contributed somewhat to its poor performance.

Tensile Strength

The membranes tensile strength and percentage elongation at break are shown in Figures 2 & 3. Both the percentage elongation and tensile strength show the same trend in that a reduction in the membrane polymer molecular weight, resulting from polymer chain scission by oxidation is noticed for the Mem#l and Mem#2 samples but not for the virgin or hydrolyzed membrane fibers. The tensile strength of the oxidized fibers could not be evaluated as the fibers were too brittle for the measurement, indicating that the membrane polymer is highly oxidized. A maximum value of tensile strength and percent elongation at break are noticed, as expected, for the virgin membrane fibers, and with comparable values for the hydrolyzed membrane. Thus proves that the latter membrane did not undergo any sort of oxidation leading to membrane polymer scission. Hydrolysis removes the pendant acetyl group and is not expected to affect the chain backbone which is responsible for the polymer strength. By comparison to virgin membrane, the tensile strength and percent elongation of Mem#2 were 48% and 60%, respectively, of that for the virgin fibers, indicating membrane polymer oxidation. Whereas the tensile strength and percent elongation of Mem#l were found to be too much lower than those of Mem#2, i. e., only 24% and 1 0% respectively of that of the virgin fibers (see Figures 2 & 3). So, it can be said that both the commercial membranes have undergone oxidation, and the Mem#1 which was in operation for nearly 5 years was highly oxidized and more so than Mem#2.

Intrinsic Viscosity

Intrinsic viscosity measurement results support the tensile strength (see Figure. 4). As expected, virgin membrane was found to have the maximum value of intrinsic viscosity (1.24 dl/g) and the oxidized membrane had the lowest value (0.30 dl/g). For hydrolyzed membrane, the intrinsic viscosity was found to be about 1.24 dl/g indicating that the membrane did no undergo any oxidation, thus confirming the results obtained by tensile measurement. Intrinsic viscosity of Mem#2 could not be measured because the fibers were not soluble in the solvent (methylene chloride / methanol mixture). This was obvious, because it is known that the solubility of cellulose acetate polymers in a particular solvent is strongly influenced by the degree of acetylation [Encycl. of Polym. Sci. & Eng., 1985]. Degree of acetylation measurement results indicated that Mem#2 had lower degree of acetylation compared to all the other membranes analyzed, thus making it insoluble in the solvent by altering the solubility property of the polymer. The Mem#l sample showed a remarkable reduction in intrinsic viscosity (0.72 dl/g) indicating that membrane polymer molecular weight was decreased by polymer chain scission due oxidation, thus supporting the tensile study that the membrane was oxidized.

CONCLUSION

By comparison to virgin sample, it was established that same CTA HFF membrane exposure to high concentration of chlorine and to high pH induced decreases in Polymer chain molecular weight and in degree of acetylation, respectively. It was found that both the commercial membranes, Mem#l and Mem#2 which were in operation for 5 years and 2 years, respectively, have undergone both oxidation as well as hydrolysis. However, polymer chain oxidation was the main cause for the degradation in performance of the commercial sample Mem#1. Whereas, hydrolysis of the pendant acetyl group was established as the major cause for the degradation in performance of commercial membrane sample Mem#2.

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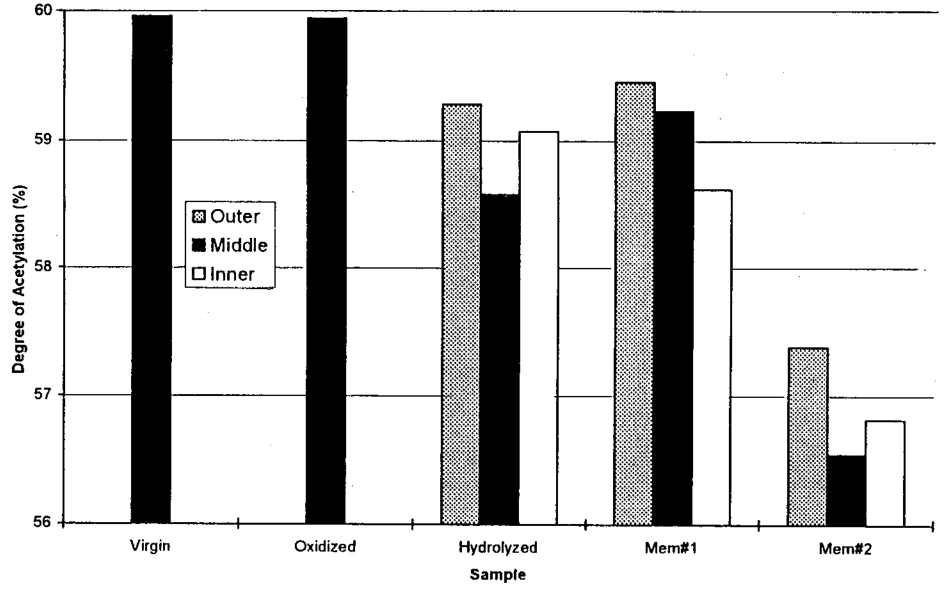

Figure 1: Degree of Acetylation of Various Membrane Samples

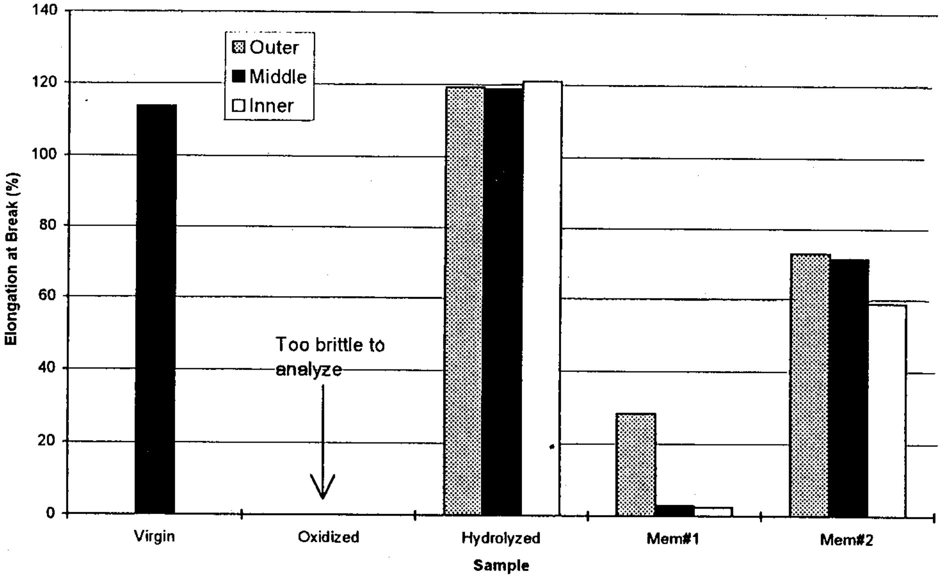

Figure 2: Percentage Elongation at Break for Various Membrane Samples

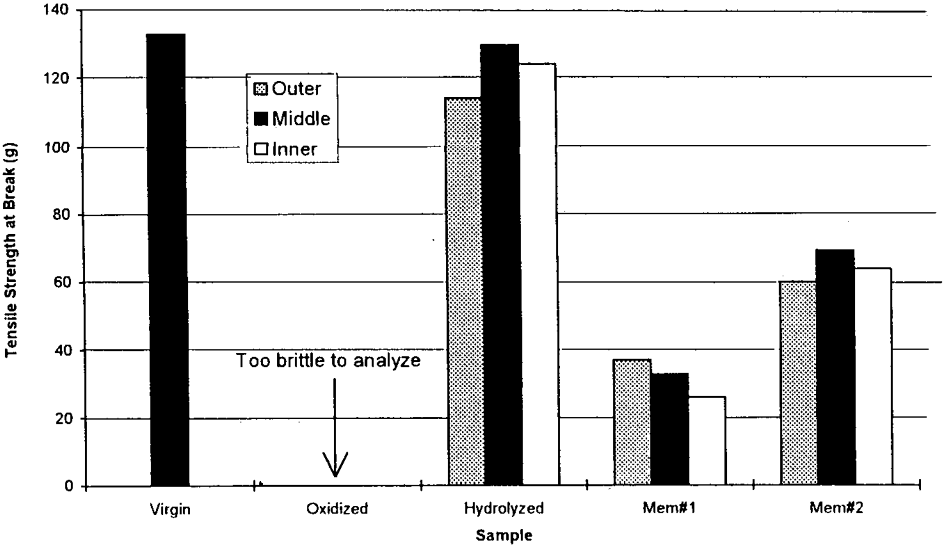

Figure 3 : Tensile Strength at Break of Various Membrane Samples

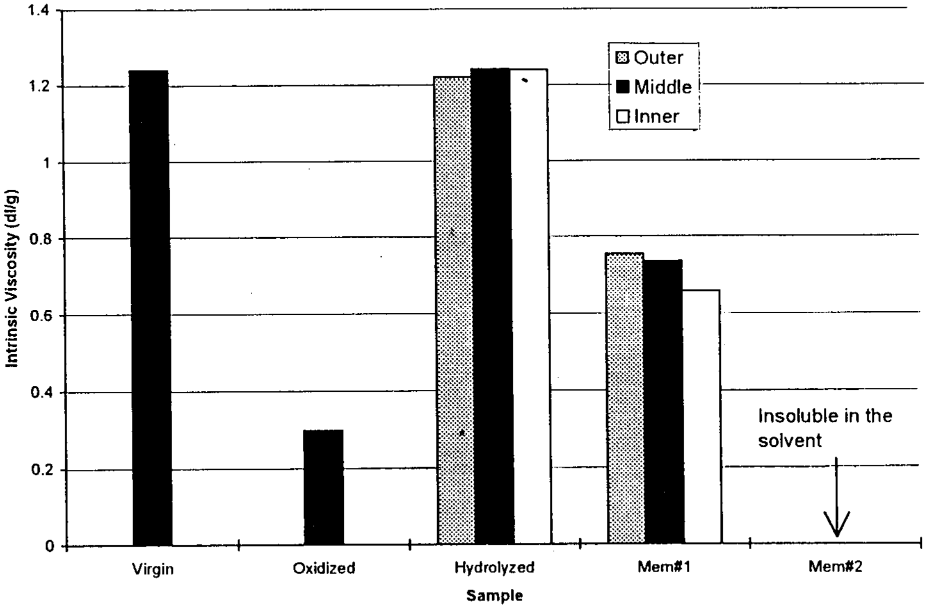

Figure 4: Intrinsic Viscosity of Various Membrane Samples

A New Approach to Membrane and Thermal Seawater Desalination Process Using Nanofiltration Membranes (Part 2)

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A NEW APPROACH TO MEMBRANE AND THERMAL SEAWATER DESALINATION PROCESSES USING NANOFILTRATION MEMBRANES (PART 2)

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ABSTRACT

In earlier publications it was reported that for the first time ever, a new approach to seawater desalination processes by integrating the nanofiltration (NF) membrane pretreatment process with one of the conventional desalination processes to form, for example, an NF-SWRO or NF-MSF or a combination thereof, such as NF-$SWRO_{reject}$-MSF, was developed at Saline Water Conversion Corporation (SWCC), R & D Center, and was successfully applied to those cases on a pilot plant scale with remarkable results. The seawater treatment first with the NF membrane removed from it turbidity and microorganism, caused significant rejection of the scale forming hardness ions, e.g., $SO_4^=$ by up to 98%, reduced TDS in Gulf Seawater by up to 65 %, and produced a new, partially desalinated seawater product, considerably different and superior to seawater in qualities and without the problems normally associated with seawater of high concentration of scale forming ions, high TDS, high turbidity and high bacteria count. The said desalination arrangements led to a significant improvement in the seawater desalination processes, for example by doubling the SMRO product water output and recovery ratio and the production of high purity permeate (TDS < 200 ppm) from one single stage SWRO. It also allowed for the successful operation at high recovery of the MSF as part of a NF-MSF unit or as part of a trihybrid NF-$SWRO_{reject}$-MSF desalination system, where the reject from SWRO constituted the make-up to the MSF unit, at top brine temperature of 120 °C, without the addition to the MSF make-up of acid or antiscalant or antifoam and its operation under those conditions without scale formation. Moreover, by this tri-hybrid process, up to 90% of the NF permeate was converted to potable water. The paper describes the effect of long term operation on the performance of those seawater hybrid desalination processes.

Key Words: NF-SWRO, NF-MSF, $SWRO_{reject}$-MSF, hybrid desalination process, NF-pretreatment of seawater

INTRODUCTION

In earlier publications [1-4] the application of "A New Approach to Membrane and Thermal Seawater Desalination Using Nanofiltration Membranes", which was developed at SWCC, was described. Use of the NF membrane as part of feed pretreatment eliminated the main problems encountered in conventional seawater desalination by removal of turbidity, microorganisms as well as removal of hardness scale forming ions, e.g., $SO_4^{=}$ ions by up to 98% and Ca^{++} ions by up to 90%, and overall reduction of feed TDS by up to 65%. By the removal of hardness ions and by lowering of feed TDS, the NF feed pretreatment also has performed a significant part of the desalination process. Prior to its NF pretreatment the nonchlorinated seawater feed was filtered through dual media filtration followed by cartridge filtration without adding the coagulant chemicals which are normally employed in SMRO feed pretreatment. Moreover, the NF pretreatment allowed for the successful operation of SWRO in an NF-SWRO hybrid pilot plant at high permeate recovery, e.g., 80% recovery at an applied pressure of 65 bars, and producing high purity water in one single stage SWRO. The NF seawater feed pretreatment also allowed for the successful operation of MSF pilot plant without its scaling at high temperature of 120°C and without addition of antiscalant or acid or antifoam in the following hybrid combinations of : NF-MSF and NF-$SWRO_{reject}$-MSF, where in the first and second cases the NF permeate and the SWRO reject from an NF-SWRO unit constituted the make-up to the MSF pilot plant. Main emphasis in this article is on the long term operation performance of those NF-seawater desalination hybrid systems.

SOME MAJOR PROBLEMS IN SEAWATER DESALINATION

As indicated in Section 1 and discussed in detail in earlier publication [1-4], the seawater is characterized by having high concentrations of (1) hardness ions of $SO_4^{=}$, HCO_3^{-}, Ca^{++} and Mg^{++}, (2) TDS, (3) Turbidity and microorganism. In seawater desalination processes, the hardness ions give rise to scale formation problem, while specially in the SWRO process high TDS means high osmotic pressure requiring an increase in applied pressure to overcome it. This leads to an increase in energy waste in form of pressure as seawater TDS is increased. The presence of turbidity and microorganisms in the feed leads to membrane fouling problems. To overcome those problems and in order to minimize their effect on seawater desalination, the seawater feed to desalination plant is pretreated to produce a certain quality feed. This is summarized in Table 1. Presence of hardness ions in feed to SWRO or Thermal plants requires removal of those ions or inhibition of their precipitation by addition of antiscalant and by the operation at correct operating conditions. The high concentration of salts in seawater requires, when possible, the lowering of TDS which results in lowering of SWRO feed osmotic pressure and, therefore, also the lowering of process energy followed by an increase in both permeate flow and recovery ratio. This is also beneficial to thermal desalination process since lowering of TDS reduces the concentration of hardness ions. Obviously, removal of turbidity and microorganisms diminishes the chance of membrane fouling.

Both the high degree of hardness and high TDS place limits on product water recovery, while depending on the process the presence of large turbidity, if not removed, tends to degrade plant performance. In the past, it was demonstrated that use of beach-well effectively removed turbidity and improved SWRO plant performance. Use of antiscalants proved effective in preventing scale formation but failed to increase significantly water recovery in both the membrane and the thermal processes. Also, the coagulation-filtration process is employed in the removal of fine particles. This process, however, does not remove the very fine particles or ions with sizes of less than 1 to 2 μm. For removal from seawater of particles with sizes less than 1 μm, microfiltration (MF), and UF have been used for separation of particle having sizes above 0.08 and 0.01, respectively [5,6,7]. Other membranes that allow for fine particle separation are the RO and NF processes. The RO process deals with separation of ionic size particles in the range of 0.001μ or less. The NF Membrane falls in-between the RO and UF separation range, and is suited for the separation of particle sizes in the range of 0.01μ to 0.001μ. In addition to the rejection of neutral particles according to their sizes, as is the case with the MF and UF membranes, the NF rejection of inorganic matter is achieved by their electrostatic interaction with the negatively charged membrane [8]. Moreover, the degree of rejection by the NF membrane is lesser for mono-valent ions, such as Cl^-, Na^+, than that for the divalent $SO_4^=$ and Ca^{++}.

The above ion selectivity allowed for the use of NF in the removal of hardness from low salinity water. It has been applied in softening of brackish water and drinking water, especially to meet the various Drinking Water Acts, water softening, removal of color, turbidity, removal of dissolved organic which are precursors to disinfection by-products (THM) [9-13]. The NF has been used in other applications to treat salt solution and landfill Leachate [14], demineralization of whey, removal of sulfate from seawater to be injected in off-shore oil well reservoirs [15-17], oil water separation, etc. The use of NF membrane as part of a combined NF-seawater desalination, to pretreat feed to seawater desalination plants in order to remove hardness, to lower TDS, to remove turbidity, has been used by SWCC, for the first time, in the pretreatment of feed to SWRO and make-up to MSF pilot plant units in the hybridization process of NF-SWRO, NF-MSF or NF-$SWRO_{reject}$-MSF [1-4]. With those processes, significant improvement in desalination processes was achieved. The long term operation effect on performance of those hybrid system is presented in the following sections.

EXPERIMENTAL

All experimental work was done on pilot plants scale (capacity 20m^3/d). A schematic flow diagram of the NF-SWRO pilot plant is given in Figure 1, while Figures 2 shows the integration of NF-SWRO with an MSF pilot plant distiller comprising 2 and 4 stages of heat rejection and recovery, respectively. Moreover this last arrangement allows for utilization of the seawater from MSF heat rejection section as the feed to the NF unit during winter season. The NF-SWRO, the SWRO, and MSF pilot plant set-ups were described in earlier work [1-4, 5,18]. The NF unit consists of a high pressure pump and five NF modules each containing two commercial NF spiral wound membrane elements (size 4"x40"). The feed is supplied to the first two modules, arranged in parallel, and the reject of each is fed to its following module which is connected to it in series. Reject from the latter two modules constitutes the feed for the final fifth module. The SWRO unit is made of a high pressure pump followed by six SWRO modules, each contains one commercial spiral wound membrane element (size 2.5"x40"), all arranged in series as shown in Figure 1. After its filtration **without coagulation,** the filtrate was passed to the NF membrane under pressure, initially of about 18 bars. This was followed by passing the product from the NF unit to the SWRO unit, or alternatively by passing the NF permeate or SWRO reject from an NF-SWRO pilot plant, to the MSF unit as shown in Figure 2. Detail of experimental procedure was given in earlier publications [2-4].

RESULTS AND DISCUSSION

NF-Trials

Table 2 lists the concentration of the various seawater ions in Gulf seawater before and after the NF treatment along with their percent salt rejection. At an applied NF feed pressure of 18 bars, the concentration of the hardness ions of Ca^{++}, Mg^{++}, $SO_4^{=}$ and HCO_3^- in NF permeate was 93 ppm, 193 ppm, 206 ppm and 46 ppm, respectively, compared to their concentration in seawater of : 481 ppm, 1608 ppm 3200 ppm and 128 ppm. Their average salt rejection were 80.7%, 87.7% 93.3% and. 63.3%. Total hardness was reduced by 86.5%. In addition to the reduction of hardness ions by the NF pretreatment, the Cl^- ions were also reduced from 22,780 ppm in seawater feed to an average of about 16,692 ppm in NF permeate or a reduction of about 26.7% with similar reduction for the Na^+ and K^+ ions. The net effect of this reduction by the NF treatment in Cl^-, Na^+ and K^+ ions together with the reduction in hardness ions caused a reduction in TDS from 44,046 ppm in seawater to an average of 27,720 ppm for the NF pretreated feed, for a reduction of 37.3%. The pH of the feed of 8.2 was also reduced to an average of 7.85 in the NF permeate. Raising the NF feed pressure to 22 bars reduced further the Ca^{++}, Mg^{++}, $SO_4^=$ and HCO_3^- concentration in the filtrate thereby increasing their salt rejection (Table 2). Moreover, reduction in seawater hardness and TDS, the latter by up to 63%, was achieved when the NF feed pressure was raised to 31 bars, although hardness reduction was less than in the previous case, at p = 22 bar, Table 2.

Long Term NF Operation

Figure 3 shows the actual and normalized NF flow of feed, permeate and product recovery, their conductivity and operating conditions which are plotted vs operation time. At an applied feed pressure of about 20 bars, the product conductivity remained steady at about 41,000 µs/cm, while product water recovery, depending on feed, was about 40 to 45%. Increasing the NF applied pressure to 31 bars resulted in further reduction of both conductivity to above 24,600µs/cm and TDS to about 16,400 ppm, Table 2. The normalized NF product flow fluctuated at about 13 l/min at the start of the operation, to about 11 l/min after 5000 hours of operation time, Figure 3. The seawater filtrate SDI was about 4.3, while the SDI of NF permeate was less than 1. The differential pressure across the NF membrane (ΔP) remained steady at 15 psi (≈ 1bar).

Figure 4 illustrates the NF permeate flow vs operation time, for about 9700 hours of operation. The data in Figure 4b are fully normalized for feed: flow, temperature and pressure, while data in Figure 4a are normalized only for temperature effect. Figure 4a shows that during the cold seasons, operation hours 6000 to 9000 hours, adjustment of feed pressure allowed for raising the feed flow by off-setting the decline in flow due to the seasonal lowering of feed temperature. Full normalization of the data, however, shows that the decline in flow occurred after the cleaning of the NF membranes by a commercial phosphate - based type detergent, DMCA-14/BIZ, made by Cheyma Inc., Monterial, Canada. Use of various cleaning and flushing procedures failed to improve the permeate flow to its level prior to cleaning.

Operation at higher pressure of 31 bars instead of 22 bars, however, increased the permeate flow to its level before membrane cleaning (Figure 4a). However, after cleaning the NF permeate flow remained nearly steady at its level after the cleaning (Figure 4). Prior to this cleaning the membrane retention coefficient (MRC) calculated from Qt/Qi was about 80%, where Qt and Qi are the quantity of permeate flow at time t = 5000 hour and initial time over the first 100 hours.

In the NF membrane process, permeation of the NF permeate through the membrane occurs by two means, either by passage through the fine pores or by the RO permeation process. In both cases the permeate flow is dependent on the applied feed pressure, where up to the membrane flux limitation, the flow increases as the pressure increases. This is illustrated in Figure 5a, where both the permeate flow and recovery increased while permeate TDS decreased as the applied pressure is increased. As shown in Figure 5b, more specifically, the observed increase in permeate flow and recovery is due to the net driving pressure (Pnet). As illustrated by equation given in Figure 5b, a good portion of the applied pressure is lost in overcoming the osmotic pressure (π). For example, at the applied pressure of 37.9 bars, the Pnet, the net driving pressure, is only 12.6 bars for a loss in the applied pressure of 25.3 bars (Figure 5b).

NF-SWRO Trials

During the early operation, 0.0 to 2200 hours at the applied pressure of 56 kg/cm^2 and feed temperature of 33 ± 1 °C, the conductivity of the NF product, which constituted the feed to the SWRO unit in the NF-SWRO hybrid system, was 41000 µs/cm. The SWRO permeate flow and recovery ratio were maintained steady at 5 l/min and about 50%, respectively. This is illustrated in Figure 6, which also shows the SWRO permeate flow, product recovery and conductivity plotted vs operation pressure. During the final phase of SWRO operation, because of a drop in NF permeate conductivity to about 25000 µs/cm, the same above SWRO permeate and recovery ratio were maintained at an applied pressure of only 40 bars as shown in Figure 6. Because of the low conductivity of the NF feed it was possible to operate the unit at 40 bars. Operation at the normally applied pressure of 56 bars increased both the permeate flow and permeate recovery each by about 40%, Figure 6. Most important, the SWRO membranes maintained, a steady high performance, which did not decline with operation time when operation was done at same operating conditions.

The effect of applied pressure on SWRO permeate flow, recovery and conductivity are illustrated in Figure 7, which for reasons of comparison shows also the same for the conventional operation of same SWRO membranes under identical conditions but without the NF pretreatment. Passage of the NF permeate to SWRO unit under pressure gave satisfying results with ΔP remaining steady and constant at 2 bars during the entire operation. As shown in Figure 7, because of the low hardness of SWRO feed (see Table 2 for ions conc. at 31 bar) it is possible to obtain a recovery of up to 80% when the pressure is raised to about 70 bars. This high product recovery was achieved with NF permeate compared to one half this value or less for normally pretreated seawater feed. The product flow (Q_p) and recovery ratio are also increased directly with the applied pressure (Figure 7). The SWRO permeate from the combined NF-SWRO desalination system were much greater than those for the SWRO alone when the two systems are operated at same pressure and temperature conditions. For example, at 40 bars the permeate flow and recovery from the conventional SWRO is 1 l/min and 16.7%, respectively, as compared to a much higher flow and recovery ratio of 4.8 l/min and 48% from the new process of NF-SWRO, i.e., for an increase of 480% in flow and by 3 folds for permeate recovery. Even at the frequently employed pressure of 56 bars (800 psi) the SWRO product flow and recovery ratios for NF-SWRO : SWRO alone are in the ratios of 2.43:1.

Moreover, the quality of the permeate product from SWRO process is 4500 µs/cm at an applied pressure 40 kg/cm^2 and drops to 2300 µs/cm at 60 kg/cm^2 as compared to less than 500 µs/cm for SWRO permeate from the NF-SWRO process obtained from the same membrane at the same pressure range. The SWRO membranes used in this test, Figure 6, are old membranes and the NF pretreatment revives its low performance. This process (NF-SWRO) is expected to extend the life of an otherwise ready to be replaced membranes.

NF-MSF Trials

In this trial NF permeate was used as an MSF make-up at a flow rate of 1.5 m^3/hour replacing normal seawater at concentration of scale forming ions of alkaline and non-alkaline types in the NF feed to the MSF pilot unit as shown in Table 2, while Table 3 shows for comparison these concentrations in the brine recycle stream of the MSF operated with NF or SWRO reject from an NF-SWRO unit as make-up or seawater as make-up (conventional MSF). Also, listed in Table 3 are pH and conductivity values of the brine recycle streams. From this table it can be seen that scaling potentials in the MSF system have been significantly reduced and it was safe to operate the MSF plant with NF make-up for over 2320 hours, or with reject SWRO from the NF-SWRO unit for 1000 hours at high temperature of 120 °C without addition of antiscalant or antifoam chemicals. At the same operating conditions, the concentration of the scale forming ions of Ca^{++}, and $SO_4^{=}$ of 168 and 410 ppm in the NF-MSF case, Table 3A, and 232 and 1020 in the SWRO reject make-up case, Table 3B, are low when compared to 882, and 5830 ppm in the brine recycle stream of conventional MSF. These observations, especially the drastic reduction in $SO_4^{=}$, Ca^{++} and Mg^{++}, are encouraging to project MSF operation at higher TBT in the range of 120 °C to 160 °C. Thus improving plant production and hence water cost. Operation of MSF plants at higher temperature should increase the gain output ratio (GOR) in $Kg_{product}/Kg_{steam}$ and the performance ratio (PR) in $Kg_{product}/1000$ Kj, while decreasing the energy consumption in Kj/Kg product. Finally, the MSF plant operation on NF make-up remained steady with operation time.

CONCLUSION AND RECOMMENDATION

The NF membrane treatment of noncoagulated dual media filtered seawater feed to desalination plants removes from it (1) very fine turbidity, (2) residual bacteria, (3) scale forming hardness ions, in some cases by up to 98% and (4) lowers its TDS, depending on operation conditions by more than 60%. With this NF feed treatment the otherwise complex conventional seawater desalination process, are simplified since the effects on seawater desalination by the above four factors, which constitute the major problems in seawater desalination by the conventional processes, are eliminated. Feeding of the said NF product to desalination plants, whether based on the thermal or the membrane processes, not only allows for their operation with less or no chemicals normally used in the conventional process (i.e., antiscalants, and in case of MSF also antifoam) and by raising significantly their permeate and distillate recovery ratios to 70% and 80%, respectively, but also by lowering their energy consumption with the ultimate benefit of lowering potable water production cost. The combination of NF with MSF process makes it possible to operate MSF plants on NF-product or $SWRO_{reject}$ from NF-SWRO unit at high distillation temperature of 120 °C to 160 °C with high distillate recovery, and again without chemical addition or at reduced level thereof. Similarly, the NF-SWRO process makes it feasible to produce high purity permeate from a single stage SWRO process without the need for a second desalination stage. Also, this process improves significantly the quality of permeate from an otherwise low performance SWRO membranes. In short, it can be concluded that operation of seawater desalination plants when combined with this new NF membrane pretreatment is superior in performance to their operation without the NF pretreatment. Further, work is required in improving the NF membrane performance at an applied high pressure of 35-40 bars. Definitely, the process should be demonstrated by its application to demonstration and commercial seawater desalination plants.

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Table 1: Pretreatment and Quality Requirements of Feed Taken from an Open Sea (Surface) Intake

Problems in Seawater Desalination Due to Seawater Characteristics	**Pretreatment and Quality Requirement of Feed to**	
	SWRO	**Thermal**
High Degree of Hardness of (Ca^{++}, Mg^{++}, SO_4^{2-}, HCO_3^-)	Requires * Removal or * Inhibition of precipitation by addition of antiscalant, and by * Operation at correct conditions	Requires * Removal or * Inhibition of ppt by adding antiscalant * Operation at correct conditions
High TDS	Requires lowering of TDS which * Lowers Waste due to π * Increases Recovery Ratio * Lowers Energy/m^3 * Lowers Cost/m^3	Lowering of TDS is not essential, but it is beneficial by reducing concentration of hardness ions
High Turbidity (TSS, Bacteria, etc.).	Requires complete removal	* Requires partial Removal * Complete removal of turbidity, however, reduces foaming and, therefore, eliminates need for addition to make-up of antifoam

Table 2: Chemical Composition and Physical Properties of Seawater, NF Filtrate, and NF and SWRO rejects at Different NF Feed pressure

Element / Parameter	**Seawater**	**NF Filtrate (5 modules)***						**SWRO Reject**	**NFReject**
	Ion. Conc.	Ion Conc.	Rejection (%)	Ion Con.	Rejection (%)	Ion. Con.	Rejected%	Ion Con.	Ion Con.
NF Feed Pressure (Bar)		18	18	22	22	31	31	31	31
A. Hardness Ca^{++} (ppm)	481	93	80.7	50	89.6	52	89.2	96	701
Mg^{++} (ppm)	1608	193	87.7	96	94.0	143	91.1	253	2200
Total Hardness (ppm)	7800	1049	86.5	520	93.3	720	90.8	1280	10800
SO_4 (ppm)	3200	206	93.3	72	97.8	230	92.8	414	4950
HCO_3^- (ppm)	128	46	63.3	30	76.6	24	81.3	42	133
B. Other Ions Cl^- (ppm)	22780	16,692	26.7	12320	46.3	9640	57.7	19570	29350
Na^+ (ppm)	(12860)	(94261)	26.7	(6904)	46.3	(5442)	57.7		
C. Total Dissolved Solids TDS (ppm)	44046	27,720	37.3	20230	54.1	16400	62.8	30,640	63640
pH	8.2	7.85		7.92		6.38		7.08	7.46
Conductivity (μs/cm)	60,000	40,470		31100		24600		43300	68600

* Module arrangement 2:2:1, each module contains 2 NF elements, arrangement equals two parallel lines each having 5 elements in series

Table 3 : *Operation and Performance Parameters of NF-MSF and NF-SWRO $_{reject}$-MSF Desalination Hybrid Systems Vs Conventional Seawater MSF Desalination System*

(Operation of MSF unit with NF product or SWRO Reject from NF-SWRO unit was done without addition to make-up of antiscalant or antifoam)

Trials	**Brine Recycle Data***						**Performance Parameters**		
	TBT (°C)	**CA^{++} (ppm)**	**SO$_4^{=}$ (ppm)**	**M-alkalinity (ppm)**	**Conductivity**	**pH**	**Make-up (m^3/hr)**	**Product (m^3/hr)**	**Recovery ratio (%)**
A. NF-MSF									
With acid	120	160	390	26	59000	8.19	1.5	0.97	65
Without acid**	120	168	410	65	62000	8.63	1.5	0.97	65
B. NF-SWRO $_{reject}$-MSF*									
Without acid	120	232	1020	72	87400	8.50	1.5	0.97	65
C. Seawater MSF									
With acid	120	882	5830	30	92000	7.99	1.5	0.97	65
With acid	90	661	4460	20	74000	7.57	1.5	0.59	39
With acid	120	561	3330	14	61000	7.50	4.0	0.92	23
With acid	90	581	4000	18	66000	7.51	2.1	0.94	45

* Brine recycle flow rate was maintained between 6.5 to 6.8 m^3/h.

** Make-up is *NF* product having Ca^{++} 52, Mg^{++} 143, SO$_4^{-}$230, M-Alkalinity 24 and TDS 16400 ppm

*** Make-up is *SWRO* reject from *NF-SWRO* unit with composition: Ca^{++}96 Mg^{++} 253, SO$_4^{-}$ 414, M-Alkalinity 42 and TDS 30640 ppm

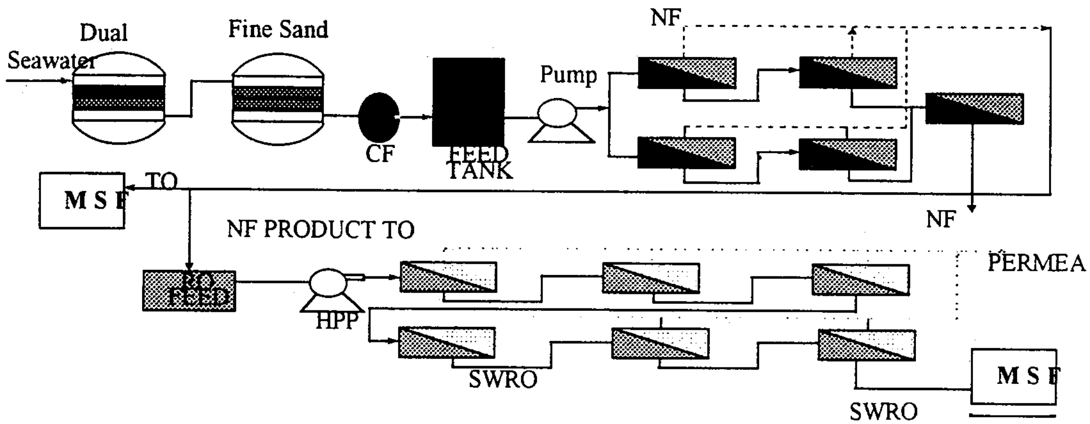


Figure 1. Schematic Flow Diagram of NF-SWRO Desalination Pilot Plant

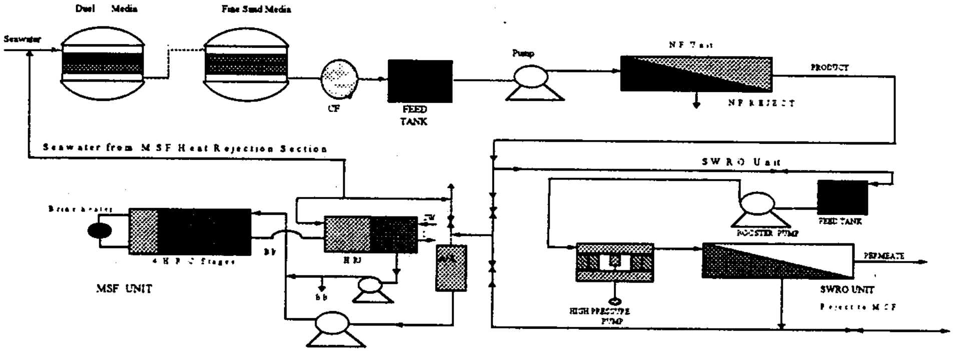
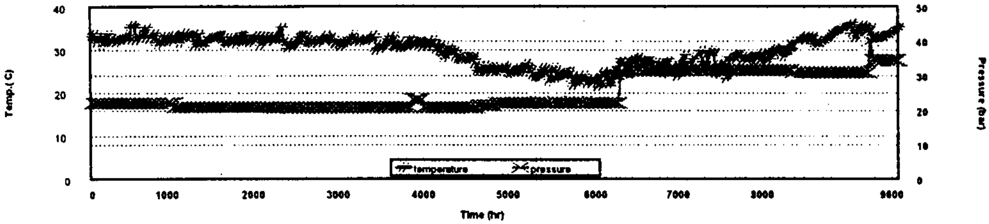
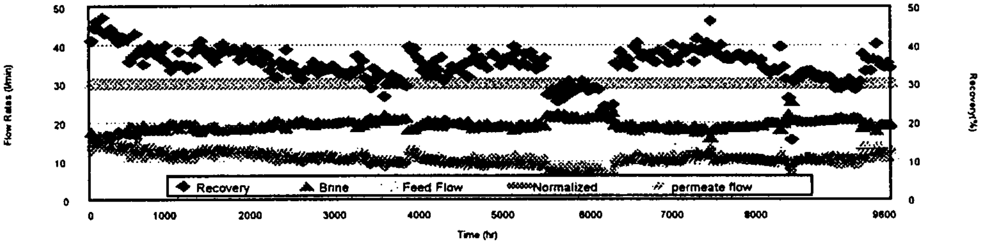


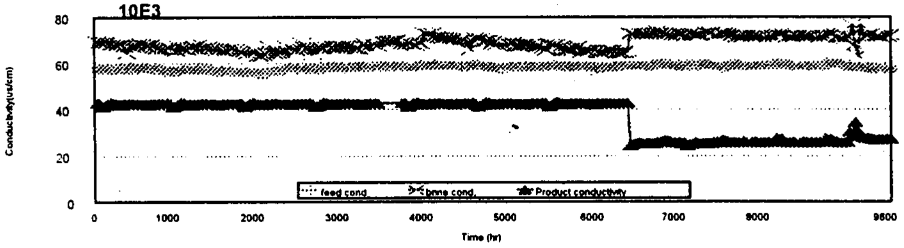
Figure 2: Schematic Flow Diagram of NF, SWRO and MSF Pilot Plants



a. Operating Conditions



b. Flow Rates



c. NF Product, Brine and Feed Conductivity

Figure 3: a. NF Operating conditions and
b. NF Treated Gulf Seawater Actual and
Normalized Flow Rates and Recovery (%)
c. NF Product, Brine and Feed Conductivity Vs Operation Time

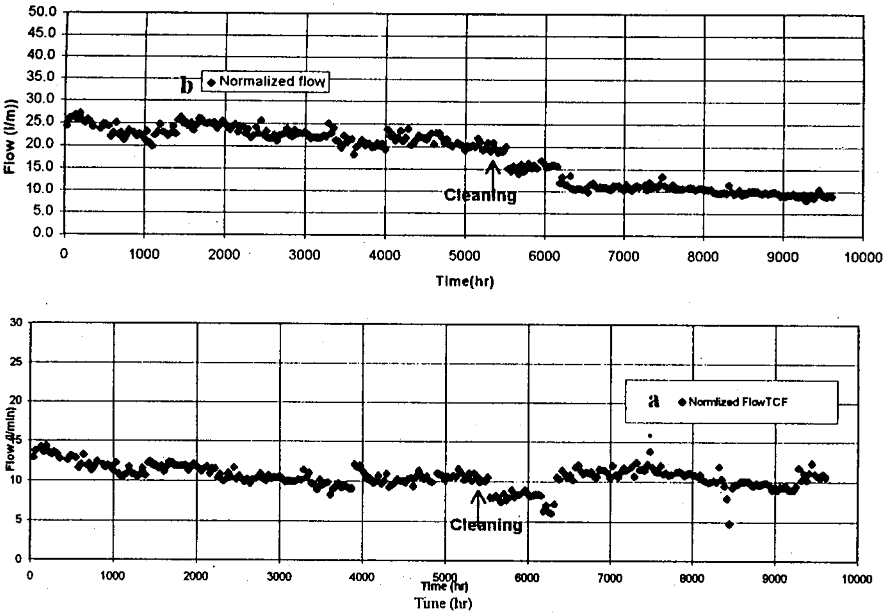


Figure 4: NF Permeate Flow Vs Operation Time

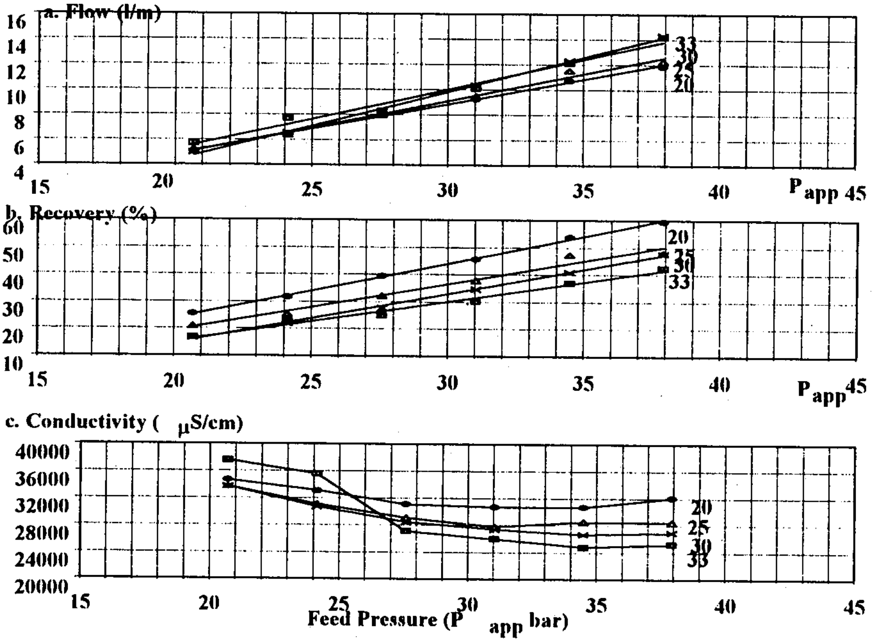


Figure 5a : NF Permeate (a) Flow, (b) Recovery and (c) Conductivity Vs Applied Pressure at Different Feed Flow Rates (l/min)

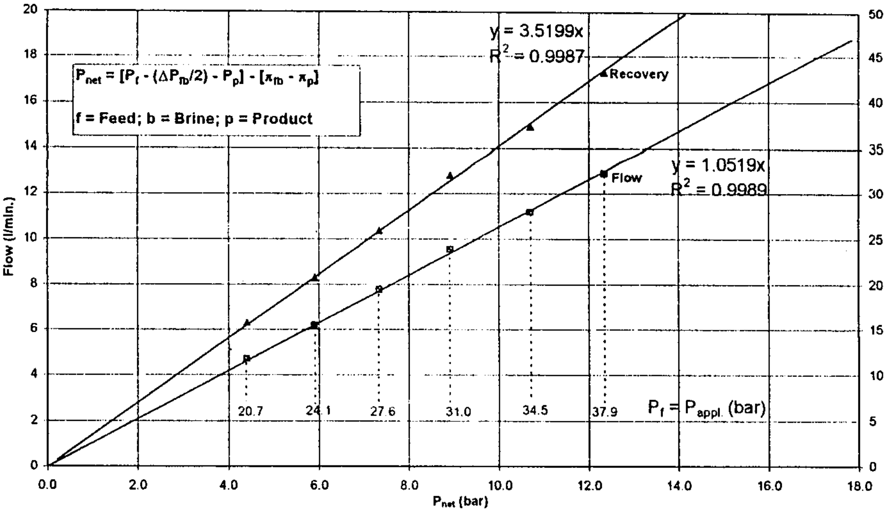


Figure 5b: Effect of Net Driving Pressure (P_{net}) on NF Permeate Flow

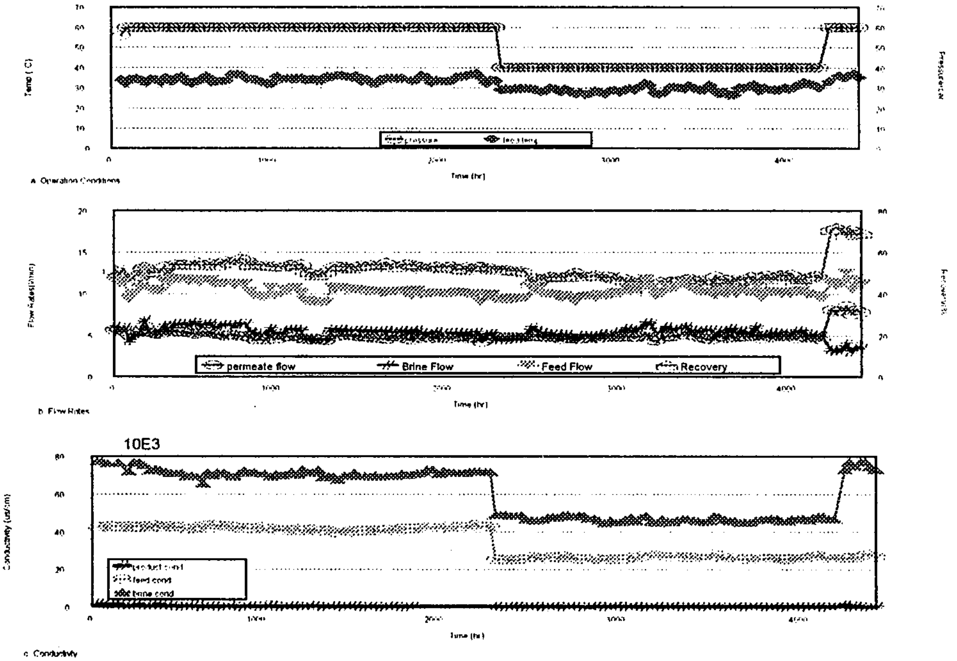


Figure 6: a. SWRO Operation Conditions b. SWRO Flow Rates and Recovery (%) and c. SWRO Conductivity of Feed, Product and Reject in a NF-SWRO Unit Vs Operation Time

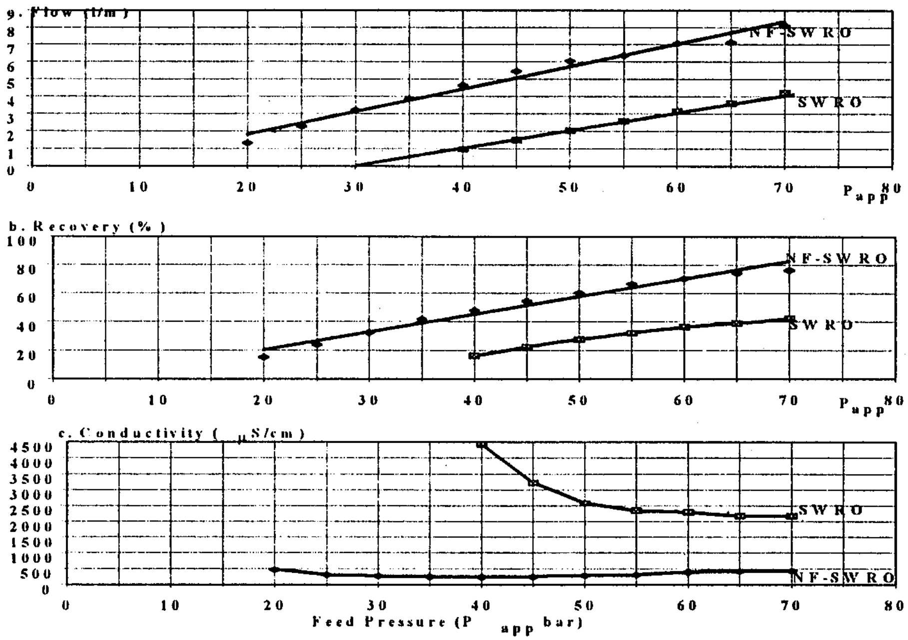


Figure 7: SWRO Permeate (a) Flow (b) Recovery and (c) Conductivity for the New NF-SWRO Process and the Conventional SWRO Vs Applied Pressure

Improving the Premeate Flux by Unsteady Operation of a RO Desalination Unit

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IMPROVING THE PERMEATE FLUX BY UNSTEADY OPERATION OF A RO DESALINATION UNIT

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ABSTRACT

Concentration polarization in reverse osmosis processes results in a lower production rate and a higher rate of fouling. In this paper, a reverse osmosis water desalination unit having a small spiral wound membrane was operated in a cyclic mode so as to reduce the effect of concentration polarization. Asymmetric square wave pressure pulses were employed in which the average pressure was always kept at a value of 25 bars. The effects of the time split (the duration of the higher operating pressure divided by the total period of the forcing wave) on the performance of the system was investigated. For the cases considered, cyclic operation lead to an increase in the permeate flux of about 6.5% over that obtained from steady state operation. The maximum permeation rate was obtained at a time split $\gamma \approx 0.6$. As for the product quality, the concentration of the salt was not affected by the periodic mode of operation.

KEY WORDS: Cyclic operation; Asymmetric forcing function; Reverse osmosis; Water desalination

INTRODUCTION

Although steady state operation is usually the standard method of operating process plants, it has been demonstrated in many experimental studies that improvements can be gained by employing unsteady operation for certain applications. The unsteady operation may be carried out by introducing cyclic fluctuations of a key parameter, such as pressure, flow rate or composition or by imposing flow disturbances. Cyclic operation has been studied for distillation, liquid extraction, membranes, heat exchangers and chemical reactors [1-6]. In all of the cases studied, the time average performance of the process with periodic operation was found to be superior to that corresponding to steady state operation. In certain reactors, periodic operation lead to performances which were up to 5 times higher than the corresponding performances obtained from steady state operation [5].

In membrane processes such as reverse osmosis and ultrafiltration, one factor which limits mass transfer is a phenomenon known as concentration polarization. It refers to the presence of a small thickness film near the membrane in which the solute concentration is much higher than that at the bulk of the fluid. Such a film of high solute concentration leads to a reduction in the production rate of the permeate and an increase in the rate of fouling and scaling. Periodic operation of the process can be used as a method to reduce the effect of concentration polarization. Most of the investigations on periodic operation of membrane systems have concentrated on micro- and ultrafiltration [5]. One of the few published works on the effects of cyclic operation on the performance of reverse osmosis membranes was carried out by Kennedy *et al.* [6,7]. They varied the flow rate of a sucrose solution to a reverse osmosis unit according to a harmonic function. For a frequency of 1 Hz, the permeation rate was reported to show an increase of 70% over that corresponding to steady state operation.

In a recent study by the authors of this work, periodic operation of a spiral wound reverse osmosis desalination unit was employed using a symmetrical square wave pressure function [8]. For the considered operating pressures, wave periods and amplitudes, it was found that cyclic operation leads to increases in the permeation rates of up to 13% over that corresponding to steady state. The objective of this work is to study the effects of varying the pressure according to an asymmetric square wave function on the performance of the RO water desalination unit.

EXPERIMENTAL SETUP AND PROCEDURE

A schematic diagram of the experimental unit used is shown in Fig. 1. This experimental setup includes a feed, permeate and concentrate PVC storage tanks with sizes of 340 liters, 45 liters and 340 liters respectively. A 1.3 kW pump draws the feed solution from the feed tank through a cartridge filter and a manual valve, V2, and delivers it at a high pressure to a FilmTec membrane which is a spiral wound thin film composite type placed in a GRP shell with 61 mm diameter and a length of 457 mm. The membrane is housed inside a hoop wound GRP tube with a cylindrical plug inserted at each end. Each of the permeate and concentrate streams passes through a rotameter to measure its flow rate before entering into its corresponding tank. In this investigation, the flow rates were also calculated from the total amount of each of the two products (permeate and concentrate) collected at the end of each run and the run duration. The operating pressure was set using a valve V1 installed in the concentrate line. The inlet and outlet pressures to the membrane were measured using Bourdon tube gauges. The concentration of the salt in each of the two exit streams was measured using a Metrohm type conductivity meter. At the end of each test a sample was taken from each of the product and concentrate tanks for conductivity measurements. The concentration of the solute was then read from available conductivity-concentration calibration curves.

The feed solution employed in the tests was a water solution having a concentration of 10,000 *mg* NaCl/*liter*. The feed solution was prepared by mixing sodium chloride with distilled water.

Initially, a steady state experiment was performed at an operating pressure of 25 bars. Note that the duration of this experiment and of each run performed in this investigation was 30 minutes. The steady state experiment was repeated three times. The results obtained from all four runs were similar. The average values of the results of these runs were employed as a reference point against which the performances of the periodic tests were compared. Periodic tests were performed using a square wave forcing function by fluctuating the operating pressure manually around an average value $\bar{P}$=25 bars as shown in Fig. 2. Such a forcing wave may be described by the following mathematical relation

$$\bar{P} = \frac{\tau^H P^H + \tau^L P^L}{\tau} \qquad (1)$$

where P^H and P^L are the higher and lower operating pressures respectively. τ is the period of the wave. τ^L and τ^H are the durations of operation at the lower and higher pressures respectively. Equation (1) may be rewritten as

$$\overline{P} = \gamma P^{H} + (1-\gamma)P^{L}$$

where,

$$\overline{P} = \frac{\tau^{H}}{\tau} \text{ and } \tau = \tau^{H} + \tau^{L}$$

As stated earlier, the parameters in the right hand side of eq. (2) have to be chosen such that $\overline{P}$=25 bars. For convenience, in all of the periodic runs performed, a value of 15 bars has been used as the lower operating pressure P^{L}. The wave period, τ, was set at 5 minutes. To complete the specification of the forcing wave, we have two more parameters to set: P^{H} and γ. One parameter will be specified freely and the other is calculated from eq. (2). In this work, in order to study the effect of the time split, γ, of the forcing wave on the performance of the process, six different sets were employed as shown in Table 1. The values of γ ranged from 1 to 0.27. Note that a value of γ=1 refers to the steady state case and γ=0.5 represents a symmetric square wave.

Table 1. Parameters of the forcing waves used. For all tests P=25 bars, τ=5 minutes and P^{L}=15 bars.

P^{H}, bar	25	28	34	40	46	52
γ	1.0	0.77	0.53	0.4	0.32	0.27
τ^{H}, min	5	3.51	2.39	2.00	1.36	1.21
τ^{L}, min	0	1.49	2.61	3.00	3.64	3.79

RESULTS

The experimental results for the various asymmetric tests are shown in Fig. 3. The results of four different sets of tests are plotted in terms of a normalized permeate flow rate, Q_N, which is calculated by dividing the permeate flow rate in the periodic test, Qp, by that from a steady state test, Q_s, i.e., $Q_N=Q_p/Q_s$. The avearge flow rate obtained from the repeated steady state runs was Q_s=13.8 l/h. The scatter observed in the results are typical of reverse osmosis membranes due to the changes that occur on the membrane surface condition. The solid line is the result of a third order polynofitting of all the data, which represents the average behaviour. The average curve shows a maximum at a time split value $\gamma \approx 0.6$. The maximum normalized flow rate, Qp, is 1.065, i.e., 6.5% improvement over steady state operation. In comparison, in an earlier investigation by the authors an improvement of 13% was obtained using a symmetric wave, γ=0.5, and an average pressure of 45 bars [8]. It should be noted that in the present work, the average pressure was kept at 25 bars. This indicates that using asymmetric periodic operation at higher average pressures will most probably lead to much higher improvements in the performance over steady state operation. Currently, further work is being carried to investigate the effects of the average pressure and other parameters of the forcing wave on the performance of the process. As for the product quality, the concentration of the salt was not affected by the unsteady mode of operation. In all of the runs performed, the value of the salt rejection was about 98%. Finally, it has to be noted that measurement of the power consumed has shown that, compared to steady state operation, periodic operation will not result in any increase in energy requirements. Measurements of power consumption by a digital energy meter [8] have indicated that the use of periodic operation results in a negligible increase in power consumption of no more than 0.05%.

CONCLUSIONS

The results have shown that the use of asymmetric operation improves the permeate flow rate for time split values in the range of 0.3 to 1, with the optimum value occurring at $\gamma \approx 0.6$. With an average operating pressure of 25 bars which was used here, a maximum improvement of about 6.5% in the permeate flow rate was obtained. Such an improvement in the production rate does not require any increase in the power consumption. The salt rejection was found to be about 98% in all of the steady state and periodic runs performed.

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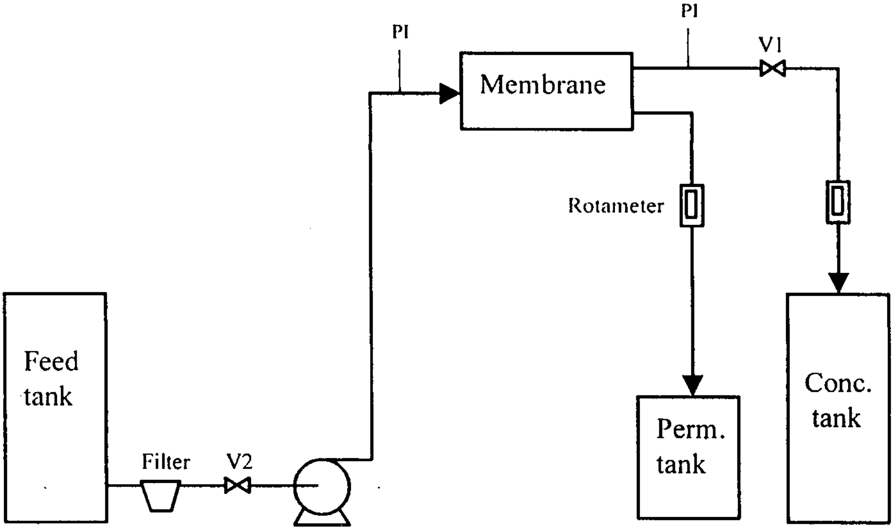

Figure 1: Schematic diagram of the experimental set-up

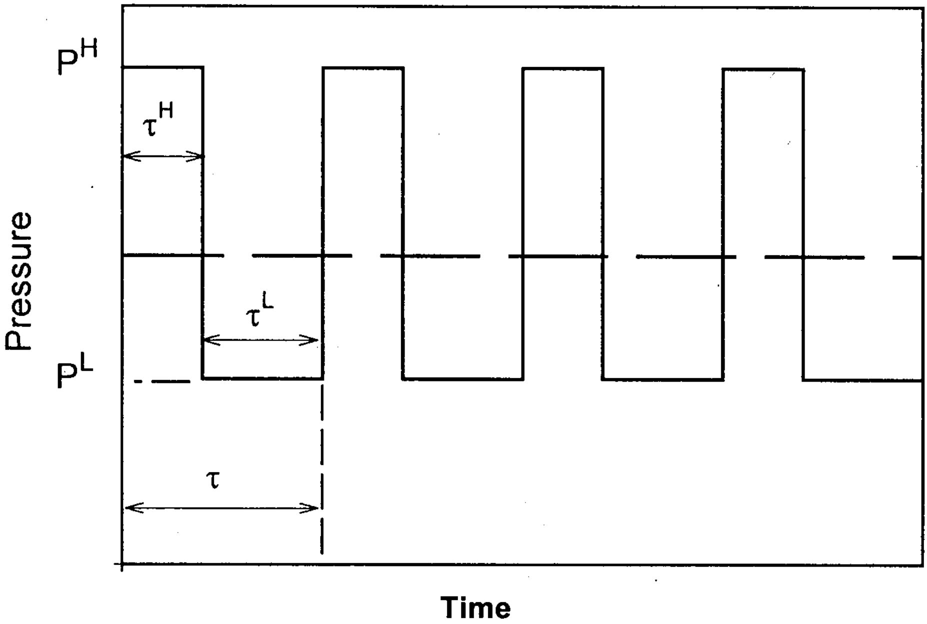

Figure 2 : Asymmetric square wave function

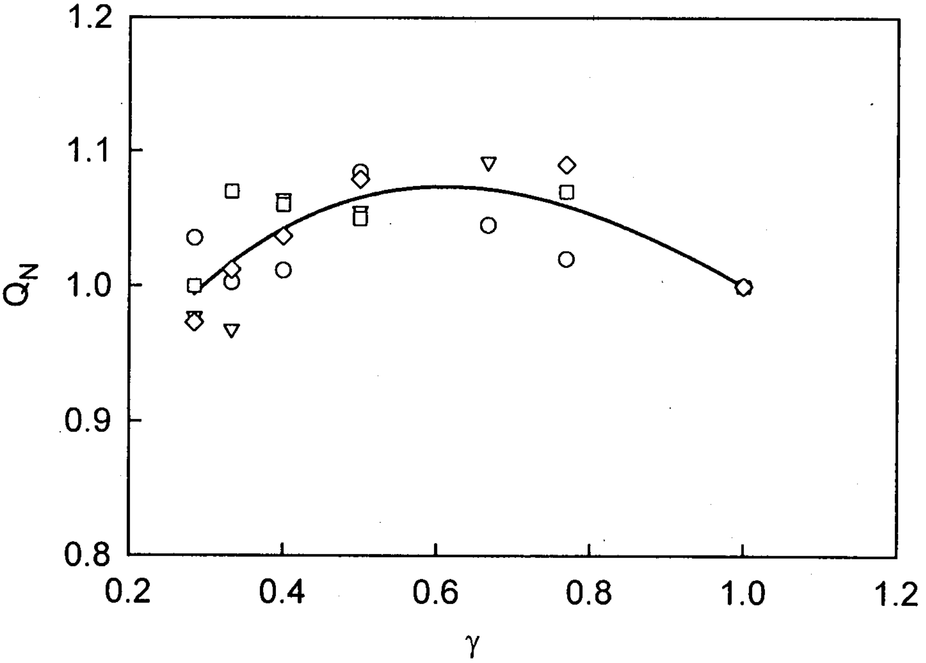

Figure 3. The normalized permeate flow rate as a function of the time split

Evaluation of Ultra-Violet Radiation Disinfection on the Bacterial Growth in the SWRO Pilot Plant, Al-Jubail

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EVALUATION OF ULTRA-VIOLET RADIATION DISINFECTION ON THE BACTERIAL GROWTH IN THE SWRO PILOT PLANT, AL-JUBAIL

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ABSTRACT

The present study summarizes findings of an experiment on the evaluation of UV- treatment for bacterial disinfection at SWRO pilot plant at Al-Jubail. Two UV-generation units were installed along the pretreatment line. One before the dual media filter and the second before the micron cartridge filter. The study was carried out at three seawater flow rates: the regular flow operation and two other flow rates below the normal flow. The investigation was carried out at five locations in the plant: raw seawater (RSW), after UV-unit (AUV), after media filter (AMF), after cartridge filter post the second UV-unit (ACF) and the brine (BR). The study showed 90-99% reduction in bacterial counts after UV- treatment as compared to raw seawater at the three flow rates suggesting relatively good performance of the first UV-sterilizing unit. However, an increase in bacterial count was noticed at AMF and ACF. Presumably, the second UV-sterilizing unit has not reduced the bacterial counts in the feed water at this site. Laboratory studies also showed that, incubation of UV-treated samples for 24h resulted in bacterial recovery and aftergrowth. AMF and ACF also registered reduced Total Organic Carbon (TOC), nitrite and phosphate levels in the feed water indicating the presence of a nutrient trap before CF. A decline in phosphate, nitrite and TOC levels were also found in 24h incubated samples indicating that bacteria had used these nutrients. The study suggests that while UV-treatment considerably reduced bacterial counts in the SWRO plant, stagnant condition of the feed water before CF has presumably led to bacterial recovery.

KEY WORDS: Ultra-Violet Radiation, Bacterial After Growth, Reverse Osmosis, Micronutrients

INTRODUCTION

In the last twenty years the use of UV-radiation technology has been developed and applied on a large scale in treatment of waste water and drinking water. The germicidal effects of UV-radiation at wavelength ranging from 227 to 329nm was known as early as 1900 (Barnard & Morgan, 1903). The action spectrum of UV- degermination exhibits a maximum at 270nm wavelength for most of the microorganisms (Gates, 1930). It is now well established that killing of cells by UV-radiation is primarily due to denaturing of deoxyribonucleic acid (DNA) (Block, 1991). The UV absorption by DNA is maximal at 260 nm so the UV-radiation at the 260nm is most effective as a lethal agent. Proteins also absorb UV-radiation, but have a peak at 280 nm. The UV-radiation has already been suggested as one of the successful disinfection practices for water treatment (Gaudy and Gaudy, 1980). In addition to bacteria, phages, viruses and organic micropllutants are also killed or degraded by photochemical wet combustion down to and below detection limits of organic carbon (Armstrong, *et al.* 1966). Therefore UV-sterilization has become a practical solution to safe disinfection of drinking water. Also, usage of medium pressure lamps with small reactors has helped to substantially reduce the capital costs of UV-treatment (Kruithof *et al.*, 1989 a).

The seawater reverse osmosis (SWRO) desalination plants face biofouling problems from bacteria originating from feed seawater. The problems are most evident in SWRO plants using surface seawater as feed. Biofouling leads to decreased SWRO membrane performance and thus affects its successful operation. Once bacteria are permitted through the pretreatment system and reach the RO membrane, they divide and colonize the membrane surface building a biofilm and causing membrane fouling.

Most of the disinfection procedures in SWRO plants can be roughly classified into: (1) chemical disinfections like chlorine, chloramine, copper sulfate, chlorine dioxide, hydrogen peroxide and ozone, (2) physical disinfections such as ultraviolet radiation, ultra sound, X-ray and ionizing radiation. The chemical disinfectants like chlorine and chloramine are widely adopted in pretreatment of SWRO plants. However, control of bacterial growth depends on factors such as chemical concentration, its nature and mode of action, density of the organisms and total suspended solid (TSS) content of feed water (Gaudy and Gaudy, 1980). These factors make it often extremely difficult to attain absolute disinfection processes. In addition, chemical disinfectants like chlorine may be hazardous to health directly or through by-products (Kutty *et al.*, 1991). Chlorine which is a widely used biocide is known to oxidize and degrade the humic substance in the seawater, thus resulting in smaller molecules which are assimilable organic carbon (AOC) (Kruithof *et al.*, 1989 b). The AOC in turn become a good nutrient source to bacteria (Applegate *et al.* , 1989) and under such situations could also result in elevated fast growth of biofilm in the SWRO plant.

On the other hand, UV-treatment offers certain advantages: in a closed system it is safe, it requires only a small space for equipment and it has got immediate germicidal effect (Haruhiko, 1985). However, incorrect application of UV- radiation in the feed water treatment can be unsuccessful (Luckiesh, 1946) and determining the right dosage for a system is therefore essential. The present study was carried out to evaluate the UV-radiation disinfection in an SWRO pilot plant at Al-Jubail.

EXPERIMENTAL

The study was carried out during June-July, 1996, a period known for its high biological activity of the microorganisms in the seawater feed at Al-Jubail (Munshi *et al.*, 1995). Samples were aseptically collected for bacterial growth studies using standard pour plate method (APHA, 1995) from the following points: 1. Surface raw seawater (RSW), 2. Seawater After UV-treatment (AUV), 3. After media filter (AMF).4. After cartridge filter (ACF) following the second UV-treatment., 5.Brine (BR) (Appendix A). Samples were serially diluted and plated using marine agar 2216. Plates were incubated at 30 ^{0}C for 48-72h (Munshi *et al.*, 1994) and the colony numbers were recorded as colony forming units (CFU). Samples were further incubated at 30 ^{o}C for 24 and 48 hours for bacterial aftergrowth studies.

In order to monitor the changes in micronutrients concentrations during the UV-treatment process and their possible effects on the bacterial aftergrowth, TOC and micronutrient measurements were carried out. The samples for Total Organic Carbon (TOC) were collected in sterilized glass bottles. TOC was determined by measuring CO_2 released by catalytic combustion of organic carbon using infrared detector. The sample was acidified and total inorganic carbon (TIC) was purged off prior to the analysis. TOC analysis was carried out using SHIMADZU TOC Analyzer using the Instruction Manual and USEPA Method (1983). The samples were also collected for analyses of phosphate, nitrite and silicate and analysis was carried out following methods of Parsons, *et al.* (1985). pH and conductivity of the samples were recorded using a pH meter (Fisher Scientific Co.) and a conductivity meter (Yellow Springs Inst. Co.).

RESULTS

The viable bacterial counts observed at different feed flow rates of pretreatment and at different locations of the SWRO pilot plant are given in Table 1. At the three flow rates (2, 4 and $7m^3/h$) reduction in bacterial counts was observed at AUV as compared to RSW. While RSW showed $1.21x10^3$ CFU/ml, AUV and AMF showed $1.37x10^1$ CFU/ml and $76.26x10^1$ CFU/ml, respectively. BR showed an increase in number of bacterial counts as compared to ACF and AMF except on June 2, 1996 when ACF, AMF and BR showed identical growth. At $4m^3/h$ and $7m^3/h$ also an increase in bacterial counts was noticed at AMF and ACF. Therefore, the results showed that UV-treatment has drastically reduced the bacterial growth immediately after the first UV-treatment unit but after the media filter and storage tanks the seawater entering the membrane system showed an increase in bacterial counts.

The percentage of reduction in bacterial counts achieved by the UV-treatment is given in Figure 1. The results showed a 90.59 to 99.99% reduction in bacterial growth after UV-treatment as compared to RSW. Out of the sixteen observations, fourteen showed reduction above 97%. June 11 and July 2, 1996 showed minimum reduction of 90.59 and 94.21% respectively. However, percentage reduction was not related to flow rates. These results demonstrated that the first UV-unit presently employed in the pilot plant before AMF has substantially inactivated the bacteria in the feed seawater at all the flow rates.

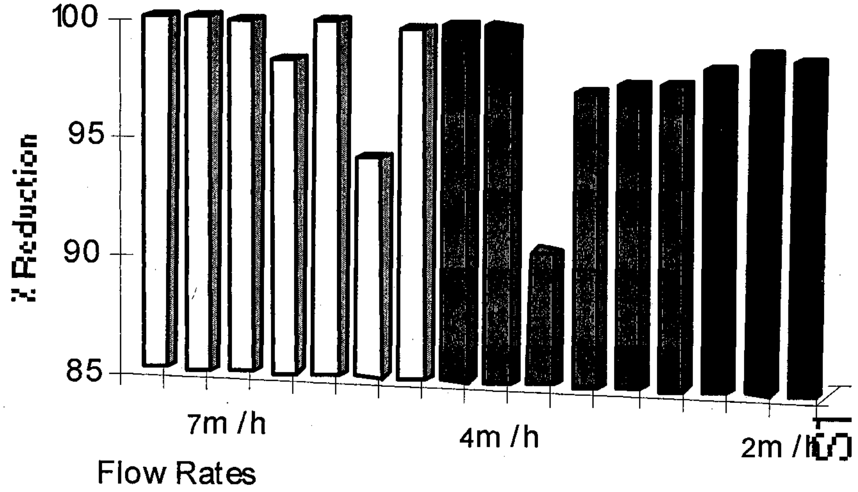

Figure 1. Percentage Reduction in Bacterial Counts After Exposure to UV-Radiation

Table 1. Viable bacterial counts (CFU/ml) in seawater at different flow rates and at different stages of the SWRO pilot plant

Sampling date	**Sampling stage**				
	RSW	**AUV**	**AMF**	**ACF**	**BR**
(2m^3/h Flow rate)					
June 2, 1996	1.47x10^3	1.53x10^1	1.84x10^3	4.60x10^3	1.26x10^3
June 4, 1996	1.00x10^3	0.80x10^1	1.63x10^2	5.12x10^2	7.76x10^3
June 4, 1996	1.17x10^3	1.80x10^1	2.85x10^2	5.46x10^2	1.35x10^3
(4 m^3/h Flow rate)					
June 8, 1996	1.15x10^4	2.51x10^1	1.609x10^2	1.62x10^3	1.39x10^4
June 9, 1996	1.36x10^3	3.08x10^2	3.98x10^2	5.10x10^1	8.06x10^3
June 10, 1996	3.34x10^3	8.94x10^1	2.01x10^2	3.78x10^1	6.77x10^3
June 11, 1996	3.32x10^3	3.12x10^2	5.47x10^2	1.03x10^2	3.15x10^3
June 16, 1996	1.33x10^4	0.40x10^1	5.33x10^3	2.03x10^2	6.77x10^3
June 17, 1996	2.10x10^3	0.60x10^1	5.20x10^2	2.05x10^2	5.95x10^3
(7 m^3/h Flow rate)					
June 30, 1996	1.46x10^4	2.50x10^1	2.18x10^4	1.43x10^4	9.05x10^3
July 2, 1996	3.87x10^3	2.24x10^2	1.35x10^4	3.32x10^2	1.16x10^4
July 6, 1996	4.25x10^3	1.27x10^2	2.57x10^3	6.68x10^3	8.94x10^3
July 7, 1996	5.22x10^3	8.88x10^2	3.28x10^3	1.03x10^3	1.02x10^4
July 8, 1996	5.72x10^3	3.88x10^2	2.70x10^3	2.18x10^3	1.91x10^4
July 9, 1996	1.99x10^4	1.18x10^2	4.76x10^2	3.82x10^3	7.28x10^3
July 10, 1996	1.43x10^4	4.51x10^1	2.99x10^2	2.49x10^5	1.38x10^4

Bacterial after growth at different locations of the SWRO pilot plant after 24 and 48 h incubation times at 30^{0}C are given in appendix B. AUV was similar to RSW at all the three flow rates tested. AMF, ACF and BR also showed similar bacterial growth at the three flow rates suggesting that the bacteria recovered to normal growth after 24 and 48 h, inspite of the UV-treatment. The second UV-sterilization unit at HFRO unit appears to have failed to control bacterial growth.

Conductivity of RSW, AUV and ACF recorded during the study period are given in Table 2. Conductivity values ranged from 59.4 to 60.0 milli seimens/cm at the RSW during the period. A slight elevation (about 1 milli seimens/cm) in conductivity was noticed after 24h incubation time of samples at 30^{0}C. In general, conductivity values were identical at RSW, AUV and ACF. pH values ranged from 8.09 to 8.26 at RSW during the period (Table 2). pH values were similar even after 24h incubation time at 30^{0}C. In general, RSW and AUV showed identical pH but a decrease in pH was found at ACF.

Table 2. Conductivity and pH of seawater at different flow rates and stages of the SWRO pilot plant (24h measurements are given in parenthesis)

Sample source	**Conductivity (milliseimens/cm)**			**pH**		
	2 m^3/h	**4 m^3/h**	**7 m^3/h**	**2 m^3/h**	**4 m^3/h**	**7 m^3/h**
RSW	59.40	59.50	60.00	8.26	8.10	8.9
	(60.40)	(59.40)	(60.00)	(8.15)	(8.17)	(8.13)
AUV	59.40	60.00	60.70	8.16	8.14	8.10
	(61.00)	(60.00)	(60.60)	(8.15)	(8.17)	(8.07)
ACF	ND	59.60	60.20	ND	7.82	8.01
	(59.80)	(59.50)	(60.30)	(7.72)	(7.62)	(7.80)

ND- No Data

TOC values ranged from 0.90 to 2.60mg/l at RSW during the study period (Table 3). A substantial reduction in TOC values were noticed at RSW, AUV and ACF after incubation of samples for 24h at 30^{0}C. In general TOC values were less at AUV as compared to RSW except at 7 m^3/h where TOC values were more at the AUV as compared to RSW. At 2 m^3/h a slight increase in TOC was found. However, at 4 m^3/h, TOC was slightly less at ACF as compared to AUV and at 7 m^3/h all the three sites showed similar TOC values. In general, TOC values declined after UV- treatment and after 24h incubation time at 30^{0}C.

Table 3. Total Organic Carbon (TOC) in seawater at different flow rates and stages of the SWRO pilot plant (24h measurements values are given in parenthesis)

Sample source	**TOC (mg/l) 2 m^3/h**	**TOC (mg/l) 4 m^3/h**	**TOC (mg/l) 7 m^3/h**
RSW	2.60	1.82	0.90
	(1.13)	(0.81)	(0.95)
AUV	1.46	1.59	1.23
	(0.48)	(0.76)	(0.76)
ACF	2.03	0.71	0.80
	(0.33)	(0.53)	(0.72)

Phosphate concentration at RSW ranged from 0.55 to 1.22 μg as PO_4 -P/l during the study period (Table 4). Phosphate values declined at 2 and 4 m^3/h when incubated for 24h at 30^{0}C. However, at 7 m^3/h the values remained the same even after 24h-incubationat 30^{0}C. In general phosphate levels were identical at RSW, AUV and ACF. Silicate levels ranged from 3.84 to 18.42 μg SiO_2 - Si/l. There was no considerable change in silicate concentration even after 24h-incubationat 30^{0}C. At 2 and 4 m^3/h RSW and AUV showed identical silicate levels but slightly high values were found at ACF. At 7 m^3/h a reduction in silicate values was found at AUV and ACF as compared to RSW. Nitrite levels were relatively low and a reduction in nitrite levels were found after 24h incubation time at 30^{0}C. Elevated nitrite levels were found at AUV. The results showed that phosphate and nitrite levels are reduced after 24h incubation time and silicate levels were similar before and after the incubation.

Table 4. Micronutrients in seawater at different flow rates and at different stages of the SWRO pilot plant (24h measurements values are given in parenthesis)

	Phosphate (μg /l Po^{3-}_4-P)			Silicate (μg/l SiO_2-Si)			Nitrite (μg/l No_2-N)		
Sample Source	2 m^3/h	4 m^3/h	7 m^3/h	2 m^3/h	4 m^3/h	7 m^3/h	2 m^3/h	4 m^3/h	7 m^3/h
RSW	1.22	1.29	0.55	3.84	4.02	18.42	BDL	BDL	BDL
	(0.14)	(0.33)	(0.50)	(4.50)	(3.94)	(15.08)	(BDL)	(BDL)	(BDL)
AUV	1.22	0.64	0.55	3.84	4.69	10.52	3.03	6.06	BDL
	(0.68)	(0.11)	(0.75)	(4.75)	(4.65)	(8.77)	(2.42)	(BDL)	(BDL)
ACF	ND	0.64	0.55	ND	6.54	11.84	ND	BDL	BDL
	(0.41)	(0.22)	(0.50)	(7.0)	(6.08)	(6.66)	(1.52)	(BDL)	(BDL)

BDL- Below Detection Limit ND- No Data

DISCUSSION

The germicidal effect of UV-radiation is dependent on several factors such as total dissolved solids (TDS), total suspended particles (TSS), conductivity, total organic carbon (TOC), total hardness, sensitivity and type of microorganisms (Gaudy and Gaudy, 1980). The present study showed that UV-treatment has reduced bacterial counts. The number of the viable bacteria count of 1.37X10^1 at a flow rate of 2 m^3/h (residence time 22.04 secconds) showed a percentage reduction of 98.87 after the UV-treatment. Other flow rates: 4 and 7m^3/h (normal operation rate of the plant) (11.02 and 6.29 seconds residence time, respectively) had also achieved 97.24 and 98.88 percentage of reduction respectively. In total, the UV-radiation treatment employed in the pilot plant before MF achieved a high percentage kill of bacteria at the three flow rates.

However, AMF, ACF and Brine samples showed an increase in the bacterial counts suggesting that the second UV-sterilization unit was not successful in controlling bacterial growth. The results showed that the source for increase in bacterial counts at the AMF and ACF appears to be due to recovery of bacteria after UV- treatment. ACF and brine samples showed similar bacterial counts. Gaudy and Gaudy (1980) reported that spore-forming and slowly growing cells are less affected by UV light than vegetative and rapidly growing cells. Since UV acts primarily on DNA, there may be no time for repair mechanisms to be put into effect when cells are rapidly replicating, whereas for less active cells, resistance to UV may be a measure of the operational repair mechanisms. Johnson *et al.*, (1978) have reported that there is increasing evidence that bacterial photoreactivation (recovery after exposure to light) may be a significant concern in UV-disinfection of secondary effluent which is commonly discharged into open receiver systems in the United States. Harris *et al.* (1987) also reported that sunlight penetrating the receiver system may reactivate a significant fraction of the UV-inactivated bacteria. Photoreactivation often repairs much of the damage to the DNA. However, viruses lack the repair enzymes for the photoreactivation processes (Jagger, 1967). Whitby *et al.*, (1984) also reported evidence of reactivation in *fecal streptococci*. Potential photoreactivation of UV-inactivated coliforms in secondary effluent has also been reported. The after growth studies also showed that 24h-incubationtime of UV-treated samples at 30^{0}C has resulted in normal bacterial counts (of untreated seawater). All samples had similar growth after 24h in spite of the UV-treatment. Therefore, it is reasonable to presume that in the SWRO plant, bacteria after the UV-treatment have received ample time to recover prior to the cartridge filter (CF). It is also possible that 1-2% live bacteria which tolerate UV-treatment could multiply in the storage tanks under favorable conditions such as availability of nutrients. The present study on the 24h-incubationtime of bacteria has shown that bacteria uses phosphate and nitrite. While phosphate concentrations were unaffected by the UV-treatment, an increase in nitrite levels after UV-treatment was noticed. However, at ACF the levels of phosphate and nitrite declined suggesting a nutrient trap prior to ACF, presumably due to consumption by bacteria in the storage tanks where the UV-treated feed water is stored. Nitrite is one source of nitrogen to organisms and only organisms possessing necessary enzymes can utilize nitrite as a source of cellular nitrogen (Gaudy and Gaudy, 1980). Fenchel and Harrison (1976) also showed that after inoculation of bacteria, the concentration of phosphate in the water rapidly decreased and the decrease was related to phosphorus consumption. It was also noticed that if the water was enriched with inorganic nitrogen source, bacterial growth was more rapid. In natural systems, the phosphate is released from bacteria when grazed by ciliates (Johannes, 1965). However, at $7m^3/h$ AUV and ACF had identical phosphate levels. Therefore at $7m^3/h$, increased flow appears to mask the difference between AUV and ACF. The laboratory results also suggested that nutrient consumption takes place during bacterial growth indicating that the observed nutrient consumption prior to ACF is due to bacteria. This showed that determination of micronutrients could be helpful to draw conclusions about the bacterial growth dynamics.

TOC values also showed a slight reduction after UV-treatment except at $7m^3/h$. However, a substantial reduction in TOC was found at ACF except at $2m^3/h$. Active bacteria exude dissolved organic matter in connection with the decomposition of organic material (Zobell, 1943). In static systems, the process consists of hydrolysis of particulate carbon using extracellular enzymes and later the dissolved organic-substrates are assimilated by the bacteria. The TOC value therefore depends on concentration of substrate, rate of bacterial metabolism including assimilation (Alexander, 1977). Jannasch and Pritchard (1972) showed that substrates at low concentrations were utilized more efficiently in the presence of inert particles such as sand but not in the absence of substrate or when it is present in high concentrations. Apparently the reduction of TOC observed at AUV in the present study is due to physical and chemical change caused by UV-radiation. Li *et al.* (1996) have also stated that UV-treatment can result in decarboxylation of compounds including humic acids. However, further decline in TOC at the ACF appears to be due to consumption by bacteria. Absence of such TOC consumption at 2 m^3/h flow rate (not an operational flow) remains unclear at this point. Results of the laboratory study also showed reduction in TOC levels associated with bacterial growth. More than 50% reduction in TOC was evident after 24h-incubation time.

The results of the present study showed that the significant increase in bacterial count at ACF is due to bacterial recovery. This data are comparable to the result from Al-Birk SWRO plant where the bacterial aftergrowth rates increased after the chlorination point and the aftergrowth rate was substantially high. Applegate *et al.*, (1989) also reported that the chlorine degrade humic acids (HA) into small molecules which are assimilable organic compounds (AOC) and since AOC favor bacterial growth this could result in bacterial aftergrowth. Li *et al.*, (1996) found that there is a slight decay of HA after UV-treatment presumably due to decarboxylation. Egil *et al.* (1991) also stated that UV-treatment of water containing humic substances may cause presence of oxidizing compounds such as single oxygen, hydrogen peroxide and OH-radicals. While details of chemical changes under UV-treatment is not clear at this point, it appears that the initial bacterial kill by the UV-treatment was followed by bacterial recovery down the pretreatment line. Therefore, mode of UV-bactericidal action may resemble chlorine oxidation. It has been observed that even at the same flow rate the percentage kill by the UV-radiation was subject to fluctuations at different sampling dates. This is to be expected because it is known that the efficacy of UV-radiation is affected by the suspended matter in water column. A turbidity increase could cause decline in the percentage of kill.

TECHNO-ECONOMIC ANALYSIS

Techno-economic analysis of UV-radiation showed an approximate cost of 0.73 SR per cubic meter of water (Table V). UV lamps require replacement after each 8000 hours and this appears to be the only maintenance cost for the UV-unit. Chlorination would cost about 0.75 SR per cubic meter of water with relatively high operation and maintenance cost to achieve similar disifection efficiency. The magnitude of bacterial recovery after initial kill of both technologies are comparable.

Table 5 Techno-economic analysis of UV-radiation vs Chlorination

Description	UV-radiation	Chlorine (approx)
Unit Cost	4348.00	2000.00
Operation Cost	Nil	4000.00
Maintenance Cost	3827.00 / Lamp	2000.00
Chemical Cost	Nil	448.00
Total	8175.00	8448.00
Cost / m^3	0.73	0.75
*Bacterial disinfection Efficiency (%)	90 - 99.9	**98.0
**Aftergrowth *(k/h)*	0.19	0.16

*Chlorine cost calculated at 2ppm dosing (hypochlorite 4.00 SR/m^3)
**(Munshi *et al.*, 1995)

CONCLUSIONS

The following conclusions were drawn from the present study.

1. The study showed a reduction of 90-99.9% in bacterial count after the UV-treatment.

2. At all the flow rates, the bacterial counts increased through the pretreatment stage specially at the AMF and ACF suggesting that the second UV-sterilization unit has failed to control bacterial growth. This also indicates bacterial recovery from UV-treatment, prior to ACF.

3. Laboratory studies have shown that bacteria recover from the UV-treatment within 24h.

4. The results of laboratory analysis of TOC and micronutrients suggest that the bacterial growth recovery following UV-treatment resembles chlorination disinfection.
5. The following are the advantages of the UV-radiation for disinfection: 1. Relatively less maintenance cost., 2. Requires less space for equipment., 3. Minimal supervision., 4. Unlike chlorination, UV-radiation does not facilitate corrosion and 5. No phase change in water and no toxic by-products.

In summary, UV-radiation appears to be a potential alternative to chlorination, if it is employed in plant with continuous flow of water. This would considerably minimize the operation and maintenance cost and avoid other plant problems allied with the chlorination, particularly discharge of toxic chlorine by-products such as trihalomethanes into the receiving water. However, the storage of UV-treated water in tanks during its passage through different stages of the plant could permit bacterial reactivation and recovery as evident in the present study. This could eventually affect the plant performance.

RECOMMENDATIONS

1. Since suspended particles could affect the efficiency of UV-radiation treatment, the UV-unit may be installed right before the SWRO membrane section.

2. To increase the performance of UV-radiation disinfection process in the SWRO pilot plant, stagnant conditions of water flow such as storage of water in the tanks may be avoided.

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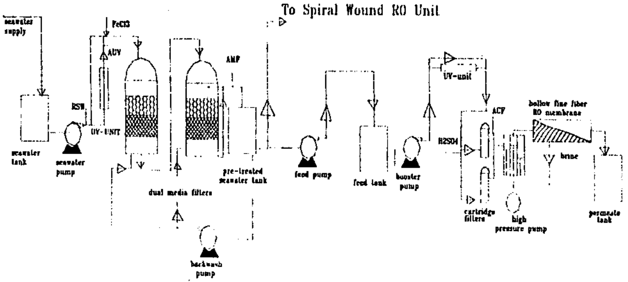
Appendix A



A Schematic Flow Diagram of The SWRO Pilot Plant Consisting of a common pretreatment and UV-Sterilization Unit

Appendix B

24 and 48h bacterial aftergrowth counts (CFU/ml) in seawater at different flow rates and stages of the SWRO pilot plant (48h aftergrowth values are given in parenthesis)

Sampling date	**Sampling stage**				
	RSW	**AUV**	**AMF**	**ACF**	**BR**
(2 m^3/h Flow rate)					
June 03, 1996	1.88x10^5	2.13x10^4	6.53x10^4	3.36x10^5	6.51x10^4
[June 04, 1999]	[NA]	[NA]	[NA]	[NA]	[NA]
June 05, 1996	9.66x10^4	TNTC	9.26x10^4	TNTC	TNTC
[June 06, 1996]	[1.46x10^5]	[4.03x10^5]	[1.41x10^5]	[1.3x10^5]	[4.06x10^5]
June 04, 1996	2.63x10^4	4.10x10^3	7.03x10^5	1.50x10^5	1.20x10^5
[June 05, 1999]	[NA]	[NA]	[NA]	[NA]	[NA]
Mean	1.04x10^5	1.27x10^4	2.86x10^5	2.43x10^5	9.25x10^4
[Mean]	[1.04x10^5]	[4.03x10^5]	[1.41x10^5]	[1.3x10^5]	[4.06x10^5]
(4 m^3/h Flow rate)					
June 09, 1996	1.24x10^4	8.53x10^3	1.23x10^4	6.68x10^5	2.66x10^5
[June 10, 1996]	[3.00x10^6]	[1.41x10^5]	[1.26x10^5]	[8.3x10^5]	[1.31x10^5]
June 10, 1996	3.65x10^4	9.80x10^4	4.70x10^4	3.20x10^5	9.96x10
[June 11, 1996]	[1.41x10^5]	[2.75x10^5]	[1.88x10^5]	[3.5x10^4]	[9.73x10^5]
June 11, 1996	3.07x10^5	3.25x10^5	5.93x10^4	8.70x10^4	1.41x10^5
[June 12, 1996]	[2.44x10^5]	[1.71x10^6]	[9.56x10^4]	[3.9x10^5]	[3.25x10^5]
June 12, 1996	5.40x10^4	1.31x10^5	3.23x10^4	3.05x10^4	5.65x10^4
[June 13, 1996]	[NA]	[NA]	[NA]	[NA]	[NA]

June 17, 1996	$7.80x10^{4}$	$2.43x10^{4}$	$2.00x10^{5}$	$3.30x10^{5}$	$1.42x10^{5}$
[June 18, 1996]	[$1.57x10^{4}$]	[$4.70x10^{5}$]	[$1.26x10^{5}$]	[$1.7x10^{5}$]	[$5.20x10^{5}$]
June 18, 1996	$8.05x10^{4}$	$4.40x10^{4}$	$7.10x10^{4}$	$1.56x10^{5}$	$1.32x10^{5}$
[June 19, 1996]	[$7.20x10^{4}$]	[$4.15x10^{5}$]	[$1.25x10^{4}$]	[$1.5x10^{4}$]	[$3.50x10^{6}$]
Mean	$1.12x10^{5}$	$1.26x10^{5}$	$8.44x10^{4}$	$3.18x10^{5}$	$1.67x10^{5}$
[Mean]	[$6.94x10^{5}$]	[$6.02x10^{5}$]	[$1.09x10^{5}$]	[$2.9x10^{5}$]	[$1.09x10^{6}$]
(7 m^{3}/h Flow rate)					
July 01, 1996	$3.04x10^{5}$	$1.40x10^{6}$	$4.60x10^{6}$	$1.32x10^{4}$	$1.27x10^{5}$
[July 01, 1996]	[TNTC]	[$5.53x10^{5}$]	[$5.11x10^{5}$]	[TNTC]	[$5.33x10^{5}$]
July 03, 1996	$1.40x10^{5}$	$6.07x10^{5}$	$3.96x10^{4}$	$2.44x10^{5}$	$8.63x10^{4}$
[July 03, 1996]	[NA]	[NA]	[NA]	[NA]	[NA]
July 07, 1996	$5.70x10^{3}$	$1.31x10^{5}$	$7.55x10^{4}$	$3.10x10^{5}$	$2.89x10^{5}$
[July 08, 1996]	[$1.30x10^{5}$]	[$1.45x10^{5}$]	[$2.40x10^{4}$]	[$4.8x10^{5}$]	[$5.36x10^{5}$]
July 08, 1996	$3.00x10^{4}$	$3.52x10^{5}$	$3.29x10^{6}$	$8.00x10^{5}$	$2.02x10^{5}$
[July 09, 1996]	[$4.00x10^{5}$]	[$1.48x10^{5}$]	[$4.45x10^{5}$]	[$6.2x10^{4}$]	[$3.40x10^{5}$]
July 09, 1996	$1.95x10^{4}$	$1.66x10^{5}$	$1.78x10^{4}$	$1.77x10^{5}$	$2.10x10^{5}$
[July 10, 1996]	[$1.05x10^{5}$]	[$2.05x10^{5}$]	[$2.00x10^{4}$]	[$1.3x10^{5}$]	[$1.35x10^{5}$]
July 10, 1996	$1.76x10^{5}$	$2.65x10^{5}$	$1.07x10^{4}$	$2.07x10^{5}$	$1.03x10^{5}$
[July 11, 1996]	[NA]	[NA]	[NA]	[NA]	[NA]
Mean	$1.12x10^{5}$	$4.87x10^{5}$	$1.34x10^{6}$	$2.92x10^{5}$	$1.69x10^{5}$
[Mean]	[$9.16x10^{5}$]	[$2.62x10^{5}$]	[$2.04x10^{5}$]	[$2.2x10^{5}$]	[$3.86x10^{5}$]

Raw Seawater and Pre-Treatment Process Investigations at the Addur SWRO Desalination Plant—The Crux to Pre-Treatment System Rehabilitation

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RAW SEAWATER AND PRE-TREATMENT PROCESS INVESTIGATIONS AT THE ADDUR SWRO DESALINATION PLANT - THE CRUX TO PRE-TREATMENT SYSTEM REHABILITATION

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ABSTRACT

The **raw seawater and pre-treatment process investigations at the Addur SWRO Desalination Plant** is, in reality, **the crux to pre-treatment system rehabilitation**.

These investigations stand fundamentally crucial in determining the quintessential basis for the final configuration of the pre-treatment process, as they play a vital role in determining the suitability of the pre-treatment design proposed by the consultant (M/s. *SWECO International*) as a solution to the existing pre-treatment problems at Addur.

The pilot plant testing and analogous laboratory analyses, which constitute the investigation work, were initiated towards the end of January 1998. The tests were jointly performed by the *Ministry of Electricity and Water* staff at *Addur*, together with the technical representative of the consultant, both in Bahrain and Stockholm, Sweden. The endeavour was to envision and formulate the selection of the new pre-treatment system design, based on the investigation results, and would seriously reflect on pre-treatment and RO membranes performance during subsequent plant operation following the rehabilitation.

In essence, the paper brings to light the details of the raw seawater and pre-treatment process investigations performed at *Addur*. The work mainly involved *laboratory tests* on chemical flocculation/flotation/dual media filtration, *pilot plant operation* on chemical flocculation/dual media filtration (with/without flotation), and *pilot plant operation* with membrane filtration. The study also went beyond to unfold the neoteric technology of *UF-membrane filtration* concept harnessed in this venture. The re-design approach has not yet been consummated, but would materialize by the time of paper presentation.

KEY WORDS : Chemical flocculation; flotation; dual-media filtration; ultra-filtration (UF) membranes for seawater pre-treatment; silt density index (SDI).

INTRODUCTION

The shortfall in pre-treatment, augmented by unfavourable environmental and seawater conditions which take place from time to time all-the-year-round, all combined have resulted in poor quality of pretreated seawater, which ultimately caused severe damage to the membranes at Addur, thus, signifying the necessity for pre-treatment rehabilitation.

Fundamentally, the investigation work included : **(1)** *laboratory tests* on chemical flocculation and dual media filtration, intended to improve pretreated water quality in terms of lower SDI; **(2)** *laboratory tests* on chemical flocculation/flotation/dual media filtration, intended for further removal of colloidal matter and fine particles, and to adapt optimum operation-conditions in the pilot plant tests; **(3)** *laboratory tests* on membrane filtration, principally micro-filtration (MF) and ultra-filtration (UF), as an alternative to conventional pre-treatment with chemical flocculation, intended to select two suitable membranes for a site *pilot plant trial-operation*; **(4)** *pilot plant operation* on chemical flocculation/dual media filtration with/without flotation (performed during January-June 1998); and **(5)** *pilot plant operation* with membrane filtration (as an alternative to conventional pre-treatment). All investigations were intended to evaluate the design parameters for a full-scale plant.

Up to the end of June 1998, there were *12 Pilot Plant Test Cycles* performed. The tests were carried-out by the Addur Laboratory staff, based on consultant directions, where the results of each cycle were examined (by the consultant) as to formulate the basis for the subsequent cycle.

Figures 1 and 2 illustrate the Flocculation/Flotation/Filtration Pilot Plant, and the UF Membrane Pilot Unit as "Membrane Pre-treatment" utilizing UF *SW-Membrane* Elements.

HISTORICAL OVERVIEW OF PRE-TREATMENT PROBLEMS

Since the early days of plant commissioning in 1990[1], the Addur SWRO Desalination Plant has been going through a series of predicaments. The most crucial was the inability of the RO System to achieve the plant design production of 10 MIGD. This was perceived as a direct reflection of the abnormal feed water conditions materializing in the high SDI of RO feed water exiting the pre-treatment system as a result of pre-treatment shortfall and its inability to cope with seasonal changes in seawater state. The persisting problem associated with the shortfall in pre-treatment was recognized, essentially, in the failure of the Dual Media Filters (DMFs) to yield the design filtered water SDI of $\leq$ **2.7**. Over the years, this problem resulted in a continual rise in ΔP across the membranes, deterioration in RO productivity and an ever-rising permeate salinity.

When the Addur plant was handed-over in June 1993, many ventures were initiated toward the improvement of the pre-treatment system in order to resolve the major problems suspected of contributing to the *high SDI* circumstance being experienced. In its entirety, the major trials performed on the pre-treatment alone involved :

i) Variation of pH across the DMFs to achieve an improved coagulation refinement.

ii) Variation of residual-chlorine concentration across the pre-treatment to assess chlorination efficacy and chlorine demand.

iii) Study of dechlorination efficiency and consequences.

iv) Variation of coagulant and coagulant-aid dosing rates to establish new guidelines for an improved coagulation/flocculation process.

v) Revising the DMFs' backwash sequence and backwash cycles.

vi) Investigation of DMFs and Cartridge Filters (CFs) performance efficacy/durability.

vii) Analytical investigations of sludge samples from the DMFs feed channels and *mud-balls* extracted from the DMFs beds.

During the final quarter of 1995, two major modifications, involving pre-treatment chemicals, were carefully studied and implemented. Firstly, the coagulant and coagulant aid dosing points were brought up-stream the

Contact Tank, essentially, to offer longer residence time for coagulation/ flocculation to transpire. Secondly, the sodium bi-sulphite (SBS) dosing point (dechlorination) was relocated from the CFs outlet to the immediate suction points of the individual high pressure pumps, which was necessary to provide an extended chlorine exposure in the piping upstream the RO Trains to delay biological re-growth that normally succeeds dechlorination.

However, all trials and modifications provided trivial improvements, in terms of pre-treatment SDI reduction, whilst pre-treatment rehabilitation was foreseen as the only solution to the problem.

OUTLINE OF INVESTIGATION TRIALS AND RESULTS

The raw seawater investigation involved *12 Laboratory and Pilot Plant Operation Test Cycles*, (performed during 19th January - 23rd June 1998), with each being concluded within a *6-day* period. The Laboratory Tests (Jar Tests) were followed by Pilot Plant Tests so as to verify the Laboratory Tests results, with the intention of investigating each parameter separately for optimization. **Tables I, II, III** and **IV** comprise the **results** and **analyses** of the **final *test-day*** of each cycle, as being most representative.

Cycle 1 (19th - 27th January 1998)

Ferric chloride ($FeCl_3$), as coagulant, cationic polymer (Superfloc C573), as coagulant aid, and sulphuric acid (H_2SO_4), for pH adjustment, were tested. Laboratory Tests were planned such that a direct comparison between chemical pre-treatment with/without flotation was possible. $FeCl_3$, H_2SO_4 and polymer dosing rates between 1.5-5.0 mg/l (as Fe^{+++}), 25-75 mg/l and 0-0.4 mg/l, respectively, were evaluated. The optimum rates were 2.5 mg/l Fe^{+++}, 25 mg/l H_2SO_4 and 0.2 mg/I polymer. SDI results were higher than expected (> **4.0**).

Pilot Plant Tests were based on the optimum $FeCl_3$ and H_2SO_4 dosing rates obtained through the Laboratory Tests. No polymer was added during *day-1*, while 0.2 mg/l was used during *day-2*. Detention time for flocculation was 15 minutes (used for all cycles), with the agitation rates in the three compartments of the Flocculation Tank set at 28/24/14 rpm. Two recycle flows for flotation were tested, 10% and 20%, at 6 Bar air dissolution pressure.

The addition of 0.20 mg/l polymer was favourable, with respect to colloidal matter removal. The SDI was **slightly lower** after **flocculation/flotation/**

dual media filtration (**3.10**) compared to corresponding conditions **without flotation** (**3.13**).

Cycle 2 (7th - 14th February 1998)

Laboratory Tests indicated that 2.5 and 2.0 mg/l Fe^{+++} showed similar results at 25 mg/l H2SO4, 0.25 mg/l polymer and 75 rpm. Optimum H_2SO_4 dosing rate was 25 mg/l with/without flotation, yet it was difficult to determine whether 0.25 mg/l polymer was better than 0.50 mg/l. Residual iron after the laboratory DMF was extremely low (<0.01 mg/l), but, conversely, the SDI value was extremely high (**5.5-6.0**). The test was repeated, and SDI reduced to **4.1-4.2**.

2.5 mg/l Fe^{+++}, 25 mg/l H_2SO_4 and 0.25 mg/l polymer were used in the Pilot Plant Tests, at 22/22/22, 33/33/29 and 44/44/29 rpm flocculation rates, 10% recycle flow and 6 Bar air dissolution pressure. The results showed significantly lower SDI values at higher rotation rates. The **lowest SDI** values obtained after **DM-filtration with/without flotation** were **3.01/ 3.13**, i.e. about **0.12** SDI-unit lower with flotation.

Cycle 3 (16th February - 1st March 1998)

Laboratory Tests performed on the agitation rates between 50-125 rpm (with 2.5 mg/l Fe^{+++}, 25 mg/l H_2SO_4 and 0.25 mg/l polymer) indicated that 75 rpm was better without flotation, but differences were insignificant with flotation. During *day-3*, 2.0 mg/l Fe^{+++} was selected, at 125 rpm and filtration through laboratory DMF. All SDI values after filtration were considerably high (> **5.2**).

Pilot Plant Tests were performed to investigate the influence of $FeCl_3$ dosing rate within 1.0-2.5 mg/l Fe^{+++}, at constant H_2SO_4 and polymer dosing rates (25 mg/l and 0.25 mg/l, respectively) and 33/33/29 rpm. 1.0 mg/l Fe^{+++} was too small, with respect to floc characteristics. The rotation rates between 22-33 rpm seemed better. The **lowest SDI** values **with/without flotation** were **2.91/3.05**, i.e. about **0.14** SDI-unit lower with flotation.

Cycle 4 (8th - 12th March 1998)

Laboratory Tests showed 1.5 mg/l Fe^{+++} being too small for flotation, but was not so for flocculation. Effect of 0.25-0.50 mg/l polymer seemed negligible for flocculation, while 0.50 mg/l was better before DM filtration.

2.0 mg/l Fe^{+++} was tested with and without 25 mg/l H_2SO_4 and 0.25 mg/l polymer, at 75 and 125 rpm; the results indicated that 125 rpm was better for flocculation, while 75 rpm was better for flocculation/flotation before DM filtration.

Pilot Plant Tests were conducted to investigate the effect of declining rotation rates (33/33/29, 56/33/29, 56/44/29 and 44/44/29 rpm), with 2.0 mg/l Fe^{+++}, 25 mg/l H_2SO_4 and 0.25 mg/l polymer. SDI measurements, after DM filter back-washing, were carried-out until a stable value was obtained, (which occurred after 4 hours filtration). The **best SDI** results (of **2.84**), after **DM filtration**, were obtained at 56/44/29 and 44/44/29 rpm.

Cycle 5 (17th - 19th March 1998)

Laboratory Tests performed with 2.0 mg/l Fe^{+++}, 25 mg/l H_2SO_4 and 0-0.50 mg/l polymer, at 125 rpm, showed humic acids (HA) concentrations between 0.39-0.74 mg/I.

Pilot Plant Tests performed with 2.0 mg/l Fe^{+++}, 25 mg/l H_2SO_4 and 0.25 mg/l polymer, at 33/33/29 and 56/44/29 rpm rotation rates, respectively, showed **SDI** results of **3.07** and **2.91 with flotation**, and **3.18** and **2.99 without flotation.**

Cycle 6 (17th March - 2nd April 1998)

Laboratory Tests were performed to evaluate different polymers, namely Magnafloc LT20 (nonionic), Magnafloc LT22S (30% cationic), Magnafloc LT25 (20% anionic), Praestol 611TR (weak cationic), Praestol 650TR (middle-strong cationic), Praestol 2530TR (middle-strong anionic) and Fennopol N200 (nonionic). Polyaluminium chloride (KEMWATER PAX-XL60S) was also tested as coagulant.

2.0-2.5 mg/l Fe^{+++} and 0.8-2.0 mg/l polyaluminium chloride were tested separately with 0.25 and 0.50 mg/l polymer, at 125 rpm. While $FeCl_3$ was tested with 25 mg/l H_2SO_4, polyaluminium chloride was tested with and without H_2SO_4, as pH-adjustment is known to be less sensitive with the latter coagulant. 0.50 mg/l Praestol 611TR showed better results with $FeCl_3$, in terms of turbidity. Polyaluminium chloride tests were performed with 0.25 mg/l Superfloc C573, and two with 0.50 mg/l Fennopol N200.

With 2.5 mg/l Fe^{+++}, 25 mg/I H_2SO_4 and 0.25 mg/l Superfloc C573, *residual Fe* was extremely high (0.12-0.13 mg/I) after flocculation and filtration through Whatman #1 filter paper, while corresponding *residual Al* was low (0.02 mg/l, on average). Optimum dosing rates were 1.4 mg/l polyaluminium chloride, 25 mg/l H_2SO_4 and 0.25 mg/I Superfloc C573, while results with 0.25 mg/l Fennopol N200 were a little better.

Pilot Plant Tests on flocculation/flotation/dual media filtration were performed for $FeCl_3$ and polyaluminium chloride in combination with the above polymers, at 56/44/29, 50/45/50, 44/44/44, 44/44/29 and 33/33/29 rpm. There was no distinct difference between the polymers tested. **Lowest SDI** values of filtered water obtained **with/without flotation** were **2.83/2.85** (for 2.0 mg/l Fe^{+++}) and **2.84/2.88** (for 2.0 mg/l polyaluminium chloride), with 25 mg/I H_2SO_4 and 0.25 mg/l Superfloc C573, at 50/45/50 rpm.

The Membrane Pre-treatment Unit, (Figure 2; Fluid Systems Ultra 8321 UF Spiral-Wound polyether-sulphone PES membrane, incorporating 3 X 8 “elements in one pressure vessel), was operated on 31st. March 1998. The operating conditions were 3 Bar feed pressure, 7 m^3/h recirculation flow (changed to 4 m^3/h from *Cycle 7* in order to decrease pressure loss), 90% recovery rate, and a retentate equaling 10% of feed flow. The total operating time during this cycle was 19 hours, while the calculated membrane flux on start-up was 39.7 I/m^2h, at 25°C, and decreased to 36.2 I/m^2h after 5.25 hours, and was 32 I/m^2h during the remainder of the test. The **SDI** value of **UF filtrate** was about **0.5**.

Cycle 7 (13th - 18th April 1998)

Pilot Plant Tests continued using 2.0 mg/l of Fe^{+++} and of polyaluminium chloride separately with 25 mg/l H_2SO_4 and 0.25 mg/l Superfloc C573, at constant rotations of 33, 44, 55 and 66 rpm, respectively. SDI results indicated that 44 was slightly better than 33 rpm, and similarly with 44 and 55 rpm, while it was the other way round with 55 and 66 rpm. With polyaluminium chloride, the difference between SDI values of filtered water with respect to rotations were smaller (**0.10** SDI-unit higher at 66 than at 33 rpm). The **lowest SDI** values for filtered water **with/without flotation** were **3.16/3.08**, with $FeCl_3$ at 55 rpm.

Membrane Pre-treatment Unit operation performed for 32 hours during 4 days. Flux decline form start-up till end of test period was 4.9 l/m^2h, at 3 Bar and 25°C, (i.e. after 53 hours of total operation time), corresponding to about 12% of the calculated membrane flux on start-up (39.7 l/m^2h).

Pressure loss calculations indicated that a *combined-feed* flow (= feed flow + recirculation flow) of 8 m^3/h not to be exceeded. Operating conditions were as in *Cycle 6*, with 4 m^3/h recirculation, and membrane flushing was carried-out for 5 minutes at the beginning of each day (using 200 litres of 10 mg/l NaOCl, at pH 10 and 4 m^3/h recirculation rate). The **SDI** value of **UF filtrate** was **0.42**.

Cycle 8 (23rd - 25th April 1998)

Pilot Plant Tests were performed with 2.0 and 2.5 mg/l Fe^{+++}, and 25 mg/l H_2SO_4 and 0.25 mg/l Superfloc C573, at 44 and 55 rpm. The results were best with 2.0 mg/l Fe^{+++} at 55 rpm, where the **average SDI** values **with/without flotation** were **3.76/3.88**. The **SDI** values obtained on the *final day* **with/without flotation** were **4.22/4.00**.

Membrane Pre-treatment Unit operation was performed for 22 hours, with the operating conditions as in *Cycle 7*. Membrane flux decreased form 34.3 to 27.6 l/m^2h, at 3 Bar and 25°C, corresponding to above 20%. The **SDI** values of **filtered UF filtrate** on each test day were **0.53**, **0.67** and **0.68**. It was evident that the abnormal seawater condition during the final day had no distinct predominance on the UF membrane performance, unlike the case with chemical pre-treatment.

It was also of interest to investigate whether or not the **SDI** is of an *equalizing-nature* when samples of different values are mixed. Hence, equal volumes of filtered water from chemical pre-treatment (SDI = **4.22**) and UF filtrate (SDI = **0.68**) were mixed. The SDI of the mixture was **3.30**, which was higher than the arithmetic average of the SDI values of both volumes (**2.45**). Therefore, the **SDI** does not seem to be an *equalizing parameter*; further investigation is required to confirm this conclusion.

Cycle 9 (29th April - 4th May 1998)

A finer sand quality, compared to the one used at Addur, was tested. The sand in one dual media filter was replaced with fine sand (used-sand) from the Ras Abu Jarjur RO Desalination Plant. SDI measurements after flocculation/flotation/dual media filtration for both columns were performed for performance comparison of both sand qualities.

Laboratory Tests, without flotation, were performed using 1.0, 1.5, 2.0 and 2.5 mg/l Fe^{+++} with 0.25 mg/l each of Superfloc C573 and Magnafloc LT20, while maintaining 25 mg/l H_2SO_4, 125 rpm, 15 minutes flocculation and

Whatman #1 filtration as constants. The results, evaluated with respect to turbidity and residual iron in filtered water (**Table V-A**), indicated 2.5 mg/l Fe^{+++} was the optimum, with little difference between the two polymers.

Pilot Plant Tests, with flotation, were performed using 2.5 mg/l Fe^{+++}, 25 mg/l H_2SO_4 and 0.25 mg/l Superfloc C573, at 33, 44, and 55 rpm. The flocculated and floated seawater was filtered through the two dual media filters (fine sand/Addur sand); on *day-1*, the SDI values were extremely high (**4.9/4.5**), and were still high on *day-2* (**4.4-4.5/3.8-3.9**), where the effect of rotation was trivial with flocs being bigger at the lowest rpm. On *day-3*, 2.0 mg/l Fe^{+++}, 25 mg/l H_2SO_4, and 0.25/0.75 mg/l Superfloc C573, was tested at 33 rpm, but there was no positive effect on SDI with 0.75 mg/l polymer (**4.4-4.5/3.9-4.1**). During the *final day*, the **SDI** results were a little better (**4.11/3.49**) with 0.25 mg/I Superfloc C573, where 38 rpm was selected as best. It was evident that fine sand has shown markedly higher SDI values of filtered seawater after flocculation/flotation than Addur sand.

Membrane Pre-treatment Unit operation performed as in *Cycle 8*. Following membrane cleaning using Fluid Systems Solution L1 (sodium tri-polyphosphate/tri-sodium phosphate and Na-salt of EDTA, at pH 10.5-11.5), increase in membrane flux was relatively small, (form 27.6 to 28.2 I/m^2h at 3 Bar and 25°C). The **SDI** was measured to be **0.70**.

Cycle 10 (9th - 12th May 1998)

Laboratory Tests were performed, without flotation, using 1.0, 1.5, 2.0 and 2.5 mg/l Fe^{+++} with 0.25/0.75 mg/l Superfloc C573 and 25/50 mg/l H_2SO_4, while maintaining 100 rpm, 15 minutes flocculation and filtration through Whatman #1 as constants. With respect to turbidity and residual iron, the results, after Whatman #1, were a little better with 50 mg/l H_2SO_4 (pH 6.22-6.24) compared to 25 mg/l (pH 6.71-6.76), while differences in coagulant dosing rates were small. Improved results through increasing H_2SO_4 dosing rate were valid for HA, with best results obtained at 2.5 mg/l Fe^{+++} and 50 mg/l H_2SO_4 (pH 6.22).

Analyses of seawater samples filtered through activated carbon (AC)/CFs outlet (**Table V-B**) showed more than 90% HA removal and complete removal of UV Absorbance, but increased bacterial and microorganisms concentrations after AC filtration.

Pilot Plant Tests were performed using 1.5/2.5 mg/l Fe^{+++}, 25/50 mg/l

H_2SO_4 and 0.25 mg/l Superfloc C573, at 40 rpm. **SDI** results after flocculation/flotation/dual media filtration were **best** with 2.5 mg/I Fe^{+++} (**3.1**), at pH 7.07-7.10 (25 mg/l H2SO4), while at pH 6.2-6.3 (50 mg/l H_2SO_4) **SDI** was **best** with 1.5 mg/l Fe^{+++} (**3.0**). The results indicated that increasing H_2SO_4 dosing rate could permit decreasing FeC13 dosing rate. During *final day*, with 1.5 mg/l Fe^{+++}, 50 mg/I H_2SO_4 and 0.25 mg/l Superfloc C573, **SDI** value was **3.17**.

Membrane Pre-treatment Unit operation was performed as in *Cycle 9*, in parallel with Chemical Pre-treatment, for 24 hours combined operation time (completing 108 hours from initial start-up). Rather than the contrary, UF filtrate flow increased with time from 26.6 to 28.2 l/m²h, at 3 Bar and 25°C. Following membrane cleaning, flux was not sufficiently restored; thus, cleaning was repeated using Solution L2 (100 mg/l CI_2 max. in a Na-salt of EDTA solution, at pH 10.5-11.5). Flux decline from the initial 40 l/m²h to about 28 l/m²h seems to be an irreversible decline that occurs to a varying degree for all new UF membranes.

On *days 2* and *3*, **SDI** values of UF filtrate were **0.65** and **0.54**, respectively. As in *Cycle 8*, SDI for an equal volumes blend of UF filtrate (**0.54** SDI) and dual media filtered water after flocculation/flotation (**3.17** SDI) was measured as **2.01**; i.e. higher than the arithmetic average SDI of the two volumes (**1.855**).

Cycle 11 (19th - 24th May 1998)

Laboratory Tests were carried-out using 1.0, 1.5, 2.0 and 2.5 mg/l Fe^{+++} with 25/50 mg/l H_2SO_4 and 0.25/0.75 mg/l Superfloc C573. Turbidity and residual iron were determined after filtration through Whatman #1, whereas HA, TOC and UV were determined after filtration through fresh AC; (**Table V-C**). HA dropped from 0.5 to 0.12 mg/l at pH 6.9, and to nil at pH 6.5, after filtration through AC.

Pilot Plant Tests were performed with flocculation/flotation, DM filtration and AC filtration, to investigate removal of organic colloids, using 1.5/2.5 mg/l Fe^{+++}, 25/50 mg/l H_2SO_4 and 0.25 mg/l Superfloc C573, at 40 rpm; (**Table V-D**).

SDI values for DM filtered water were lower at 2.5 mg/l Fe^{+++} and independent of pH, whereas the corresponding values for AC filtered water were considerably lower and relatively independent of $FeCl_3$ dosing rates. With 2.5 mg/l Fe^{+++}, 25 mg/l H_2SO_4 and 0.25 mg/l Superfloc C573, at 23-40 rpm, the SDI value after DM filtration was much higher (**4.0**) compared to that after AC filtration (**3.16**) at 26 rpm. On final day, at 40 rpm, the **SDI** value for **DM filtered water** still high (**3.91**), and increased to **3.69** for **AC filtered water**, which may be attributed to deterioration of the activated carbon with time.

Membrane Pre-treatment Unit operation performed as in preceding cycles, for 36 hours within 4 days. Membrane flux development, at 3 Bar and 25°C, was 29.9 l/m^2h at start of cycle (i.e. after 108 hours of total membrane operation) and was 28.3 l/m^2h at end of cycle (i.e. after 144 hours of total membrane operation). On final day, the **SDI** value of **UF filtrate** was **0.45**, while the value for an equal volumes blend of UF filtrate (**0.45** SDI) and DM filtered water (**3.91** SDI) was measured as **2.81**, i.e. higher than the arithmetic average of SDI values of both volumes (**2.18**).

Cycle 12 (13th - 23rd June 1998)

Pilot Plant Tests were performed with $FeCl_3$ and polyaluminium chloride to investigate more extensively the effect of pH (7.0, 6.0 and 5.0) and agitation rates on the removal of organic matter (HA).

With 2.5 mg/l Fe^{+++} and 0.25 mg/l Superfloc C573, the results showed lower SDI value of filtered water at pH 7.0 while the influence of rotations (40-60 rpm) was small. On the final day, using the same conditions at 60 rpm, the SDI value of DM filtered water was **3.13**. It was apparent that some colloidal matter could not be possibly removed by chemical flocculation, probably due to small colloidal size. With polyaluminium chloride, similar results were also obtained at pH 7.0, while the SDI value decreased from **3.28** to **3.05** when the polyaluminium chloride increased from 1.50 to 2.50 mg/l.

Membrane Pre-treatment Unit operation was performed as in preceding cycles (*6* to *11*), with an extended operation, up to 26th July 1998, for 86 hours (from 144 to 230 hours of total membrane operation), while membrane cleaning was carried-out using Solution L2 after 175 hours, on 16th June 1998. Membrane Flux improved from 27.1 to 28.6 l/m^2h, and dropped to 27.2 l/m^2h after 230 hours. On final day, the **SDI** value of **UF filtrate** was **0.75**.

At the end of *Cycle 12*, first UF membrane cartridge was sent for autopsy, and new UF membranes (*Trisep*, Spiral-Wound, Cut-off = 100,000) were installed for further trials.

OBSERVATIONS AND RESULTS INTERPRETATION

The following are the most vitally incisive observations and results on the overall Pilot Plant testing :

- All test cycles involved alterations of Pilot Plant chemicals' dosing rates (i.e. $FeCl_3$, H_2SO_4 and polymer), as well as the agitation rates for flocculation, for the main objective of optimising these parameters to achieve best results in terms of SDI.

- Results of FeCl3 and polyaluminium chloride tests were similar, with 2.0-2.5 mg/l Fe^{+++} and 1.4-2.0 Al^{+++} evaluated as optimum. The effect of polymer was relatively minor, with 0.25 mg/l being an optimum; none of the polymers tested showed better results than Superfloc C573 (which is in-use at Addur), providing insufficient improvement in terms of colloidal matter removal. Optimum pH for flocculation was around 7.0, corresponding to 25 mg/I H_2SO_4, with 15 minutes detention time being best. The optimum agitation rate for flocculation (in the three Flocculation Tank compartments) seemed to be around 40 rpm, with constant rates being more effective than declining rates, in terms of filtration-effect.

- Seasonal changes in weather and seawater conditions, which normally have adverse effect on the performance of the existing plant, were also affecting the performance of the chemical pre-treatment Pilot Plant in a similar manner, but had negligible influence over the UF Membrane performance.

- The sand and anthracite used in the Pilot Plant testing was *used-material* from Addur and the fine sand was also *used-sand* from the Ras Abu Jarjur RO Desalination Plant, while only activated carbon was fresh.

- As SDI is the most important parameter in the whole investigation, **Table VI/Chart 1** illustrate the **SDI** values obtained during the **Pilot Plant *final test-day*** of each cycle, as being the most representative and most relevant throughout the trials. The following are some interpretations of the tabulated results :-

    - On the final day of *Cycles 8* and *9*, a maximum SDI value of > **4.0** was obtained, which was a **reflection of unfavourable raw seawater condition** during those periods, while the **SDI** values of the **Cartridge Filters outlet** at the existing plant were **4.40/4.46**.

- Throughout *Cycles 1* to *12*, the highest and the lowest SDI values of filtered water obtained **with/without flotation** were **4.22/4.00** (*Cycle 8*) and **2.83** (*Cycle 6*)/**2.84** (*Cycle 4*), respectively. Thus, with **flocculation, flotation and dual media filtration Pilot Plant processes**, it was **only on a few occasions**, and with difficulty, that the **SDI** was **marginally reduced** to less than **3.0** (not even < **2.8**), which was achieved after enormous effort and numerous trials.

- Based on **chemical pre-treatment SDI** results during the final day of each cycle, the **SDI** values were ***40% of the time*** below **3.0 without flotation**, and ***29.4% of the time*** below **3.0 with flotation**, corresponding to an *overall 33.3%* of the time below **3.0**. On the other hand, based on **UF membrane pre-treatment SDI** results (*Cycles 6* to *12*), the **SDI** values of **UF filtrate** were ***100% of the time*** below **3.0**.

- The **average SDI** value of *Cycles 1* to *12*, final-day, was **3.261** after **flocculation, flotation and dual media filtration**, and **3.113** after **flocculation and dual media filtration**, i.e. the **average SDI** value was **higher with flotation**. **Final day SDI** measurements of only ***Cycles 1***, *2*, *3*, *5* and *6* have demonstrated that the **Flotation Unit** was **able to reduce the SDI** by an average of **only 0.077** SDI-unit.

DISCUSSIONS AND CONCLUSIONS

- The major objective of the Pilot Plant testing at the Addur SWRO Desalination Plant was to prove the adequacy of the Flocculation/ Flotation/Dual Media Filtration pre-treatment design proposed by the consultant to reduce the SDI value of raw seawater to a value ≤ **3.0**. The results showed that **only on a few occasions** the **SDI** was marginally brought down to < **3.0** (not even < **2.8**), as they were < **3.0** during *40%* of the time without flotation and *29.4%* of the time with flotation, corresponding to an *overall 33.3%* of the time being < **3.0**. These **SDI results** have **evidently indicated** that the **proposed design was inadequate** to satisfactorily achieve the desired objective, providing insufficient evidence to conclude the suitability of this design.

- The Ultra-Filtration (UF) membrane pre-treatment, introduced in *Cycle 6* as an alternative to chemical pre-treatment, has very sharply brought down the SDI of the non-pretreated/chlorinated seawater from the normal seawater SDI of about **16-19** to a value of < **1.0.** Even though, the operating and chemical cleaning/flushing aspects of this technique require further examination and research. Nonetheless, from an **SDI standpoint**, the **extremely low SDI** results of **UF filtrate** (*Cycles 6* to *12*, where the SDI was *100%* of the time < **3.0**), is worth giving the **concept of UF membrane pre-treatment process** some serious consideration as an **alternative solution to chemical pre-treatment at Addur.**

- **Rough seawater conditions**, which **adversely affect the existing plant performance**, were also perceived to **affect the Pilot Plant performance in a similar manner** (*Cycles 8* and *9*). It is, therefore, viable to conclude that the **Pilot Plant design**, as it stands, **cannot be considered as a solution to Addur pre-treatment problems**. Fundamentally speaking, **regardless of seawater state, any pre-treatment design worthy of consideration has to stand unaffected against all types of raw seawater conditions, as it is the case with the UF membrane pre-treatment.**

- ***Final-day*** **SDI** values of ***Cycles 1*, *2*, *3*, *5*** and ***6*** have shown that the **Flotation Unit** was **barely able** to **reduce the SDI** by an **average of 0.077 SDI-unit**. On the other hand, the ***final-day*** **average value of SDI** values of ***Cycles 1*** to ***12*** was **0.148 SDI-unit higher with flotation**. Hence, the **Flotation Unit** has **demonstrated** that **its contribution** to the **reduction** of **flocculated raw seawater SDI** value was **insignificant**.

- **Alterations** of **$FeCl_3$**, **H_2SO_4** and **polymer dosing rates**, ensued in the Pilot Plant trials, were in a **similar manner** to that **performed** on the **existing plant** during **intensive trials** one and a half years ago; even then, it was **not much of a success.**

- According to the Pilot Plant test results, the **optimum chemicals dosing rates** were evaluated as **2.0-2.5 mg/l Fe^{+++} ($FeCl_3$), 1.4-2.0 Al^{+++} (poly-$AlCl_3$), 0.25 mg/l polymer (Superfloc C573)** and **25 mg/l H_2SO_4** (with an **optimum pH for flocculation** of 7.0).

- The **agitation rates** in the **Flocculation Unit did not** seem to **have a consequential influence** on the **flocculation process,** in terms of SDI reduction. However, **40 rpm** appeared **reasonably best**, at a **flocculation detention time** of **15 minutes**, with **constant rates** being **more effective** from a **filtration-effect** perspective.

- The Pilot Plant Trials also involved testing of fine-sand and fresh AC. As foreseen, the **fine-sand** was **not much of a success**, where the **SDI** value for a **flocculated/floated/filtered water** sample was **0.62 SDI-unit higher** than that with the **Addur sand** (*Cycle 9, final-day* results). In the case of **fresh AC** (*Cycle 11*), **initial** test results showed **slight drop** in **SDI** value for a **flocculated/ floated/AC-filtered water** sample (just below **3.0**), but as the **AC** grew **saturated**, the **SDI** value became **similar** to **DM filtered water**. **Based on** these **results**, it is **concluded** that the **employment** of **AC** for **seawater pre-treatment at Addur** is **not promising.**

- Investigations performed on the efficiency of organic matter and microorganisms removal using AC-filtration, at lower pH, have shown improved HA removal over MD-filtration, of up to 75% (*Cycle 11*, Jar Tests results). However, **by virtue** of the **fast saturation tendency** of **AC**, the **resulting high SDI** value **does not validate** the **technique** as a **complete success**.

- While the general perception that the colloidal matter in the raw seawater at Addur is of an *inorganic nature*, the SDI results of the AC-filtered water sample suggest that it may be of an *organic nature*; still, it could very well be a combination of both.

- TOC analysis indicated that organic matter in raw seawater was inconsistent during the trial period, in the range 0.17 to 4.23 mg/l. It was noted that TOC content rapidly increased to 3.90 mg/l, at the beginning of May (*Cycle 9* results), as raw seawater temperature increased. It was clear that, on average, **TOC removal** was about **twice** as much **with UF membrane** compared to that **with flotation**.

- As **raw seawater temperature** was **increasing** during the trial period, from 16.8 to 33.6°C (*Cycles 1 to 12*, respectively), there was an **increasing trend** in **Total Bacterial Population** (AODC, total) from 29,090 to 127,700 Cells/ml, respectively.

EPILOGUE

In essence, this paper has been intended to bring to light the most essential results of the **Raw Seawater and Pre-treatment Process Investigations** performed **at The Addur SWRO Desalination Plant**, perceived as **the Crux to Pre-treatment System Rehabilitation**.

Over and above the outcome of the Pilot Plant investigations, the entire work has provided all participants with enhanced knowledge and experience in the fields of coagulation, flocculation, flotation and filtration. The *trial-study* went beyond to unfold a neoteric technology involving the **employment** of the **UF membrane filtration concept** harnessed as an **alternative** to **chemical pre-treatment** for **redesigning** the **pre-treatment system** at **Addur**.

The **consistent low SDI** values, of **< 1.0**, of the **UF membrane filtrate**, whilst **rough seawater conditions** pausing **insignificant dominance** over its **performance, firmly justify** granting the **UF membrane pre-treatment process serious consideration**, as it may **hold within its frontiers** the **mastery** of the **final configuration** of the **Addur future pre-treatment**. It is **envisioned** that the **UF membrane filtration technology** will have a **decisive role** in **seawater pre-treatment** within the **Arabian Gulf region**.

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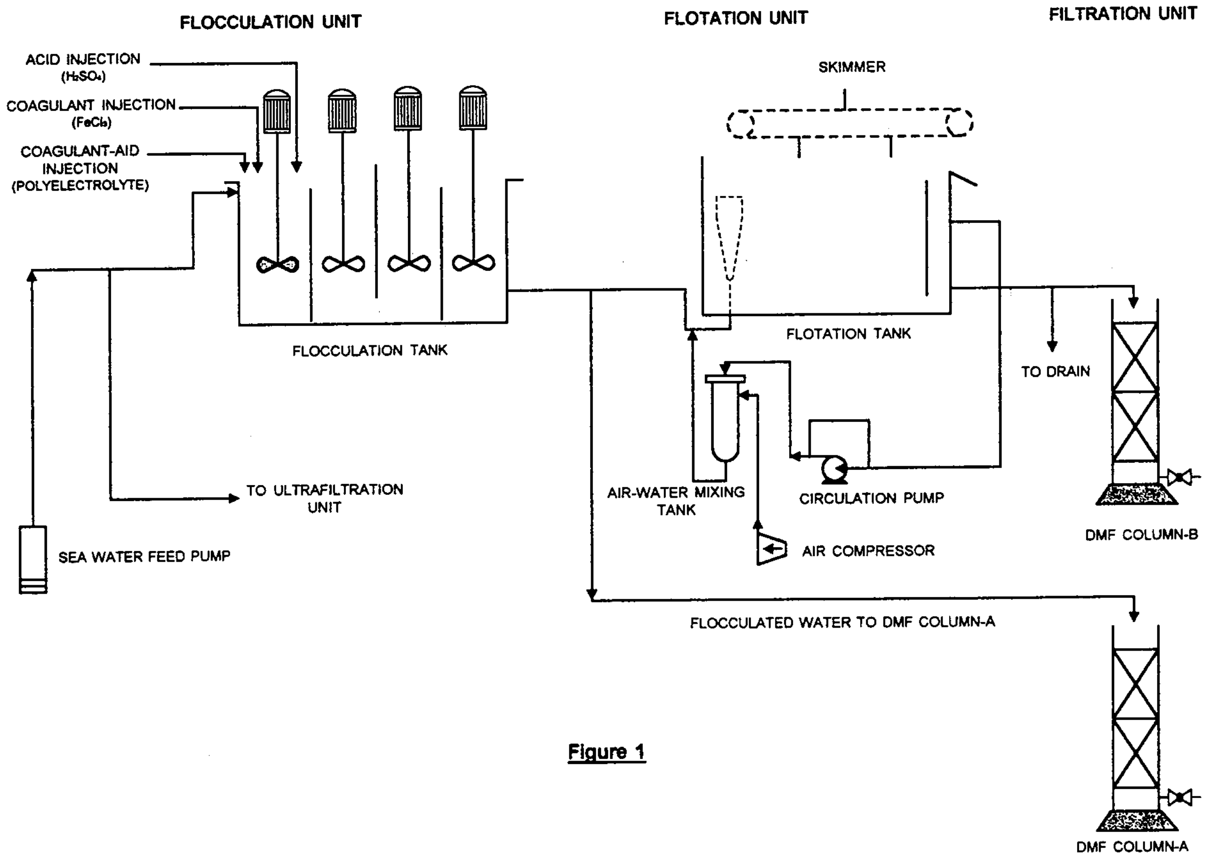


Figure 1

Figure 1: Flocculation, Flotation and Filtration Pilot Plant

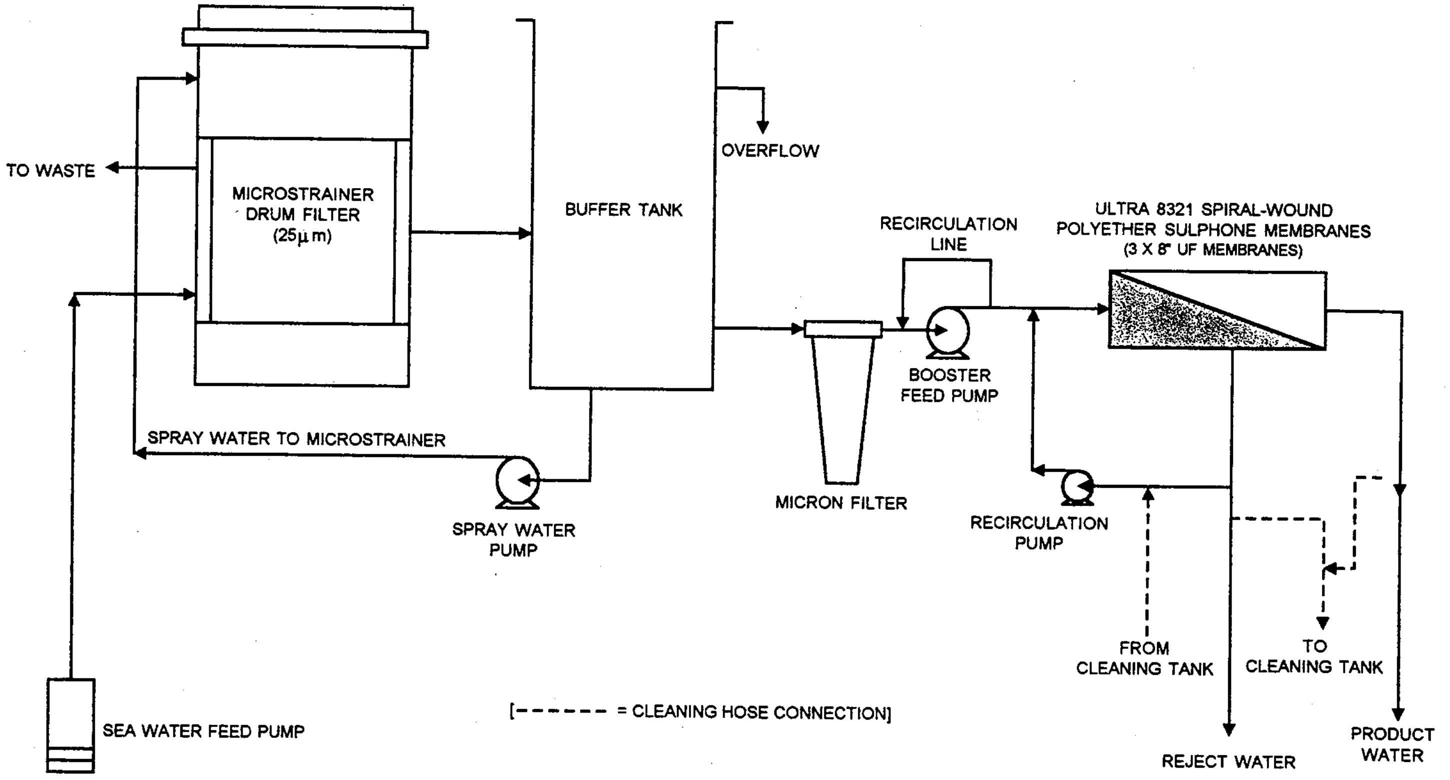


Figure 2: UF Membrane Pilot unit as "Membrane Pre-treatment" utilizing UF SW-Membrane elements

Table I: Raw Seawater Test Results and Analyses of Plant Samples Raw Seawater Investigations at Addur, 1998

	CYCLES											
DATES	1	2	3	4	5	6	7	8	9	10	11	12
PARAMETERS	26.01	14.02	01.03	12.03	19.03	02.04	18.04	25.04	04.05	12.05	24.05	23.06
Temperature (°C)	16.8	18.9	19.6	19.5	21.6	22.4	25.9	25.1	26.5	29.2	30.6	33.6
pH	8.21	8.18	8.27	8.21	8.24	7.97	8.07	8.18	7.99	8.07	8.14	8.29
Conductivity at 25°C (µS/cm)	63,600	63,100	64,000	64,700	63,700	64,100	64,200	64,100	63,900	64,600	65,900	65,800
TSS (mg/l) at 105°C	5	5	3	3	12	3	1	3	3	3	4	4
at 220°C	3	3	2	2	11	3	<1	3	2	2	3	1
at 550°C	1	2	1	1	10	1	<1	1	<1	1	1	<1
Turbidity (NTU)	2.1	2.25	2.60	2.40	28	4.10	2.10	5.5	2.30	2.40	2.40	2.80
Residual Chlorine (mg/l)	1.0	1.20	1.05	1.20	1.30	1.10	1.50	1.65	1.80	1.75	2.00	2.15
SDI	17.5	17.8	17.2	18.3	18.8	18.9	18.5	18.7	18.5	18.8	18.0	18.2
Total Iron (mg/l)	<0.01	<0.01		0.02	0.04	0.02	0.02	0.02	0.01	<0.01	<0.01	<0.01
TOC (mg/l)					2.11	1.11	0.90	(0.17)	3.90	3.2	4.23	
Humic Acid (mg/l)				0.35	0.59	0.37	0.43	0.51	0.37	0.45	0.49	0.50
Absorbance UV 254 nm	0.032	0.011	0.018	0.015	0.040	0.028	0.011	0.012	0.019	0.024	0.025	0.020
TBC, 96 hours (CFU/ml)	200	<100	<100		<100	<100	<100	200	<100	<100	<100	<100
AODC (Cells/ml)	29,090	42,820	43,630	54,940	44,440	50,900	47,670	59,790	56,560	71,100	79,990	127,700
INT (Cells/ml)	808	3,230	6,464	5,656	1,212	11,310	4,850	4,850	6,460	6,460	11,310	16,160

Table II: Results and Analyses of Test Samples "Without Flotation" Raw Seawater Investigations at Addur, 1998

	CYCLES														
	1	2 Jar/Pilot	3	4	5* Jar/Pilot	6**	6***	6 UF	7	7	8	8	10	11	12
DATES PARAMETERS	26.01	14.02	01.03	12.03	19.03	02.04	02.04	02.04	18.04	18.04	25.04	25.04	12.05	24.05	23.06
Temperature (°C)	18.3	-/20.9	21.2	22.1	22.2/ 22.2	25.9	25.1		28.6		25.5			31.4	
pH	6.78	7.60 7.04	6.75	6.89	6.90 6.94	6.79	6.67		7.08	8.16	7.02			8.14	
Conductivity at 25°C (µS/cm)															
TSS (mg/l) at 105°C							1		0.3		8				
at 220°C							1		0.2		3				
at 550°C							<1		<0.1		1				
Turbidity (NTU)	0.27	0.25 0.29	0.28	0.26	0.25 0.31	0.18	0.24		0.27	0.24	0.35	0.26	0.23	0.23	0.31
Residual Chlorine (mg/l)		0.15 0.15	0.20	0.20	0.20 0.25		0.25		0.40		0.40	1.0		1.4	1.5
SDI	3.13	4.20 3.13	3.05	2.84	3.18 2.99	2.85	2.88		3.08	0.42	4.00	0.68	0.54	0.45	0.75
Total Iron (mg/l)	<0.01	<0.01 <0.01		<0.01	-/<0.01	<0.01		<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
TOC (mg/l)					1.21/ 1.25		0.56	1.07	0.83	0.07	0.470	<0.01	0.61	3.51	3.04
Humic Acid (mg/l)				0.49	0.32/ 0.39		0.47	0.29	0.40	0.38	0.59	0.30	0.55	0.52	0.42
Absorbance UV 254 nm		0.011 0.011	0.014	0.08	0.014/ 0.015		0.008	0.020	0.013	0.004	0.002	0.001	0.008	0.012	0.018
TBC, 96 hours (CFU/ml)		39,800/ 2,300	600		-/100		100	<100	<100	<100	<100	<100	<100	<100	200
AODC (Cells/ml)		43,630/ 31,510	4,848	21,010	-/ 26,660		19,390	20,200	26,660	5,660	37,170	9,700	12,120	8,080	44,440
INT (Cells/ml)		2,424/ 2,424	Nil	8,080	-/8,890		8,080	6,460	6,460	1,620	2,420	Nil	2,420	2,420	8,480

* Rotation rates 33/33/29/56/44/29. **$FeCl_3$. ***Poly-$AlCl_3$

Table III: Results and Analyses of Test Samples "Without Flotation" Raw Seawater Investigations at Addur, 1998

	CYCLES															
	1	2 Jar/Pilot	3	4	5*	6(a)	6(b)	7	8	9(c)	9(d)	10	11(1)	11(2)	12(3)	12(4)
DATES PARAMETERS	26.01	14.02	01.03	12.03	19.03	02.04	02.04	18.04	25.04	04.05	04.05	12.05	24.05	24.05	23.06	22.06
Temperature (°C)	18.3		21.2	22.1	22.2/22.1	25.9	25.1	28.6	25.6	27.1	27.0	29.8	30.8	30.8	34.7	34.2
pH	6.75	7.59 7.04	6.74	6.88	6.89 6.93	6.79	6.66	7.05	7.00	6.98	6.98	6.22	6.95	7.30	6.98	7.01
Conductivity at 25°C (µS/cm)																
TSS (mg/l) at 105°C							<1	0.3	0.7	1	4	2	3	2	0.7	2
at 220°C							<1	0.1	0.3	<1	4	2	2	1	<0.1	2
at 550°C							<0.1	<0.1	<0.1	<0.1	3	1	1	0.3	<0.1	<1
Turbidity (NTU)	0.27	0.25/0.29	0.23	0.26	0.29/0.27	0.17	0.23	0.26	0.23	0.27	0.29	0.27	0.25	0.25	0.31	0.29
Residual Chlorine (mg/l)		0.10/0.10	0.20	0.20	0.20/0.20		0.25	0.35	0.20	0.35	0.15	0.45	Nil	Nil	0.50	0.30
SDI	3.10	4.10/3.01	2.91	2.84	3.07/2.91	2.83	2.84	3.16	4.22	3.49	4.11	3.17	3.91	3.69	3.13	3.05
Total Iron (mg/l)	<0.01	<0.01 <0.01		<0.01	-/<0.01	<0.01		<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
TOC (mg/l)					1.91/1.57		1.23	0.76	<0.01	3.2	3.4	2.9	3.27	2.23	3.28	3.54
Humic Acid (mg/l)				0.45	0.27/0.32		0.38	0.41	0.22	0.31	0.82 (1.53)	0.44	0.77	0.29	0.41	0.46
Absorbance UV 254 nm		0.009/0.005	0.008	0.01	0.010/0.012		0.008	0.012	0.001	0.018	0.007	0.007	0.008	0.002	0.016	0.011
TBC, 96 hours (CFU/ml)		35,700/1,900	1,900		-/1,400		<100	<100	<100	<100	3,700	<100	5,300	1,200		<100
AODC (Cells/ml)		32,320/48,480	10,500	15,740	-/19,390		12,120	17,780	16,970	12,120	25,050	14,540	37,170	29,900		20,200
INT (Cells/ml)		7,272/4,848	5,656	5,656	-/8,080		4,850	4,040	1,620	1,620	2,420	3,230	5,660	5,660		7,270

Note: * Rotation rates 33/33/29/56/44/29. (a) Fecl$_3$ (1)Dual media filtration (Addur sand) (b) Poly-AlCl$_3$ (2) Activated carbon filtration
(c)Addur Sand (3) as (1) with FeCl3 and Superfloc C-573 (d) Fine Sand (4) as (1) with Poly-AlCl3 and Fennopol N200

Table IV: Filters Outlet Test Results and Analyses of Plant Samples Raw Seawater Investigations at Addur, 1998

DATES PARAMETERS	CYCLES											
	1	2	3	4	5	6	7	8	9	10	11	12
	26.01	14.02	01.03	12.03	19.03	02.04	18.04	25.04	04.05	12.05	24.05	23.06
Temperature (°C)	16.8	18.7	19.8	19.6	21.7	22.6	25.9	25.1	26.5	29.5	30.0	33.3
pH	6.27	6.38	6.29	6.25	6.21	6.26	6.29	6.38	6.21	6.24	6.32	6.41
Conductivity at 25°C (µS/cm)	63,400	62,800	63,900	64,500	63,100	63,900	63,500	63,900	63,700	64,500	65,600	65,400
TSS (mg/l) at 105°C	3		3	2	0.3	1	0.5	3	3	3	1	3
at 220°C	2		2	1	0.1	<1	<0.1	2	2	2	<1	2
at 550°C	1		<1	<1	<0.1	<1	<0.1	0	<1	1	<0.1	1
Turbidity (NTU)	0.32	0.26	0.29	0.20	0.36	0.25	0.28	0.28	0.25	0.25	0.23	0.32
Residual Chlorine (mg/l)	0.55	0.55	0.50	0.40	0.50	0.50	0.50	0.50	0.50	0.60	0.65	0.55
SDI	4.13	3.44	3.89	3.62	3.68	3.59	3.75	4.40	4.46	3.74	4.23	3.43
Total Iron (mg/l)	<0.01	<0.01		<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
TOC (mg/l)						0.85	<0.01	0.06	3.43	3.48	3.73	
Humic Acid (mg/l)				0.23	0.30	0.27	0.34	0.32	0.27	0.50	0.64	0.48
Absorbance UV 254 nm	0.019	0.007	0.010	0.007	0.013	0.10	0.003	0.002	0.012	0.009	0.015	0.015
TBC, 96 hours (CFU/ml)	<100	<100	<100		200	<100	<100	<100	<100	<100	<100	<100
AODC (Cells/ml)	16,160	20,200	4,848	4,848	14,540	23,430	21,010	13,740	26,360	15,350	29,900	39,590
INT (Cells/ml)	Nil	4,040	2,424	808	4,040	5,660	3,230	2,420	4,850	1,620	4,850	4,850

Table V-A Cycle 9 : Laboratory Test Analyses (Performance Evaluation of Two Polymers)

Variables		Polymers	
		Superfloc C573	Magnafloc LT20
HA	mg/l	0.45	0.36
TOC	mg/l	4.96	4.65
TBC (96 hours)	CFU/ml	<100	<100
AODC	Cells/ml	27,470	13,740
INT (Direct Microscopic Technique)	Cells/ml	8,080	4,040
Turbidity	NTU	0.29	0.26
residual Fe	mg/l	0.026	0.026

Table V-B Cycle 10 : Laboratory Test Analyses (Laboratory AC-Filter vs. Existing CFs Outlet)

Variables		Filtrate Type	
		AC Filtrate	CFs Outlet
HA	mg/l	0.03	0.05
TOC	mg/l	2.15	3.48
TBC (96 hours)	CFU/ml	3,600	<100
AODC	Cells/ml	22,630	15,350
INT (Direct Microscopic Technique)	Cells/ml	10,500	1,620
UV Absorbance (at 254 nm)		Nil	0.009

Table V-C Cycle 11 : Laboratory Test Analyses (Polymer Performance at Different pH)

Dosing Rates, mg/l			Floccul-ation		After Filtration (mg/l)				
H_2SO_4	Fe^{+++}	Polymer	rpm	pH	NTU	Fe^{+++}	HA	TOC	UV
25	2.5	0.25	100	6.92	0.24	0.010	0.115	2.3	0.00
50	1.0	0.75	100	6.55	0.26	0.018	0.001	4.2	0.01

Table V-D Cycle 11 : Pilot Plant Test Analyses (Evaluation of Filtration Media, in terms of SDI)

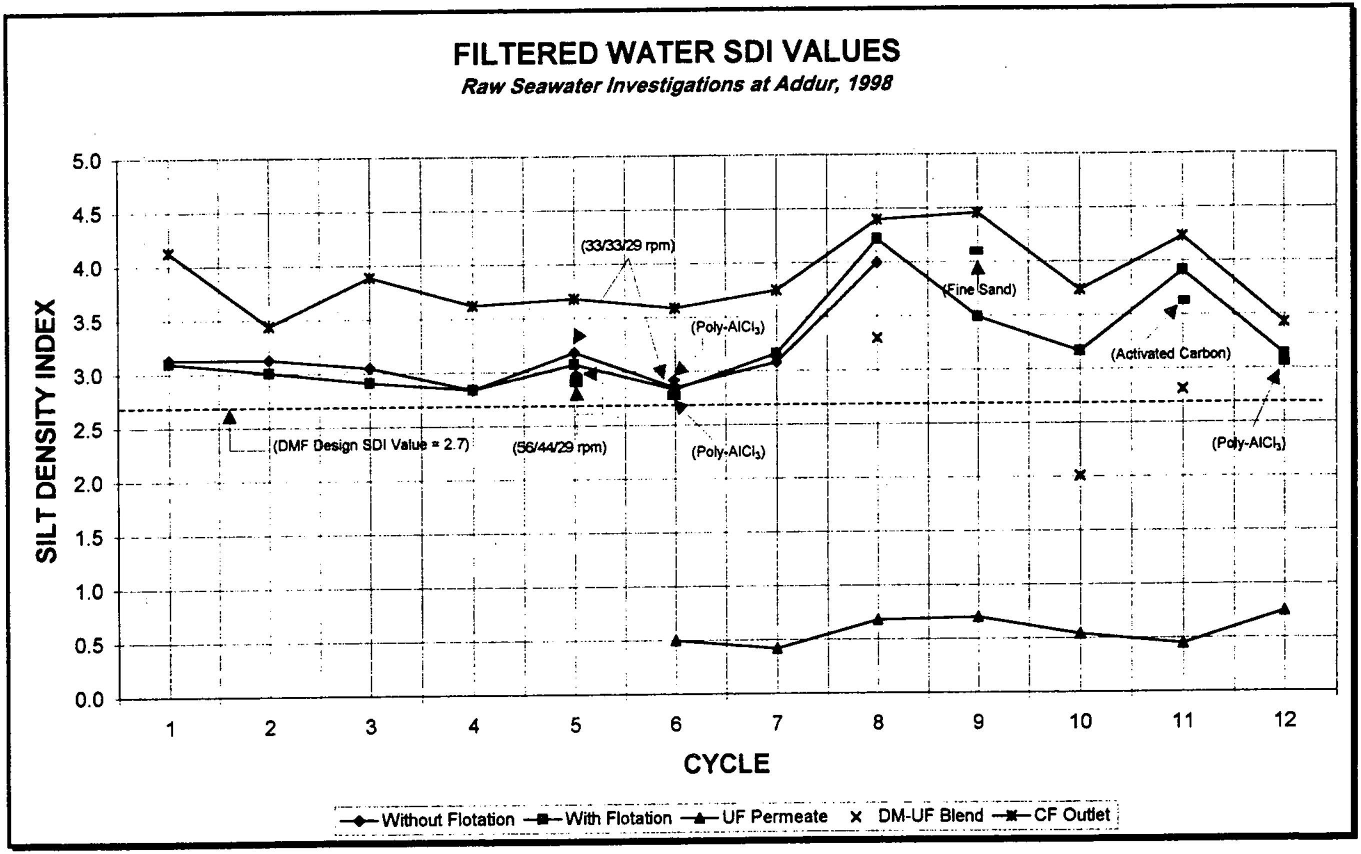
Day	Dosing Rates, mg/l				Filtered Water SDI	
	H_2SO_4	pH	Fe^{+++}	Polymer	DM	AC
1	25	6.95	2.5	0.25	**3.53**	**2.93**
	25	7.00	1.5	0.25	**3.75**	**3.02**
2	50	6.20	2.5	0.25	**3.58**	**3.36**
	50	6.20	1.5	0.25	**3.79**	**3.27**

Table VI Cycles 1 to 12
Pilot Plant Final Test-Day Filtered Water SDI Values
Raw Seawater Investigations at Addur (19th January to 23rd June, 1998)

Cycle	**Filtered Water SDI Values (Final Test-Day)**				
	Without Flotation	**With Flotation**	**UF Filtrate**	**DM-UF Blend**	**CFs Outlet**
1	**3.13**	**3.10**	-	-	**4.13**
2	**3.13**	**3.01**	-	-	**3.44**
3	**3.05**	**2.91**	-	-	**3.89**
4	**2.84**	**2.84**	-	-	**3.62**
5	**3.18** (1) **2.99** (2)	**3.07** (1) **2.91** (2)	- -	- -	**3.68** **3.68**
6	**2.85** (3) **2.88** (4)	**2.83** (3) **2.84** (4)	**0.5**	-	**3.59**
7	**3.08**	**3.16**	**0.42**	-	**3.75**
8	**4.00**	**4.22**	**0.68**	3.30	**4.40**
9	-	**3.49** (5) **4.11** (6)	**0.70**	-	**4.46**
10	-	**3.17**	**0.54**	**2.01**	**3.74**
11	-	**3.91** (5) **3.69** (7)	**0.45**	**2.81**	**4.23**
12	-	**3.13** (3) **3.05** (4)	**0.75**	-	**3.43**

(1) 33/33/29 rpm
(2) 56/44/29 rpm
(3) $FeCl_3$
(4) poly-$AlCl_3$
(5) Addur Sand
(6) Fine Sand
(7) Activated Carbon

Chart 1 : Filtered Water SDI Values Raw Seawater Investigations at Addur, 1998



Productivity Increase of High Salinity Brackish Water Reverse Osmosis Plant— The Case of Bahrain

Dr. Khalid A. Burashid and Khalid E. Almansour

PRODUCTIVITY INCREASE OF HIGH SALINITY BRAKISH WATER REVERSE OSMOSIS PLANT – THE CASE OF BAHRAIN

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ABSTRACT

Through the extensive studies and experimentation, the operational recovery of Ras Abu Jarjur reverse osmosis plant was increased from 65% to 68%, with one R.O train at 70% for the first time since plant's commissioning in 1984. As a result, the permeate production increased by one million gallon per day (1 MGPD).

During may 1998, the first operation of Ras Abu Jarjur with eight trains commenced via parallel operation of the standby train simultaneously making it possible to achieve an additional 1.5 – 1.7 million gallon/day of permeate water, thus increasing the total possible throughput of this facility from 10 MGPD to 12.5 – 12.7 MGPD. Thus, the added permeate capacity increased by 2.5 – 2.7 MGPD. Future prospects for further production increase and upgrade strategy will also be discussed.

INTRODUCTION

The plant employs Du Pont's *B-10* Permasep Permeators for the desalination of high brackish ground-water, containing *13,000 mg/l* TDS, obtained from aquifer "C" . (Appendix 1)

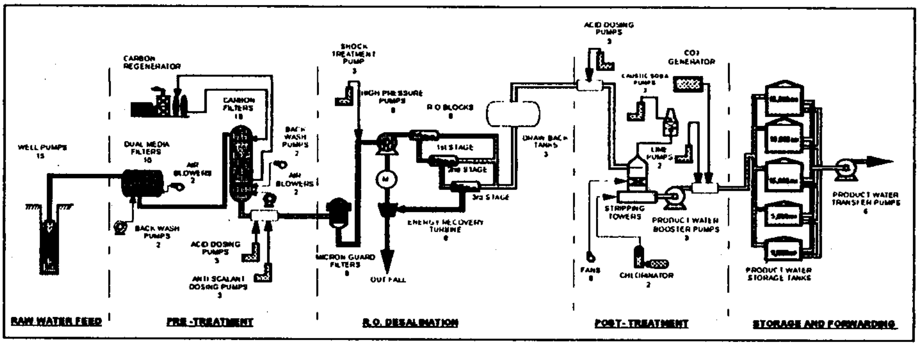


Figure 1: Simplified schematic of Ras Abu Jarjur R.O plant

Feed water is passed through activated carbon filters for the adsorption of soluble hydrocarbons, followed by sulfuric acid and scale inhibitor for the prevention of carbonates and sulphates scaling. The feed is then passed through *8* micron guard filters wherein suspended particles, down to *5 μm*, are removed.

At the RO, filtered-treated water is fed through *7* High Pressure Pumps (plus one standby), each feeding an RO Block comprising 3 Brine-Staging [*5:3:2* Ratio].

The control system of the plant is based on three local programmable controllers linked to a computer – based control and monitoring system.

During the early years of commissioning, the plant had developed a history of biofouling problems (basically the rapid proliferation of bacteria) which affected the overall plant performance. The SHMP, which was used as an antiscalant at that time, was then replaced by a polymer antiscalant, which resulted in a significant improvement of plant performance and availability. [3,5]

POSSIBLE ALTERNATIVES TO INCREASE WATER PRODUCTION

Increase of operating recovery

Increasing the over all operating recovery at the operating feed/ brine set-up is the most cheapest alternative to boost the production. Such increase can be implemented gradually under a very close monitoring to insure safe operation below the threshold limits of fouling/scaling. The recovery increase was examined under two-phase scenarios:

A: Increase Recovery from 65% to 70% :

An incremental increase of plant's throughput is expected to reach approximately 8%, i.e. a total production of 50,000 m^3/day or an additional 0.8 MGPD based on actual throughput of 280 m^3/h per train.

One of the concerns would be the increase in the salt passage and slight increase in energy requirement. The calculations showed that an additional 19 - 22 permeators/ train would be required.

B: Increase Recovery from 65% to 75% :

The increase of the operating recovery to as high as 75% would result in an incremental increase of production by 15.4%, i.e. 53100 m^3/day (~ 11.7 MGPD). It is essential that S & DSI values should be very closely monitored at such high recovery, though precipitation is likely to occur in which case the antiscalant / acid dose level should be adjusted accordingly to alter the saturation levels of some salts to safeguard the membranes from scaling and maintain the same cleaning frequency. A slight energy increase is foreseen and an additional 51 permeators / train will be required.

It is recommended that should a decision be made in future to explore the 75% recovery, an extensive pilot tests shall be conducted to fully assess the performance of the permeators under such conditions.

Use of second pass stream - The 4th stage

In attempt to arrive at the most techno-economical solution and with conservation of feed water in mind, various approaches have been experimented to use the brine reject as a feed water to a second pass, and as far as system design and water chemistry will allow. In this analysis, the brine pressure leaving the third stage will be considered as the feed pressure in order to minimize the costs. Two scenarios were examined, however, with 8 trains in operation.

1) One additional pass using Dupont 6845 T permeators for each R.O. block :

The projection calculations were based on eight R.O blocks operating at the same time (individual pass per train).

The over-all production is expected to increase from 53700 m^3/day (11.82 MGPD) to 60,700 m^3/day (13.36 MGPD). Here, an additional 5952 m^3/ day (1.54 MGPD) of permeated water will be produced without the need to install additional wells and equipment in the pre-treatment area (2 wells, MGF, transformers, high pressure pumps and associated switch gears, etc.), thus less over all costs and faster execution period.

The drawback of such scenario is that a complete additional R.O building shall be constructed to accommodate the new trains, instead of simply extension of the existing building.

It is foreseen that a total of 416 permeators will be required (52 per train).

Due to high saturation level of sparingly soluble salts, an additional dosing system (acid + antiscalant) shall be installed.

2) Two R.O block utilizing the common brine and same membrane type.

The projected permeate throughput is expected to be around 60624 m^3/day (13.35 MGPD). i.e. an additional 1.53 MGPD over the 11.82 MGPD produced from 8 trains.

This option appear to be more feasible than the above, requiring less complex control system and possibly lesser costs of the membrane assembly infrastructure.

The requirement of the chemical dosing system is still valid.
An additional 418 permeators (209 x 2) will be required, almost similar to that in case 1 above.

In the options 1 & 2 above, the use of the modified twin membrane may be considered if considerable saving in capital investment can be achieved.

The projection calculation for option 2, utilizing different membrane type was examined and the initial results showed somewhat higher productivity 64800 m^3/d (14.27 MGPD), i.e. an additional 2.49 MGPD over the 11.78 MGPD produced from the 8 trains. The number of additional permeators were also calculated.

This may, depending on membrane cost, contribute to further reduction in capital investment. However, utilizing the same permeators as the existing will obviously ease operation & maintenance procedures and spare parts.

It is recommended that extensive pilot tests are carried at site out on different membrane types to assess the performance with the existing feed water chemistry.

PRODUCTIVITY INCREASE THROUGH INCREASE OF OPERATIONAL RECOVERY:

The feed water quality at the source " Um Er Radhuma" acquifer remained almost stable over 12 years, though expert opinion at the time of modeling this aquifer indicated that salinity will rise from 13000 ppm to 19000 ppm after about eight years from initial commissioning. A slight increase of H_2S level has been observed (6-7 ppm) in the current years, however, much below the plant design figure of 20 ppm. With this credit, the H_2S striping towers will still operate very efficiently necessitating no additional chlorine requirements other than residual amount required for potabilization.

In the beginning of year 1997, a decision was made to explore the increase of production via increase of operational recovery. Due to the fact that large quantity of permeators were relatively old and exceeded 7 years of continuous operation. It was recommended that the increase of operating recovery be carefully attempted.

The first step to achieve this target was to develop a computer program to study the maximum permissible recovery for the permeators under the prevailing feed water situations and to verify the flow adequacy and scaling potential in the existing brine staging configuration. (an example is shown in Appendix 2)

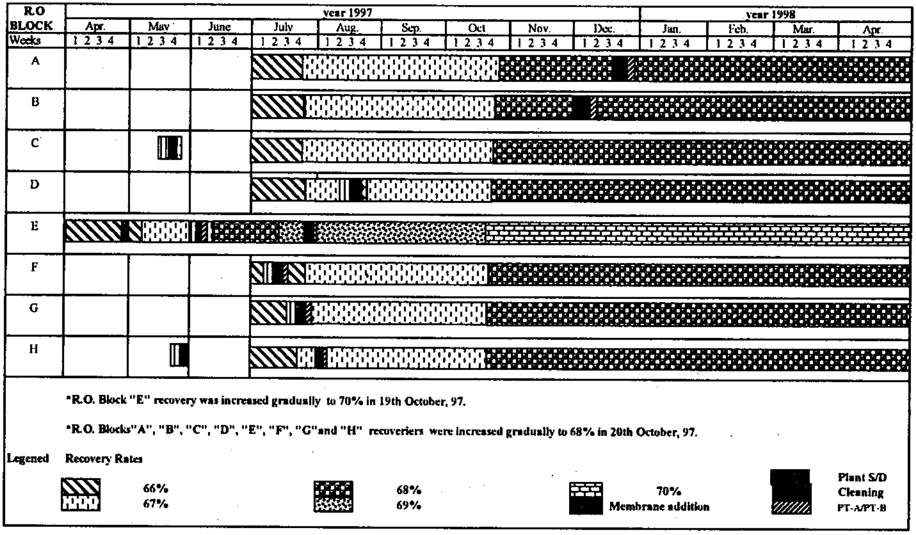
The computed values clearly indicted that recovery may be safely increased to 72% based on new permeators. However, with the presently aged permeators, it was decided to increase the production recovery gradually by 1% increment over 6 – 10 weeks to assess the overall performance of the membranes and collect maximum possible performance data under very close monitoring.

To minimize possible risks, it was decided to start the operational recovery increase on one train only "E". Until after a sufficient time, the remaining 6/7 trains shall undergo recovery increase from 65% to 66% by the time that train "E" reaches 68%.

During each phase of recovery increase, the program for permeators addition was continuously revised and modified (Fig 2) followed by a few days shut down activity to replace and/or add permeators. The aim was to keep the maximum permissible pressure of 60 bars at the permeators as a limiting factor.

In parallel, various other studies were also performed to confirm the adequacy of the chemical dosing system and expected efficiency of stripping towers to meet the new operational conditions.

The program for increased recovery commenced on 1st April 1997 on train "E" being the most stable train to ensure continuos reliable run. The operation of this R.O. train initially commenced with 66% recovery for a period of 4 weeks.



•R.O. Block "E" recovery was increased gradually to 70% in 19th October, 97.

•R.O. Blocks"A", "B", "C", "D", "E", "F", "G"and "H" recoveriers were increased gradually to 68% in 20th October, 97.

Figure 2: Program for recovery increase at Ras Abu Jarjur R.O plant

The average production increase was about 200 m^3/d for each recovery increment. The normalization results were extremely satisfactory.

By the second week of May 97, the recovery increased to 67% for three weeks. All performance data including salt passage were also satisfactory.

By mid June 97, the operational recovery of train E was further increased to 68%.

In view of encouraging results obtained at 68% recovery run of train E and before the end of 68% recovery phase, all other trains were set to commence 66% recovery by 1 st July 97 for a period of 3 weeks only. It was felt the shortest possible period to assess the overall performance.

The program for recovery increase (Fig. 2) was modified seven times to cater for membrane additions and/or cleaning. The presence of the standby train "H" has helped smooth execution of the recovery increase trial operation.

As some of the trains contain relatively old permeators with old pressure vessels, a decision was made to extend the 67% recovery period for at least two months to enable sufficient monitoring under higher recovery rates. The results of the membrane performance were found to be extremely successful. As the pressure at the permeators have reached almost 60 bars, the operating recovery was increased to 68% and continued at this rate to allow a longer monitoring of the performance and to ensure safe operation of the plant during summer. It was recommended that attempts to further increase the recovery is made after a long period of operation at 68% for about 6 months or until major membrane replacement is carried out by early 1999.

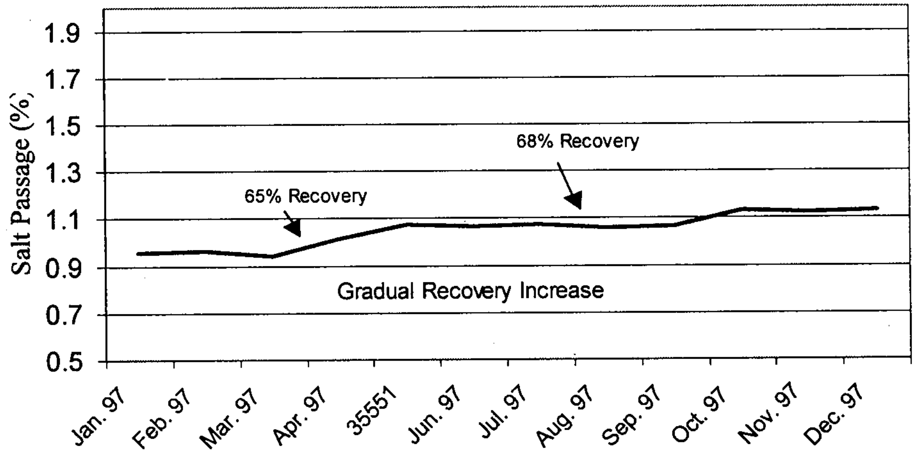


Figure 3: Variation of salt passage (overall) during 1997

TRAIN 8 PARALLEL RUN

This train was originally introduced as a standby to compensate the production loss due to the outage of any of the 7 trains, during maintenance and/or chemical treatment activities. [2]

The 8th train was commissioned in January 1994 and since then its performance has been very satisfactory, contributing to an increased availability of the plant. The process computer software was modified in such a way that any of the 7 trains can run simultaneously. Eight R.O. trains can not operate simultaneously since the interlock circuit limits the operation of total trains to 7 only which is also installed in the software.

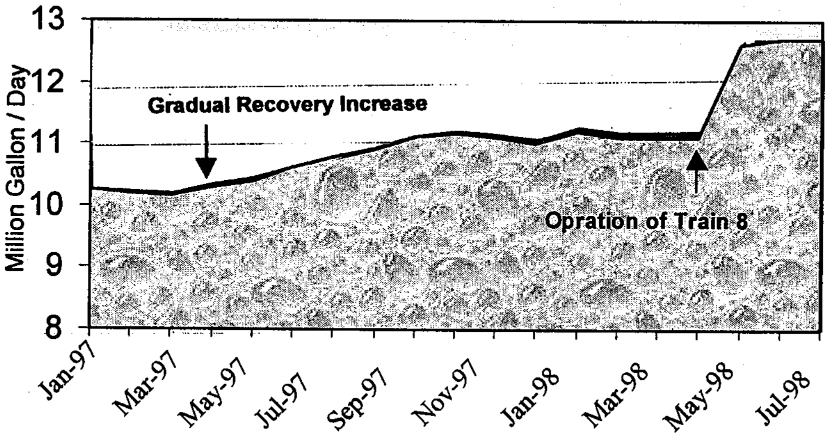


Figure 4: Permeate production from January 97 – August 97

The discussions of operating train No.8 in parallel with the other trains in parallel mode began in 1996 when the water demand in the Island was too high and the additional 1.5 MGPD would make a very useful contribution. However, it was realized that major modifications in the software would be required to place train 8 in full automatic operation mode. Further hardware expansion to the existing computer system was also required to cater for the amount of data need to be processed.

The cost for full development of the software and the additional hardware was very high and did not warrant the investment due to the fact that the main computer system (and related hardware) is relatively old.

In May 1997, the idea of running plant simulation came about, to confirm the possibility of manual control operation of train 8 when the interlock relay circuit of LCP 202 & 203 is removed. After various discussions with Sasakura Engineering Co., a request was made to Mitsubishi Electric Corporation through Sasakura Engineering Co. to carry out a simulation test. The aim was to check the possibility of operating the 8th train on Remote manual/ Local control mode if the interlock is removed via bypassing the interlock signal relay. However, maintaining the option for any selection of the 7 trains to run on the auto mode. The simulation test showed that additional software modification is required to permit the operation of train 8 in Remote Manual Mode.

Further studies on various plant items were also carried out (by Sasakura Engg. Co. and in-house) to confirm the viability of train 8 parallel run, these studies included the following:

- Well collector and pumps requirement
- Available electrical power
- Carbon filters
- Micron guard filters
- Stripping towers adequacy
- Chemical dosing system

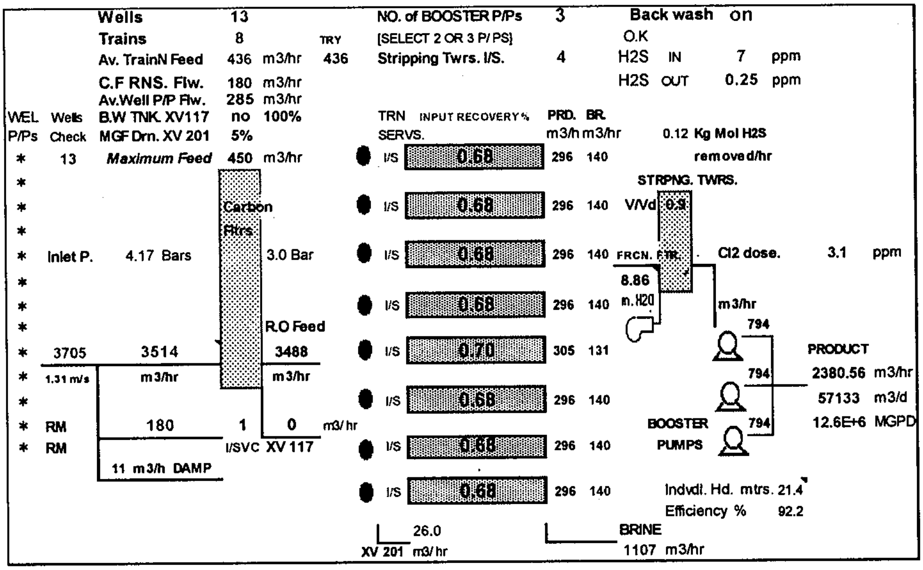


Figure 5: Modeling of Ras Abu Jarjur R.O

PROCESS MODELING :

A simple model developed by the authors of this paper was developed

(Figure 5), to study various plant parameters when operating at 7 or 8 trains.

The aim was to predict the overall plant throughput using the existing membranes at various operating recoveries and plant conditions, i.e. during carbon filters backwash and filling of rinse backwash tank, explore the optimum number of operating wells and feed flow rates.

The model also predicts the performance of the booster pumps at different production rates (whilst operating at 2 or 3 pumps), H_2S stripping towers performance and required chlorination levels.

During the plant outage in March 1998, the modification works pertaining to the removal of the interlock and amendments to the software were carried out by the plant's staff. The plant operation procedures were then modified after various consultation with Sasakura Engg. Co. Some modifications to the procedures were left until the actual trial run. The parallel operation of train 8 was successfully attempted on 14th June 1998, increasing the permeate production to 57,000– 57,700 m^3/d (12.5 – 12.7 MGPD).

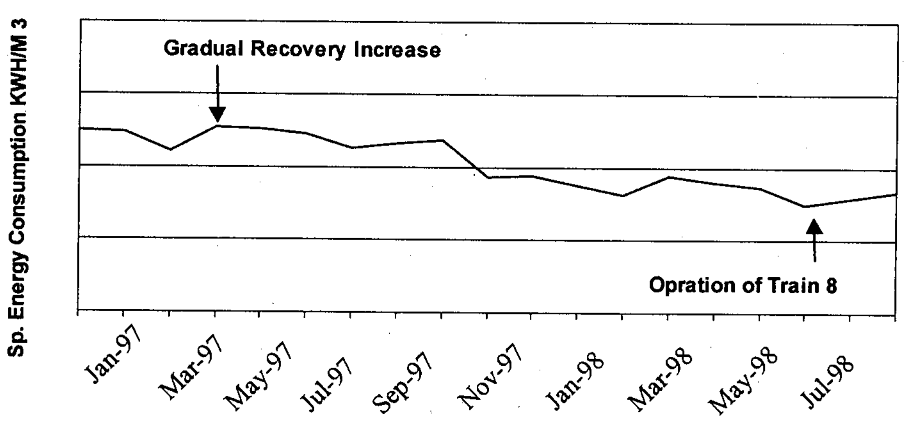


Figure 6: Specific Energy Consumption

ELECTRICITY CONSUMPTION

It was very difficult to observe the changes in the electricity consumption during the course of recovery increase due to non availability of measurement facility for individual trains. The changes, however, were very minimal due to slight recovery increase. The average total consumption was 7,281,500 Kwhr for 7 trains in operation and 8,180,400 Kwhr for 8 trains operation @ 65% - 68% recovery.

However, the specific energy consumption was reduced from 5.2 Kwhr/m^3 @ 65% to 4.9 Kwhr/m @ 68% recovery (Train E @ 70%) .

WATER COST

The unit water cost from Ras Abu Jarjur showed considerable decrease over the past few years as a result of high availability and various optimization programs for cost reduction [1,3]. Further considerable reduction in unit cost can be achieved with increased production of 11 & 12.5 MGPD (Fig 7.) The computed costs found to be BD 0.25/ m^3 and BD 0.23/ m^3 for 11 & 12.5 MGPD respectively.

BASIS OF COMPUTING WATER COST (8 TRAINS)

The following parameters were taken into consideration :

- Constant feed water salinity (13,000 ppm)
- Plant average availability with 8 trains operation at 93 %
- Number of well pumps in service = 13/14
- Constant H_2S level in the feed water at 7 ppm

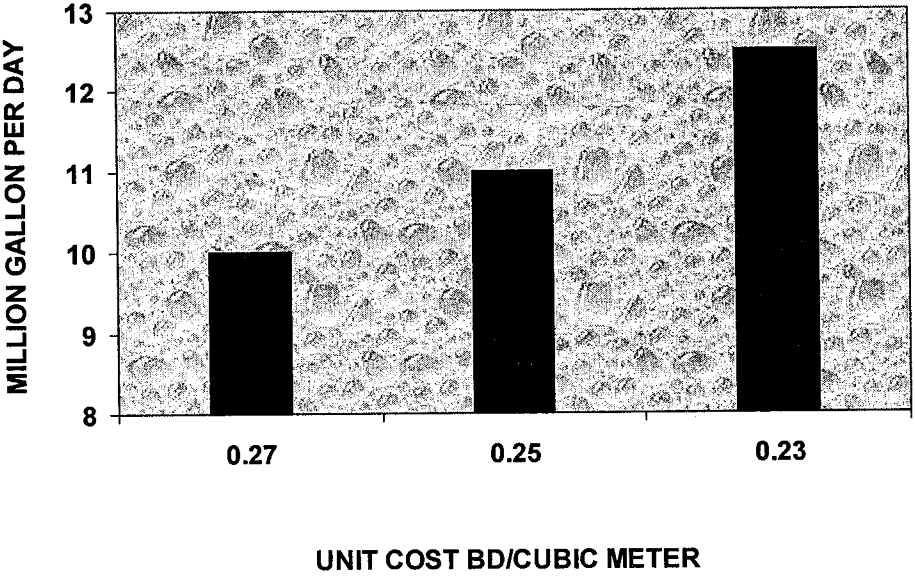


Figure 7: Unit water cost of permeate versus production

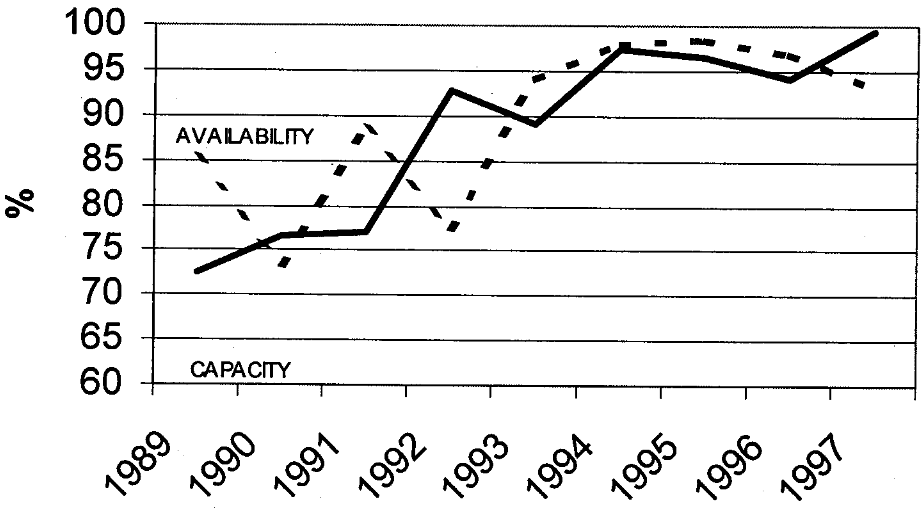


Figure 8: Plant availability and capacity factors

Depreciation has been calculated on the basis of 10 years and counts for one full replacement of the permeators in the 8th train with 4% replacement/ additional factor for the period 1997 - 2000 and 4.5% for the period 2002 - 2007.

Figure 8 shows that the capacity factor recorded in 1997 reached 97% based on 46000 m^3/ day (10 MGPD), due to the increase of permeate production. It is therefore necessary that the calculations of the capacity factors be based on 12.5 MGPD with 8 trains in simultaneous run at 68% - 70 % recovery. On the other hand, despite all efforts made to maintain the plant on base load conditions, the availability figures – based on 8 trains - showed slight decrease since 1995 due to problems associated with the aging of the well field collector piping and the control system.

FUTURE PROSPECTS FOR FURTHER PRODUCTION INCREASE AND UPGRADE STRATEGY

1. Placing of one train (No. 9)

Entails addition of two wells & well pumps, full infrastructure of R.O train (pump, motor, modules, etc.), control panels for well/R.O & pre treatment, one Micron Guard filter vessel, additional CPU, UPS panels and two operator stations. One additional stripping tower will be required to cater for the H_2S removal. The existing CO_2 generation is not sufficient to post treat the product water. The CO_2 plant shall either be replaced with a higher capacity plant or install a smaller unit similar to the existing. An additional 6 MVA transformer and related switch gear/MCC/ well pump panel shall also be added. The R.O building shall be extended along with HVAC system.

2. Placing of two additional trains (No.s 9 & 10)

The addition of two extra trains at the plant will boost the production by another 3-3.5 Million Gallon /day through a large scale modifications. Feasibility studies showed that this option is more cost effective as compared with train 9 only. All the above modifications are required. Two additional wells will be required (i.e. a total of 19 wells), control panel and all associated electromechanical equipment. A full train infrastructure, including a Micron Guard Filter vessel and one additional stripping tower.

No changes are foreseen for the computer system, CO_2 plant, main transformer and R.O building (as all shall be incorporated within train 9 modification).

Should 10 trains become operational, the velocity in the well field collector pipe will increase from 1.88 m/s to 2.05 m/s which requires further assessment for the entire well field collector piping.

It is foreseen that if plant extension is carried out in near future for 10 trains operation, the design shall be based on 19000 ppm feed water salinity. The following plant items need to be considered:

- Requirement of four additional wells
- A new over-ground well collector main with anti surge system, running parallel to the existing.
- Additional stripping tower to cater for any future increase in H_2S level.
- Additional booster pump.
- Additional CO2 plant utilizing existing LPG storage.
- Supply of one additional 6 MVA transformer with associated switch gear, bus-bars and breakers for powering the two additional high pressure pumps. No bus-tie panel connection is provided to these new pumps.
- One additional CPU and LCP panels (with CRT display in Main Control room)
- One additional LCP panel in the pretreatment and one additional LCP panel & one MCC panel in product control area.

No changes are anticipated in the plant piping, activated carbon filters & feed water pretreatment.

The ultimate desire for the upgrade strategy would be as follows:

- A new Central Processing Unit based on a 32-bit processor. (all existing I/O units would be used)
- UPS module would be incorporated into the CPU panels.
- Formats of Alarms screen display & Event Printout shall be improved.
- All sequencer units to be replaced with newer type.

Alternative scenario incorporating state of the art technology/latest developments shall be considered, taking into account maximum utilization of the exiting equipment and facilities. An example would be to utilize the reject brine as a feed for new train(s).

Availability of spares - especially for process Computer, sequencers, Multi-echo system shall be guaranteed for 10-15 years plant operation period.

RECOMMENDATION

1) Commence with increased recovery for all trains to 70% maximum, under close monitoring . The total through put is expected to reach 13 MGPD.

2) A more limiting parameter in expanding Ras Abu Jarjur beyond the operation of the 8 trains will be the water salinity table at the existing extraction aquifer "C" lens. Though two wells have recently exhibited fouling contaminants (Iron Sulphide possibly due to casing failure), the water quality from the 15 wells remained constant over the past 11 years. Testing on these wells has been recently conducted by Water Resources. The water quality condition from the existing system shall be further assessed before any conclusion is made to commission additional wells in the area. In this juncture, whether additional expansion will go ahead or not, it is highly recommended to update the previous zone 'C' model produced in 1983.

CONCLUSION

Due to the presence of train 8 as a standby at Ras Abu Jarjur, the availability record of the plant has risen to around 98% in 1995 (97% in 1997). Optimizing various operational philosophies at the plant has also been a factor for such high availability record.

The unit water cost produced by 8 trains - running simultaneously - has been calculated at BD 0.23 /m^3, even at reduced availability of 93% and increased energy and depreciation figures. Thus modification of the plant to allow for continuous operation of the 8th train appears to be the next promising opportunity for increasing the desalinated water from water production facilities.

The problems associated with the aging of the buried GRP pipework, the computer system and the associated fiber optic control cables linking the control of 15 well pumps to the main computer system remain the main problems affecting the high availability figures. Thus, the expected reliability whilst 8 trains running simultaneously is expected to fall below 85%. It is anticipated that the scope of work in the forthcoming rehabilitation of this plant will consider all these issues to guarantee a reliable operation and parts availability for the next 10 - 15 years.

The addition of trains 9 &10 comes timely with mid life assessment plan of this facility. Some areas mainly the computer system has already reached a stage where some parts are becoming obsolete and with view of extending the operational life till year 2010 (or beyond) such critical area of the plant needs to be fully assessed for possible replacement /upgrading.

The available information of different membrane types appear to give a slight increase in production, at lower recoveries for the second pass. It is also worth exploring.the economics of the use of twin Dupont membranes for plant's expansion beyond train 8.

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APPENDIX 1

Average Water Analysis for Feed, Product & Brine

			56% Recovery		70% Recovery	
		Plant Inlet	Product	Brine (Av. All	Product	Brine (Train E)
Temperature	C	30.6	30.6	32.5	32.0	33.6
PH	-	6.85	5.48	6.68	5.7	6.7
Turbidity	NTU	0.21	0.17	0.20	0.20	0.30
E. Conductivity	Us/cm	19600	185	50000	279	57400
Total Dissolved Solids	PPM	12240	87	35020	155	41060
Silt Density Index	-	1.94	-	-	-	-
Dissolved Oxygen	PPM	0.08	-	-	-	-
Total Bacteria Count	CFU/ml	240	1.0	600	1.0	1.5
T. Hardness CaCO3	PPM	2786	5	8000	17	9740
Calcium as Ca	PPM	637	1.00	1926	5	2252
Magnesium as Mg	PPM	290	0.6	774	1.3	999
Sodium as Na	PPM	3300	35	9270	45	10168
Potassium as K	PPM	138	1.2	440	1.5	464
Strontium as Sr	PPM	14.5	0.1	45	0.10	41
Total Iron as Fe	PPM	0.04	-	0.02		0.03
Manganese as Mn	PPM	< 0.03	<0.03	<0.03	<0.03	0.05
Copper as Cu	PPM	< 0.03	<0.03	<0.03	<0.03	0.05
Zinc as Zn	PPM	0.03	<0.03	<0.03	<0.03	0.12
T. Alkalinity as CaCO3	PPM	159	26	356	29	397
Bicarbonate as HCO3	PPM	194	31	434	35	484
Chloride as Cl	PPM	6670	42	18590	62	21777
Sulphates as SO4	PPM	553	1.00	1692	1.00	1927
Total Sulphide as S	PPM	5.5	4.5	7.40	4.8	7.25
Silica as SiO2	PPM	17.7	0.14	56	0.14	60
Floride as F	PPM	1.3	<0.1	1.35	<0.1	2.10

Appendix 2

Recovery Increase Calculations – Ras Abu Jarjur R.O Plant

1.) INPUT DATA	English units		Metric units	
Feed Flow (1st stg)	gpd	2,318,062	439	S
Overall Plant Capacity, Qpo	gpd	1,692,185	320	m^3/hr
Min.(Design) Temp.	°C	30		
Max. Temp.	°C	35		
Overall Conversion, Y0	---	0.73		
Feed TDS, as analyzed	mg/l	13,000		
Feed TDS, as NaCl, Cf_1 [CQ]	mg/l	10,839		
Osmotic Pressure, p [CQ]	psi	130	8.97	bar
Initial Perm. Capacity, Q_i	gpd	7,000	1.32	m^3/hr
Feed Pressure, Pf_1	psi	930	64.14	bar
Product Pressure, Pp_1	psi	20	1.38	bar
Flow Balancing dP, 1st stg	psi	20	1.38	bar
Flow Balancing dP, 2nd stg	psi	20	1.38	bar
Flow Balancing dP, 3rd stg	psi	35	2.41	bar
Stage Component 1	---	5		
Stage Component 2	---	3		
Stage Component 3	---	2		

2.) STAGE RECOVERY "Y" ESTIMATION

Estimated	---	0.39
Estimated	---	0.33
Estd/Calc	[CQ]	0.34
Target Y_0	---	0.73
Qf_1	gpd [CQ]	2,318,062
Qp_1	gpd [CQ]	904,044
Qb_1	gpd [CQ]	1,414,018
$Qf_2 = Qb_1$	gpd [CQ]	1,414,018
Qp_2	gpd [CQ]	466,626
Qb_2	gpd [CQ]	947,392
$Qf_3 = Qb_2$	gpd [CQ]	947,392
Qp_3	gpd [CQ]	321,515
Qb_3	gpd [CQ]	625,877

3.) Stage No. 1 [calculated quantities]

Estimated Y_1	---	0.39
Cb_1	mg/l	17,769
Ave Cf_1Cb_1	mg/l	14,304
πf_1b_1	psi	172
1/2dPf_1b_1 (estimated) INPUT	psi	3.0
PCF_1		1.59

P_{1-1}	psi	1,000
P_{1-n}	psi	930
P_{1-2}	psi	900
$MFRC_{1-1}$		0.590
$MFRC_{1-n}$	[CQ]	0.627
$MFRC_{1-2}$		0.643
$MFRC_1$	[CQ]	0.627

Design T	°C	30
TCF_1	[CQ]	1.15

MFRC Chart
PEM '92/2020
Table No.1

Qp_1	gpd	8,013
Qb_1	gpm	8.70
Qf_1	gpm	14.27
Ave Qf_1b_1	gpm	11.49
dPf_1b_1	psi	6.0

Qb limits for B-10 6845T		
Max	28	gpm
Min SW	4.2	gpm
Min HBr	7.0	gpm
Note:	Brine flow is ok	

4.) Stage No. 2

Estimated Y_2	---	0.33
$Cf_2 = Cb_1$	mg/l	17,769
Cb_2	mg/l	26,521
Ave Cf_2Cb_2	mg/l	22,145
πf_2b_2	psi	269
1/2dPf_2b_2 (estimated) INPUT	psi	3.1
Pf_2	psi	904
PCF_2		1.32

P_{2-1}	psi	1,000
P_{2-n}	psi	904
P_{2-2}	psi	900
$MFRC_{2-1}$		0.590
$MFRC_{2-n}$	[CQ]	0.641
$MFRC_{2-2}$		0.643
$MFRC_2$	[CQ]	0.641

Design T	°C	30
TCF_2	[CQ]	1.15

MFRC Chart
PEM '92/2020
Table No.1

Qp_2	gpd	6,823
Qb_2	gpm	9.62
Qf_2	gpm	14.36
Ave Qf_2b_2	gpm	11.99
dPf_2b_2	psi	6.2

Qb limits for B-10 6845T		
Max	28	gpm
Min SW	4.2	gpm
Min HBr	7.0	gpm
Note:	Brine flow is ok	

5.) Stage No. 3 [caculated quantities]

Estimated Y_3	—	0.34
$Cf_3 = Cb_2$	mg/l	26,521
Cb_3	mg/l	40,144
Ave Cf_3Cb_3	mg/l	33,332
πf_3b_3	psi	409
1/2dPf_3b_3 (estimated) INPUT	psi	3.1
Pf_3	psi	863
PCF_3		0.93

P_{3-1}	psi	900
P_{3-n}	psi	863
P_{3-2}	psi	800
$MFRC_{3-1}$		0.643
$MFRC_{3-n}$	[CQ]	0.663
$MFRC_{3-2}$		0.697
$MFRC_3$	[CQ]	0.663

Design T	°C	30
TCF_3	[CQ]	1.15

MFRC Chart
PEM '92/2020
Table No.1

Qp_3	gpd	4,965
Qb_3	gpm	6.71
Qf_3	gpm	10.16
Ave Qf_3b_3	gpm	8.44
dPf_3b_3	psi	4.4

Qb limits for B-10 6845T		
Max	28	gpm
Min SW	4.2	gpm
Min HBr	7.0	gpm
Note:	Adjust conversion rate	

6.) Verification of Brine Staging [calculated quantities]

Prod. Flow, T. Staging, Qp_s	gpd	70,462
Feed Flow, T. Staging, Qf_s	gpd	96,523

7.) Verification of Recovery

Y_1	[CQ]	0.42
Y_2	[CQ]	0.36
Y_3	[CQ]	0.28

8.) Number of Permeators Required to Produce Qp_o

Nos. of Permeators	---	240	[calculated quantity]

9.) Number of Permeators & Est. Prodn., Stagewise [calculated quantities]

Stage 1	---	120	962,154	gpd	182	m^3/hr
Stage 2	---	72	491,554	gpd	93	m^3/hr
Stage 3	---	48	238,478	gpd	45	m^3/hr
	Total Production		1,692,185	gpd	320	m^3/hr

CQ : Caculated Quantities

Effect of Various Forms of Iron in Recycle Brine on Performance on Scale Control Additives in MSF Desalination Plants

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Saad Al-Sulami, K. Sahul and Radwan Al-Rasheed

EFFECT OF VARIOUS FORMS OF IRON IN RECYCLE BRINE ON PERFORMANCE OF SCALE CONTROL ADDITIVES IN MSF DESALINATION PLANTS

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ABSTRACT

Scale control additives (SCA) or antiscalants are used in multistage flash desalination plant to prevent fouling problems. Effective dose rates of some of these SCA are found to vary from plant to plant. Efficiency of antiscalants could be influenced by its chemical nature, plant configuration and design, heat transfer rate and its surface and antiscalant reaction with brine constituents. To understand the variation in concentration of SCA and effect of brine constituents on efficiency of SCA, some bench top experiments at 95 °C with brine were conducted using the threshold effect technique. Two SCA, a polyphosphonate based (PPN) and polymaleic acid based (PMA) were used for these evaluations. The effect of Fe^{3+}, Fe^{2+}, Fe_2O_3, $Fe(OH)_3$, and rust on efficiency of SCA was determined.

The threshold effect method which measures the total alkalinity and calcium content of supernant to evaluate efficiency of SCA was found to be an effective technique. Recycled brine constituents and corrosion products, especially Fe and its various chemical forms affect the efficiency of SCA at 2 ppm concentration and at 95 °C in 1.4 times concentrated seawater brine. Various chemical hydrated and hydroxylated forms of Fe present in brine, viz., $Fe^{3+}(H_2O)_6$, $Fe(OH)_3$, Fe^{2+}, Fe_2O_3, rust, etc., all showed detrimental effect on efficiency of SCA. Among all the species of iron, $Fe(OH)_3$, reduced the efficiency of SCA by maximum of 40%.

Among the two antiscalants, PPN showed 10-12% better efficiency in the presence of various iron species compared to PMA under similar bench top experimental conditions. The effect of brine constituents on PMA was more pronounced compared to PPN perhaps due to better complexing ability of the former.

Key words: Antiscalant, Polyphosphonate, Polymaleic acid, Scales, MSF, Iron, Brine

INTRODUCTION

Saline Water Conversion Corporation (SWCC) being the largest desalted water producer in the world, produces about 20% of world production of desalted water and thus is also the biggest consumer of scale control additives (SCA) [1-2]. Scale control additives or antiscalants which are used in multistage flash (MSF) desalination plants to minimize scaling problems, act nonstiochiometery but still cost several million dollars every year. Thus to make SCA consumption more economical and at an optimum level, it would be of great significance to evaluate various parameters that affect the efficiency of scale control additives. It is known that optimum dosages of SCA often differ from plant to plant and location to location [2]. The root causes for such variation has not been understood. However, it appears that the variation in concentrations and/or efficiency of antiscalants could be influenced by its chemical nature, plant configuration and design, heat transfer rate and its surface and antiscalant reaction with brine constituents.

The exact chemical composition of scale control additives is a closely guarded secret of the manufacturer. In this respect, Sham-El-Din and Mohammad carried out physico chemical studies of antiscalant additives in relation to their functional groups [3, 4]. Further, the role or mechanism of antiscalants reactivity is still not clear in spite of their popular use. It is believed that either it alters the crystal morphology preventing scale growth or it may be acting by what is known as the threshold effect or both [5]. Interaction of brine constituents with SCA may be either advantageous to prevent scale formation or may be detrimental to its performance. In seawater some transition metal ions are present in trace concentrations which can preferentially coordinate with these agents thus reducing their efficiencies. However, the conditions inside distillers and condensers vary due to the construction materials used. The generation of oxides of iron, copper and nickel due to corrosion is unavoidable. A variety of species may circulate through the evaporators as corrosion products such as Fe^{2+}, Fe^{3+}, $Fe(OH)_2$, $Fe(OH)_3$, Fe_2O_3 and rust. These ferrous and ferric ions may readily complex with scale control additive agents and thus in turn will reduce the availability of antiscalant for scale control reaction thereby resulting in an increase in SCA consumption. On the other hand $Fe(OH)_2$ and $Fe(OH)_3$ may form good a suspension in saline water, providing nucleation sites for alkaline scale formation thereby accelerating the rate of scaling. It is this property of iron species along with its electrical property in aqueous solution which make Fe^{3+} an effective coagulant in water treatment processes [6]. In short, iron species may affect the efficiency of an antiscalant agent due to their characteristic complexing and coagulant abilities. This paper describes the results of an investigation on possible influences of iron species that may be present in recycle brine on the performance of antiscalants. The major objective of these investigations is to study the effect of main corrosion related species namely Fe^{2+}, Fe^{3+}, $Fe(OH)_2$, $Fe(OH)_3$, Fe_2O_3 and rust on antiscalant efficiency. The study indicates that there is a specific range of iron species that appreciably affect the threshold effectiveness of SCAs. Moreover, comparison of efficiency of various SCA in the presence of different species of iron shows the necessity of choosing the right type of SCA in a highly corroded environment.

EXPERIMENTAL

Reagent

The specification of the reagents used is as follows :

(a) Sodium carbonate (1N) :- 53 gm Na_2CO_3 dissolved in a liter of water

(b) Antiscalant solution (1000 ppm) :- 0.5 gram of each antiscalant was dissolved in 500 ml of deionized water to give 1000 ppm.

(c) Brine

  (i) Chemical used for preparation of brine
  Chemical A - NaCl
  Chemical B - instant ocean salt (commercially available mixture salt with almost all seawater components)

  (ii) In order to prepare brine with concentration of 1.4 time that of seawater, 8.8 gm NaCl and 6.4 gm instant ocean were dissolved in one liter of seawater, M.alkalinity of final solution was 180 ppm.

(d) Brine Constituents

  (i) Fe^{3+} was taken in the form of $FeCl_3$ solution.

  (ii) $Fe(OH)_3$ and $Fe(OH)_2$ were freshly precipitated from solution containing known concentrations of Fe^{3+} and Fe^{2+} ions. Precipitate was then centrifuged and transferred to reaction vessel as a slurry in seawater.

  (iii) Fe_2O_3 used was powdered Analar grade.

  (iv) Rust was collected from some rusted parts of the plant which was further grounded before use.

Apparatus

Figure (1) shows the schematic diagram of set up used for evaluation of performance of the SCA. It consists of three necked round bottom flask of 1 liter capacity fitted with water condenser, thermometer, sampling port and chemical injection port. Heating mantle equipped with stirrer was used to heat up brine the solution. The condenser is connected to a water bath where the temperature of condensing water is maintained at 10 to 15 °C. To create dynamic flashing in the brine at temperature at 95 °C, vacuum is applied using a pump.

Test Conditions

(a)	Test temperature	:	95 °C
(b)	Brine M-Alkalinity	:	180 ppm
(c)	Concentration of antiscalant	:	2 ppm
(d)	Retention time	:	40 minutes maximum

Test Procedure

Brine 500 ml was measured accurately and spiked with 1 ml of antiscalant stock solution in order to make 2 ppm of antiscalant in the final brine solution. The brine was charged to the reaction vessel and was heated to the desired temperature under refluxing. When the solution reached the required temperature, it was refluxed for 15 minutes to reach equilibrium and to observe any possibility of scale formation. Because there was no precipitation in the presence of antiscalant, 7.5 ml of 1 N Na_2CO_3 was added to break the supersaturation point and to form the scale. The moment at which Na_2CO_3 was added, was considered to be time T = 0. Successive samples (30 ml each) were withdrawn from solution at time intervals of 5, 10, 15, 20, 30 and 40 minutes. The samples were filtered through 0.45 micron filter paper and filtrate was analyzed for Ca, Mg and M-Alkalinity. Total alkalinity (M-Alkalinity) as function of time intervals was plotted to get the efficiency of antiscalant (a scaling threshold diagram).

RESULTS AND DISCUSSION

The efficiency of the two antiscalant additives used viz., polyphosphonate based (PPN) and polymaleic acid based (PMA) was determined in the presence and absence of various iron species such as Fe^{3+}, $Fe(OH)_3$, $Fe(OH)_2$ and rust by what is known as the threshold effect. The scaling threshold is defined as the "maximum permissible concentration factor of water samples at fixed temperature before nucleation begins under equilibrium conditions" [5]. Thus scaling threshold diagrams are drawn between concentration factor and temperature. In MSF plants it is the practice to fix both temperature and concentration factors at optimum values to avoid or minimize scaling problems. In the present study, the net total alkalinity (TALK) after super saturation due to the addition of sodium carbonate as function of time (threshold diagram) was taken as a measure of the efficiency of antiscalant. The comparison of the scaling threshold diagram of antiscalant in the presence and absence of chemical constituent of brine leads to the understanding of the behavior of these antiscalants. Figure (2) shows the loss of TALK with time for distilled water (DW), Gulf seawater and brine, with and without antiscalant. One can see that the saturation of Ca^{2+} and CO_3^- in case of DW is lower than seawater and brine indicating that the high salt content in brine significantly increases the saturation of Ca^{2+} and CO_3^- above their solubility product value $Ksp = 4.8 \times 10^{-9}$. The addition of antiscalants PPN and PMA at 2 ppm concentration raised the solubility of Ca^{2+} and CO_3^- by manifold compared to undosed. This property of certain chemical agents in raising the supersaturation determines the scale control performance of SCA. It is commonly referred to as threshold effect or scaling threshold of that agent. The scale control performance of antiscalant agent is affected by the presence of foreign constituents in the brine solution especially various forms of iron. Figure (3) shows the effect of Fe^{3+} on the performance of PPN and PMA. It is clear from the plot of TALK vs time (Fig. 2) that PPN has got a higher scale control property than PMA. Secondly, up to 0.3 ppm of Fe^{3+} concentration, the scale control performance is unaltered under the conditions of a flashing temperature of 95±0.5 °C, a brine concentration factor of 1.4, an antiscalant concentration of 2 ppm and pH = 8.2. However, as the concentration of Fe^{3+} was varied from 0.3 to 5 ppm, a pronounced effect on the performance of antiscalant was seen. Figures (4) and (5) show the effect of varying the concentration of Fe^{3+} on PPN and PMA respectively. It can be seen that the scaling threshold alkalinity gradually dropped as the Fe^{3+} concentration increased. The decrease in alkalinity is more in the case of PMA than PPN. These experimental results could be explained on the basis of the chelation of active center of SCA with Fe^{3+} which is stronger in PMA than in PPN [7].

When Fe^{3+} is added to an aqueous system several hydrated and hydroxylated species coexist in equilibrium with each other. In acidic solution, hydrated species dominate while in alkaline solution, like seawater pH 8.2, the hydroxylated species are predominate. The equilibrium could be written as follows;

$$Fe^{3+}(H_2O)_6 \rightleftharpoons Fe(OH)^{2+} \rightleftharpoons Fe(OH)_2^{+} \rightleftharpoons Fe(OH)_3$$

These hydrolyzed species may have a significant influence on the scale control performance of SCA. The PMA which contains $COOH^-$ group, may chelate with Fe^{3+}. The chelation of Fe^{3+} with COO^- form a strong bond in complexes such as Ferric ethylene-diaminetetacetate and Ferric nitriloacetate [7, 8]. With the high possibility of strong bond formation with PMA, the active centers concentration will deplete with net result of reduced scale control performance. However, from Figure (4), it can be seen that even at a concentration of 5 ppm of Fe^{3+}, the scale control performance is still more than 20% which indicates that chelation has a minor effect on SCA performance. Hence other effects such as nucleation site offered by $Fe(OH)_3$ and hydrolyzed chain of ferric ions may be playing a role in retarding scale control performance. In fact as the Fe^{3+} concentration is increased from 1 ppm to 5 ppm the ferric species is extensively hydrolyzed into $Fe(OH)_3$ and polymeric hydrolytic chains [8]. Retardation of scaling threshold could therefore be related to the combined effect of the presence of Fe^{3+}, $Fe(OH)_3$, $Fe(OH)_2^+$, $Fe(OH)^{2+}$, $Fe(OH)_n^{m+}$ species. It is evident from the Figures (4) and (5) that alkalinity is steady up to 0.5 ppm Fe^{3+} after which loss of TALK is very sharp as the Fe^{3+} concentration increases. It appears that the drop in alkalinity is due to the formation of $Fe(OH)_n^{m+}$ and $Fe(OH)_3$ species at higher Fe^{3+} concentration. $Fe(OH)_3$ may disperse in the brine and offer enormous nucleation site for $CaCO_3$ precipitation. This kind of coagulating tendency is more pronounced in ferric species and thus ferric is extensively used as coagulant in water treatment [6, 9]. It can also be seen from the Figures (4) and (5) that TALK beyond 2 ppm remains almost constant which means the scale control performance is unaltered and unaffected by the species. Furthermore, from the figures it can be seen that the effect of Fe^{3+} is very similar in both PMA and PPN except that PPN is slightly less affected than PMA. Table (1), which shows the efficiencies of antiscalant in the presence of 2 ppm of iron species at treatment time of 15 minutes, also leads to the same conclusion. This may be because the -P=0 group in the PPN may not chelate with Fe^{3+} as effectively as COOH in PMA.

The effect of $Fe(OH)_3$ on the efficiency of SCA was studied by adding freshly prepared $Fe(OH)_3$ to brine solution containing SCA. Figures (6) and (7) show the effect of $Fe(OH)_3$ in the presence of PPN and PMA. The two figures also include the performance of the antiscalants in the presence of 1 ppm of Fe^{3+}. It can be seen that 1 ppm Fe^{3+} has almost the same effect as that of 0.6 ppm $Fe(OH)_3$. Further retardation of scale control performance was observed in both PMA and PPN almost to the same extent. The results thus suggest that the nucleation effect rather than chelation plays a significant role in the scale control process.

Highly porous $Fe(OH)_3$ may loose water molecules to form denser Fe_2O_3

and may also offer nucleation sites. Figure (8) shows the loss of TALK with time in the presence of Fe_2O_3. Its effect on retardation was much less in comparison to $Fe(OH)_3$ and Fe^{3+}. The effect of 2 ppm Fe_2O_3 was almost negligible, but 10 ppm it was significant and was almost equal to that of 2 ppm $Fe(OH)_3$. This may be due to the fact that Fe_2O_3 is more compact and may offer fewer nucleation sites with lesser electrical property and surface area than $Fe(OH)_3$.

In actual plant conditions, corroded surfaces offer more nucleation sites as compared to smooth metal surfaces. The outer corroded surface may have metal oxides (rust). Figure (9) shows the effect of rust on SCA performance. The retardation effect by rust is very low at 2 ppm and only at 10 ppm, loss of TALK is appreciably high. However, the extent of retardation is less compared to $Fe(OH)_3$. Once again this is due to the high density of rust, (i.e., higher compactness in comparison to $Fe(OH)_3$) and thus offers less surface area for nucleation.

Corrosion is a process whereby the Fe is oxidized first into Fe^{2+} at seawater pH = 8.2, then Fe^{2+} is further oxidized to Fe^{3+} by O_2. However, in desalination units, O_2 is removed in the deaerator and the O_2 level is brought down to less than 0.02 ppm. Thus, the probability of oxidation of Fe^{2+} to Fe^{3+} is reduced under these conditions and Fe^{2+} may be hydrolyzed to $Fe(OH)_2$ before being oxidized to $Fe(OH)_3$. Thus to study the effect of $Fe(OH)_2$, freshly prepared $Fe(OH)_2$ was used. Figure (10) shows the results of these studies. It can be seen that the loss of TALK in these experiments was less significant, yet similar to that observed in the case of $Fe(OH)_3$.

Summaries of the results obtained on the efficiencies of antiscalants are shown in Table (1). The table gives efficiencies of PMA and PPN in the presence of 2 ppm of various species of iron in brine at 95 °C with a treatment time of 15 minutes. These results indicate that PPN efficiency was less affected by iron species in the brine compared to PMA.

CONCLUSIONS

From the foregoing discussion it is evident that various forms of iron in the recycle brine tend to retard the efficiency of Scale Control Agent (SCA). When 2 ppm of iron species (as ferric) is present in the brine, the performance of PMA was reduced to 37% by Fe^{3+}, 32.3% by $Fe(OH)_3$, 84.1% by Fe_2O_3, 85% by rust and 73.1% by $Fe(OH)_2$, while that of PPN was reduced to 46.4% by Fe^{3+}, 40.5% by $Fe(OH)_3$, 90.6% by Fe_2O_3, 88.1 by rust, and 84.5% by $Fe(OH)_2$.

These conclusions can be elaborated on as follows:

1. The method of adding 1000 ppm Na_2CO_3 to 1.4 times concentrated seawater brine and determining the scaling threshold by measuring total alkalinity in the brine as a test for scale control performance of SCA, was found to yield satisfactory results. It is thus recommended that this method can be adopted for preliminary screening of antiscalants for alkaline scale control.

2. Using the above test under laboratory conditions PPN efficiency was found to be better than PMA antiscalant. PMA efficiency is 88% of PPN (in the absence of iron).

3. In the presence of 0.5 ppm Fe^{3+}, SCA efficiency is reduced to 93% while at 5 ppm Fe^{3+}, it is reduced down to 30%.

4. Ferric hydroxide reduced the efficiency of SCA as compared to Fe^{3+}. At 2 ppm of $Fe(OH)_3$, the efficiency of PPN is 40% and the efficiency of PMA is 32%.

5. Fe_2O_3 and rust showed significant reduction of SCA performance only at 10 ppm concentration.

6. The efficiency of antiscalant was reduced by the presence of brine constituents in the following order :
$Fe(OH)_3 > Fe^{3+} >> Fe(OH)_2 > Fe_2O_3 >$ Rust.

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Table (1) Percent Efficiency of Antiscalant in Presence of 2 ppm Iron Species

Iron Species / Antiscalant	Fe^{3+}	$Fe(OH)_3$	Fe_2O_3	Rust	$Fe(OH)_2$
PMA	37%	32.3%	94.1%	85%	73%
PPN	46.4	40.5	90.6%	88.1%	84.5%

(Treatment Time 15 Minutes)

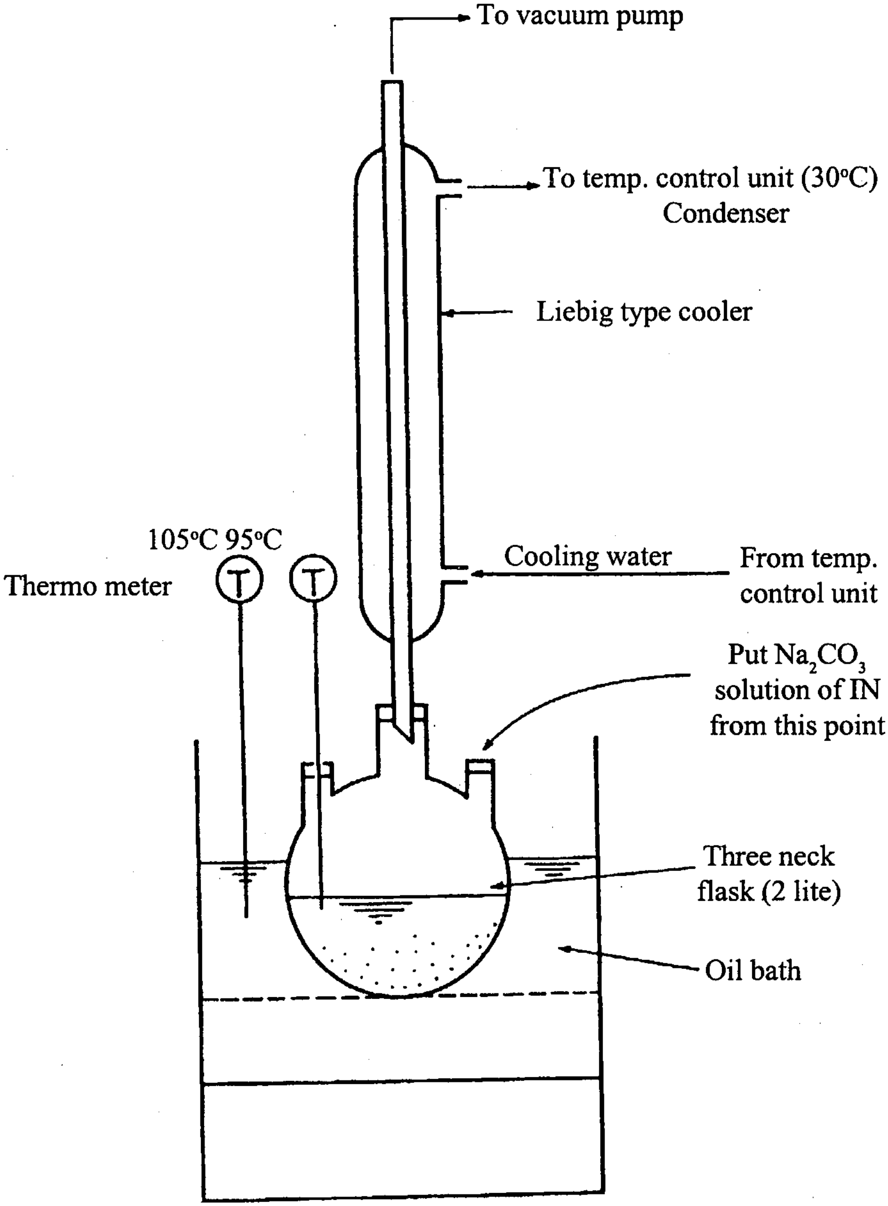

Figure 1: Schematic diagram of experimental set up for determination of scaling threshold of antiscalants

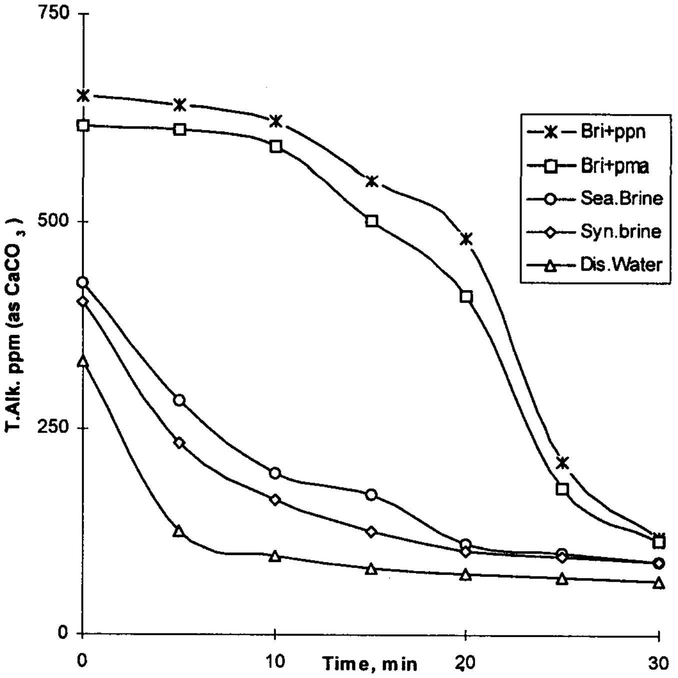

Figure 2: The scaling threshold in Brine with and without scale control additives

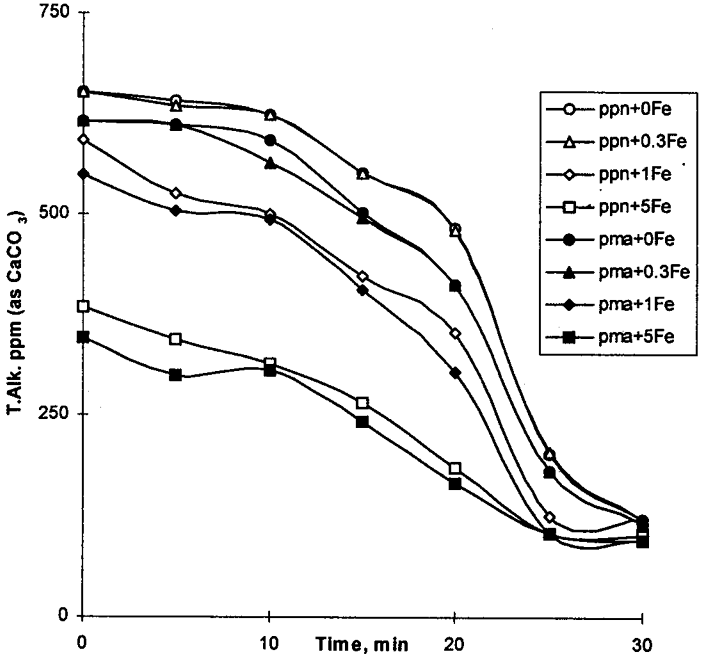

Figure 3: The effect of Fe^{3+} on the performance of PPN and PMA

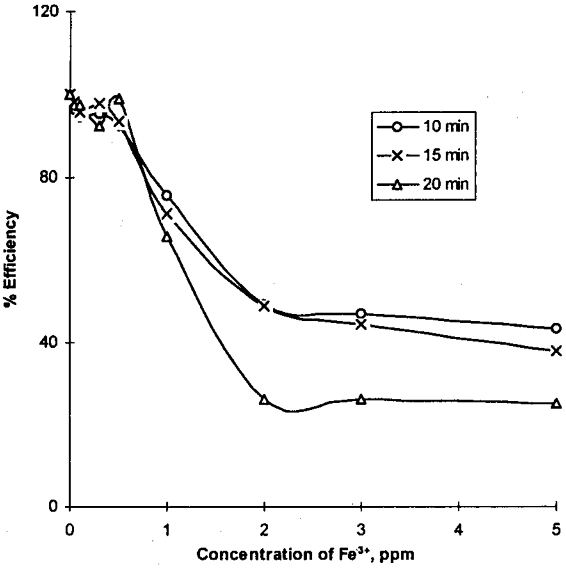

Figure 4: Percent efficiency of PPN in the presence of Fe^{3+} at different time intervals

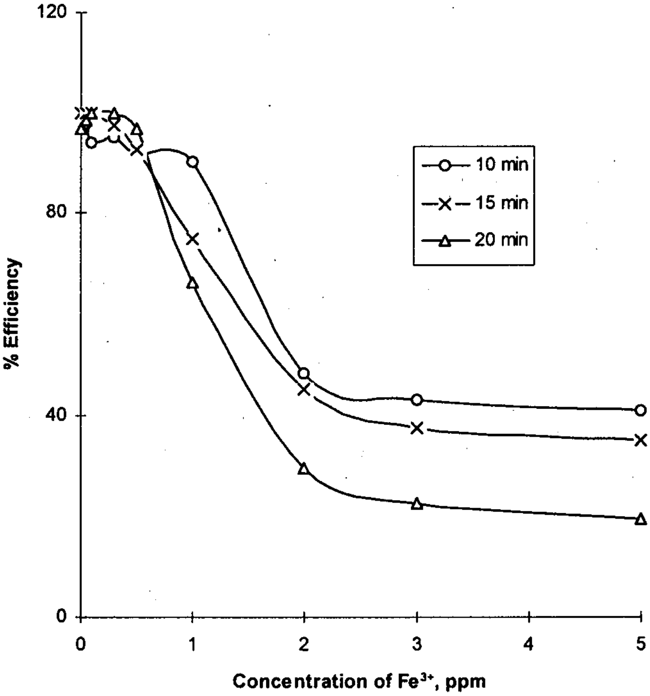

Figure 5: Percent efficiency of PMA in the presence of Fe^{3+} at different time intervals

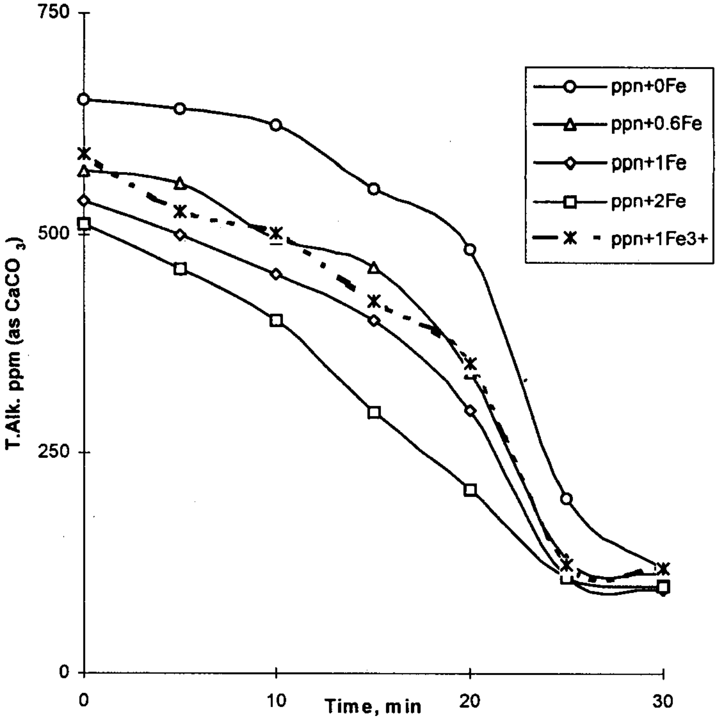

Figure 6: The effect of $Fe(OH)_3$ on the performance of PPN

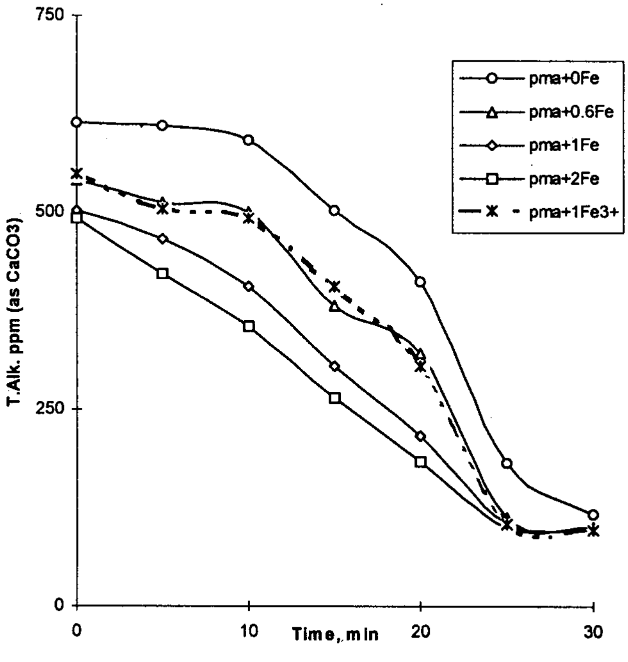

Figure 7: The effect of $Fe(OH)_3$ on the performance of PMA

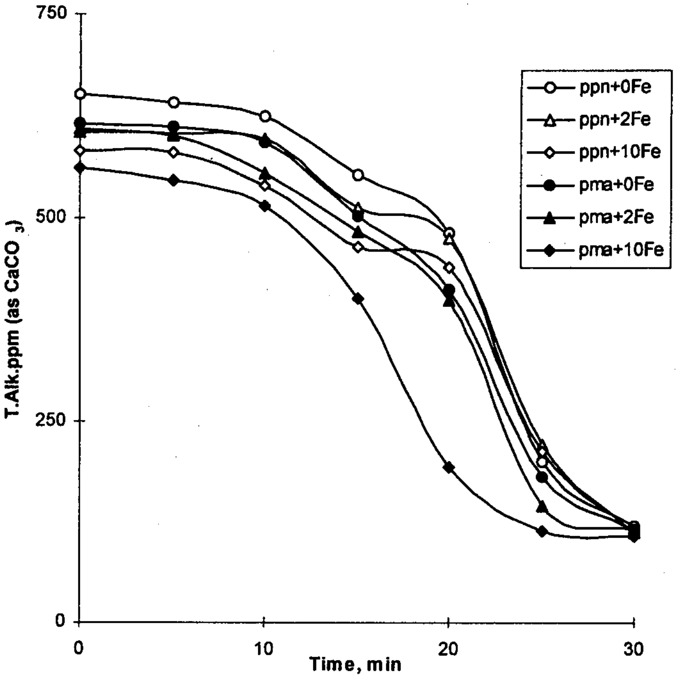

Figure 8: The effect of Fe_2O_3 on the performance of PPN and PMA

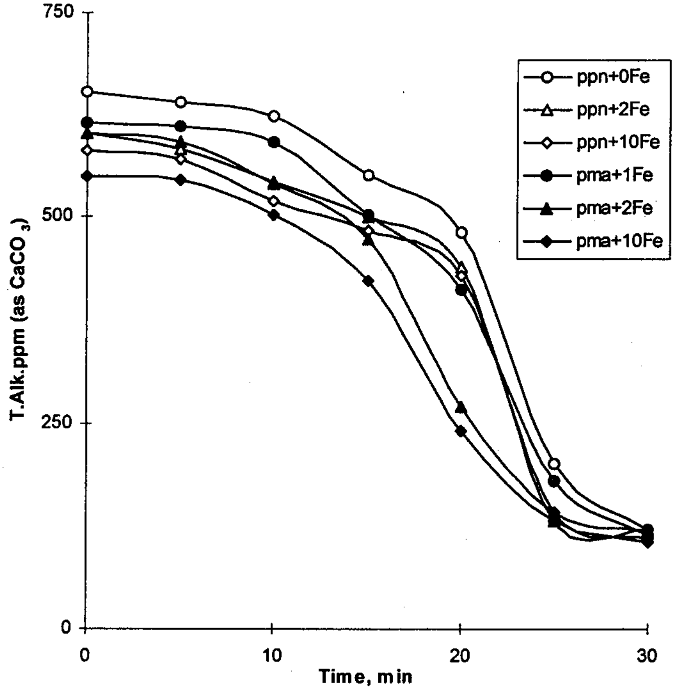

Figure 9: The effect of Rust on the performance of PPN and PMA

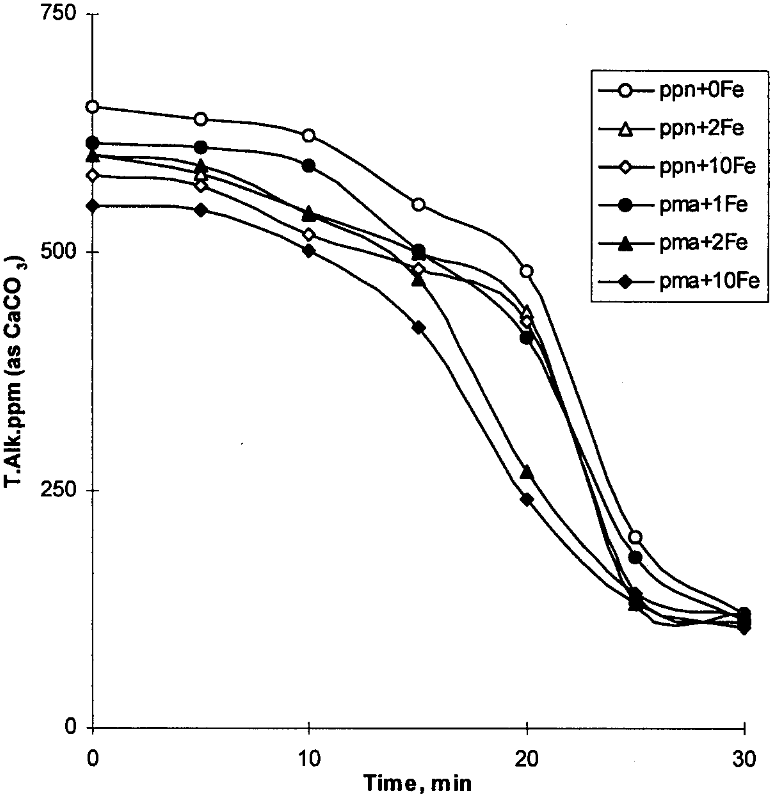

Figure 10: The effect of $Fe(OH)_2$ on the performance of PPN and PMA

The Performance of Different Antiscalants in Multi-Stage Flash Distillers

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THE PERFORMANCE OF DIFFERENT ANTISCALANTS IN MULTI-STAGE FLASH DISTILLERS

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ABSTRACT

The Saline Water Conversion Corporation (SWCC) is an internationally recognized authority on scale control in multi-stage flash (MSF) distillers. This international recognition came about as a result of many years of work on this topic. There are publications of SWCC on MSF scaling phenomena over the past couple of decades. During the last decade, there was a monumental growth in SWCC's installed MSF plant capacities. The majority of SWCC MSF distillers were designed to operate with additive scale control method rather than the previously predominant method of carbonate depletion through chemical reaction with acid, e.g., sulfuric or hydrochloric.

The Research and Development Center (RDC) in Al-Jubail has been involved during the last three years in the evaluation of available commercial scale inhibitors. A total of eight tests have been carried out at RDC MSF pilot plant, Jeddah Phases III & IV and Al-Jubail Phase II to evaluate the effectiveness of three antiscalants. These tests by antiscalant types and places are (i) polycarboxylates in Al-Jubail II, RDC pilot plant and Jeddah IV, (ii) polymaleic acid in Jeddah IV and RDC pilot plant and (iii) polyphosphonate in Jeddah III and RDC pilot plant. Moreover, a run of acid operation carried out in the pilot plant is included for comparison.

This paper presents qualitative and quantitative performance results of three different antiscalants as well as acid dosing tests for scale control. The threshold effect of these antiscalants was tested on laboratory scale. On pilot and commercial plants the variation of the fouling factor with time is examined and quantified. The evaluation tests conducted on the commercial plants revealed that all the examined antiscalants were successful in inhibiting alkaline scale formation and improving plant performance at top brine temperature ranging between 105 and 110°C and with respective dosing rates of 1.5 and 3.0 ppm.

Keywords: Desalination, Multistage flash, Alkaline scale, Antiscalants and Fouling Factor.

INTRODUCTION

Control of scale formation on heat transfer surfaces is one of the basic problems in the distillation processes. Formation of scale on heat transfer surfaces impedes the rate of heat transfer. The main scale forming constituents of sea water are calcium, magnesium, carbonate, bicarbonate and sulfate. On heating, bicarbonate will yield carbonate which can precipitate with calcium if the saturation limit is exceeded. As a result magnesium hydroxide will also form. In practice, the formation of alkaline scales are controlled by either lowering the pH through bicarbonate depletion or by threshold additive chelation through special chemicals called threshold agents[1]. Conversely, $CaSO_4$ formation is primarily controlled, so far in commercial plants, by operating at top brine temperature (TBT) that are lower than its appreciable precipitation limits, i.e., TBT < 120 °C in view of the raised salt concentration in MSF brine streams.

Mixtures of sodium tripolyphosphate dispersing agent have been used to inhibit alkaline scale deposition in seawater evaporators since the 1950s [2]. The major problem with polyphosphate based inhibitors was found to be its thermal degradation at temperatures above 90°C and the subsequent loss of its threshold effect. This restriction in TBT to 90°C limited the thermal efficiency of evaporator designed for threshold treatment from mid 1950's through 1960's into a good part of 1970's.

Acid dosing was introduced in the 1960s as a means of overcoming the temperature limitations and the poor performance of polyphosphate. As problems associated with acid operation became dominant, the need for high temperature scale control additives to replace acid while giving commensurable performance was felt towards the end of the seventies into the eighties (1970 & 1980s) [2]. Low molecular weight polymeric/ carboxylic acid and phosphorous base alkaline were developed as high temperature additives. It has become a common belief that these chemicals distort the lattice structure and minimize the rate of growth of the alkaline scales. Polyphosphonate "phosphorous base alkaline group" does not hydrolyze as easily as the polyphosphate group due to the greater stability of the c-p bond in phosphonates as compared with the p-o bond in phosphates.

The dosing rate of antiscalant is one of the most important operating parameters. Under dosing leads to scale formation while overdosing is believed to enhance sludge formation [3]. It is thus essential to establish an optimum dose rate. Although a number of chemical reagents have been used to minimize or prevent the formation of alkaline scales [4], no commonly accepted method for avoiding the formation of scales due to calcium sulfate salts at high temperatures and brine concentrations in large commercial MSF plants is available. The only commercialized approach so far, i.e., currently used to prevent the formation of calcium sulfate; is to operate the plant below the solubility limits of calcium sulfate. Yet the known ones are by sulfate depletion or reaction removal of calcium and/or sulfate content of the seawater make-up. Most recent promising approach was addressed last couple of years in international and regional conferences and is expected to be addressed in this conference. In the recent work, membrane nanofiltration has been successfully used in almost total sulfate depletion [5-7].

Saline Water Conversion Corporation (SWCC) has been actively involved in the inhibition of scale formation in order to improve the performance of its distillers. A direct comparison on the performance characteristics and economics of a number of additives tested on the MSF distillers of Al-Jubail phase I has been reported [8,9]. SWCC also conducted optimization of scale control additive dose levels at different top brine temperatures in Al-Khobar II and Al-Jubail II desalination plants[10]. Moreover, the performances of polymaleic acid (PMA) and polyphosphonate (PPN) antiscalants at Al-Jubail phase II desalination plant were verified[11]. Gained operational experiences on Al Khobar Phase II plant with only one acid cleaning during the period 1982-1992 was presented in detail by Al-Sofi, et al.,[12].

BASIC APPROACH

This paper reports the results of the evaluation tests of scale control additives performed by Al-Jubail Research and Development Center during the last three years. Testing of scale inhibitors characteristics and effectiveness has been carried out at three levels: laboratory, pilot plant and finally testing on commercial plants. At the laboratory level the threshold effect of different antiscalants was first tested on a bench scale equipment. At the pilot plant level the effectiveness of polymaleic and polycarboxylic acids was tested [13]. The MSF pilot plant is of 20 ton per day capacity and its schematic diagram is shown in Figure 1. It consists of a brine heater, four heat recovery and two heat rejection stages. The plant is also equipped with external deaerator as well as decarbonator (for TBT of 120 °C acid operation) and on-line ball cleaning facilities. At the commercial levels, tests were conducted in Jeddah III and IV also Al-Jubail II to evaluate the performance of polycarboxylic acid, polymaleic acid and polyphosphonate as scale control additives. Design characteristics and parameters of tests carried out in these commercial plants are summarized in Table 1.

Table 1: Design Characteristics and Test Parameters of the Commercial Plants

S. No.	Design & Test Parameters	Jeddah IV Unit 19	Jeddah IV Unit 15	Jeddah III Unit 10	Jubail II Unit 8
1	Top brine temperature	110	110	108	105
2	Concentration ratio (RB)	1.4	1.3 - 1.32	1.4	1.4
3	Antiscalant dose rate, ppm	1.8	2	3	1.5
4	Ball cleaning , times/week	2	2	14	21
5	No. of ball cycles/ per shift	9	9	3	10
6	Ball to tube ratio	0.4	0.4	0.75	0.3
7	Production, MIGD	5	5	5	5.75
8	Number of stages	24	24	16	22
9	Gain output ratio (GOR)	7	7	7	9.0
10	Flash range, °C	72	72	70	69
11	Design fouling factor BH (m^2K/kW)	0.395	0.395	0.325	0.2
	HRS (m^2K/kW)	0.176	0.176	0.178	0.15

RB = Recycle Brine, BH = Brine Heater, HRS = Heat Rejection Section

RESULTS AND DISCUSSION

LABORATORY TESTS

Antiscalants have two principal effects: threshold and crystal distortion. Threshold effect is responsible for preventing scale initiation and inhibiting or slowing down the deposition of soft scale consisting mainly of calcium carbonate and magnesium hydroxide. Crystal distortion effect is responsible for suppressing the adhering ability of scale. Antiscalant addition results in the distortion of scale crystals into spheroid structures which have less contact surface than a flat shaped untreated scale crystal hence less adherence ability to metals. Spheroid shaped particles tend to rotate around their own axis while moving with the liquid stream along pipe axis. This combined, i.e., revolution as well as traversial motion results in a radial force that causes the particle to migrate towards the pipe center leaving particles free pipe wall.

The threshold characteristics of three different antiscalants were examined. These were a modified blend of polyphosphonates (PPN-M), polymeric maleic acid (PMA) and polymeric carboxylic acid (PCA). The method of confirming the threshold effect was performed through the addition of a certain quantity of inhibitor in a supersaturated brine solution. The ability of the scale inhibitor to prevent the deposition of scale was checked by observing the change of M-alkalinity with time. The threshold effect of the three examined antiscalants was carried out in laboratory at a constant 2 ppm dosing rate and at two different temperatures of 95 and 110°C. The change of M-alkalinity as $CaCO_3$ (ppm) with time in minutes is shown in Figures 2 and 3. Both figures show that the alkalinity of the supersaturated solution with no scale inhibitor (blank), which can be considered as a reference state, decreased rapidly with time. At 10 minutes time the alkalinity dropped to 140 ppm at 95 °C and 80 ppm at 110 °C.

Figures 2 and 3 also reveal that all scale inhibitors have the ability to prevent the deposition of scale when they contact supersaturated brine solution but their effects decrease with the increase in retention time. Modified polyphosphonate and polymaleic acid were found to exhibit the highest threshold effect compared to polycarboxylic acid especially at conditions of 95 °C with a retention time boundary of 20 minutes and 110°C with retention time boundary of 10 minutes. All scale inhibitors lost their threshold effect within 30 minutes at 90 °C and 15 minutes at 110 °C under the above laboratory conditions.

PILOT PLANT TESTS

Two different antiscalants PCA and PMA were tested on the RDC MSF pilot plant. Tests were conducted with TBT range of 110-112 °C, concentration ratio of 1.38 -1.4, antiscalant dose rate of 2 ppm and cyclic ball cleaning. The increase of fouling factor in the brine heater with time was calculated to demonstrate the thermal effectiveness of antiscalants. Figure 4 shows the time dependence of the fouling factors. It shows that both antiscalants are quite effective in their performance and maintaining fouling factors well below the design values. Small variance is seen where the slope of PCA line is slightly less than that of PMA and where PMA starts at a lower value than PCA, yet after 700 hours they are almost the same. Results of visual inspection conducted before and after these tests revealed that there was a very thin layer of soft sludge on the tubes. The increase of the fouling factor in the brine heater (R_D, m^2 K/kW) with time (t, hours) is quantified. Using a regression analysis the following linear relationships are obtained assuming that the rate of fouling remains constant.

For Polycarboxylic acid, $R_D = 0.00002\,t + 0.0349$ .... (1)
For Polymaleic acid $R_D = 0.00004\,t + 0.0217$ .... (2)

The linearity of fouling factor with time indicates that the rate of increase in fouling factor is very small which is reflected in the low values of the time coefficients. Assuming the linear behavior of fouling factor, the allowable periods required for the fouling factor to reach design value and before acid cleaning becomes necessary, are 375 and 483 days for polycarboxylic and polymaleic acids, respectively. Based on the experience especially those reported in reference [12] such periods are found to be very short. This could be attributed to the fact that the rate of change in fouling was only linear during those limited duration tests. However, for longer operating periods the asymptotic fouling model could be better representative to the behavior of fouling curve [14]. Moreover, in view of cyclic ball cleaning this asymptotic behavior is expected to follow the saw-tooth trend.

Two additional runs were conducted in the MSF pilot plant distiller operating with a once-through mode rather than the conventional recirculation mode to briefly show the effect of total concentration and alkaline scale forming ions content on heat transfer performance. Figure 5 shows how far polyphosphonate dose rate can be reduced when only seawater of 128 ppm alkalinity is used to flow through the brine heater i.e. once through operation. In this test antiscalant dose rate was reduced down to as low as 0.5 ppm at an elevated TBT of 115°C. Although the quantity of seawater which has to be pretreated in the once-through mode of operation is 2.6 times that of the recycle mode, the required quantity of antiscalant has been reduced down by a factor of two. In essence, the pretreatment cost is halved.

COMMERCIAL PLANT TESTS

Tests were conducted in Jeddah III and IV and Al-Jubail II to evaluate the performance of polycarboxylic and polymaleic acids and polyphosphonate as antiscalant agents. In these tests pre, interim and post trial visual inspections were conducted as frequent as was deemed essential and as production schedules could have allowed. During each test relevant flow rates, temperatures and pressures were monitored. Calculations were then made to determine the heat transfer characteristics induced by each examined antiscalant. Design characteristics and test parameters of the four tested commercial units are summarized in Table 1.

Results of the performance evaluation of polycarboxylic acid at Jeddah IV unit 15 commercial plant are shown in Figure 6. The fouling factors of both the brine heater and heat recovery section are well below the design values which indicates that the distiller was working satisfactorily. Analytical results of pH values, M-alkalinity, turbidity, chlorine and total dissolved solid indicate that the unit was working smoothly with no abnormal behavior. Analysis for residual antiscalant in the samples taken before flashing shows that the antiscalant was active throughout the brine recycle passage. The assumed linear regression analysis reveals that the allowable operating periods for the brine heater and heat recovery sections, before acid cleaning becomes essential are 403 (compared to 375 days in the pilot test) and 867 days, respectively, while dosing 2.0 ppm of PCA at a TBT of 110°C.

Results of the performance of polymaleic acid at Jeddah IV unit 19 commercial plant are shown in Figure 7. Trends of fouling factors in heat recovery and heat input (i.e., the brine heater) sections show an increase with time but remain below their respective design fouling factors. The rates of increase of fouling factor for this unit (as well as unit 15 discussed in the above paragraph) are very low indicating minimum depositions inside the brine heater and heat recovery tubes. These findings are supported by the interim and post trial inspections of brine heater and the first three stages of the recovery section which show that tubes were in good clean condition. These results are found to support the validity of the asymptotic behavior of scale buildup. Post inspection at the end of this trial run showed that demister fouling in the first stage had blocked all the pads. In the next two stages, demisters were less fouled compared to the first stage and the scale was also soft. These results are found to support the proposed asymptotic growth behavior in Jeddah tests.

Results of performance of polyphosphonate at Jeddah III unit 10 commercial plant are shown in Figure 8. The fouling factor of heat recovery section remained at almost a constant value during the whole period of the test which indicates negligible deterioration in its thermal performance. This was confirmed by the visual inspection which revealed that only a very thin layer of scale had been deposited in the first stage and a dusting of scale inside the tubes of the second stage was apparent. The GOR remained virtually unchanged throughout the trial period. The fouling factor of the brine heater was increased from 0.05 to 0.1 m^2K/kW during 2100 hours which was well below the design value of 0.325 m^2K/kW. Post trial visual inspection of the brine heater revealed that there was no evidence of excessive sludge and only a thin layer of scale was observed on the tubes inner surfaces. The demister pads in the first three stages showed some depositions, similar in degree to that experienced under normal operation. The assumed linear regression analysis predicts that it will take 4.37 and 3.2 years for the fouling factors of the brine heater and heat recovery sections to reach the design fouling factors at a dose rate of 3 ppm and a TBT of 108°C.

Evaluation tests were performed in Al-Jubail II to demonstrate the effectiveness of polycarboxylic acid at a TBT of 105 °C. The test results are depicted in Figure 9 which shows slight increase of fouling factor with time. The distiller was also inspected before and after test trial. The post trial inspection revealed that no sludge was found in tubes of brine heater and heat recovery section. The demister pads were found in good and clean condition. Thus both the calculated test results and visual inspection confirmed that the tested antiscalant is highly effective in controlling scale formation and its deposition thereafter. The linear regression analysis revealed that it will take 415 days for the fouling factor of the brine heater to reach the design value.

CONCLUSIONS

1. Literature search revealed that one of SWCC's main achievements in controlling alkaline scale formation in commercial MSF distillers was successful operation at low antiscalant dose rates.

2. The threshold laboratory tests revealed that modified polyphosphonate and polymaleic acid inhibitors are having higher threshold effects compared to polycarboxylic acid. All scale inhibitors lost their threshold effectiveness within 30 minutes at 90°C and 15 minutes at 110°C during laboratory testing.

3. MSF pilot plant distiller tests revealed that the rate of rise in fouling factors for PCA & PMA are quite comparable. Another test in this pilot plant indicated the effect of concentration of total and alkaline scale forming ions, in allowing operation at drastically reduced antiscalant dose rate and increased TBT of 115°C. Last test conducted in the pilot plant revealed the effect of alkaline scale forming ions

of bicarbonate origin through reaction with acid, in allowing operation at a reduced brine recycle flow rate hence lowered tube velocity through the brine heater from 2.34 meters per second down to 1.8 meters per second at TBT of 110°C (as shown in Figure 5).

4. Evaluation tests on polyphosphonate, also polymaleic and polycarboxylate polymers performed in Jeddah and polycarboxylate polymer in Al-Jubail desalination plant showed that these antiscalants are very effective and successful in controlling scale formation at a TBT range of 105 and 110°C and with dose rate in the range of 1.5 and 3.0 ppm. The fouling factors of brine heater and heat recovery section suppressed by the three antiscalants were always well below the design values. The linear regression analyses of the tests results reveal that the allowable operating periods for the brine heater and heat recovery sections before acid cleaning is required, are generally more than a year. These findings were confirmed by post trial visual inspection which revealed in all cases that brine heater and heat recovery tubes were normally maintained in clean conditions.

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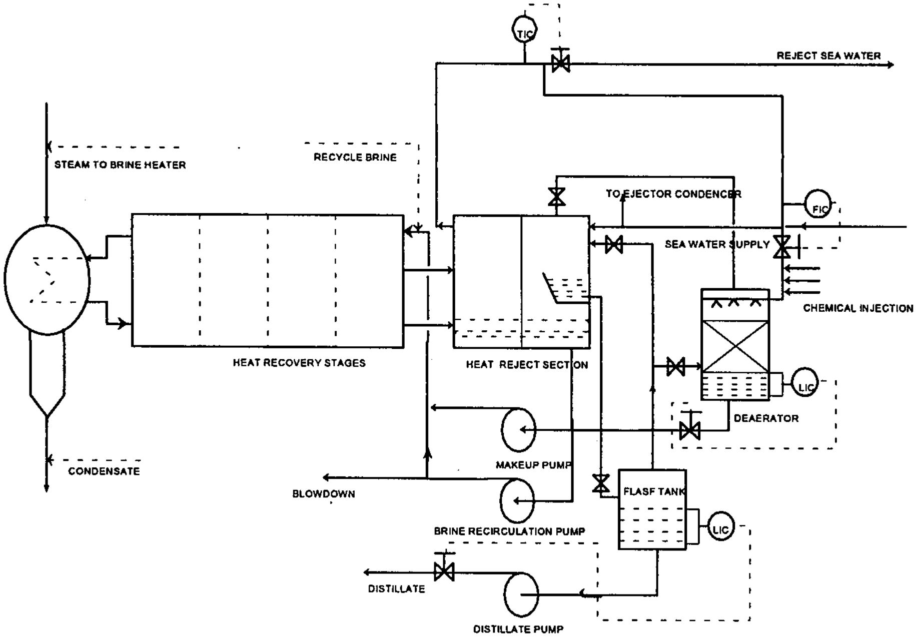

Figure 1. Schematic Diagram of the Pilot MSF Plant

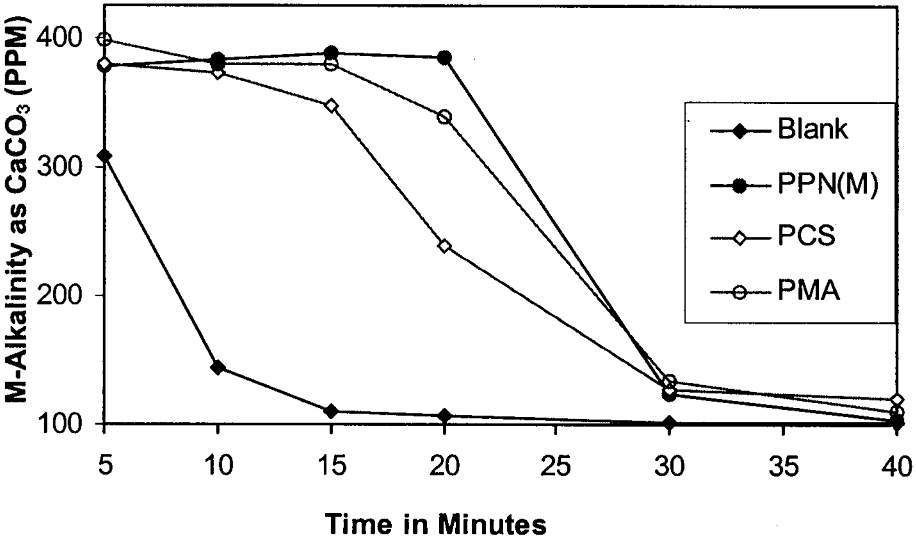

Figure 2. Laboratory Test of Various Antiscalants for their Threshold Effect at 95°C & 2 ppm Concentration

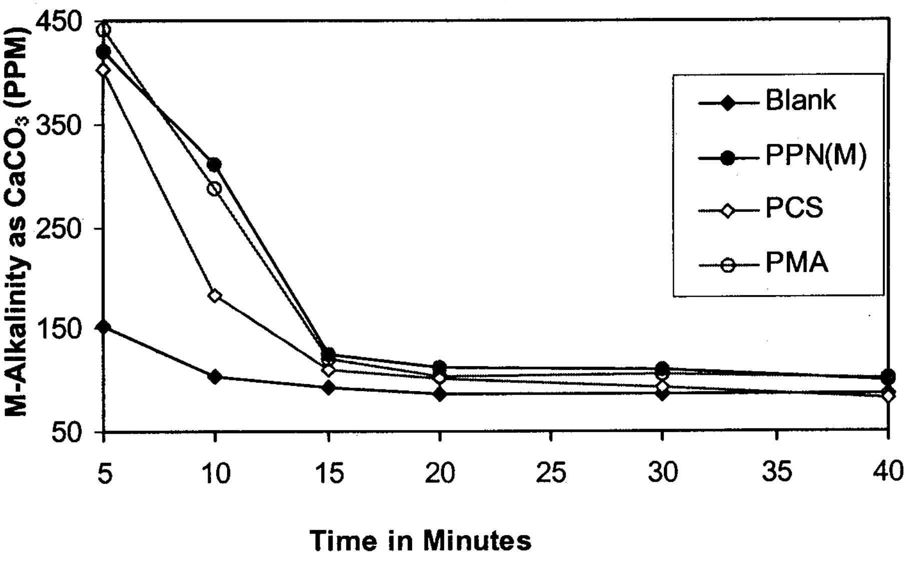

Figure 3. Laboratory Test of Various Antiscalants for their Threshold Effect at 110°C & 2 ppm Concentration

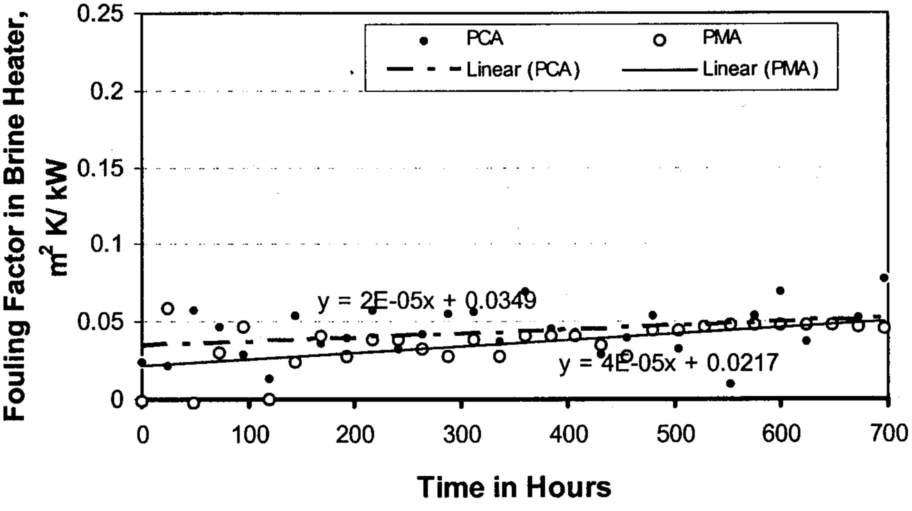

Figure 4. Pilot Plant Tests of Polycarboxylic & Polymaleic Acids at 112 °C and 2 ppm

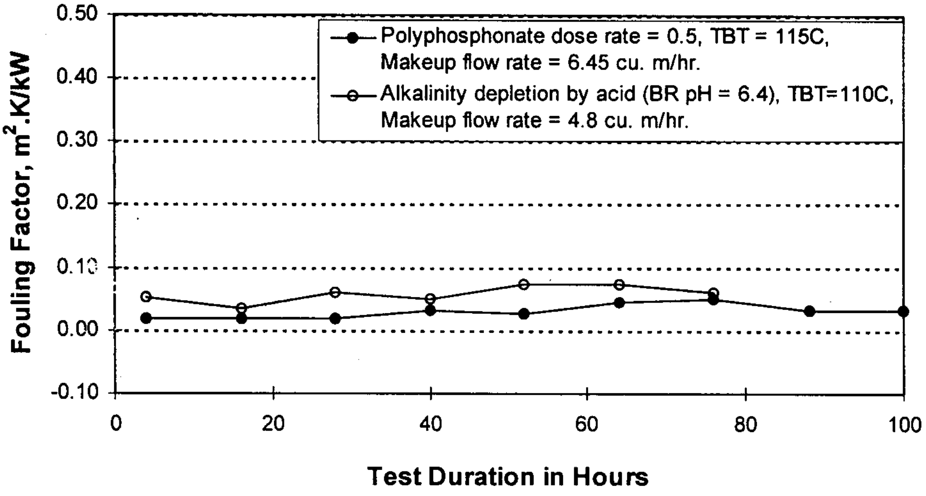

Figure 5. Once-Through Operation in Pilot Plant

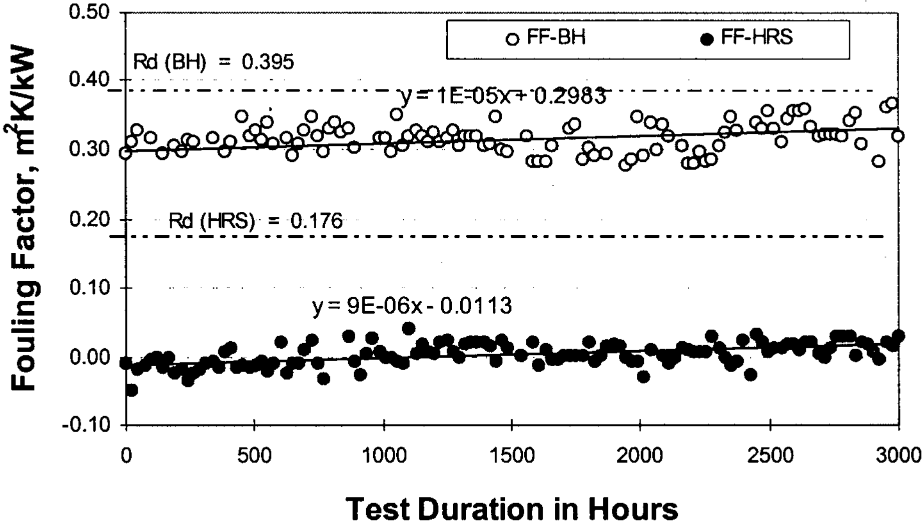

Figure 6. Test of Polycarboxylic Acid in Jeddah IV Unit # 15 at 110 °C and 2.0 ppm

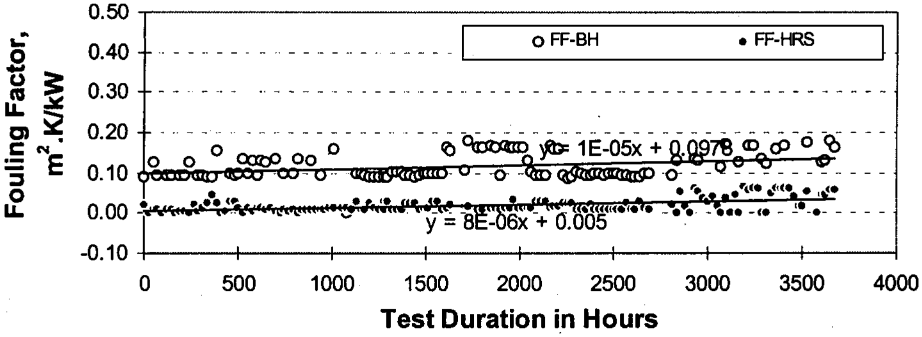

Figure 7.Test of Polymaleic Acid in Jeddah IV Unit # 19 at 110 oC and 1.8 ppm

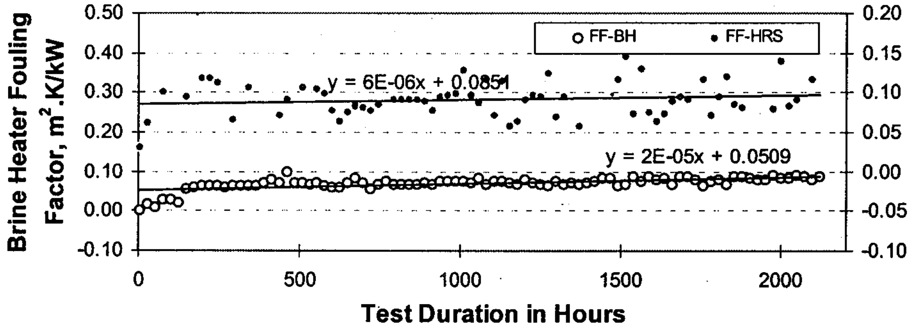

Figure 8. Test of Polyphosphonate in Jeddah III Unit # 10 at 108 °C and 3 ppm

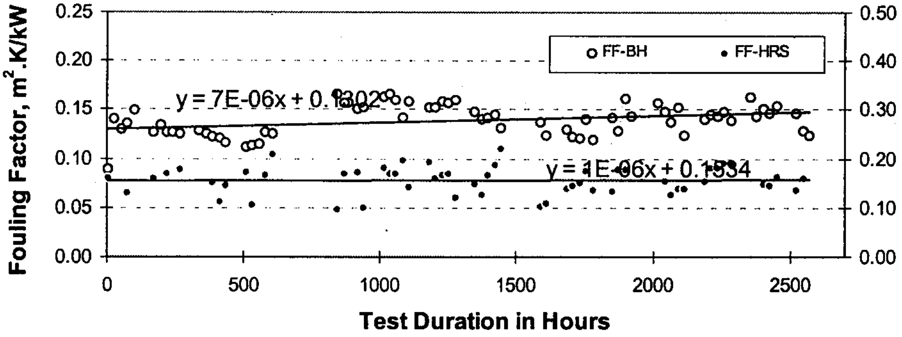

Figure 9. Test of Polycarboxylic Acid in Al-Jubail II Unit # 8 at 105 °C and 1.5 ppm

A Small Size MSF Model for Field Experimentation with Ratio of Dimensions Based on Number of Tubes

Salah A. Abunayib

A SMALL SIZE MSF MODEL FOR FIELD EXPERIMENTATION WITH RATIO OF DIMENSIONS BASED ON NUMBER OF TUBES

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ABSTRACT

This paper presents an idea of designing a small size MSF model. The ratio of dimensions is based on the ratio of number of tubes of the model to that of the main plant. The tubes of the model are of the same diameter, thickness and length as the main plant. For better aspect, each stage tube is bent or cut into a reasonable number of pieces, which are connected in series. The pressure & temperature differences across any one tube are the same like the main plant. Hence the model tubes have the same flow pattern and heat transfer rate as the main plant. The dimensions of the flash chamber, distillate tray and distillate trough are based on the same ratio of number of tubes. By connecting this model to main plant, many variables can be studied without interrupting the production or quality of production of the main plant.

INTRODUCTION

During the second and third Gulf Water Conferences and the IDA World Congress held in Abu Dhabi, one has seen a considerable number of studies based on practical experience with MSF desalination technology. Unfortunately most of these studies were only observations which were not necessarily concluding with solutions to the problems faced in field of practice. This, of course, is mainly because in practice the plants are for commercial production of water and not for study. Even in cases when courageous modification decisions were taken to improve performance, the observations were taking long time. Also the results in most cases can not be generalized to suit other plants. In practice a plant can have its unique problems which can be inherent or due to particular local operating conditions which may not be same for some other plant.

The following is an idea of designing a small size model with relative dimensions based on the ratio of number of tubes in the model to that in the full size (or main) plant. The model is of one or a small number of tubes/stage with full tube length cut (or bent) into a reasonable number of pieces connected in series and with full tube diameter and full thickness. This model can be connected to the main plant to help study different parameters without interrupting the production or the quality of production.

THE RATIO OF NUMBER OF TUBES.

Normally MSF desalination plants have a large number of tubes. At Ghubrah, Sultanate of Oman, the long tube type unit has 9000 tubes in the brine heater and 10072-tubes/ stage in the heat recovery and heat rejection sections. The other five cross-tube type units have each 2044 tube in the brine heater, 2044 tubes/ stage in the heat recovery section and 2398 to 2404 tubes/ stage in the heat rejection section. Practically speaking, the flow through any stage is equally divided among the tubes with the pressure and temperature differences across any one tube are the same as for the total flow through the stage. In other words any one tube can be thought of as a flow stream tube with similar flow pattern, velocity, heat transfer rate and same fluid with same concentration is flowing through like in all other tubes. Now if we think of a model plant with one or a small number of tubes/ stage and with the same overall pressure and temperature differences as the main plant and with the same brine concentration, it is then expected that the model plant will behave similar to the main plant.

This requires that the number of tubes in the model is a whole number. But main plant may not have the same number of tubes in the brine heater, heat rejection and heat recovery sections. The heat recovery section, however, has the biggest number of stages and the highest temperature elevation with the problems of scaling, corrosion and fouling and the required chemical treatments mainly taking place in it. The heat recovery section is hence taken as the reference for the ratio. The ratio (r) of the number of tubes of the model heat recovery section to that of the main is taken as the ratio for the physical dimensions of the model. If in the main plant the number of tubes in the brine heater or heat rejection is different from the heat recovery section, the resulting respective number of tubes in the model will be a whole number plus a fraction of a tube. The fraction of a tube is a full-length tube with a flow area ratio equal to the fraction. It is difficult to maintain conditions in a fractional tube close to actual. This is because while the ratio of area of flow is equal to the fraction, the ratio of surface area is equal to the square root of the fraction. Also, in a fractional tube the flow pattern will be different from the actual. It is, hence, important to decide the number of tubes for the model heat recovery section carefully so that the resulting fraction in other sections, if any, will be as close to unity as possible.

A FIVE- TUBES MODEL.

Let us consider a 5-tube model for the 6 MGPD cross-tube type MSF plants of Ghubrah. Table No. 1 shows the tube details of Ghubrah plants Ref.[1,2,3,4].

Table No.1 : Tubes in Ghubrah Desalination Plants No. 2 to 6

Plant No.	**Section stages**	**Number of tubes/stage**	**Number of (mm)**	**Outside Diam.**	**Thickness (mm) (mm)**	**Effective tube length**
2, 3 & 4	Brine Heater	1	2044	35	1.2	17400
	Heat Recovery	13	2044	35	1.2	15446
	Heat Rejection	3	2398	31.75	1.2	15446
5 & 6	Brine Heater	1	2044	35	1.2	17400
	Heat Recovery	13	2044	35	1.2	15350
	Heat Rejection	3	2404	31.75	1.2	15350

Now if we consider a model with five tubes in the heat recovery section and with reference to above table then: -

- * The ratio of number of tubes = r = 5: 2044
- * The number of tubes in the model brine heater = 2044 (5/2044) = 5 tubes.
- * The number of tubes in the model heat rejection Section for plants No. 2,3 & 4 = 2398 (5/2044) = 5.86 tubes.
- * The number of tubes in the model heat rejection Section for plants No. 5 & 6 = 2404 (5/2044) = 5.88 tubes.

The fraction is 0.86 for No. 2,3 & 4 plants and 0.88 for No. 5 & 6 plants, which in both cases can be approximated to a full tube. Hence, the model heat rejection will be of six tubes. Anyway seawater flow can be adjusted within the flow ratio (r).

STAGE DIMENSIONS:

Using the above ratio (r = 5: 2044), the stage dimensions for the model plant can be worked out as follows: -

* First let: -

- X ≡ The geometrical dimensions in the direction perpendicular to the flow of the flashing brine in the main plant .
- Y ≡ the geometrical dimensions in the direction parallel to the flow of flashing brine
- Z ≡ The height dimensions.

First consider the lower part of the stage (below the demisters). This is the part in which the flashing brine is flowing. In order for the relative amount of vapor to be flashed and for the temperature to drop the same amount as actual, the residence time must be same. The residence time is a function of the specific brine flow per width and of the geometry of the stage, namely the length Ref.[5]. Taking this into consideration the flow of the flashing brine in the model stage is best thought of as a slice of the main stage flow cut with a width of (r X) and running the full-length (Y). But this will be too narrow and too long a slice that may restrict the flow and result in loss of equilibrium. To overcome this the slice can be cut into a reasonable number of strips (Fig. 1a) as follows: -

Area of the slice = r X Y
Take number of strips = 5

Then: -

Modified length = Y/5
Hence modified width = 5 r X

For the reduced length (Y/5), the brine specific flow per width is also reduced as the width is increased by five times. The specific flow is namely reduced from 779661-kg./hr. m. (which is the main plant flow) to 155932 kg./hr.m. for a model stage length ranges from 0.53 to 0.76m. This is still within the range for low output plants in practice Ref.[5]. The orifice opening will also be modified (Fig. 1b) so that its area is the same. The orifice opening can be made adjustable for further practical adjustments. The weir dimensions and position to be decided practically in order to maintain level of brine close to actual.

Now for the relative volume of this part of the stage to be of the same ratio (r) the height must be kept full, since the area is already of ratio *r*. In other words the height of the demisters above the surface of the brine will be same like actual. This is in fact required, as for similar conditions the carry over will be similar.

Same like stage bottom area, the area of demisters in each stage will be of ratio *r* and the thickness will be full.

Elevation, as mentioned earlier, will be full with some allowance for required adjustments depending on the practically resulting brine depth. The demisters are shifted to the middle of the stage (Fig. 1c) for better overall arrangement of the model.

For the upper part of the stage the relative volume is of the same ratio *r* and the relative heat transfer surface is that of the full length and diameter tubes of the model stage. This part is composed of two spaces. One is the space above the demisters with its area equal to demister area and height is full. The other space is the tube bundle and the distillate tray space. For this latter space to be of reasonable tube length let us cut (or bend) each tube into a reasonable number of equal pieces (say 5 pieces).

Then: -

Model tube length	= (full tube length)/5
Volume of bundle space	= r (full tube length) (full space cross sectional area)
	= $5r$ [(full length)/5] (full area)
Model bundle X-sectional area	= (volume)/ (Model tube length)
	= $5r$ (full area)
Hence the ratio of areas	= $5\ r$

The square root of the area ratio $\sqrt{(5r)}$ is used to get the relative dimensions c & d in Fig 2a

Keeping the same relative volume and allowing some distance above demisters for any required height adjustments, the space above the demisters is modified to match the bundle as shown in Fig. 2b. To minimize the overall size of the model, the tubes are placed parallel to the direction of brine flow in the flash chamber. View of the complete model stage with interconnecting headers is shown in Fig.3.

The distillate trough will run outside the stages connecting the distillate trays (Fig.3) The square root of the ratio *r* will be used for the relative dimensions of the cross-sectional area and for the opening that connects the distillate tray to the trough. The dimensions of the orifice openings in the trough will also have the same ratio ($\sqrt{r}$).

BRINE HEATER

Same as for the stages, each tube of the brine heater will be cut (or bent) into 5 equal pieces. This results in a cross-sectional area ratio of ($5r$) then:-
Diam. of the Brine Heater Shell = (Full diam.) $\sqrt{(5\ r)}$
The hot well can be treated in same manner. The height will be 1/5 of the full and the diam, is $\sqrt{(5r)}$ times the full diameter.

VENTING SYSTEM

If the ratio *r* is used for the vent openings & vent lines then the lines will be of too small diameter. Reasonable diam. can be used with throttling valves on each line from each stage and these valves to be adjusted during initial running of the model plant till close to actual stage conditions are attained. All stage vents connect via the throttling valves to one header that connects to the highest vacuum point in the main plant (Fig 4).

BRINE RECIRCULATION

For the ratio used the brine recirculation flow will be around 28 m^3/hr. only but the pressure must be same like main plant. A small pump can be selected to satisfy this requirement. Otherwise brine can be taken from the main plant brine recirculation pump discharge as shown in Fig. 5. The point of connection is just at the discharge of the pump, which ensures enough pressure, and it is also upstream to the chemical dosing and the ball cleaning injection points. This is required as chemicals, chemicals dosing rates and ball cleaning are some of the main experimental variables. A suitable connection is a T-joint on the discharge vent line. The ratio of diameter of this line (80mm.) to the main discharge line (1500mm.) is slightly more than the square root of the ratio *r*. The model brine recirculation pump, if required, or the supply from the main plant are connecting to a header fitted with pressure and temperature gauges & flow meters to which each of the five model tubes is connected. At the outlet from the brine heater again each tube connects to a common header and then via a pressure control valve, the brine enters the first stage flash chamber & flows through to the last stage. The model pump suction is from last stage and blow down is from discharge of the pump via a valve to control the level of the last stage. In case pump is not required then connection from last stage should be made to the suction vent line (80mm.diam) of the main pump and again this line to be provided with a valve to control last stage level. In this latter case there is no separate blow down line. The concentration is controlled via the same level control valve.

SEA WATER AND MAKE -UP

Sea water supply pump is not required. Seawater will be taken from any suitable drain or vent connection on the sea water line between the tempering point and inlet to the heat rejection section of the main plant. Sea water supply line will be fitted with tapping for measuring the temperature, pressure and flow. Each of the six model heat rejection tubes connects to a common header and at outlet again the tubes will join on a common header fitted with a flow control valve and tapping for measuring temperature and pressure.

Make-up can be taken from the main plant before or after the anti-foam dosing point to the deaerator of the model.

STAM AND CONDENSATE

The required amount of steam will be around 0.44 t/hr. The steam will be taken from a suitable drain trap or vent on the LP steam supply to the main brine heater. The line will be fitted with a flow control valve and tapping for measuring the pressure, temperature and flow. The condensate connects back to the suction of the main brine heater condensate pump via a line fitted with level control valve. An optional flash vessel can be used to dump the condensate.

DISTILLATE

The product distillate (≅ 2.8 t/hr.) can be pumped out from the distillate sump in the last stage via a pump with level control valve on the discharge line and with tapping for flow and quality measurements. The product distillate can be connected to the main production line or dumped to drain.

ON- LOAD TUBE CLEANING

On-load tube cleaning should be designed and operated carefully, as the number of tubes is very limited. The injection of sponge rubber balls should be one at a time and in only one tube. When the ball is caught in the outlet of the brine heater then only the next ball in the same or a different tube can be injected.

INSTRUMENTATION AND CONTROL

As the model plant is small it is possible to control the flow, levels, pressure and temperature by manually adjusting the respective control valves. Any way for measurements the instruments should be accurate and provision for sample collection should also be possible. In order to allow comparative tests each of the model tubes in each stage should be provided with tapping, probes and pockets at inlet and outlet for measurement of temperature and pressure. These tapping should be made so that the measuring equipment can easily be fixed or removed so as to limit the number of measuring equipment. It will not be necessary to always monitor all readings at all points same time. The stages also should be provided with probes, tapping and pockets for measurement of different parameters above and below the demisters and at the distillate trays and troughs. Level gauge glasses and inspection windows are also required.

MATERIALS

The tubes and demisters material will be same as in the main plant unless it is required to test some different materials. For the body of the evaporator and as the volume is small thin metal sheet (e.g. 3mm. painted M. S or SS) can be used. Insulation material can be same as main material and its thickness will also be same since the temperatures are same as for the main plant. The piping, the headers and the impulse lines can be SS.

COST ESTIMATION

No actual cost survey is made in this study but a rough estimation of the total cost is in the same ratio 5:2044 of the main plant cost. In this estimation it is assumed that the cost of the evaporator is mainly tube cost.

CONCLUSION

Having five tubes and a full number of stages the model allows to carry out different tests simultaneously or at different times. The following are some possible examples of tests: -

- Tubes of different materials, thickness or shape.
- Different chemicals (scale or corrosion control chemicals).
- Chemical dosing at different rates in different tubes.
- Tests of different on-load tube cleaning cycle duration, frequencies or different types of balls.
- Different stage wall materials, claddings or paints.
- Demisters of different materials, configuration, and different heights.

A combination of the above tests can be carried out simultaneously at different parts of the model. For example one can carry out tests on some anti-scale chemical while another can carry out tests on demisters. It is also possible to by-pass any number of stages if not required in the test by changing the connection points between the model and the main plant. Some stages can even be detached and connected to some other plant.

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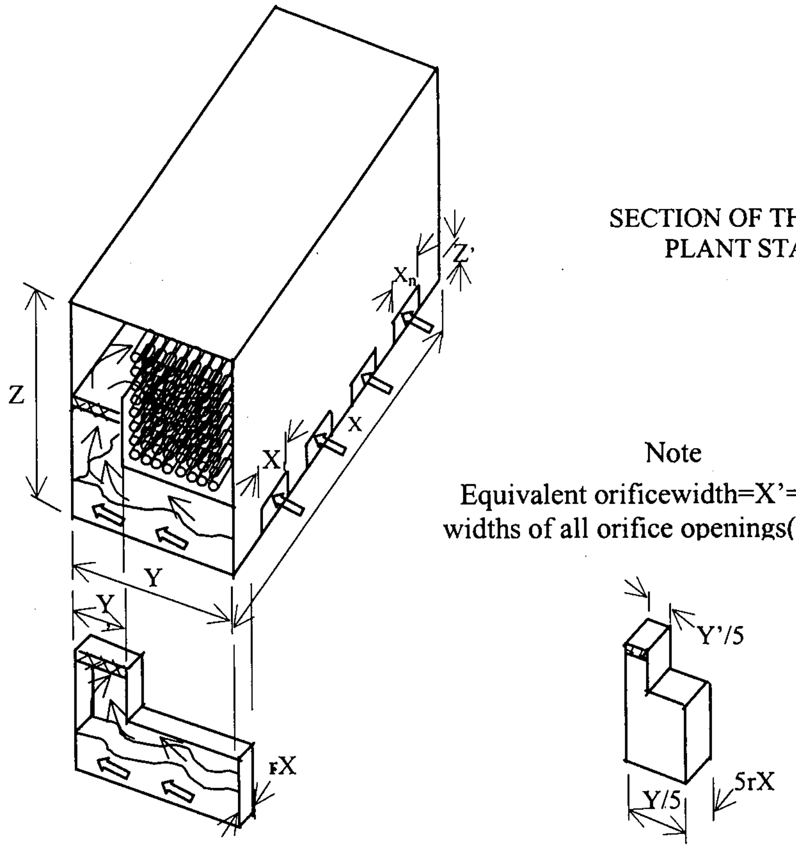


SECTION OF THE MAIN PLANT STAGE

Note

Equivalent orificewidth=X'=Sum of widths of all orifice openings(X_1 to X_n)

A SLICE FROM THE BOTTOM PART OF THE MAIN STAGE

THE SLICE WITH MODIFIED AREA

Figure 1a

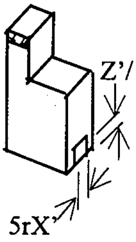


Fig.1b
MODIFIED ORIFICE
OPENING

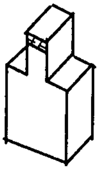
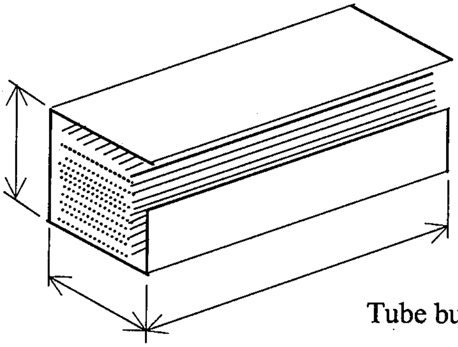
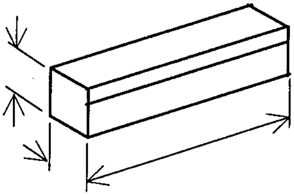


Fig.1c
DEMISTERS SHIFTED TO
THE MIDDLE

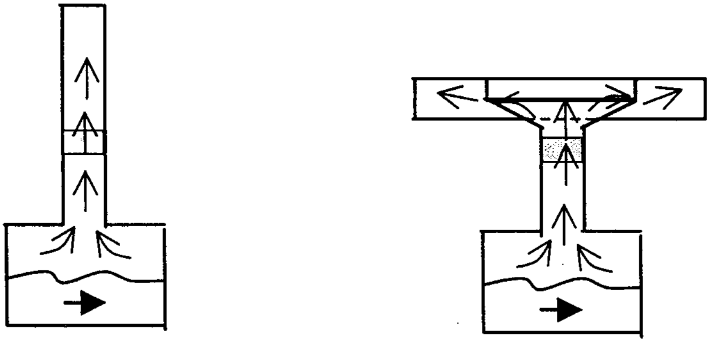


Tube bundle & distillate tray in
main stage



Model tube bundle &
distillate tray

Figure 2a



Model flash chamber

Space above demisters modified to suit tube bundle

Figure 2b

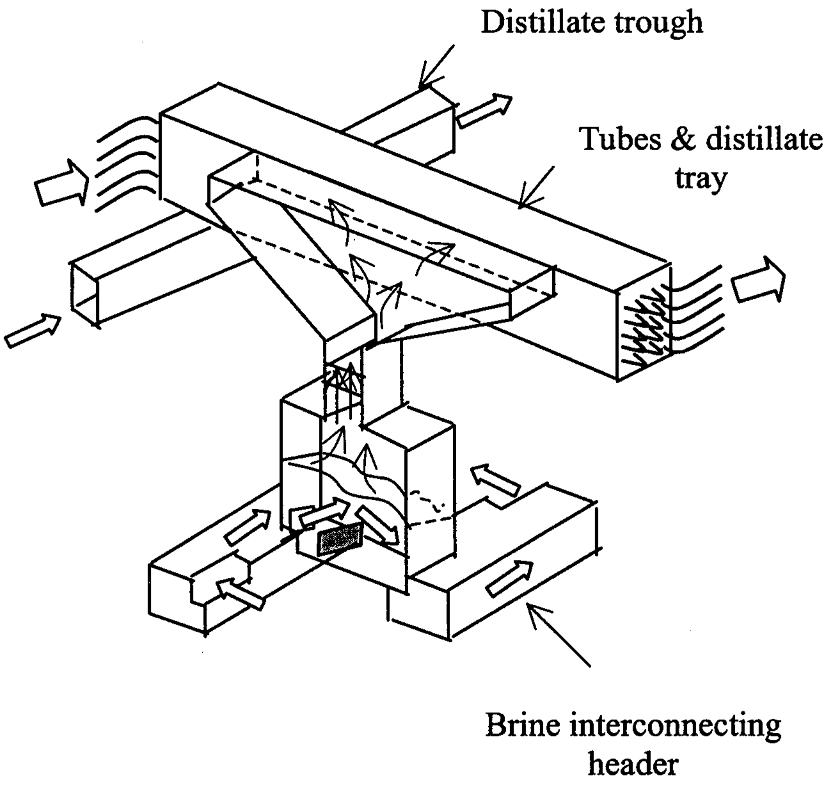


Figure 3

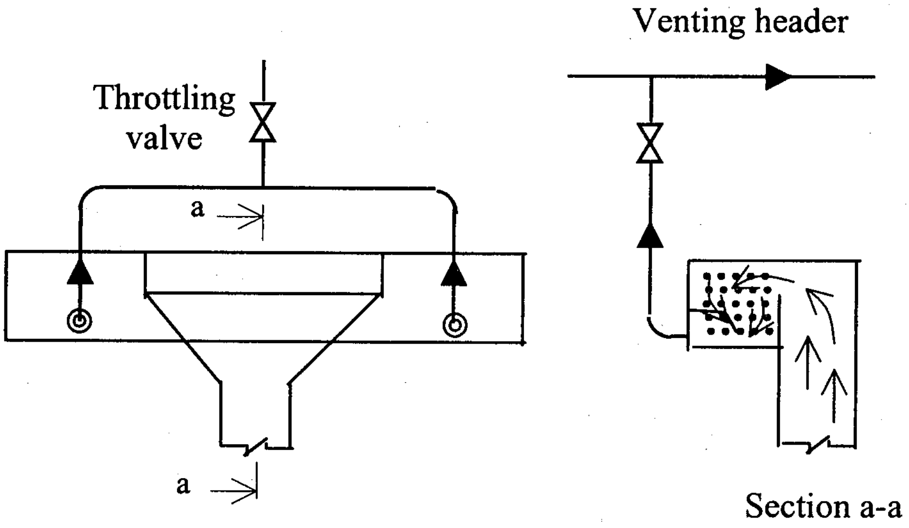


Figure 4

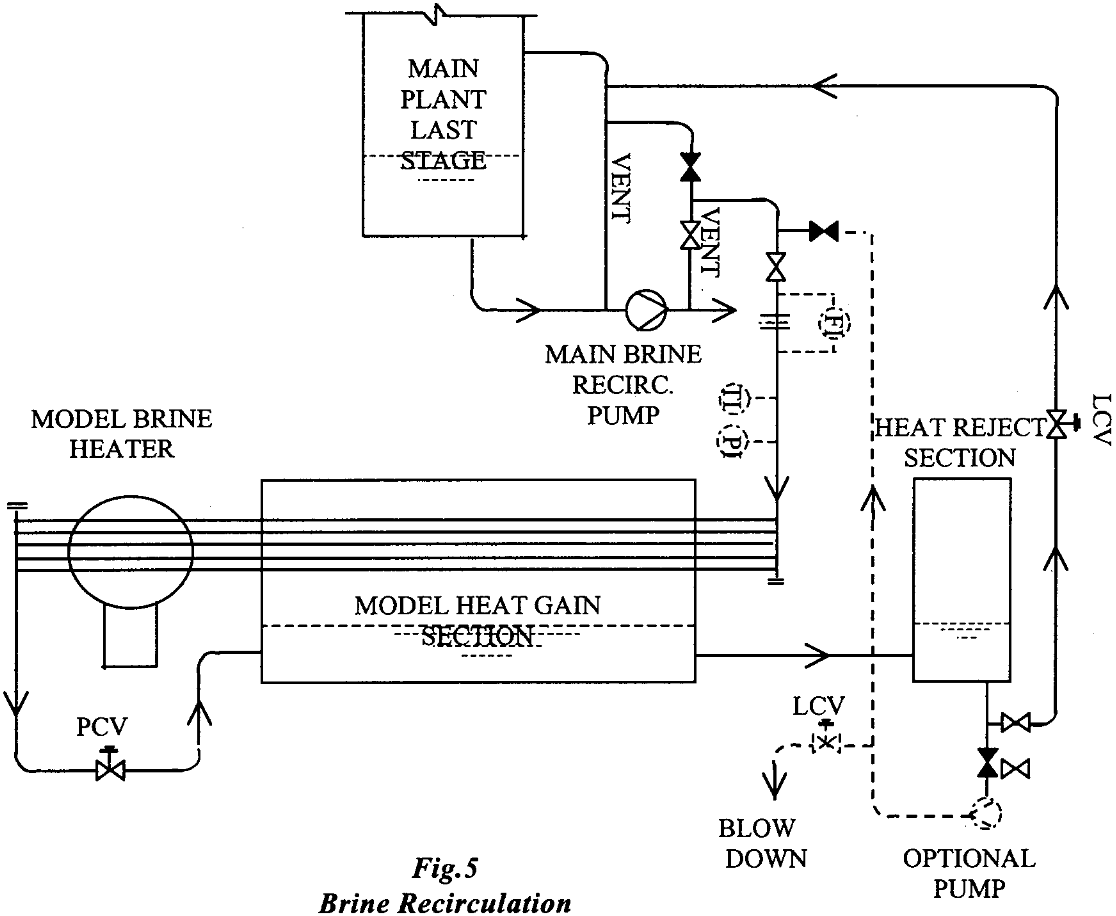


Fig.5
Brine Recirculation

Computer Package for Design/Rating of Thermal Desalination Processes

Hisham M. Ettouney and Hisham El-Desouky

COMPUTER PACKAGE FOR DESIGN/RATING OF THERMAL DESALINATION PROCESSES

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ABSTRACT

A computer package is developed for design and rating of thermal desalination processes. This is motivated by unavailability of such packages in the literature or on a commercial scale. The package includes models for various systems of the single effect evaporation (SEE), the multistage flash (MSF), and the multi effect evaporation (MEE). The MSF systems include brine circulation, brine mixing, once through, and thermal vapor compression. The MEE configurations include parallel and forward feed systems. The SEE and MEE systems include the stand-alone and the vapor compression units. Vapor compression systems include mechanical, thermal, absorption, and adsorption heat pumps. All models are based on well-developed set of material and energy balance equations as well as correlations for evaluation of physical properties, heat transfer coefficient, and thermodynamic losses. All mathematical models have been developed and previously tested and validated by authors against available industrial and literature data. All models generate the design and rating variables, which have the strongest effect on the unit product cost. These include the thermal performance ratio, heat transfer area, and cooling water flow rate. The computer package includes startup menus and windows for design, rating, flow charting, performance calculations, and graphing of performance curves. The package allows for parameter selection, opening, saving, and printing of data files, printing of forms, and selection of graph parameters. In addition, the package performs checks on parameter range before performing in calculations. Also, all modules have error handlers for overflow with warning for attempting to perform calculations outside applicable ranges. The package can be used in undergraduate and graduate courses, in training programs, and in research. Also, it will be proved to be a great tool for practicing engineers and other desalination staff.

KEYWORDS: Simulation package, seawater desalination, thermal process, multiple and single effect evaporation, multistage flash, vapor compression.

INTRODUCTION

Desalination is a highly specialized industrial field. The process involves separation of the intake seawater or brackish water into two streams of nearly salt-free fresh water and a brine stream with high salt concentration. The separation can be thermally based or performed by membrane separation. Figure 1 shows various types of thermal and membrane separation processes. As is shown, the thermal processes are based on evaporation, flashing, or freezing. Steam heating or the solar energy can be used to derive evaporation processes. In the heating processes, fresh water vapor is boiled or flashed (free boiling) of the brine stream. While, in the freezing process, heat is removed from the seawater and ice crystals of fresh water are formed in the brine solution. The membrane-based processes involve use of a selective barrier to separate fresh water and brine. In reverse osmosis, fresh water is pressurized through membrane leaving behind a brine solution with high salt concentration. The opposite form of separation is performed in electro-dialysis, where the salt ions are transported across the membrane under the influence of external electric charge.

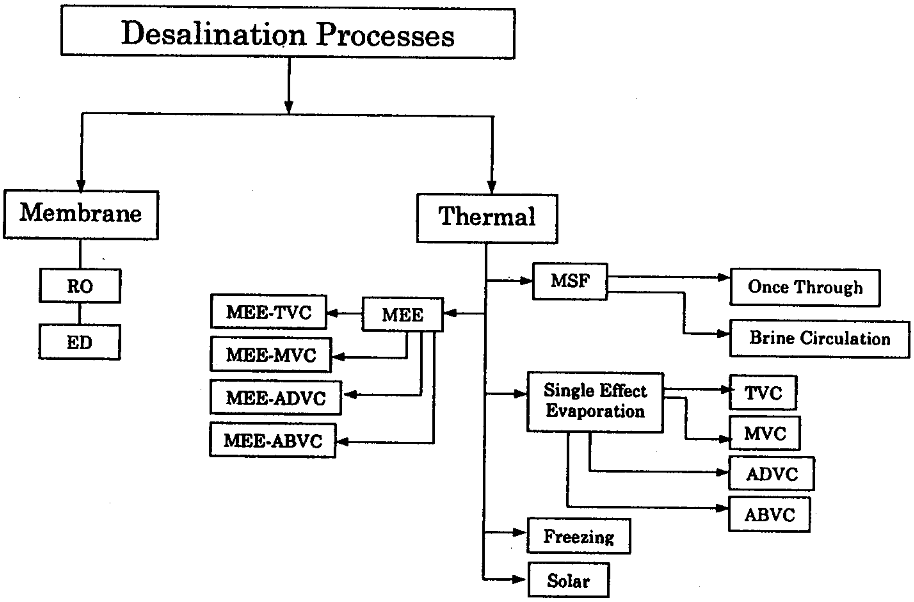


Figure 1: Thermal and membrane desalination processes

On a global scale, the desalination industry has not reached its full production capacity or the asymptotic value. Silver, 1988, assumed that the industry asymptotic production capacity is being approached with a value of 15×10^6 m^3/d. However, 10 years later the volume of the desalination industry is more than 22.8×10^6 m^3/d in 1997, Wangnick, 1998. The following facts can be attributed to this continuous increase in the desalination industry:

- The desalination industry has not yet reached its full potential in a large number of countries.
- In large producing countries, which includes the Gulf States, the desalination industry remains to expand as a result of increase in population, societal progress, and expansion in the agriculture and industrial sectors.
- In industrial states, limited resources of fresh water and continuous decrease in the desalination unit production cost made the desalination option highly attractive as a source of fresh water for selective industries and communities.
- Even in countries with limited financial resources, adoption of various desalination technologies is becoming a must to secure a clean and sustainable source of fresh water to the population. For many countries in this category, desalination is becoming the only viable and practical.

An essential part in desalination research is the development of mathematical models capable of accurate prediction of the design, rating, optimization, and operating performance of the process. Since inception of the desalination processes on a commercial scale, mathematical models became an integral part of the process development. A good example on such development can be found in modeling literature of the MSF process. Mass and energy balance models of the MSF were developed and solved to generate the necessary design data, which includes heat transfer areas, temperature profiles, stream salinity, and flow rates. Review of the developments in the MSF design and rating models include the following. Silver, 1970, developed a simple model to express dependence of the MSF performance ratio on the thermodynamic losses. El-Dessouky et al., 1985, showed the effect of the thermodynamic losses on the system parameters that affect its final cost, i.e., the performance ratio and the specific heat transfer area. Detailed mathematical models, which are based on stage to stage calculations and require computer iterative solution, are developed by Omar, 1983 and Khan, 1986. A tri-diagonal matrix solution to improve convergence of the computer iterative procedure and to reduce computation time is developed by Helal et al, 1986. Husain, 1988, showed that fitting of the model results to the plant data should take into consideration heat losses in each stage, vapor leak, presence of non-condensable gases, and evaporation from the distillate trays. Darwish, 1991, performed thermal analysis of the MSF system to arrive at quantitative assessment of the effect of design and operating parameters on the system performance. El-Dessouky et al, 1995, developed a model which takes into consideration the effect of non-condensable gases, losses to the surroundings, dependence of the thermophysical properties on stage properties. The model includes detailed account for evaluation of the heat transfer coefficients in the brine heater and the stage condensers. Results are found to compare favorably against industrial data. A fixed-point iteration scheme is developed by El-Dessouky and Bingulac, 1996, to simplify problem formulation and computer coding. El-Dessouky et al., 1998 presented a step by step mathematical development for the origin of the MSF process. The development starts at evaluation of the performance of a single stage flash unit and goes through evaluation of the once through, brine mixing, and the conventional MSF systems.

Similar developments in system modeling and analysis can be found for other thermal and membrane desalination processes, however, developments are limited to improvements in the physical model and techniques for analytical or numerical solution. For the thermal desalination processes, the computer packages were kept sealed inside consulting firms and large manufacturing companies for design and rating of newly and existing contracted units. This was dictated in part by competition among the designer, consulting firms, and construction companies. As a result, these are not available to many engineers in the field. Also, these packages are not suitable for education and training purposes. As discussed above, competition in the thermal desalination field is caused by the high capital of these processes. In practice, large industries or central governments order and operate most of the large desalination installations. The capital cost for large-scale installations, above 6 mgd, for either technology is well within $\$100 \times 10^6$. However, very small RO installations, which can be used by individuals, small workshops, and laboratories, are quiet common and their cost, although still relatively high (close to $5000), is still affordable. This fact motivated several manufacturers of the RO modules to develop computer packages for design and simulation of these units. Of course each of these packages includes data base on the membrane characteristics produced by the manufacturer (recovery, salt rejection, module area, module size, operating temperature).

The following facts motivated development of the thermal desalination computer package:

- The large increase in the size of the desalination industry.
- The large number of engineers involved in various process sectors

(design, construction, operation, maintenance, management, etc.).
- The unavailability of integrated computer packages for thermal desalination, which can be used by engineers and for education, research, and training purposes.

The following sections include main elements of the computer package and a case study on the single effect mechanical vapor compression, or MVC.

ELEMENTS OF THE COMPUTER PACKAGE

Computer packages are developed to perform general process flow sheet calculations or a very specific system design. Development of a fully integrated computer package should take into consideration the following:

- Availability of various systems and unit processes to construct or select the desired flow diagram.
- Ability to select parameters for the calculations and various types of graphs.
- Limiting the value of input parameters within specified ranges.
- Ability to move within the package various windows and displays to adjust selections and parameter values.
- Availability of help and tutorial files.
- Capability to open, save, and print data files and graphs.
- Capabilities for various error handling.

Such features simplify the use of the computer package and allow for further improvement in the package in future developments.

The thermal desalination package includes the following main elements, Figure 2,

- Help files,
- Selection of the thermal desalination process,
- Choice of the calculation type (design, rating, flow chart, or performance), and
- Selection of various types of output data.

The process selection feature includes three choices, which are the single effect evaporation (SEE), the multiple effect evaporation (MEE), and the multistage flash (MSF). Sub-selections are included for the SEE, MEE, and MSF selection. For the SEE, the sub-selections are:

- Single effect evaporation with no vapor compression (SEE),
- Single effect mechanical vapor compression (MVC),
- Single effect thermal vapor compression (TVC),
- Single effect absorption vapor compression (ABVC), and
- Single effect adsorption vapor compression (ADVC).

The MEE sub-selections are parallel and forward feed MEE. Each of these selections has five sub-selections, which include the following:

- Multiple effect evaporation with no vapor compression (MEE),
- Multiple effect thermal vapor compression (MEE-TVC),
- Multiple effect mechanical vapor compression (MEE-MVC),
- Multiple effect absorption vapor compression (MEE-ABVC), and
- Multiple effect adsorption vapor compression (MEE-ADVC).

The MSF process has five sub-selections, which include the following:

- Single stage flashing (SSF),
- Once through multi-stage flashing (MSF-OT),
- Brine mixing multi-stage flashing (MSF-M),
- Brine circulation with thermal vapor compression (MSF-TVC), and
- Brine circulation multi-stage flashing (MSF).

The selection of a specific process leads to choices for the calculation type, which includes design, rating, process flow chart, and performance calculations.

CASE STUDY FOR THE MVC DESALINATION PROCESS

To demonstrate various features of the computer package, the case of MVC desalination process is considered in this section. The description of the MVC system is given in Appendix A. Also, the mathematical model and solution parameters used in the computer package are given appendices B and C. Since its development in the late 60's, Matz and Zimerman, 1985, progress have been achieved in system design and operation. The specific power consumption for early designs were as high as 18 kWh/m^3, Matz and Fisher, 1981. Development in the efficiency of the compressor and evaporator driving force reduced the specific power consumption to lower values between 6-8 kWh/m^3. Earlier the single unit capacity with limited

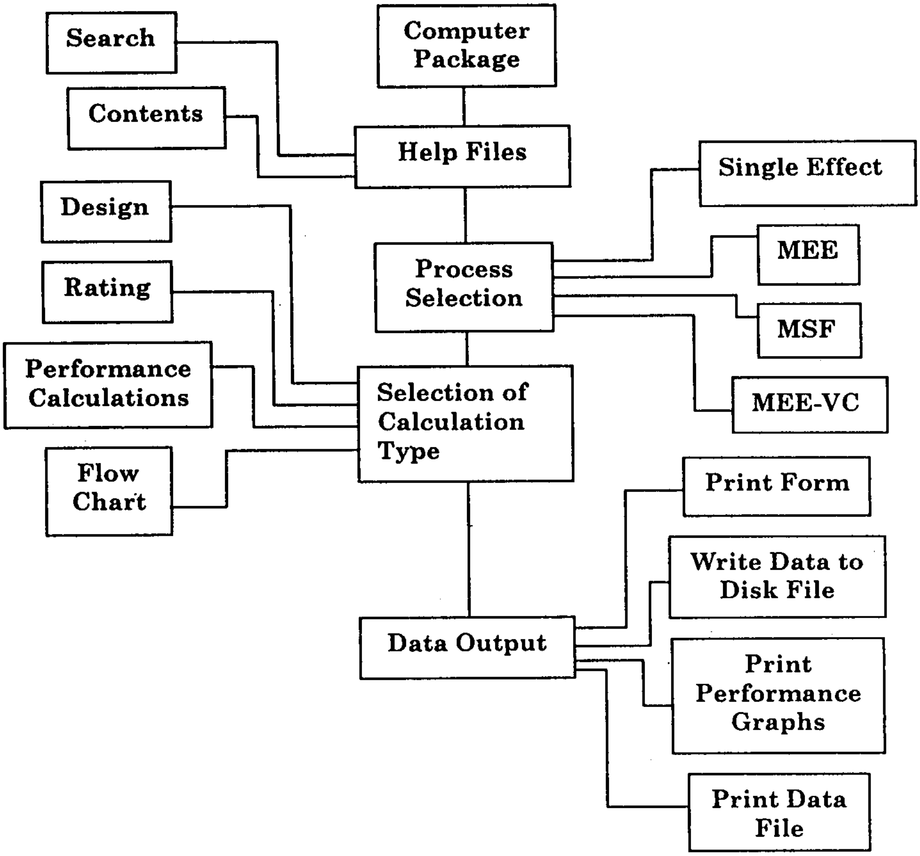


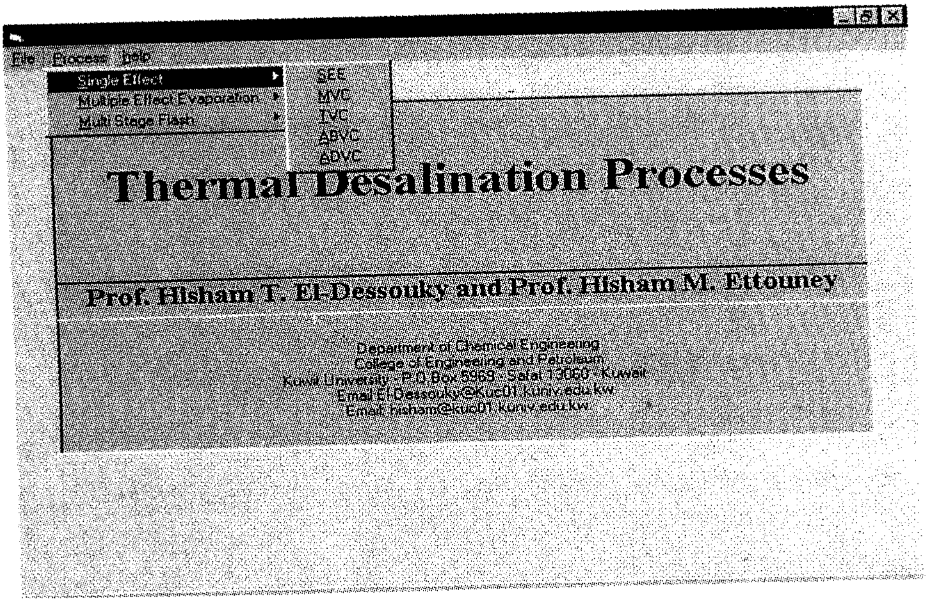
Figure 2: Elements of the Computer Package

to production rates below 500 m^3/d. This drawback was met by adopting the larger capacity multi effect vapor compression system with production rates of 3000 m^3/d, Lucas and Tabourier, 1985. Today, developments in compressor capacity allows for single unit operation with a production rate of 5000 m^3/d. Plant factor is high with averages of 90%, Veza, 1995. The number of operating units is more 200, Zimerman, 1994, which operate in single or multi-effect modes. Unit production cost is below \$0.46/m^3, Zimerman, 1994.

Modeling and performance evaluation of the MVC system can be found in a number of studies by Darwish, 1988, Al-Juwayhel et al., 1997, Ettouney and El-Dessouky, 1998, and Ettouney et al., 1998. The mathematical model by Ettouney et al., 1998, used in the calculations of the MVC routine, includes two additional features over the previous studies. These features are 1) evaluation of the salt concentration of the rejected brine as a function of the calcium sulfate solubility and the rejected brine temperature and 2) thermodynamic evaluation of the vapor superheating during compression. Startup of the calculations shows the main title window, Figure 3. Selection of MVC leads to its main display, which includes selections for design, rating, process flow chart, and performance calculations. The design and rating forms have a similar layout, where the input parameters are displayed in a separate area from the output data. Also, each form has a menu for file management and calculations. The file menu includes print commands and an option to return to the main window of the MVC process. Selecting the solve command results in loading and solution of the MVC model. Upon completing the solve command, results are then shown in the output display boxes for the heat transfer areas, flow rates, and temperatures, Figure 4. The flow chart window performs calculations similar to the design window. However results are displayed upon pressing each of the unit processes shown on the diagram, Figure 5. Calculation parameters for the design and process flow chart include:

- The intake seawater temperature is limited to a range of 5-35 °C. The lower limit corresponds to wintertime temperatures in cold climates and the upper limit represents the summer time temperatures in equatorial regions.
- The limit on the intake seawater salinity is 32,000-54,000 ppm, which represents open oceans and the gulf seawater in the summer time, respectively.
- The brine boiling temperature is limited to a range of 50-120 °C.
- The temperature difference between condensate and boiling brine is limited to a range of 1-4 °C.
- The product flow rate is not limited to a specific value. A value of 1 kg/s is used and is recommended, since all system design parameters are proportional to the product flow rate, i.e., specific heat transfer, specific power consumption, and flow rate of brine and intake seawater.

The window for the performance calculations include listings for the system parameters, which include the intake seawater temperature, the intake seawater salinity, the brine boiling temperature, and the temperature difference of the condensate and boiling brine, Figure 6. Each of these parameters has three input cells for the low limit, upper limit, and step. In addition, each parameter has a check box, which upon marking result in inclusion of the parameter in the performance calculations, where the system characteristics are evaluated over the parameter range, i.e., between the lower and upper limits at the specified step. Performance calculations



Thermal Desalination Processes

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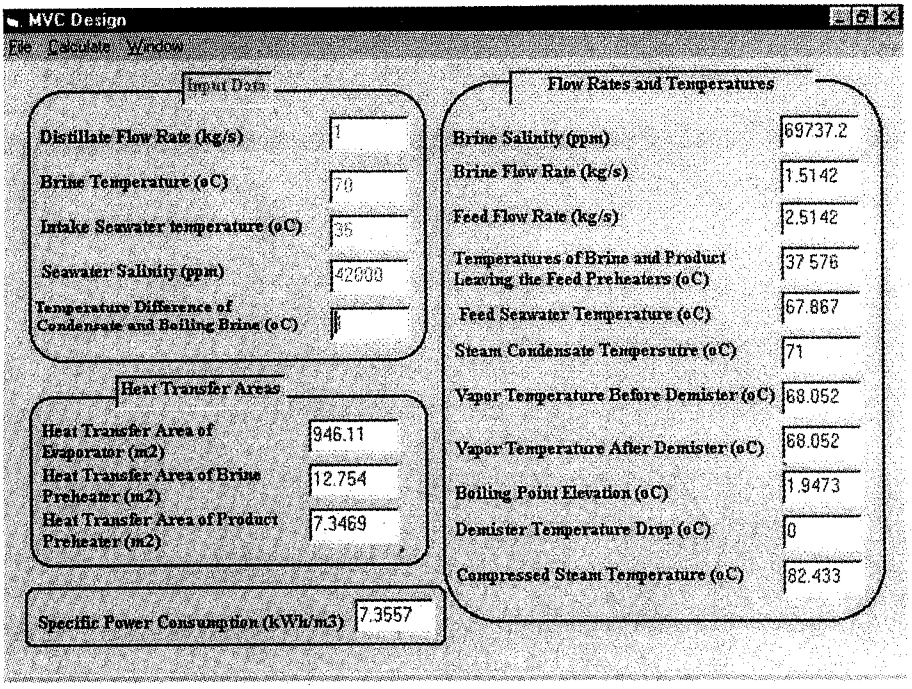
Figure 3: Package main form with menus for file, process selection, process sub-selection, and help.



Input Data

Distillate Flow Rate (kg/s)	1
Brine Temperature (oC)	70
Intake Seawater temperature (oC)	35
Seawater Salinity (ppm)	42000
Temperature Difference of Condensate and Boiling Brine (oC)	1

Heat Transfer Areas

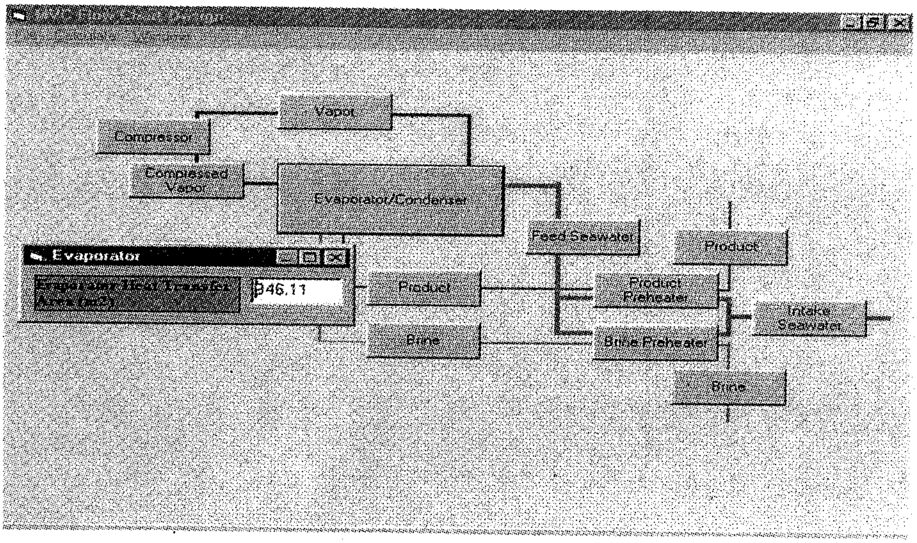
Heat Transfer Area of Evaporator (m2)	946.11
Heat Transfer Area of Brine Preheater (m2)	12.754
Heat Transfer Area of Product Preheater (m2)	7.3469

Specific Power Consumption (kWh/m3)	7.3557

Flow Rates and Temperatures

Brine Salinity (ppm)	69737.2
Brine Flow Rate (kg/s)	1.5142
Feed Flow Rate (kg/s)	2.5142
Temperatures of Brine and Product Leaving the Feed Preheaters (oC)	37.576
Feed Seawater Temperature (oC)	67.867
Steam Condensate Tempersutre (oC)	71
Vapor Temperature Before Demister (oC)	68.052
Vapor Temperature After Demister (oC)	68.052
Boiling Point Elevation (oC)	1.9473
Demister Temperature Drop (oC)	0
Compressed Steam Temperature (oC)	82.433

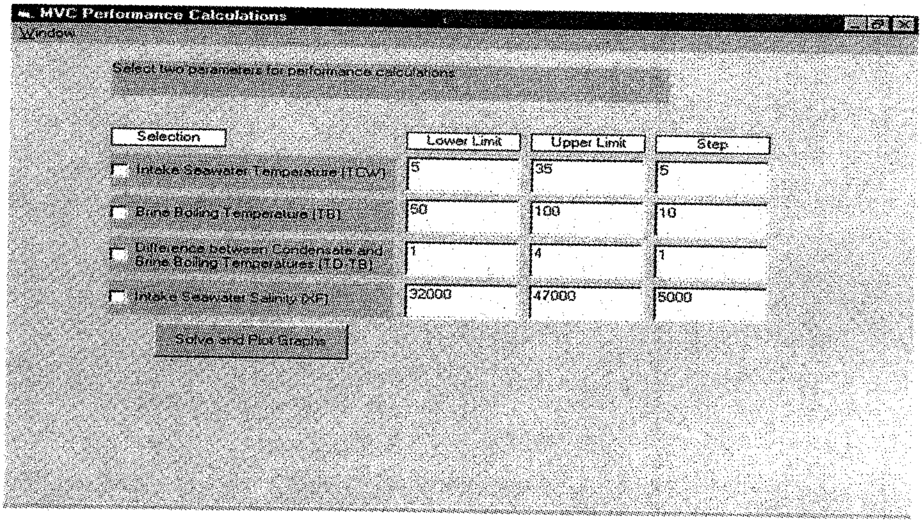
Figure 4: MVC design display



Evaporator

Evaporator Heat Transfer Area (m2) 346.11

Figure 5: Process flow chart display


MVC Performance Calculations

Select two parameters for performance calculations

Selection	Lower Limit	Upper Limit	Step
☐ Intake Seawater Temperature (TCW)	5	35	5
☐ Brine Boiling Temperature (TB)	50	100	10
☐ Difference between Condensate and Brine Boiling Temperatures (TD-TB)	1	4	1
☐ Intake Seawater Salinity (XF)	32000	47000	5000

Figure 6: MVC process, flow chart, menu, and calculation results

generates a sequence of graphs for the variations in the system parameters, which includes the specific power consumption and the specific heat transfer areas for the evaporator and feed preheaters.

CONCLUSIONS

A computer package is developed for design, rating, and evaluation of the performance of various thermal desalination processes. Although, a number of computer packages are available for the reverse osmosis processes, none is found for the thermal desalination processes. Therefore, the computer package presented here is the first of its kind to be made available to the desalination community. The features of the developed thermal desalination computer package are summarized below:

- The package includes models for the MSF, MEE, and SEE.
- The MSF models include the once through, brine mixing, and brine recycle.
- The MEE models include the forward and parallel feed with/without vapor compression.
- The SEE models include the stand-alone system as well as four vapor compression systems.
- The package includes calculations for design, rating and performance.
- The package allows for printing of forms, data files, and charts.

Features of the computer package are demonstrated for the case of MVC. As is shown, the package allows for single point evaluation for design and rating calculations. Also, results can be viewed on a flow chart to assist in understanding of relation of various design and operating parameters. The performance calculations generate large sets of data to be viewed on a sequence of graphs for the design variables versus the system parameters. All of these features are also available for selections of other thermal desalination processes.

LIST OF SYMBOLS

A	Heat transfer surface area, m^2.
BPE	Boiling point elevation, °C.
C_p	Specific heat at constant pressure of seawater kJ/kg °C.
d	Distance between plates in preheaters, m.
De	Equivalent diameter, m.
h	Heat transfer coefficient, kW/m^2 °C.
H	Enthalpy of product water.
k	Thermal conductivity, kW/m °C.
L	Demister thickness, m.
LMTD	Logarithmic mean temperature difference, °C.
M	Mass flow rate, kg/s.
P	Pressure, kPa.
ΔP	Pressure drop in the demister, Pa.
Pr	Prandtl number.
Q	Heat transfer rate, kW.
Qc	Specific power consumption, kWh/m^3.
r	Tube radius, m.
R	Gas constant, kJ/kgmole °C.
Re	Reynolds number.
R_f	Fouling Resistance, m^2 °C/kW.
s	Salinity, gm/kg.
sA	Specific heat transfer surface area, $m^2/(kg/s)$.
S	Entropy, kJ/kg °C.
t	Temperature of seawater, °C.
T	Temperature of vapor and brine, °C.
ΔT_c	Temperature difference of condensing steam and formed vapor, °C.
U	Overall heat transfer coefficient, kW/m^2 °C.
v	Velocity, m/s.
V	Specific volume of vapor, m^3/kg.
w	Width of plates in preheaters, m.
W	Compressor specific work, kJ/kg.
X	Salt concentration, ppm.
Z	Compressibility factor.

Greek Letters

δ	Plate thickness, m.
ρ	Density, kg/m^3.
η	Compressor efficiency.
μ	Dynamic viscosity, kg/m s.
λ	Latent heat of vaporization, kJ/kg.
γ	Compression ratio.

Subscripts

b	Brine.
cw	Cooling seawater.
d	Distillate vapors and product water.
e	Evaporator.
f	Feed seawater.
i	Inside tubes.
ℓ	Liquid phase.
n	Adiabatic process.
m	Polytropic process.
o	Outside tubes.
p	Demister.
r	Reference state.
s	Compressed vapor.
v	Vapor state.
w	Tube wall.

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APPENDIX A: DESCRIPTION OF MVC

The MVC system contains five major elements, which include mechanical vapor compressor, evaporator/condenser heat exchanger, preheaters for the intake seawater, brine and product pumps, and venting system.

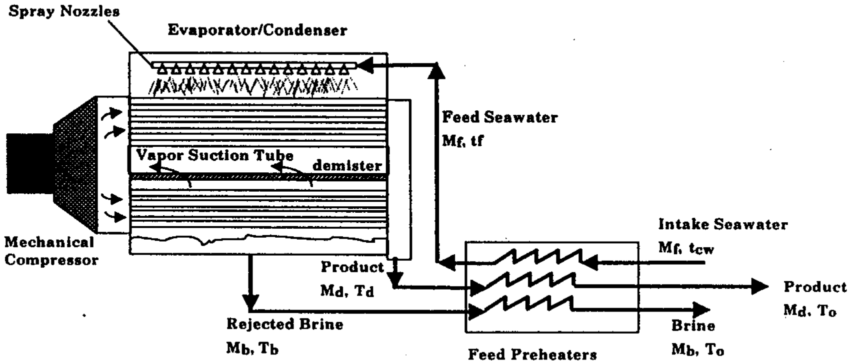

Figure A.1: Mechanical Vapor Compression (MVC) Evaporator-Desalination Process

Figure A.1 shows a schematic diagram for the system, where the compressor and evaporator/condenser heat exchanger form one single unit. The evaporator/condenser contains falling film horizontal heat exchange tubes, spray nozzles, vapor suction tube, and wire-mesh mist eliminator. The feed preheaters are plate type heat exchangers, which operates on the intake seawater and the hot liquid streams leaving the evaporator.

The feed seawater enters the evaporator and is sprayed over the horizontal tube bundle. The spray forms a falling film over succeeding tube rows. Formation of the thin film enhances the heat transfer rate and makes the evaporation process more efficient. The temperature of the formed vapor is lower than the boiling temperature by the boiling point elevation, BPE. The vapor transfers from the evaporator section to the compressor through the vapor suction tube, which is guarded by a wire-mesh mist eliminator. This is necessary to separate brine droplets, which may be entrained with the vapor stream and would result in damage of the compressor blades and reduction of the product quality. The demister pressure losses cause further drop in the vapor saturation pressure and consequently decrease the vapor saturation temperature. The vapor flows tangentially through the compressor, where it is superheated. Upon compression, the vapors are forced inside the horizontal tubes, where it loses the superheat temperature difference and its temperature drops to the saturation temperature. Condensation takes place and the released latent heat is transferred to the brine film. The balance of energy within the system is maintained by recovery of the thermal energy of the rejected brine and product streams. This is achieved in the feed preheaters, which are plate type heat exchangers. As heat is exchanged between the intake seawater, the product, and the rejected brine the temperature of the seawater is increased and the temperatures of the rejected brine and product streams are reduced.

APPENDIX B: PROCESS MODELING

The MVC mathematical model includes the material and energy balance equations as well as and the heat transfer rate design equations. The model predicts the temperature and salt concentration of various streams, the heat transfer areas, and the specific power consumption. These variables have strong effect on the cost of product water. The following is a list of the model equations. Further details and discussion of the model can be found in the study by Ettouney et al., 1998.

Material Balances

$$M_f = M_d + M_b \tag{B.1}$$

$$M_f X_f = M_b X_b \tag{B.2}$$

Feed Preheaters Energy Balance

$$\begin{aligned} Q_e &= M_f C_p (t_f - t_{cw}) \\ &= M_d C_p (T_d - T_o) + M_b C_p (T_b - T_o) \end{aligned} \tag{B.3}$$

Evaporator Energy Balance

$$\begin{aligned} Q_e &= M_f C_p (T_b - t_f) + M_d \lambda_d \\ &= M_d \lambda_d + M_d C_{p_v} (T_s - T_d) \end{aligned} \tag{B.4}$$

Evaporator/Condenser Heat Exchanger and

$$\begin{aligned} A_e &= \frac{M_d \bar{\lambda}_v + M_f C_p (T_b - t_f)}{U_e (T_d - T_b)} \\ &= \frac{M_d \lambda_d + M_d C_{p_v} (T_s - T_d)}{U_e (T_d - T_b)} \end{aligned} \tag{B.5}$$

Preheaters Heat Transfer Area

$$A_d = \frac{M_d C_p (T_d - T_o)}{U_d (LMTD)_d}$$

$$= \frac{\alpha M_f C_p (T_f - T_{cw})}{U_d (LMTD)_d} \tag{B.6}$$

$$A_b = \frac{M_b C_p (T_b - T_o)}{U_b (LMTD)_b}$$

$$= \frac{M_d (X_f / (X_b - X_f)) C_p (T_b - T_o)}{U_b (LMTD)_b}$$

$$= \frac{(1-\alpha) M_f C_p (T_f - T_{cw})}{U_b (LMTD)_b} \tag{B.7}$$

The $(LMTD)_d$ is defined as:

$$(LMTD)_d = \frac{(T_d - T_f) - (T_o - T_{cw})}{\ln \dfrac{T_d - T_f}{T_o - T_{cw}}} \tag{B.8}$$

The $(LMTD)_b$ is defined as:

$$(LMTD)_b = \frac{(T_b - T_f) - (T_o - T_{cw})}{\ln \dfrac{T_b - T_f}{T_o - T_{cw}}} \tag{B.9}$$

The total specific heat transfer area

$$
\begin{aligned}
sA &= \frac{A_e + A_d + A_b}{M_d} \\
&= \frac{\lambda_d + C_{p_v}\left(T_s - T_d\right)}{U_e\left(T_d - T_b\right)} \\
&\quad + \frac{C_p\left(T_d - T_o\right)}{U_d\left(LMTD\right)_d} \\
&\quad + \frac{\left(X_f/\left(X_b - X_f\right)\right)C_p\left(T_b - T_o\right)}{U_b\left(LMTD\right)_b}
\end{aligned}
\tag{B.10}
$$

Specific Power Consumption

The following sequence of equations is used to calculate the specific power consumption, Q_C,

$$Q_C = W\ r_d/3600 \tag{B.11}$$

$$W = H_S - H_V \tag{B.12}$$

$$\eta = \frac{W_m}{H_s - H_v} \tag{B.13}$$

$$\frac{W_m}{W_n} = \eta\left[\frac{\left(P_s/P_v\right)^{(\gamma-1/\gamma\eta)} - 1}{\left(P_s/P_v\right)^{(\gamma-1/\gamma)} - 1}\right] \tag{B.14}$$

$$\gamma = \frac{1}{1 - \left(1 + X\right)^2\left(ZR/Cp_v\right)/Y} \tag{B.15}$$

$$W_n = H_n - H_V \tag{B.16}$$

$$H_S = H_d + Cp_v\ (T_S - T_d) \tag{B.17}$$

APPENDIX C: SOLUTION PARAMETERS

Solution of the mathematical model requires specification of the following set of parameters:

- The distillate flow rate, M_d, is 1 kg/s.
- The intake seawater temperature, t_{cw}, is 30°C.
- The range for the condensed vapor temperature, T_d, is higher than T_b by 1 to 4 °C.
- The range fot the brine boiling temperature, T_b, is 50-110 °C.
- The feed seawater salinity, X_f, is 42000 ppm.
- The specific heat at constant pressure of the vapor, Cp_v, is 1.884 kJ/kg °C.
- The polytropic efficiency of the compressor, η, is 0.76 (ASHRAE, 1988).
- The plate width, w, in the preheaters is 0.5 m.
- The plate spacing, d, in the preheaters is 0.003 m.
- The plate thickness, δ, in the preheaters is 8.89×10^{-4} m.
- The thermal conductivity for the stainless steel plate preheaters is 0.0155 kW/m °C.
- The fouling resistance for seawater or brine in the preheaters is 8.805×10^{-3} m^2 °C/kW (Threlkeld, 1972).
- The fouling resistance for product water in the preheaters is 1.761×10^{-3} m^2 °C/kW (Threlkeld, 1972).

Rehabilitation of Multi-Stage Flashing Air Derence Jeddah Plant

Nassir Al-Tweergi

REHABILITATION OF MULTI-STAGE FLASHING AIR DEFENCE JEDDAH PLANT

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ABSTRACT

The Air Defense plant in Jeddah is Multi-Stage Flash plant long tubes with 20 stages (two units). The units are built as one block. The plant was commissioned in the year 1979 by Bosco company. The tube bundle of the first stage (448 tubes) was replaced in the year 1998 for one unit. Also general maintenance including replacement of the damage metal in the evaporator body was completed. After rehabilitation it was observed improvements in the production and the water quality. The paper describes the study of the corrosion in the different area of the brine side and the vapor side.

Keywords: Corrosion, Non condensable Gases, Plugged Tubes, and Air Leakage's

INTRODUCTION

The Air Defense plant is 20 years old and has suffered some corrosion. The Air Defense Command (ADC) MSF plant consists of two units of 300000 US gallons per day, each made by Bosco Co. Italy and both units have been in operation for 20 years.

The units are built as one block with 20 stages in 5 rows; each containing 4 stages. The tubing is aligned parallel to the brine flow and the tubes are the length of 4 stages (5m long). There is a separate deaerator in each unit and the dissolved oxygen level is nil but not continuously measured. The units are additives dosed for antiscalent, which is Belgard EVN.

The maximum brine temperature is 110 C. The venting to the steam ejecor occurs at stages 1 and 2, and from the deaerator. (see Fig. 1 for plant overview)

MATERIALS OF CONSTRUCTION

* All areas in contact with flashing brine - type 316SS.
* All areas in contact with vapor - C-steel.
* Distillate throughs - type 316SS.
* Water boxes - 90/10 CuNi.
* Heat recovery tubing and tube plates - 90/10 CuNi.
* Rejection section tubing - Ti.
* Rejection section tube plates - Al-bronze.

MATERIALS PERFORMANCE IN ADC PLANT

Stainless steel

The experience shows that corrosion is essentially zero when deaeration is good. The main risk of corrosion to stainless steel is when the plant is shutdown and air (oxygen) has access when brine is present. Under these conditions pitting can occur quite rapidly. However it should be noted that general corrosion does not occur, even under these conditions. Refer fig. 2, 3 & 4.

Table 1: Summary of the measured data on Stainless Steel in contact with flashing brine

LOCATION	RANGE OF DEPTH OF ATTACK (MM)	LOCATION OF MAXIMUM DEPTH OF ATTACK
Brine floor	0.4-1.4	Stage 2
Boundary wall	0.6-2.0	Stage 17
Distillate tray-bottom side	0-0.7	Stage 6,7,15
Distillate tray-vertical side	0-1.0	Stage 17

CARBON STEEL

C-steel is frequently used in the vapor space, particularly in the plant of this age. The corrosive environment is carbon dioxide although this may also contain oxygen due to air in leakage in the vacuum stages. With good venting to remove these inconensible gases corrosion rate in carbon steel can be quite low. However if these gases can accumulate in poorly venting regions severe corrosion can occur, particularly if oxygen in leakage has occurred. Refer fig. 5

Table 2: Summary of the measured data on C-steel in vapor space

LOCATION	RANGE OF DEPTH OF ATTACK (MM)	LOCATION OF MAXIMUM DEPTH OF ATTACK
Evaporator roof	0.2-10.5	Stage 15
Evaporator roof (new sheet)	12.7 (in 9 months)	Stage 7
Internal intrerstage walls in the same row	0.2-1.7	Stage 2
Internal walls between rows	0-2.1	Stage 3

The corrosion rates measured on the original C-steel structure are in the main quite acceptable. The very high rates experienced on the new material on the roof in stages 6 and 7 because of unusual level of incondensible gases such as CO_2 and Oxygen must have been present.

90/10 CuNi HEAT EXCHANGER TUBING

90/10 CuNi has high resistance to corrosion in dearated brine numerous tube failures on the brine side. The practical step is replacement of the plugged tubes.

Ti HEAT EXCHANGER TUBING

This tubing in the rejection section where no corrosion problems have been reported:

90/10 Cu Ni AND AL BONZE TUBE PLATES

No corrosion problems have been reported on the tube plates of this materials.

90/10 CuNi WATER BOXES

No corrosion has been reported on these components.

DISCUSSION

The main areas of corrosion concern in this unit are:

a) Corrosion of stainless steel in the brine in the brine space (pitting in splash gate).
b) Corrosion of the C-steel in the vapor space and interface with chamber.
c) Failures of 90/10 CuNi tubes in stages 1-4

CONCLUSION

The twenty years of operating experience on these MSF units manifest couple of points. These are (i) the right choice of materials of construction would much justify their initial cost. This is clearly evident from the prolonged problem free operation as a result of selectiong 316 stainless steel for flashing chambers. On the other hand (ii) the modest performance of carbon steel vapor space and that of 90/10 Cu-Ni tubes in high temperature bundles could have been avoided.

RECOMMENDATION

Based on the prolonged experience in these two MSF distillers, it is recommended that extreme care taken in material selection. Most appropriate choices are found to pay back quite well. The choices for the most important parts of an MSF distiller are recommended to be:

(1) Stainless steel lining/solid for flash chambers, vapor space and other internal parts. To avoid extravagance, these could be as follows:
   (a) SS 317L lining for floors,
   (b) SS 316L lining for walls,
   (c) SS 304L lining for roofs,
   (d) Solid SS 317 for demisters &
   (e) Solid SS 316L or SS 317L for distillate throughs and trays also for interstages walls and stage to stage fluid transfer assembly.

(2) Modified 66-30-2-2 Cu-Ni -Fe -Mn alloy for high temperature stages and the brine heater tubes.

(3) Other heat transfer tubes are to be as described in these two MSF units, i.e., 90/10 Cu-Ni in the rest of heat recovery and titanium in the heat rejection section.

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[3] AL HOTY STANGER "Inspection report of Air Defence Command MSF plant "- Dec. 1995

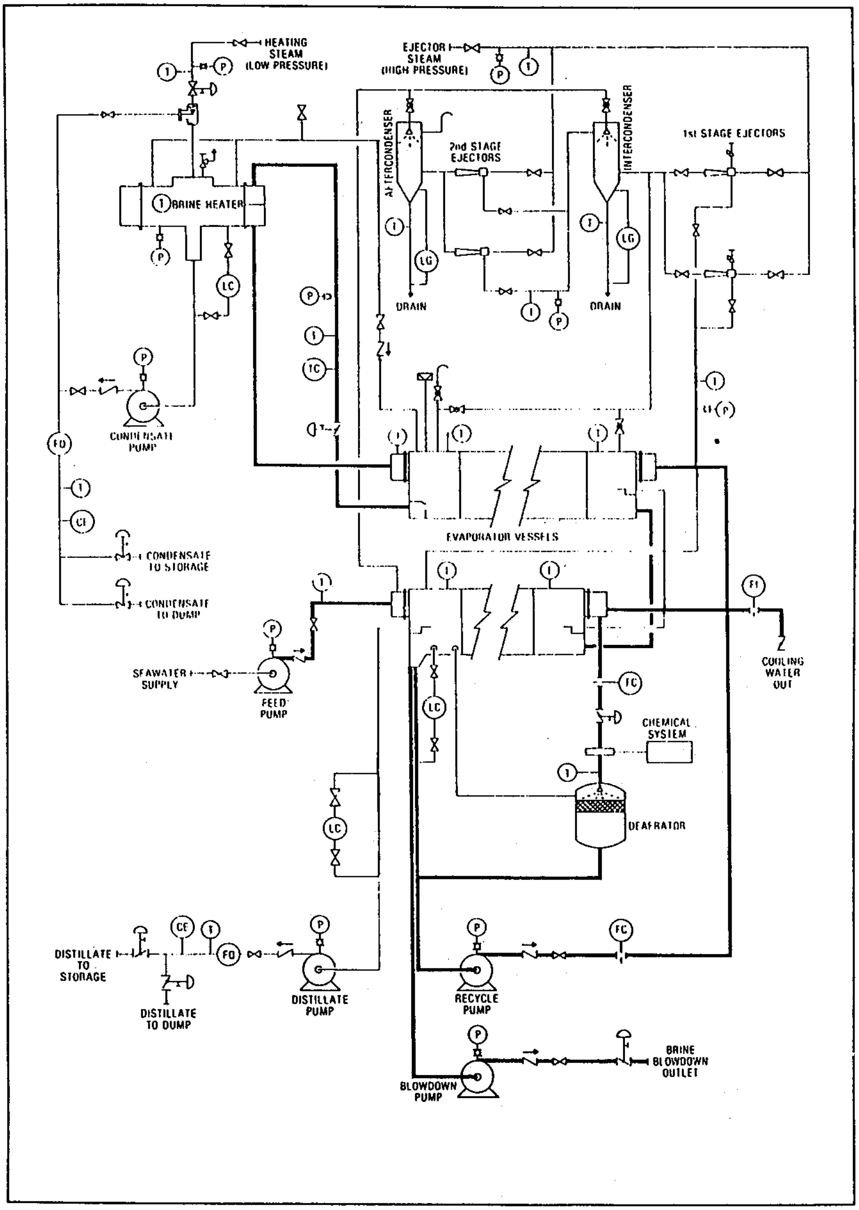

Figure 1: Flow Diagram Brine Recirculation

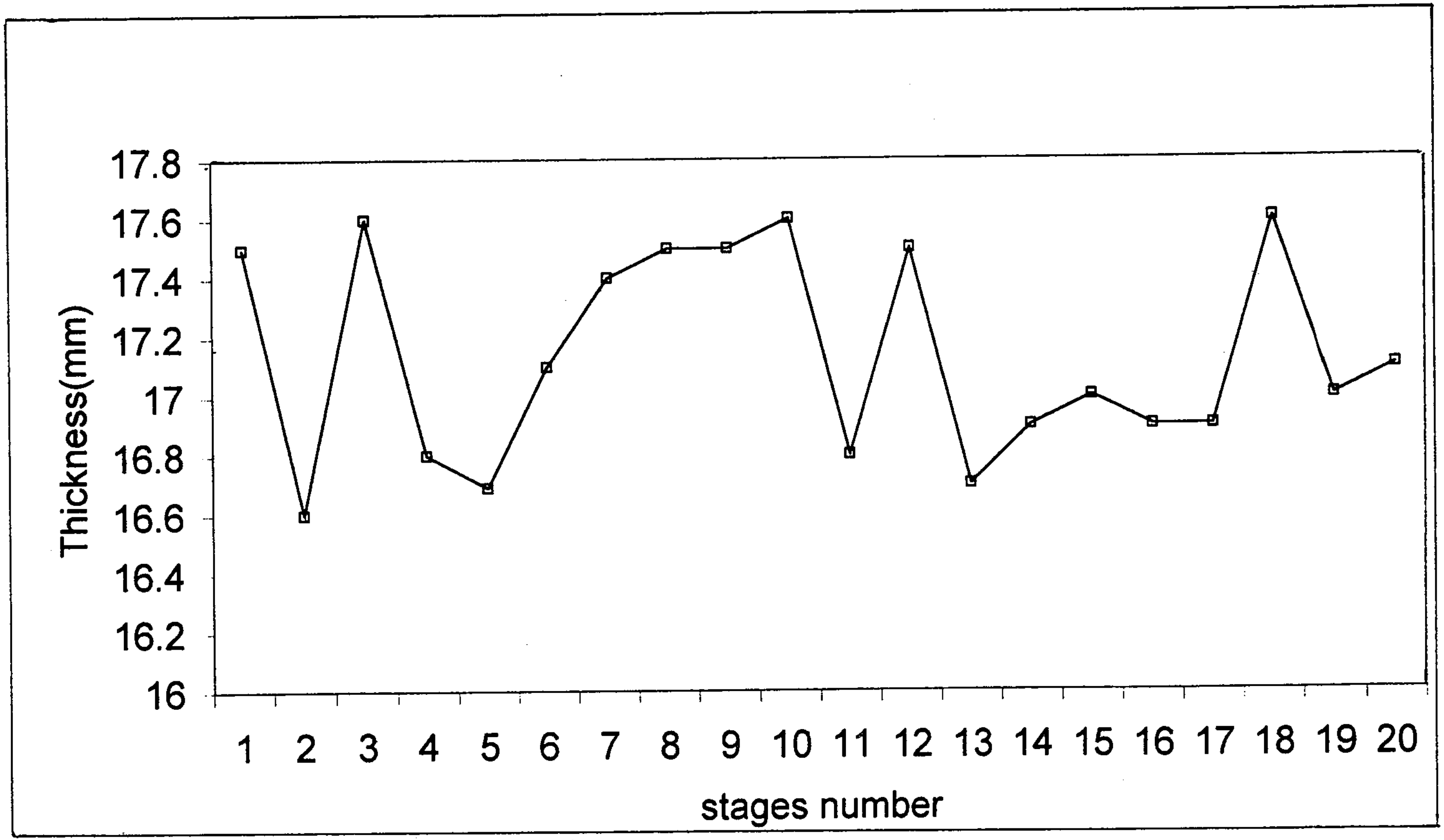

Figure 2. Thickness of brine floor

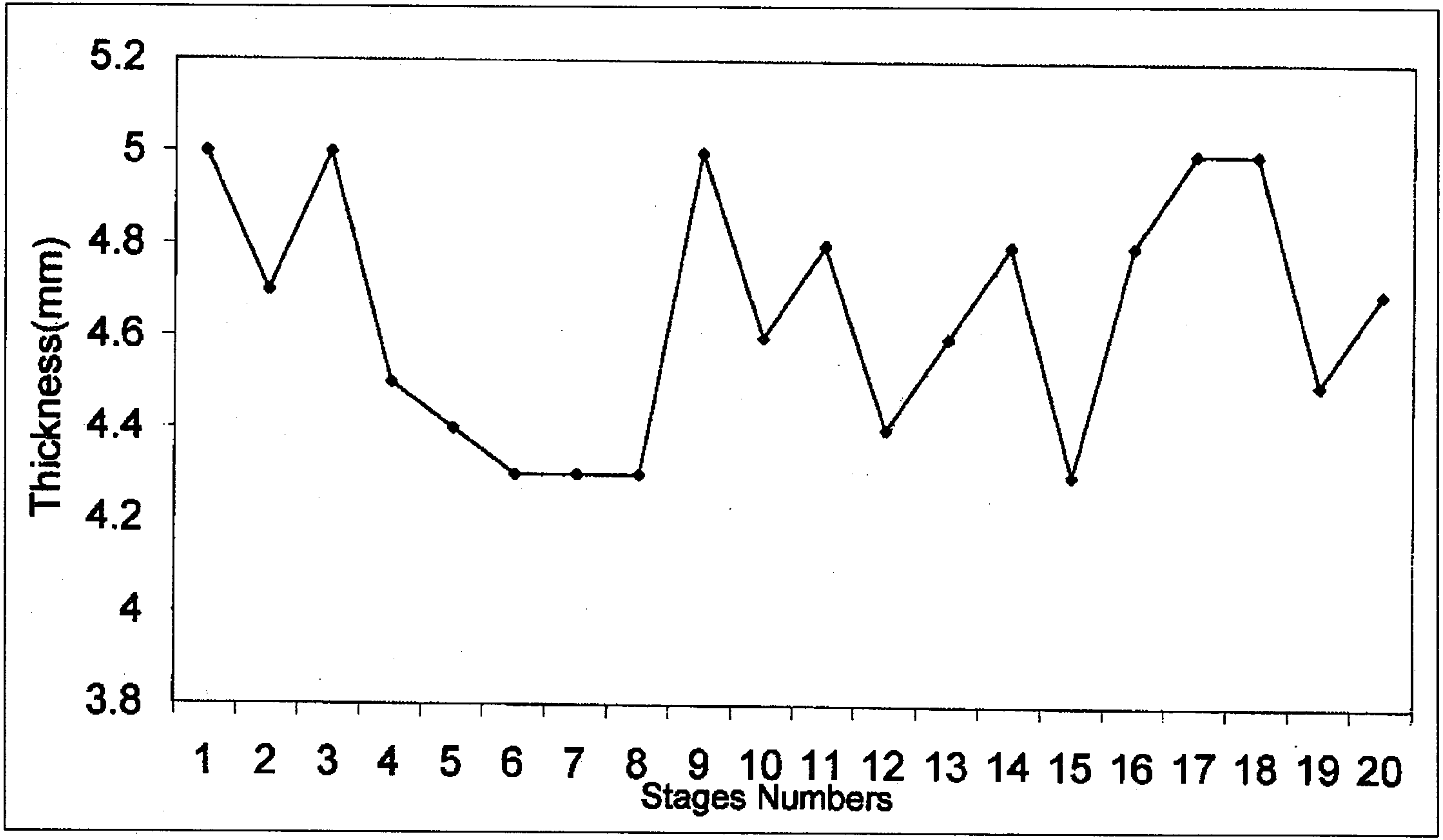

Figure 3. Evaporator Distillate Tray (bottom side)

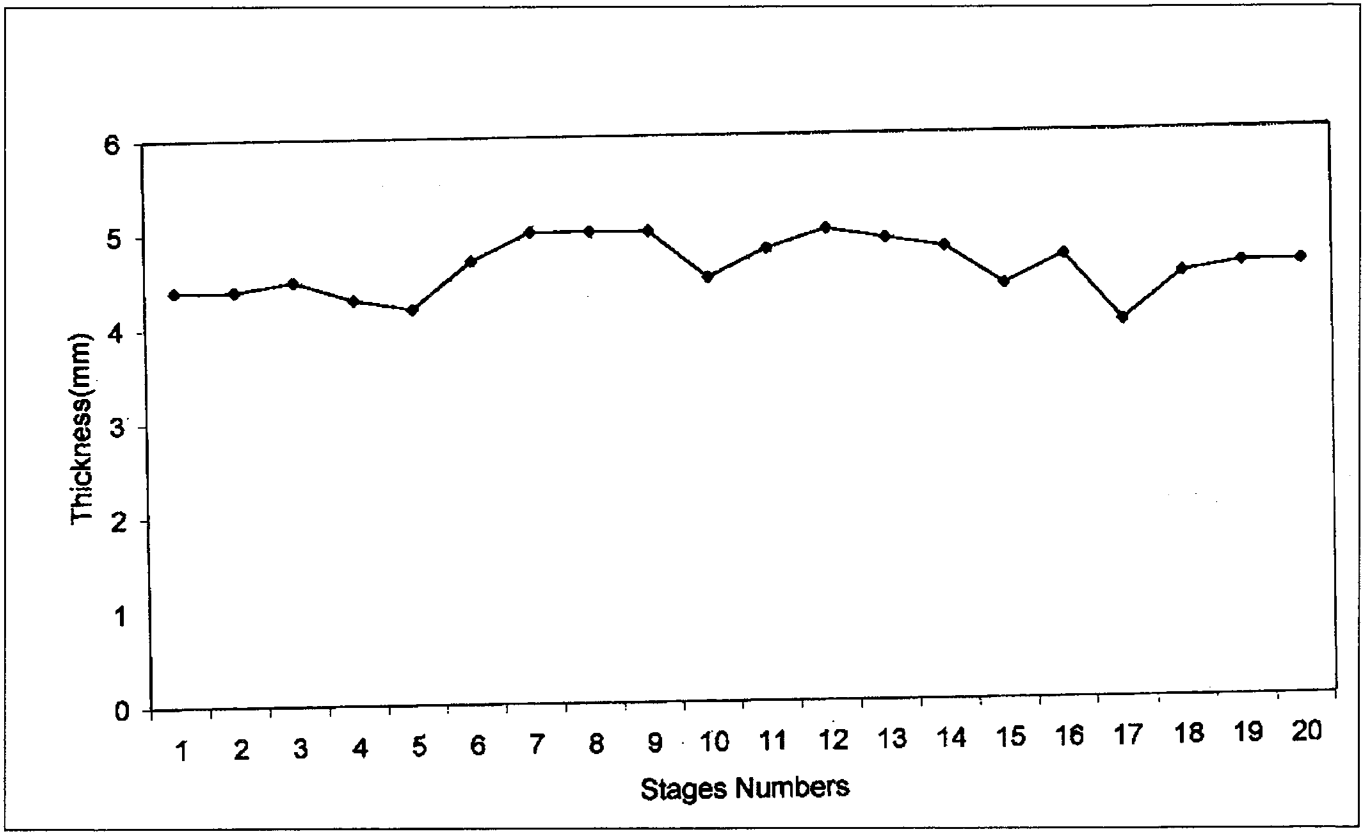

Figure 4. Evaporator Distillate Tray (vertical side)

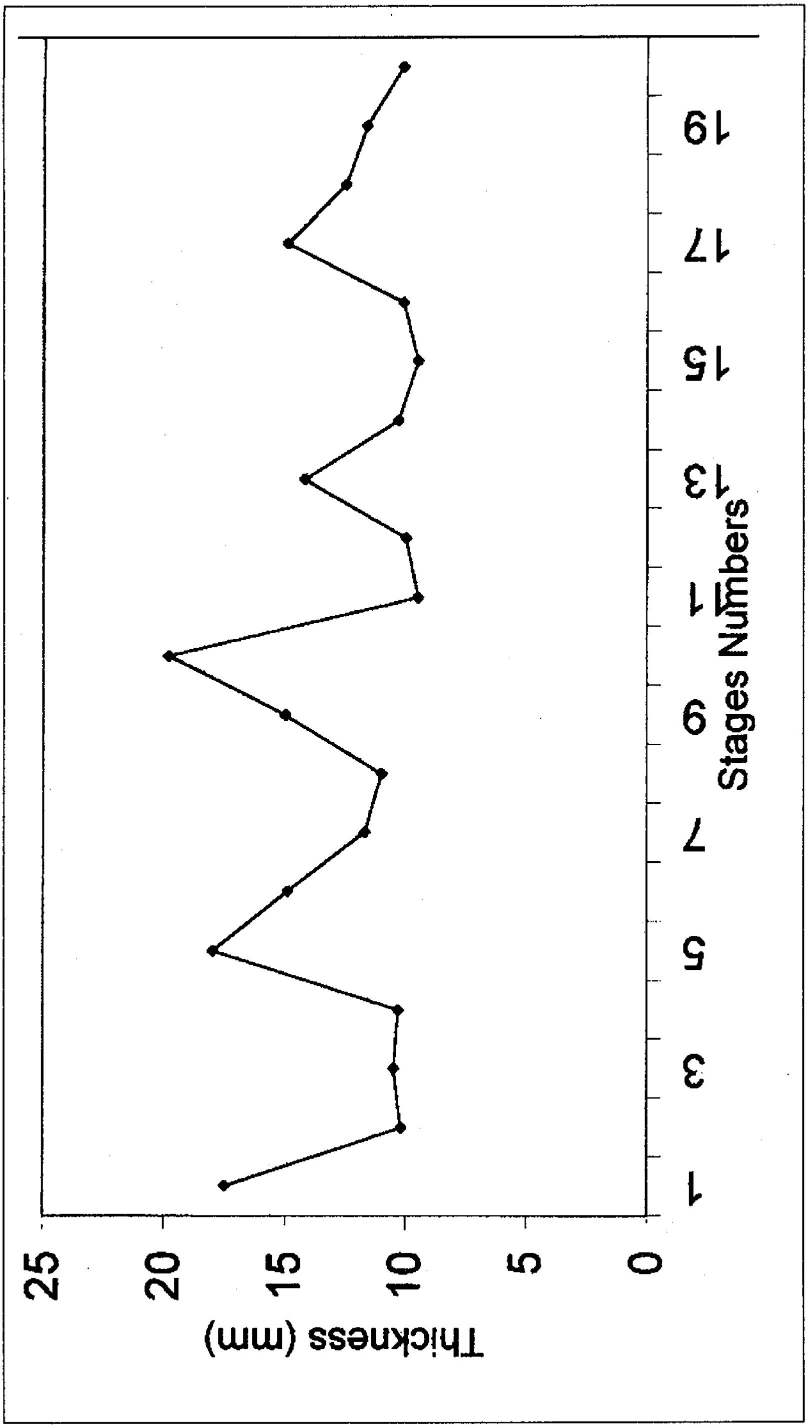

Figure 5. Evaporator Roof Thickness

Investigating Intake System Effectiveness with Emphasis on Self-Jetting Well-Point (SJWP) Beachwell System

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Ellen Abart and Robert Lovo

INVESTIGATING INTAKE SYSTEM EFFECTIVENESS WITH EMPHASIS ON SELF-JETTING WELL-POINT (SJWP) BEACHWELL SYSTEM

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ABSTRACT

A well-designed subsurface SWRO intake not only yields good quality feed that requires none or minimum of pretreatment, but also when used instead of surface type intake results in a significant saving in plant capital investment and in cost of plant O&M. An SJWP subsurface intake system was tested at SWCC Al-Jubail site, onshore, at high-tide seawater line and mid-tide point at depths of 3.5 m, 3.5 m, and 4.5 m, respectively. The project objective is to investigate the suitability of this relatively new subsurface SJWP as an intake for some SWRO plants. The best SDI values with an average of $SDI_{ave} = 2.9$, which compares reasonably well with $SDI_{ave} = 2.6 \pm 0.2$ for conventionally pretreated surface seawater feed, were obtained when the well-points were fixed at sea mid-tide point at depth of 4.5 m. The filtrate TDS, total hardness and pH values were lesser than those of seawater indicating mixing of seawater with underground or surface water. At zero incubation time, bacteria count was in the range of $5x10^2$ to $4x10^3$ colony forming units (CFU), but high aftergrowth rates and short bacteria generation time were observed in the SJWP feed after 24 and 72 hours of incubation. Based on SDI measurements, as well as on other physical and chemical measurements, a good quality SWRO feed can be derived from subsurface well-point beachwells located on a clean non-polluted site, which is expected to diminish the high aftergrowth rate and short generation time of bacteria found in the present case. Accurate information on this point can be established by testing the effect on SWRO membrane performance of feed derived from SJWP beachwell located on a clean non-polluted site which is to be conducted under Part 2 of this program.

1. INTRODUCTION

To avoid membrane fouling, which leads to degradation in SWRO plant performance, feed to SWRO plants is pretreated to remove from it all potential membrane foulants, ideally leaving only the dissolved solids in the feed. To a large extent, the success of SWRO plant operation is dependent on the feed quality and the feed pretreatment. An ideal pretreatment is designed to cause a minimum or no membrane fouling at the lowest possible cost. This not only results in longer membrane life but also improves the overall plant efficiency and reduces the cost of fresh water production. The degree and type of pretreatment required are dependent on the raw seawater quality, in particular its content of suspended solids, debris, biological and organic matters, pollutants, residual heavy metals, etc. However, the water quality, is very much dependent on type of **intake** used in SWRO plants.

The design and selection of seawater intake system are very crucial not only to the optimization of both the pretreatment, and the overall plant performance but also to lowering of cost of water production. A fault in intake design and/or intake site selection could make it difficult, if not impossible, for the pretreatment and the SWRO plant as a whole to function properly. On the other hand, a proper intake system that yields high quality feed requires minimum pretreatment. A good example, **a well-designed seawell or beachwell** type intake produces high quality water that requires no further pretreatment. The natural filtration of feed through the porous subsoil strata cleans the feed from nearly all contaminants normally present in feed taken from an open sea. In addition to low oxygen content, the feed is characterized by its constant physical and chemical properties; also unlike surface feed it is unlikely to be affected by changes in weather or sea conditions. In this case, the seawater intake feed can be supplied directly to the membrane system (only with the addition of the necessary antiscalant), thus avoiding the coagulation filtration step, the addition of coagulant, coagulant-aid and disinfectant or other chemicals to the feed stream. Moreover, many of the SWRO plant problems associated with feed supplied from an open sea intake, some of which are caused by the added chemicals, can be eliminated by drawing the feed from sea subsurface regions. Most obvious example is the elimination of chlorine. Chlorine when present in the feed could react with organic present in the same feed to form the harmful THM compounds. It could also degrade organic, mainly humic acids, to form assemble organics which serve as nutrients to bacteria leading to their fast multiplication and thus membrane biofouling. Moreover, many of SWRO membranes are oxidized and, therefore, are degraded by chlorine. Other well-designed subsurface type intakes could work equally good as seawell and are likely to produce high quality feed.

By contrast to the minimum or no pretreatment required to treat feed taken from subsurface intakes, an extensive, complex pretreatment with array of equipment, chemicals and chemical dosing systems is required to clean the surface feed from all the contaminants and membrane potential foulants. The complexity in pretreatment is dependent on type, nature and site of intake as well as on quality of feed required by the different membrane system, i.e., hollow fine fiber, spiral wound or plate and frame. All this is in addition to a higher level of dissolved oxygen which leads to the higher growth and multiplication of bacteria and other microorganism attached to the membrane as well as it may lead to a slow membrane oxidation. Moreover, it is well-known that both the expected inherent decline in membrane flux and salt passage are significantly greater when the feed is taken from an open sea than that derived from seawell.

In this paper results obtained from the application of a relatively new subsurface self-jetting well-point intake system are discussed. This work which is done at SWCC RDC, Jubail, is part of a joint R&D program on SWRO between SWCC and USBR.

2. SWRO INTAKE SYSTEM

The various intake systems that are now in use or may be utilized in the future to supply seawater feed to SWRO plants either from an open sea or from subsurface shore sites can be classified into two major intake systems: (a) Surface Intake consisting of: Pipe systems or Open Channel, Canal Lagoon, (b) Subsurface which can be classified into seawell (beachwell), Well-point, seawater gallery with gravel packing, seabed filtration with filter media packing, and Ranney Collector system.

The submerged pipe tends to be the dominant choice when the SWRO plant feed is taken from an open sea. This system is utilized in all SWCC SWRO plants located on the Red Sea shores with capacity (including SWRO plants under construction) of 87 mgd. Depending on coastal line geology and topography, the intake depth varies from one site location to another for SWCC plants on Red Sea shores the intake depth varies from 4 to 19 m [1,2,3]. This arrangement which places the SWRO intake deep in the sea and away from the coastal line allows for the supply of the SWRO plant with a relatively better quality feed than that obtained from an open channel intake. The open channel intake is rarely used to supply feed to SWRO plants. This type of intake is more suited to deliver feed to MSF plants where, unlike SWRO plants, no extensive feed pretreatment is required.

Seawells are the most used intakes among the subsurface systems. It is employed in many parts of the world to supply feed to a relatively large number of SWRO plants. All the SWRO plants in Malta, capacity 31 mgd, utilize beachwell intake at well depth of about 50 to 60 m [4]. Many of the SWRO plants in Canary Islands [5] as well as in UAE [6,7] and Caribbean Islands are supplied seawater feed from beachwells. Other examples of beachwells and wells with high salinity water TDS > 12000 ppm are described in literature [8,9,10]. Another form of seawell is the seawell infiltration system which bas been reported only for two cases at the Island of Lanzarote, Canary Islands [13]. Seawells or beachwells are drilled onshore at varying depth below the sea floor. The design and construction of seawells are similar to those of artesian wells of same dimension. Central part of the well double casing is perforated with large number of slots of about 0.3 to 0.5 mm for the interior casing as compared to about 0.7 mm for the exterior one. The depth of the well is determined by the soil geology, soil permeability and the quality of water desired from the well. Water quality tends to vary from one subsoil stratum to another. Seawell water turbidity and SDI are about 1 NTU and less than 2, respectively, as compared to a much higher values for surface feed turbidity of up to 20 NTU and SDI values of up to 6.7 or above. The high quality feed expected from seawell and other subsurface intakes constitutes their major advantage over other surface type intakes. This situation not only eliminates the pretreatment process but also leads, as mentioned earlier to lowering in plant capital cost by up to 25% and plant operation cost by up to 15% [12]. The improvement in the quality of seawater derived from wells has also a great influence on plant availability which could exceed 95% and product water recovery ratio reaching 45% [5].

Although feed with constant chemical composition and physical properties is derived from seawells its composition is not necessarily the same as that of seawater. The composition of feed derived from a well or subsurface intake is always influenced by the composition of the subsoil strata from which the seawater flows into the well. Thus, the selection of both site and depth of subsurface intakes are to be thoroughly investigated in advance of well construction by drilling of bore holes at various locations to establish the site suitability for subsurface intake construction and to make sure that the intake can supply high quality feed to SWRO plants at constant feed flow without water depletion.

Directional drilling now allows for drilling of seawells in a horizontal position which has the advantage of increasing the flow from such seawells. A new porous polyethylene well pipe that is now available since 1995 is reported to require no additional external media packing for long-term operation as that required with conventional seawells. The polyethylene pipe porous structure acts as both well screen and packing media. This product is claimed to make the horizontal drilling of wells under the seafloor feasible and economical for the first time [13].

The subsurface intake systems of: seawater gallery with gravel packing, seabed filtration with filter media packing and the Ranney Collectors have one feature in common in that in these cases feed water is collected through a number of horizontally laid PVC screen pipes at certain depth in the subsoil strata. Seabed and gallery filtration systems are similar in design and construction except for their packing system. Instead of crushed stone packing used in the seawater gallery, filter media packing material is used in the seabed filtration system. This way the seabed filtration system functions as both seawater gallery and sand filter giving it the advantage for use in SWRO intakes over the seawater gallery intake system. The three subsurface intake systems of seawater gallery, seabed filtration and Ranney Collectors appear to be promising, and good candidates for SWRO plant intakes with the advantage of possibly supplying greater quantity of feed than that taken from a single seawell, nevertheless, they have not been reported in literature as a feed source to SWRO plants. They are discussed in literature only as a possibility for SWRO intakes. For more information on those subsurface intakes see reference [14].

The well-point beachwell system tested here as a SWRO intake is shown in Figure 1. The well-point itself is similar in design to the well-point system used in ground dewatering, mainly in the coastal areas where water table is high (Figure 2). However, this system offers a finer degree of water filtration than that delivered by the well-points ground dewatering. The well-points are constructed in a fashion similar to that of seawell construction although the well-points are much smaller in diameter. Central part of the double casing is perforated with large number of very fine slots (less than 0.5 mm) in the interior casing. Outer casing consists of a screen normally with media packing in-between the two screens.

The well-point intake system combines features from both the beachwell and the other subsurface intake systems. Like the beachwell intake, the well-points are vertically jetted but at several points on or off shore at shallower depth than that of seawells. Like the subsurface intake of seawater gallery, seabed filtration and Ranney Collector, the well-point system collects the feed from many screen pipes which are vertically laid in the subsoil as opposed to their horizontal burial in the latter three systems. The well-point system, however, differs from all other subsurfaces intake systems in that it can be easily self-jetted at site without major construction as required by other subsurface intake systems. The system, as the name indicates, is self-jetting, the one and same pump used to pump feed to SWRO plant is also used to jet the well-points to the desired underground depth. Unlike other intake systems, which once constructed remain fixed to the site, the well-points can be easily moved for use at other sites.

The well-point intake system has been used by the US Army as a quick SWRO intake system for providing water for mobile SWRO plants in remote areas. The system has been in continuous operation in Port Hueneme, California and San Nicholas Island, USA, since 1993 [13, 15].

3. EXPERIMENTAL

3.1 Site Selection and Bore Holes Drilling

Because of its proximity to SWCC R&D Center, a coastal area about 100m north of Al-Jubail MSF plant intake, was selected as project site. A local company [16] was contracted to drill three onshore test bore holes at about 30m away from the sealine, 30m apart and each at depth of about 10 to 11m. Subsoil samples were collected from the surface (i.e., depth of zero) also from 3, 6 and 9m below ground level. The samples were analyzed by the contractor at his laboratories.

3.2 The Self-Jetting Well-Point Beachwell Intake

The SJWP system used in this investigation (Figure 1) consists of two well-points of 1.5" diameter and 3' in length, followed by three 316 stainless steel risers, each of 1.5" in diameter and 3' in length. The risers are connected to a common header of PVC pipe terminating at the pump. A control valve controls the water flowing through the system.

The beachwells were constructed at the selected site by using the same pump in jetting the well-points slowly through the subsoil to the desired depth, 3-5m, without encountering any difficulties. The well-points were slowly driven in the porous ground by pumping the water at low pressure through its central pipe. The water flowing out (from inside the well-point) fluidize the porous soil around it and allows for the drilling of a shallow beachwell to the desired depth below the sea floor. Based on core sample analytical results, (see section 4.1) it was possible to drive the well-point to a depth of 4.5m without difficulty. Once the well was drilled the system was put into service (using the same pump in reverse). This way, naturally filtered water through the sand layers starts to flow into the beachwell and well-points from which it was continuously pumped. To allow for good formation of the beachwells 15 to 20 minutes were required per jetting of each well-point. Trial tests were made at three location at the selected site where test bore holes were made: onshore at well-point depth of 3.5m, at high-tide seawater line at depth of 3.5m and at mid-tide seawater line at well-point depth of 4.5 m.

3.3 Water Quality Analyses

Various chemical, physical and biological analyses were made according to standard procedures for various samples collected at each of the three site locations. Main emphasis was on the measurement of beachwell filtrate: temperature, pH, conductivity and SDI. Complete chemical and biological analyses were carried out on selected samples. Analyses of residue collected on SDI filter paper were also performed by the scanning electron microscope (SEM) and the energy dispersive X-ray (EDX) method. Prior to bacteria count in CFU/ml each of the samples was incubated at 30°C for different periods of 0, 24, 72 and 96 hours, after culturing on marine agar by the pour plate method for 96 hours, at 30°C.

4. RESULTS AND DISCUSSION

4.1 Soil Structure Analysis

Results of soil analysis done by the contractor at his laboratories showed that the upper subsoil layer at a depth of about 7m consists of sand with silt and in some cases extends to a depth of 9m [16]. Below the sand-silt layer, there was a layer of a very weak lime stone moderately weathered and closely fractured.

4.2 Water Quality Criteria. for Well-point Beachwells

When the saline water is intended for use as feed to a SWRO plant it must meet certain quality requirements. The feed with the highest possible quality standard is a seawater that does not contain any impurities e.g., bacteria and other microorganisms and suspended solids including colloidal matter, organic carbon, that may cause harm to the membrane or other SWRO process components. For certain membranes the case of no chlorine or no oxygen in the feed is also a rigid requirement. Use of the disinfectant and coagulant chemicals which are essential to the pretreatment of surface seawater feed may have an adverse, damaging side effect on the membrane. Dissolved oxygen when present in the feed is also a process variable. It may lead to a slow but progressive damage to the membrane in addition to its influence on speeding up bacteria growth and multiplication which may give rise to membrane biofouling. The ideal feed case is rarely realized in practice, especially if the feed is derived from surface type intake, although quality of feed from a well-designed seawell or subsurface intakes or membrane treated feed could approach this idealized situation. In this investigation the quality of feed derived from the well-point beachwell at Al-Jubail site was evaluated as a possible feed for a SWRO plant mainly by measuring the feed SDI, pH, TDS, biocount and measurements of other relevant chemical and physical parameters.

Tables 1 and 2 list, respectively, the chemical and biological analyses parameters measured for samples collected at the same site. For reasons of comparison the chemical composition of surface seawater is also included in Table 1. Depths are measured below ground level at the location. The SDI values for the filtrate collected at different sites are plotted versus operation time in Figure 3.

4.2. 1 Well-Point Filtrate Chemical and Physical Properties

The best SDI values with an average SDI_{ave} of 2.9 were obtained when the well-points were fixed at mid-tide seawater site at a depth of 4.5 m. This value compares reasonably well with the average SDI_{ave} values of about 2.6±0.2 obtained from SDI measurements made on conventionally pretreated feed derived from an open sea at Al-Jubail. The pretreatment consisted of coagulation with $FeCl_3$ and dual media filtration [17]. Filtrate water conductivity ranged between 18,880 to 26,800 µs/cm at the onshore site and rose to over 40,000 to 49,000 µs/cm at the other two site locations at the hightide and mid-tide water lines. In all cases the conductivity of filtrate from the well-point beachwells is lower than that of surface seawater of about 60,000 µs/cm. Total hardness, is also less in beachwell water than that in seawater. Variation in pH with site location was noticed and in all cases it ranged between 7.1 to 7.9 and was less than the seawater pH of 8.2. Total organic carbon (TOC) of 1.2 to 2 ppm is of the same order, as in seawater of TOC ≤ 2 ppm. Temperature of filtrate collected from the well-point beachwell also differs from one site location to another. The lower pH, hardness and conductivity values of filtrate samples than those of seawater indicate seepage of surface water or the mixing of ground water with seawater at these three site locations, with a greater mixing ratio of surface water to seawater at the onshore site. This also tends to be true for seawell water from Kuwait beachwell drilled to a depth of 30m where the water TDS and pH are about 38,850 ppm and 7.7, respectively [18].

The H_2S water smell, which was noticeable when the water samples were taken from onshore site, disappeared when the well-points site was shifted toward the sea and was fixed at lower subsoil level of about 1m and 2m at the high-tide line site and mid-tide site, respectively. The presence of H_2S smell in the well-point beachwell filtrate indicates the presence of unaerobic sulfur reducing bacteria in the water at onshore underground at depth of 3.5m. It also indicates lack of sufficient oxygen in that water.

As was established from EDX spectrum the main residues collected on the SDI filter paper are dried seawater salts and ions, mainly Na^+, Cl^-, Mg^{++}, Ca^{++}, S^+ and K^+, most of which disappeared from the EDX spectrum upon washing gently the filter paper with deionized water. From the EDX analysis it appears that no major residue or total suspended solids (TSS) matter are present in the water derived from well-point beachwell at depth of 4.5m below mid-tide sea floor. This, however, does not rule out the presence in the well-point beachwell filtrate of soluble elements such as humic acid and other soluble organics which serve as nutrients to bacteria.

4.2.2 Well-Point Filtrate - Biological Properties

At zero hour of incubation, bacterial count in CFU/ml in filtrate from well-point beachwell and surface seawater are similar in number. At the zero hour, the bacteria count was about $1.8x10^3$ CFU/ml to $4.04x10^3$ CFU/ml in surface seawater and about $5.0x10^2$ CFU/ml to $3.79x10^3$ CFU/ml in subsurface water (Table 2). (Strangely enough and unexpectedly, these numbers compare favorably well with $1.2x10^3$ to $1.3x10^3$ T. count (CFU/ml) in water derived from two beachwells drilled at depth of 30m near the beach in Doha, Kuwait [18]! However, they (bacteria) are not necessarily the same type or possess the same properties, mainly in their ability to attach themselves to membrane surfaces. At 24 and 72 hour aftergrowth, however, the bacteria aftergrowth were faster in filtrate from beachwell than their aftergrowth in surface seawater. This is evidenced by the time in hours required for the bacteria generation (gt) to double in number, Table 2.

In the case of subsurface water where samples were collected from onshore beachwell it is noticed that only 0.207 hour is required for the bacteria to double its population after 72 hours of incubation as compared to more than twice this time of 0.484 hours when the sample is collected from surface seawater and 0.226 hours for subsurface water samples collected from beachwells located at the mid-tide point. The latter value increased to 0.235 hours after 96 hours of incubation indicating the possibility of decreasing nutrients in the sample, as shown in Table 2. This situation, indicates that bacteria aftergrowth is proceeding at a faster rate in subsurface water obtain from this site than in open seawater. Moreover, it indicates the availability of sufficient nutrients in the present subsurface water which partially originates from the accumulated organic seawater residues, e.g., humic acid and is partially due to the residues present in the surface water seeping from the surrounding areas. The data suggest the possibility of a polluted site, which was established to be the case after through survey of the site. Surface and some waste water was seeping to the site from neighboring grounds. A better quality feed with SDI ≤ 2, containing low bacterial count and presumably having low level of nutrients can be obtained by the selection of a non-polluted site.

5. CONCLUSION

From the above discussion it can be concluded that based on SDI, as well as on other physical, chemical and biological measurements (Tables 1 and 2, also Figure 3) a good SWRO quality water with low TSS content can be derived from subsurface well-point beachwell. The high aftergrowth rate and short generation time of bacteria found in the feed may be attributed to the site itself, which was found to be polluted by the seepage of surface and waste water. Changing the site to a clean non-polluted site could alleviate this effect and may result in good quality feed to SWRO plants and in consequence to good plant performance. To a certain degree, however, this should depend on type of bacteria and nutrients found in SJWP beachwell feed and the bacteria ability to condition and to attach themselves to membrane surface. These conditions are essential for the formation of biofilm [19]. Accurate information on this point can be established by testing the effect on SWRO plant performance of feed derived from SJWP beachwell located on a properly selected clean non-polluted site. This work is to be conducted soon.

6. RECOMMENDATION

From the trial tests made with this subsurface SJWP beachwell intake good quality water with an $SDI_{ave} = 2.9$ was obtained but high aftergrowth rates and short generation time of bacteria may be expected to give rise to a decline in SWRO plant performance. This latter factor can be corrected by the selection of a clean non-polluted site. Moreover, the experiment done so far was exploratory in nature and it is, therefore, recommended: (1) to conduct further experiments on a clean non-polluted site; (2) to work on optimization of the SJWP beachwell intake by improvement of yield and filtrate quality; (3) to determine design criteria for small to medium size SWRO plants; and (4) to perform process technoeconomic analysis. These recommendation were accepted by SWCC and a future evaluation of the system is under way.

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Table 1: Chemical Composition of Well-point Beach Well Water Samples From On-Shore, High-Tide and Mid-Tide Test Areas

Parameters		**Seawater**	**On-Shore**	**High-Tide**	**Mid-Tide**
Temperature °C			25	24	23.5
pH		8.2	7.5	7.62	7.3
Conductivity at 25 =°C	(μS/CM)	60000	1900	40100	41000
T.D.S. at 105 °C	(ppm)	46900		31289	31430
Total Hadness (as CaCO3)	(ppm)	6738		5250	5850
Calcium as Ca^{++}	(ppm)	519		480	480
Magnesium as Mg^{++}	(ppm)	1326		984	1130
Sodium as Na^{+}	(ppm)	12860		7888	9000
Potassium as K^{+}	(ppm)	450		344	392
Sulfate as $SO4^{-}$	(ppm)	3265		5400	5514
Chlorides as Cl^{-}	(ppm)	23000		13650	16661
TOC	(ppm)	2		1.2	2

Table 2: Bacterial Count (CFU/ml) and generation time in hours/double generation, (hrs/2g) in Well point water at two different locations: on-shore and at mid-tide seawater line

Sample	**0h**	**24 h**		**72 h**		**Locations**
	CFU/ml	**CFU/ml**	**hrs/2g**	**CFU/ml**	**hrs/2g**	
Raw seawater	$1.8x10^3$	$1.1x10^5$	0.404406	$5.6x10^4$	0.483831	Onshore
Well water	$1.3x10^3$	$3.3x10^5$	0.300394	$4.0x10^6$	0.20708	
Raw seawater	$4.04x10^3$					Mid-tide point
Well water	$3.79x10^3$					
Raw seawater						
Well water	$5.9x10^2$	$4.86x10^5$	0.247727	$9.3x10^5$	0.225892	Mid-tide point

Generation time = t x 0.693/ (Ln Nt - LnNo) : Where: t = time; Nt = Bacterial count at time t hrs; No Bacterial count at zero hrs

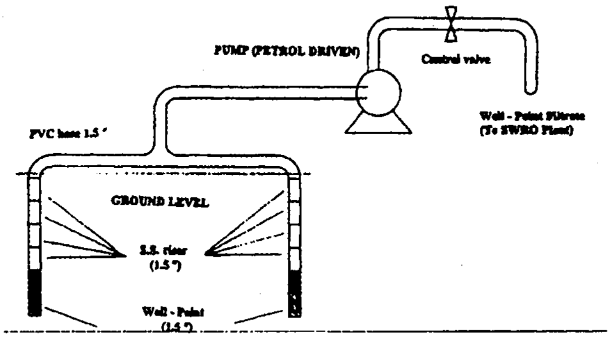
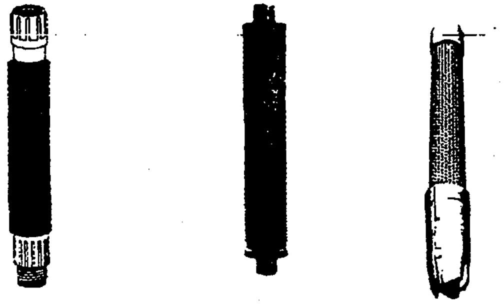


Fig. 1: Two Well-Point Beach Well Layout



Sand Packed
For SWRO Intake

Gravel Packed
For SWRO Intake

C- Dewatering
For Underground Waters

Fig. 2: Fine Water Filtering Well-Points

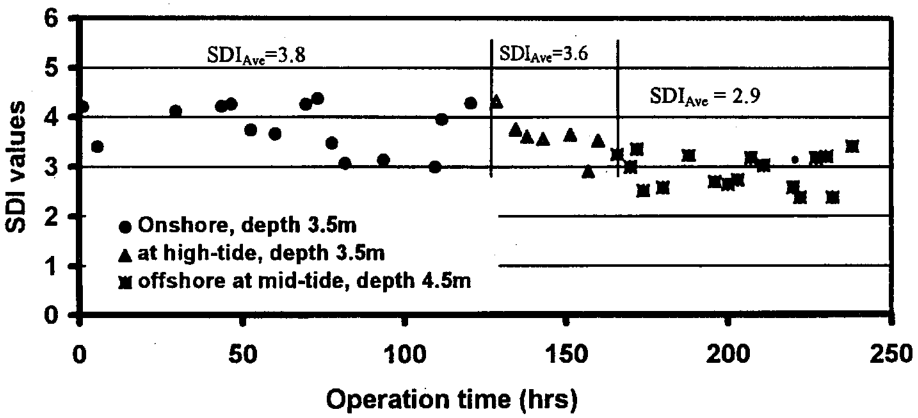


Fig. 3: Quality of water at different well-point depth as well as different locations onshore and offshore

Corrosion Behavior of Steels in Gulf Seawater Environment

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CORROSION BEHAVIOR OF STEELS IN GULF SEAWATER ENVIRONMENT

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ABSTRACT

Due to the aggressive environment prevailing in and around the Gulf seacoast, the materials of construction are subjected to corrosion of varying degree. There are many factors influencing the initiation of one or several corrosion processes. These factors include nature of material, surface finish, temperature, humidity, wind directions, rainfall, etc. Steels are the predominant constructional materials in seashore structures including desalination plants which are subjected to general or localized corrosion of varying degree. A complete systematic information regarding corrosivity of materials in Gulf seawater is lacking. Keeping in view the importance of the corrosion data of materials particularly steels in Gulf seawater, a systematic study on this subject had been initiated in 1990 and was completed in 1996. This paper provides the results of this study concerning the corrosion behavior of steels, in Gulf seawater. The steels include carbon steels, stainless steels and super stainless steels (austenitic, ferritic and duplex). The study encompasses the factors influencing general, pitting and crevice corrosion of materials. The effects of TDS in general and chloride concentration in particular and degree of chlorination have been discussed in detail. The role of dominant alloy addition on the corrosion behavior of steels has been emphasized.

Weight loss and electrochemical techniques have been employed in determining the corrosion rates of materials. Potentiodynamic polarisation technique has been used to evaluate parameters related to localized corrosion. Both exposure and accelerated tests have been carried out to study crevice corrosion attack. Effect of surface finish on crevice attack has been investigated in detail. Rank orders of corrosion rates of steels under different conditions have been presented and their importance concerning materials selection in desalination plants is discussed.

Keywords: Pitting, Crevice Corrosion, Stainless Steel, Breakdown Potential, Pitting Potential, PRE_N and Chlorination

INTRODUCTION

Steels are the most extensively used structural materials in industry. Mild steel is the most versatile general purpose material due to its good mechanical strength, easy fabricability, formability, weldability, abundance and low cost. In corrosive enrionments, mild steel structures can be protected by coating and / or cathodic protection. However, in more severe or aggressive environments such as marine or seawater, mild steel may not withstand and austentic stainless steels such as 304 and 316 have round applications as construction materials. The excellent corrosion resistance, good mechanical properties and reasonable cost are the contributing factors for their application as outstanding structural material. In seawater process industries which also include desalination plants, 316 type alloys have been formidable material for construction of evaporators, distiller pipes, pumps valves and condensers. These alloy are the most dependable structural materials under dynamic or flow conditions and virtually show no corrosion even on very long exposures. However, these alloys are subject to localized corrosion in presence of chloride ions and under static or stagnant conditions. The pitting of 316 L stainless steel is often encountered in desalination plant and appears to be the most effective factor responsible for corrosion failure in plants. The pits also provide active crevices for a formidable corrosion attack. In most cases, pitting is precursor of crevice corrosion and there are obvious similarities between crevice and pitting corrosion mechanisms. High chromium containing Ni-base alloys (Cr-Ni-Mo) were used as possible replacement for stainless steels to combat localized corrosion. However, the prohibitive cost of Ni-base alloys has severely restricted their application. Based on the fact that Cr and Mo has synergic effect in enhancing the pitting resistance of steel in seawater, 6% Mo containing stainless steels so called high alloy stainless steels were introduced in sixties and the following decades. The application of conventional and high alloy stainless steels in marine environment have been dealt in a number of review and articles [1-6].

Electrochemical techniques have been the principal tool for characterizing pitting and crevice corrosion. The pitting potential, Epit is a fundamental parameter in evaluating the susceptibility of different materials in different environment to localized corrosion [7]. Some authors [8,9] asserted that due to similarities between crevice and pitting corrosion mechanisms, it would be more appropriate to refer to localized corrosion by a new parameter called breakdown potential, Eb, which is defined as potential where pitting or crevice corrosion or both will initiate or propagate.

Temperature and velocity are the two important factors influencing the corrosion of SS superaustenitic and Ni- and Co- base alloys in seawater [8]. Electrochemical studies explicitly demonstrated the effect of increased temperature in facilitating premature breakdown of passive film on all materials. The effect of high velocity impinging flow was to further shift the passivity breakdown potential to active values but not necessarily to result in greater depth of attack. In general, increase in temperature caused a decrease in pitting potential and an increase in corrosion rate and pitting tendency [10, 11]. However, at a certain temperature, the decrease in Dissolved Oxygen (DO) level could result in reversal of the process.

The Cr, Mo and N contents of the alloy greatly influence their pitting and crevice corrosion behavior. The synergic effect of Cr and Mo in resisting pitting was first shown by Lorenz and Madwar [12] in 1969 and was further investigated by Herbsleb [13] in 1982 who also considered nitrogen. The pitting resistance equivalent, PRE_N is represented by the equation

$$PRE_N = \%\ Cr + 3.3\ \%\ Mo + 16\%\ N$$

A PRE_N above 38 is supposed to provide resistance to marine corrosion. Crevice corrosion of highly alloyed SS ($PRE_N > 40$) exposed to seawater, once initiated can propagate at temperature far lower than the initiation temperature. Repassivation properties of the material are thus important with respect to material selection and control.

This paper incorporates the results of various studies concerning with the corrosion behavior of conventional and high alloy stainless steels in Gulf seawater. A number of studies carried out in recent years signified the importance of PRE_N in assessing the ability of alloy to resist to pitting. Interesting yet useful relationship have been shown between PRE_N and several corrosion parameters such as pitting potential, Epit, pitting induction time, ti, corrosion rate (CR), etc., [13]. Alloy addition such as Mn appeared to negate the resistance to pitting [14].

Chlorination of seawater is the normal practice to destroy microorganism or microbiological growth, the residual chlorine influences significantly on the corrosion behavior of materials. The results of a number of studies concerning with pitting and crevice corrosion tendencies of conventional and high alloy steels in chlorinated seawater are reported in literature [15, 16]. High alloy stainless steels appear to have good resistance to residual chlorine. In such study [17] it has been shown that besides 254 SMO which did not corrode in seawater (25,000 to50,000 ppm Cl^- and 0-10 ppm residual chlorine) all other alloys developed crevice and suffered some degree of corrosion.

Critical crevice temperature and critical crevice index (CCI) have been used for measuring crevice corrosion tendencies of superaustenetic and duplex steels in chlorinated seawater [18] Critical Crevice Temperature (CCT) was found to be in the same range. No significant difference in CCT for metal-metal and metal-gasket crevices was found. Even the Ni-base alloys which were supposed to have formidable resistance against crevice attack in seawater have been found to be vulnerable. Creviced Ni-Cr-Mo-Fe alloys (alloy 625 and C 276) immersed in two different types of seawater were found to contain concentrated amounts of dissolved Ni^{2+}, Cr^{3+}, Mo^{3+} and Fe^{2+} in crevices showing that both the alloys undergone crevice corrosion [19].

EXPERIMENTAL

15 different commercial grade stainless steels, viz., 304L, 316L, 317L, 904L, 17-4PH, 254 SMO, 654SMO, HMO 1925, HMO 3127, 20Cb3, Remanit 4565, Remanit 4575, Monit 44635, Martensitic 410 and Duplex 2205 were used during the studies. Table 1 lists the composition of the alloys.

Aqueous chloride solutions of varying chloride concentration, seawater (Arabian Gulf seawater) and chlorinated seawater were used during the experiments. The composition of the seawater is given in Table 2. The chlorinated seawater was prepared by dozing with sodium hypochlorite and the residual chlorine was measured by HACH DREL/IC portable kit. For potentiodynamic polarisation studies, circular specimens of ~15 mm diameter and 3-5 mm thickness were machined from the sheet and were abraded sequentially on 180, 320 and 600 grit SiC papers. Rectangular coupons of the size 10 x 20 x 3-5 mm were cut from the sheet and surface furnished at 0. 12 μm rms were used for open circuit potential measurements.

Experimental details of crevice corrosion measurements and potentiodynamic polarisation and cyclic polarisation experiments carried out in this study are described in previous publications [2, 7, 10, 13].

Open Cricuit Corrosion Potential (OCP) of alloys was measured in normal and chlorinated seawater. The residual chlorine was monitored everyday. A special cell was designed and fabricated in the laboratory. Two multiples chart records of 6 and 3 pens were electrically connected to the specimens. The change in voltage against SCE as reference was plotted vs time. Every day the data were transferred into the computer and combined graph was drawn to study the change in potential for each of the alloys.

Accelerated Tests

Salt spray tests were carried out in Salt Spray fog Chamber following ASTM B17-73. Test coupons of 12 cm^2 were cut from the sheet and abroded to 180 grit SiC paper. Accelerated testing for resistance to crevice corrosion initiation was also carried out. The crevice corrosion assembly consisting of the specimen (rod or sheet form) attached with Teflon crevice former was mounted in epoxy former to provide a crevice free mount. The experimental details of salt spray and crevice corrosion accelerated tests are given in previous papers [7, 20].

RESULTS AND DISCUSSION

Accelerated Tests

The conventional (316 L and 904L) and high alloy stainless steels (SMO 254 and 654 hMo 1925 and 3127, 20 Cb3, Remanit 4565 and 4575, Monit 44635) were subject to salt-salt fog environment in a salt spray fog chamber for testing under accelerated conditions. Results of 3024 and 5000 hours exposure tests indicate very low weight losses corresponding to corrosion rates in the range of 0.001-0.005 mpy. This leads to a foregoing conclusion that conventional and high alloy stainless steels are resistant to general corrosion under severest environments.

A study of the potentiodynamic polarisation curves for the high alloy stainless steels in seawater indicate that little or no corrosion current was noticed on chloride additions showing thereby the outstanding resistance of these steels towards chloride. For getting measurable corrosion current, polarisation experiments involving most aggressive conditions were carried out. In these experiments, such conditions were produced by bringing down the pH to low values (0-3) plus addition of NACl solution at 50°C. Critical corrosion current value representing as active peak height (APH) is taken as the measure of critical crevice solution (CCS). The CCS of the alloy is taken as the value corresponding to APH of 10 $\mu A\ cm^{-2}$. Conventional alloys such as 304L and 316L and high alloy stainless steels such as 3127hMo and 654SMO have low CCS pH values of 0.481 and 0.50, respectively (Fig. 1). These results indicate excellent crevice corrosion resistance of high alloy stainless steels under aggressive conditions. Plots of critical crevice index (CCI = % Cr + 4.1% Mo + 27%N) and CCS pH for conventional and high alloy steels show a liner relationship (Fig. 2) indicating a strong but negative dependence of critical crevice solution aggressivity on Cr, Mo and N content.

Effect of Surface Finish

The surface finish has a considerable influence on the localized corrosion behavior of stainless materials in seawater. Under surface finish conditions of wheel ground, as received, 180 grit SiC ground, 600 grit SiC ground, sand blasting and HNO_3 etching, except 304 L and 316L, none of the alloys corroded to measurable depth of attack at 25°C thus indicating that crevices do not act as sites for corrosion initiation for high alloys stainless steels. Stainless steels 304L and 316L provide small yet measurable crevice corrosion depth. At 50°C, under aforementioned conditions, most of the high alloy steels either do not corrode or corrode surficially. The surficially corroded alloys showed very small crevice corrosion buildup. However, after crevice formers were removed and sample were cleaned, no perceptible depth of attack was found when examined microscopically. This has been attributed to high propagation resistance of the alloy due to the strong passivation of the protective passive film on the metal surface. Therefore, the crevice corrosion initiation of an alloy does not necessarily relate to its propagation rate. Solution immersion tests indicate that alloy SS 304 and 316L show mild attack at 25°C, the attack is more pronounced at 50°C, the former was attacked more severely.

Effect of Temperature

The temperature greatly influences the corrosion behavior of steels in seawater as has been indicated by the values of corrosion parameters from electrochemical studies. Pitting potential of the conventional and high alloy steels measured in the temperature range of 25-90°C indicate a shift to more active values up to the temperature of 65°C.

Above 65°C, Epit becomes constant. The constance in the Epit values has been attributed to the decrease in dissolved oxygen level due to increase in temperature. This factor plays more prominent role at higher temperature (> 65°C). A liner relationship (Arrhenuis plot) exists between logarithm of corrosion rate and reciprocal of absolute temperature (Fig. 3). The slop of line provides activation energy for the process in the temperature range of 25°C to 90°C. In general, with same exposure time, the severity of corrosion attack increases with increases in temperature.

Effect of Chloride Concentration

The effect of chloride concentration on corrosion rates of stainless at 50°C as studied by electrochemical liner polarisation resistance technique indicate that the corrosion rates do not follow a regular pattern but in majority of the alloys, the corrosion rate increase with increasing chloride content.

Whilst the corrosion rates of conventional stainless steels such as 304L, 316L and 317L in seawater are higher than in aqueous chloride solutions containing up to 5000 ppm chloride, the high alloy steels in general appear to have slightly lower corrosion rates in seawater (Fig. 4). The lower corrosion rates in seawater have been attributed to the formation of a stable biofilm which acts as a barrier against corrosion.

The pitting potential, Epit which is a measure of pitting tendency of a material under specific environment show a significant active shift with increasing chloride contents. Due to the formation of a strong and more resistance passive oxide film, the high alloy steels have much higher pitting potential than conventional stainless steels. This is further manifested by the strong dependence of Epit on PRE_N, the pitting resistance equivalent. Alloys having greater PRE_N have lower tendency to pitting, requiring much higher potential (Epit) to initiate pitting in the alloys. Two separate liner relationships exist for conventional and high alloy stainless steels.

Effect of Chlorination

Results from potentiodynamic cyclic polarisation studies carried out on conventional and high alloys stainless steels under chlorined and normal conditions indicate that the breakdown potential, Eb is lowered on chlorination (Table 3). The lower Eb values indicate more chances of breakdown of passive film easily and early start of pitting and crevice corrosion. Low decrease in E_b was observed for 3127 hMO, 1925 hMo, Remanit 4565 and 654 SMO. Therefore, these alloys have less chances of crevice and pitting corrosion in seawater and chlorinated seawater. SS 316L and 317 at 25°C can suffer from general as well as crevice and pitting corrosion due to low E_b and high I_{max} values. Plots of E_b vs PRE_N for high alloy SS show that the alloys having PRE_N values of 30 or above are quite resistant to chlorinated seawater (Fig. 5).

The open circuit potential of the alloys increases with increase in residual chlorine at 25°C. Whilst E_b values of superstainless steels at different levels of residual chlorine are much higher than conventional stainless steels, the open circuit potential values of superstainless steels markedly lower than the conventional stainless steels.

Open Circuit Corrosion Potential (OCP) Measurement

Results from OCP measurements can be very well analyzed from corrosion potential vs time plots. These plots indicate 4 distinct stages:

(i) Increase of potential in the noble direction resulting from thickening or healing of the passive film.

(ii) Constancy in corrosion potential although at time fluctuations in active direction characteristic of local breakdown of passivity may be observed.

(iii) Steep or gradual fall in corrosion potential from the passive region representing pit initiation stage.

(iv) Corrosion potential sustains low values during propagation stage.

The OCP vs time plots provide information about induction time for pit or, crevice initiation, t_i, which is indicated by a constancy in potential. Due to the presence of a strong passive film, high alloy stainless steels have much larger induction time than the conventional steels. In high alloy stainless steels induction time increases with increasing PREN whereas with conventional stainless steels an opposite effect is found. The latter behavior indicates that conventional steels are susceptible to pitting in much shorter time and pitting behavior is greatly influenced by low Cr + Mo t_i contents. This behavior is also manifested by two separate liner plots of Epit vs ti [Fig.6] obtained for the two categories of alloys indicating strong dependence of Cr + Mo contents on the pitting behavior of the alloys.

CONCLUSION

(i) The corrosion rates of conventional austenitic steels and high alloy stainless steels in seawater, under static conditions and at 25°C and 50°C, as determined from immersion and salt spray tests, are extremely low (0.001 - 0.005 mpy). The corrosion rates as determined from electrochemical polarisation technique are low (0.01 - 0.1 mpy) but are about one order of magnitude higher than from immersion tests. This has been attributed to the time lag for stabilisation of protective oxide film.

(ii) High alloy SS when creviced in seawater at 25 and 50°C, show very small corrosion build up but no perceptible depth of attack indicating absence of propagation stage during crevice attack.

(iii) In the temperature range of 25-65°C, a linear relationship exists between logarithmic of corrosion rate and reciprocal of absolute temperature (Arrhenius plot).

(iv) Pitting potential, Epit shows a significant active shift with active chloride contents. Due to the formation of strong and more resistance passive oxide film, the high alloy stainless steel have much higher Epit than conventional high stainless steels.

(v) With the exception of 3127hMo, all other alloys show lowering in breakdown potential, E_b on chlorination. A low shift in E_b indicates a lesser vulnerability of the alloy towards crevice and pitting corrosion in seawater.

vi) Pitting Resistance Equivalent, PRE_N, appears to be a function of Epit, ti, E_b and CSPH. In general, two separate relationships exist for conventional (304, 316, 317, 904) and high alloy SS (254 and 654 SMO, hMO 1925 and 3127, Duplex 2205, etc.).

(vii) For seawater applications, conventional and high alloy SS can be used without risk of localized corrosion in crevice and deposit free systems. However, for crevice forming system, only high alloy SS appear to be satisfactory materials for seawater applications including desalination plants.

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Table 1: Composition of Stainless Steels

UNS No.	Alloys	Product	Fe	Cr	Mo	Ni	C	Cu	Mn	Others
S30403	304L	-	Balance	18	-	10	0.03	0.2	1	0.45P, 0.02S, 1.0si
S31603	316L	-	Balance	16	3	11	0.02	0.2	1	0.045P, 0.02S, 1.0SI
S31703	317L	-	Balance	18	3	13	0.02	0.2	2	0.045P, 0.02S, 1.0si
N08904	904L	AVESTA	Balance	20	5	24.5	0.02	1	1.5	0.045P, 0.02S
S31254	SMO 254	AVESTA	Balance	20	6	18	0.02	0.7	-	0.2N, 0.002Si
S32654	SMO 654	AVESTA	Balance	25	7	22	0.015	0.4	0.5	0.5N
N08925	hMO 1925	VDM	Balance	21	6	25	0.01	0.9	0.9	0.2N, 0.002Si
N0828	hMo 3127	VDM	Balance	27	6.5	32	0.02	1	2	0.003S, Nb, Pb
-	Remanit 4565	THYSSEN	Balance	22-25	4-6	16-18	0.02	0.1	5	0.5N, 0.003Si, 0.1Nb
S44635	Monit 44635	AVESTA	Balance	25	4	4	0.025	0.2	0.4	0.035N, Ti, Nb, Pb, Si
-	Remanit 4575	THYSSEN	Balance	27-29	2-3	3-4.5	0.01	-	0.1	0.025N, Ti, Pb, Si, Nb
S31803	Duplex 2205	AVESTA	Balance	22	3	6	0.02	0.2	1.5	0.15, 0.6 (Cb+Ta)
S41000	410	-	Balance	13	-	-	-	-	1	0.04P, 0.035S, 1Si
S17400	17-4PH	ARMCO	Balance	16	-	4.5	0.07	4	1	0.3Cb, 0.04P, 1Si
N08020	20Cb3	CARPEN-TER	Balance	19	2	33	0.02	3	0.4	0.4Si, 5Cb+ Ta

Table 2: Gulf Seawater Composition, pH 8.2

Ions		Concentration mg/L
Calcium	Ca^{++}	508
Magnesium	Mg^{++}	1,618
Sodium	Na^{+}	13,440
Potassium	K^{+}	483
Strontium	Sr^{++}	17
Bicarbonate	HCO_3^{-}	176
Chloride	Cl^{-}	24,090
Sulphate	SO_4^{--}	3,384
Bromide	Br^{-}	83
Fluoride	F^{-}	1
Total Dissolved Solids	TDS	43,800

Table 3: Breakdown Potential (E_b) of Stainless Steels in Gulf Seawater at 25°C

S.No.	Alloys	E_b (mV) Normal Seawater	E_b (mV) Chlorinated (0.20-0.25 ppm)
1	316-L	228	197
2	317-L	524	300
3	904-L	1016	905
4	3127hM$_o$	950	905
5	1925hM$_o$	968	989
6	254SMO	918	911
7	Duplex-2205	1010	1003
8	Remanit 4565	1039	952
9	654 SMO	1001	948

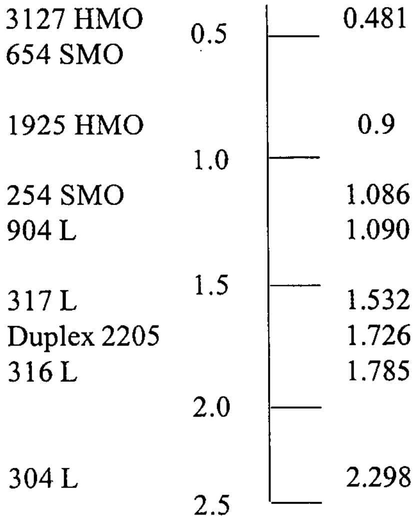

Figure 1 Predicted ranking of conventional and high alloy stainless steels.

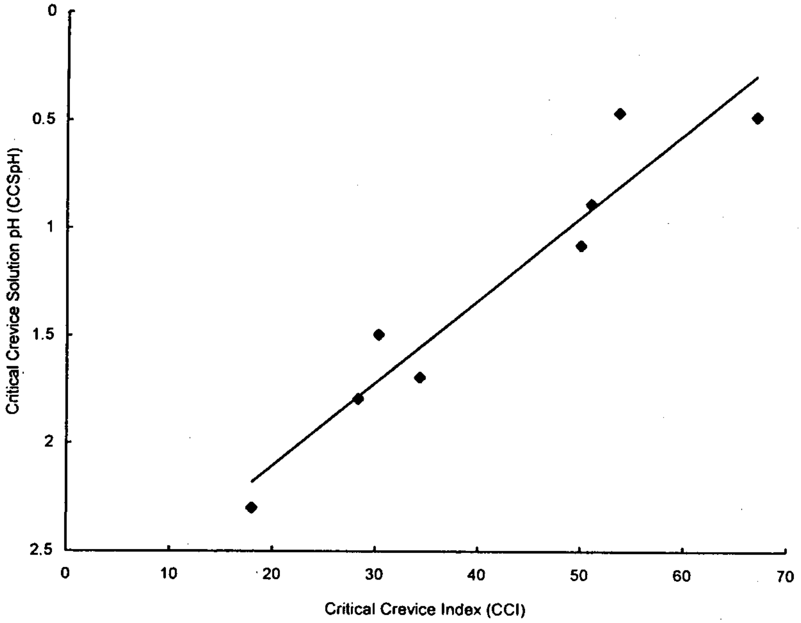

Figure 2 Plots of Critical Crevice Index (CCI) vs Critical Crevice Corrosion pH(CCSpH)

pH (CCSpH)

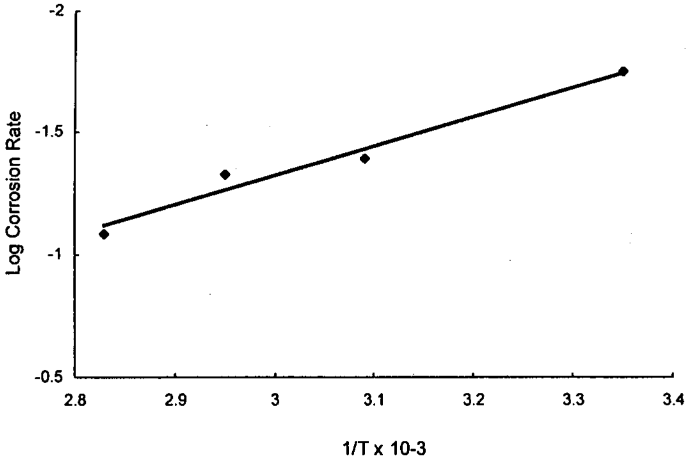

Figure 3 Effect of temperature on corrosion rate of SS316L - Arrhenius Plot.

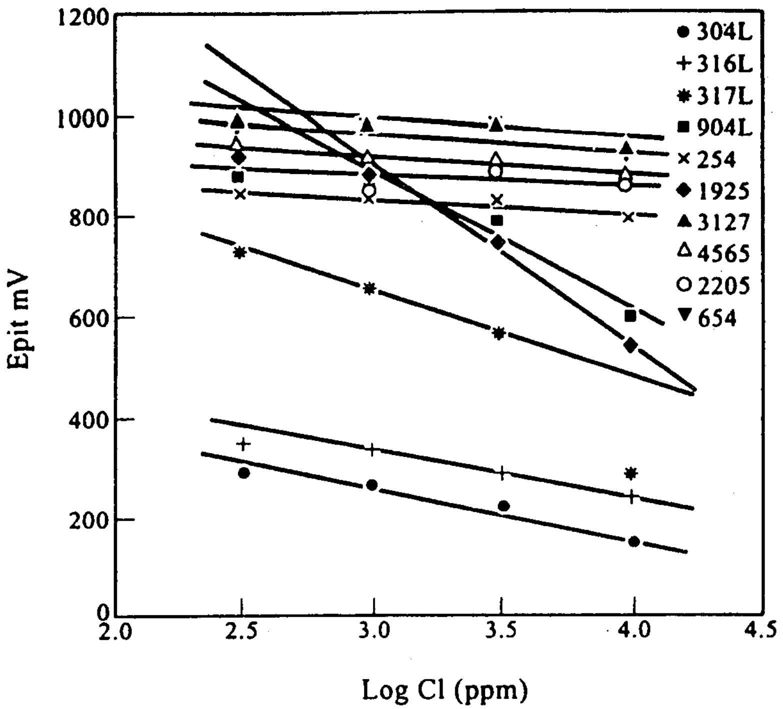

Figure 4 Plots of pitting potential E_{pit} vs Log $[Cl^-]$ for different steels in chloride containing aqueous solutions

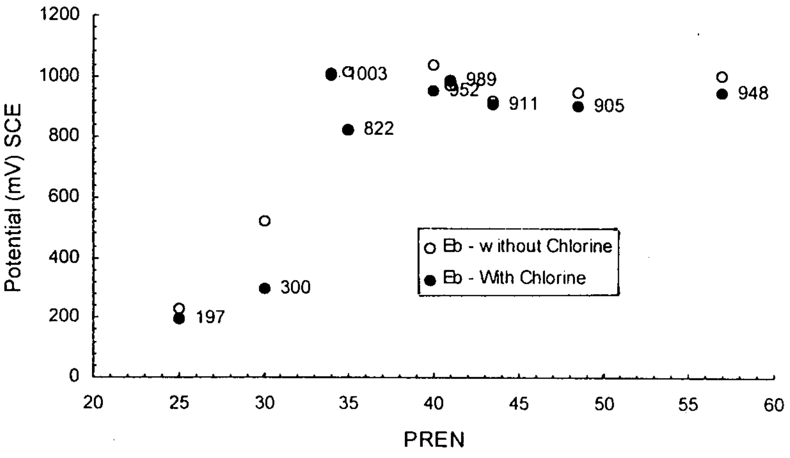

Figure 5 Graph showing variation in break down potential E_b with PRE_N values with and without chlorination (0.2 ppm) at 25°C

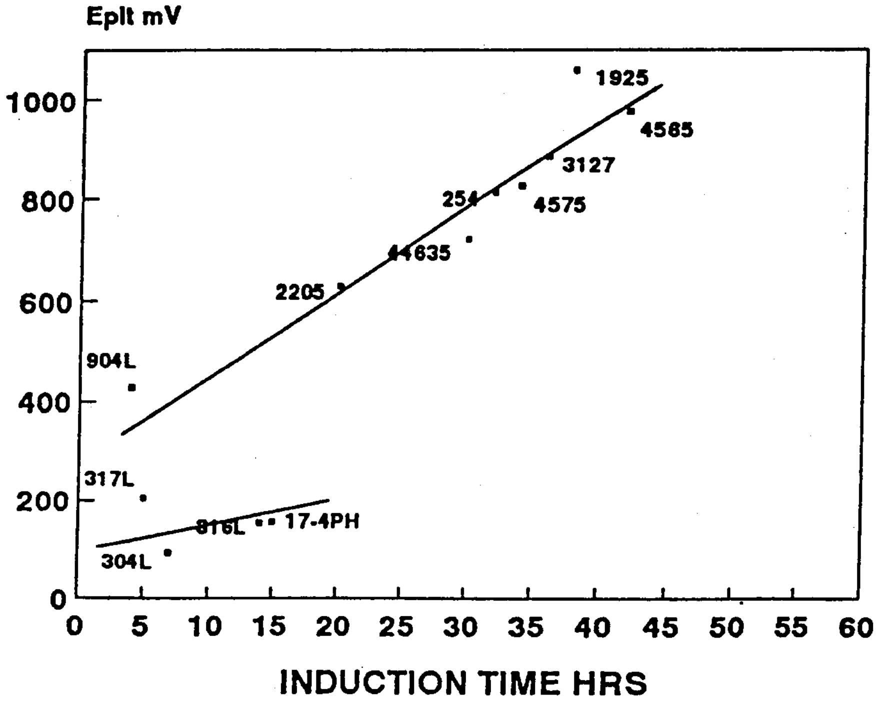

Figure 6 Plot of induction time. t_i and pitting potential, $E_{pit.}$ Two seperate linear plots represent conventional and high alloy stainless steels, respectively.

Vapor-Side Corrosion Monitoring in Al-Khafji Desalination Plant

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VAPOR-SIDE CORROSION MONITORING IN AL-KHAFJI DESALINATION PLANT

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ABSTRACT

The corrosion problem in vapor side is aggravated due to uncertainties in concentrations of oxygen, carbondioxide and other non-condensable gases. In such a situation, vapor-side corrosion could be a limiting factor in a MSF plant life, if not studied properly. This paper presents the results of a vapor-side corrosion monitoring study carried out at Al-Khafji desalination plant. Corrosion behavior of 70/30 Cu-Ni alloy, carbon steel and AISI-316 have been studied in vents and in the vapor-side of evaporator for about 9000 hours. Corrosion rates have been determined by weight loss method. The corrosion products have been analyzed by Energy Dispersive X-ray Analyzer (EDAX).

Analyses of the experimental data indicate that in addition to O_2, H_2S evolved during flashing of seawater has reacted more with Cu-Ni alloy as compared to carbon- and Stainless steels. Contrary to other reported works, ammonia and bromine, have not been found in corrosion products. Explanations for comparatively higher corrosion of alloys exposed to the vapors of early (1-3) and middle (11-13) stages of MSF plant have been given with experimental results. Stainless steel 316L has been found as the most resistant alloy to vapor-side corrosion in all stages of distiller.

Keywords : Vapor-side Corrosion, Non-condensable Gases, MSF Plant, Seawater, Oxygen Ingress.

INTRODUCTION

SWCC manages seawater desalination plants with a total capacity of 520 MIGD of potable water to meet the drinking water requirements of 70% population of the Kingdom. Some new plants are under construction to increase the capacity to 800 MIGD [1]. Failures of components may result in unscheduled shutdown of the plant and consequently in the reduced production of water.

Multistage flash (MSF) evaporation and reverse osmosis (RO) are the two main techniques used for producing potable water from seawater. The former is predominantly employed for the production of desalinated water. An MSF plant is much more prone to corrosion due to its operation at higher temperatures. Corrosion monitoring of the component/material in an operating desalination plant is a profoundly important task because it provides a wealth of information about the condition of plant in terms of construction materials, components and operational parameters. All these information can be useful in evaluating the performance and behavior of materials and efficiency of the plant.

In MSF plants, the corrosion of construction materials in flash chamber and water boxes due to flashing brines has been studied by several investigators [2-7]. However, the major problem in both acid and additive MSF plant is vapor-side corrosion. While in acid dosed plants, almost all the CO_2 is removed by the decarbonator, CO_2 is present along with O_2 as non-condensable gases in additive treatment plants. Other gases like H_2S, NH_3, Br_2 and Cl_2 may also present in significant concentrations and could be responsible for corrosion attack.

Oldfield and Todd [8] studied vapor side corrosion in an MSF plant. It was concluded that in acid dosed plants and probably also in additive plants the vapor-side corrosion is caused primarily by the presence of oxygen, this was coming not from the deaerated brine but from air-in-leakage into the plant. The results of studies on the corrosion behavior of stainless steels were reported [5]. In air ejector system of an MSF plant using 316L and 904L steel pipes, while 316L failed by SCC, the latter grade appeared to be quite resistant to corrosion. Failure of SS 316L steel was attributed to the presence of Br_2 in vent gases [9]. Lee, Oldfield and Todd [10] reported failure of 316L components in the venting system of MSF in Qatar due to bromine. The presence of ammonia in seawater above the normal background level is necessary for this to occur. After 25 years of operation, Abu Dhabi desalination plant experiences corrosion problems in its 23 large distillation units [11]. Substantial copper oxide deposits were found in the distillate product during the start up of the distiller following long and/or short shut down. The deterioration of the protective oxide film appeared to be the main cause of the problem which could be solved by improving the efficiency of the deaerator and the venting system. In another case [12], a group of 24 MSF distillers experienced failure of tubes in the first few high temperature stages after 50,000 to 100,000 hours of operation. It was concluded that the problem of tube failure was due to the presence of pockets of CO_2 gas near the tube, which redissolves in distillate to form an aggressive solution. The installation of an auxiliary vent to reduce the size of the gas pocket allowed a significant increase in distillater pH and reduction in dissolved copper.

Keeping in view the importance of vapor side corrosion in MSF plants, a monitoring study was carried out at Al-Khafji plant. This paper presents the results of that study.

SWCC Al-Khafji Phase II MSF desalination plant was commissioned in 1985. It has a designed capacity of producing 22727 m^3/d (5 MIGD). There are 2 desal units (# 100 and #200), each unit has 22 stages. In this plant seawater make-up is not treated with chemicals for scavenging of oxygen.

EXPERIMENTAL

Coupons of carbon steel, AISI 316L SS and 70 Cu/30 Ni were fixed above the demisters in the flash chamber # 1, 4, 10, 14 and 18 of unit # 100. Coupon of the aforementioned materials were also fixed with the help of retractable coupon holders, in the vent lines of stage # 3, 8, 13, 17 and 22 of unit #100. The arrangement of coupons in vent line and flash chambers of Al-Khafji plant unit #100 is shown schematically in fig. 1.

Coupons fixed in the vapor phase inside the flash chambers were taken out after 9,000 hrs, while coupons fixed in vent lines were taken out after 2,000 or 4,000 hrs of exposure. Weight loss coupon technique was used for corrosion rate measurements following ASTM G 90 [13].

RESULTS AND DISCUSSION

As shown in fig. 2 in evaporator, weight losses shown by Cu/Ni alloy and SS316L are by and large same in all the stages. In vent lines (Fig. 3), weight loss decreases steeply from stage #3 reaching a minimum at #8 and then increases showing a maximum at #13. Also these figures show that the weight losses of SS 316L specimens are much higher (about one magnitude) in vent lines than that in flash chambers although the magnitude of the weight loss is quite low. In case of vapor side corrosion of cupronickel exposed in vent line and flash chambers, weight losses after 2,000 hrs in vent line are more or less same as in flash chambers after 9000 hrs. Tables 1 and 2 indicate that in vent line Cu-Ni alloy corrode much faster than in flash chambers. Table 1 shows that corrosion rates of mild steel are highest in stage #3 (4 MPY). In comparison, corrosion rates of cupronickel are about one magnitude lower to carbon steel. The corrosion rates of AISI 316L in vent line are extremely low at all the stages and are at least one order of magnitude lower than cupronickel alloy.

Figures 4A to 5A show some typical scanning electron microgrpahs of cross-section of mild steel coupons exposed for 9,000 hrs in different stages of flash chambers. All the microstructures have a common feature that the scales are copious and adherent. The inner scales are generally separated from alloy interface. The energy profiles obtained by EDAX for the exposed alloy indicate the presence of Fe and O in the outer layers and that of chloride in the inner layers in all stages (Fig. 5B). In some cases presence of Al, Si and S is also indicated. Chloride is invariably concentrated at the alloy/scale interface.

Figures 6A shows a photograph of carbon steel, 316L SS and 70/30 Cu-Ni alloy, respectively, exposed to vapor in evaporator #100. Heavy corrosion is observed in carbon steel specimen. After removal of scale pitting can be seen on the surface of alloys (Fig. 6B), the pitting is more extensive in later stages. No or very little corrosion is observed on 316L specimen. Figure 7A shows a photograph of cupronickel alloy specimens exposed to non-condensable gas in the vent line for 2,000 hrs, appreciable corrosion is found on the specimens. Fig. 7B shows a photograph of carbon steel exposed for the same period. Heavy corrosion product formation is noticed in all the stages.

Table-2 provides corrosion rate values of all the three alloys exposed to non condensable gases in flash chambers. The corrosion rates for carbon steels are higher in stages # 10 onwards than those observed in the initial 4 stages. Corrosion rates of cupronickel are highest in stage # 1 (0.1 MPY) and lowest in stage # 18 (0.007 MPY). These corrosion rates are much below the accepted limit. Corrosion rates of AISI 316L in all stages are extremely low (0.001 to 0.002 MPY) showing virtually no corrosion occurring in flash chambers by non condensable gases.

It appears that a decrease or increase in corrosion rates is directly related to build up of non condensable gases in vents and subsequent cascading. As the non condensable gases concentration increases corrosion rates are also increased. With cascading, corrosion rate goes down but increases again from stage to stage as non condensable gases concentration builds up. In early stages, corrosion rates are high due to elevated temperatures and high concentration of corrosive gases in the vapor phase.

In flash chambers, the weight losses progressively increase due to vapor phase attack up to stage #10 when it becomes constant. In mild steel, the weight losses in flash chambers are much higher (about one order of magnitude) than vent line. In flash chambers, the vapor phase is mainly consisted of moisture and non condensable corrosive gases. This provides favorable conditions for rust formation and could be the reason for high corrosion rates in flash chambers. In vent lines, mild steel is attacked by non condensable gases but not so aggressively as in flash chambers.

Considering, vapor-side corrosion of cupronickel 70/30 exposed in vent line and flash chambers, the corrosion rates in vent line are much higher than in flash chambers. In vent lines, the cupronickel alloy is attacked much more severely due to the presence of predominantly corrosive gases. In flash chambers, there is predominant concentration of water vapors and relatively low concentrations of non condensable gases and in consequence, attack is not so severe. The corrosion rates of stainless steel 316L in vents are extremely low (0.04-0.07 MPY) but still one order of magnitude higher than flash chambers (0.001-0.002 MPY). In either case, the corrosion rates are not alarmingly significant and the 316 L can be safely used in vent lines.

An interesting feature of the corrosion products from the carbon steel specimen exposed to non condensable gases in flash chambers or vent lines is that, sulfur is invariably present though in small concentrations. Copper and bromine are present in smaller but in perceptible concentrations and only in the corrosion products of steel coupons exposed in vent lines. Sulfur and bromine are undoubtedly originated from un-condensed gases but there is no plausible explanation for the existence of copper in the scales or corrosion products of the carbon steel coupons.

Considering the corrosion behavior of materials in vent line and evaporator at different stages. The shape of the weight loss vs. no. of stages curves can be explained broadly as follows:

(i) Steep increase in weight loss (or corrosion rates) is due to breaking up or spalling of scales when fresh alloy surface comes into contact with gases.

(ii) Decrease in weight loss could be due to progressive formation of adherent protective scales.

(iii) Presence of plateau (no change in weight) could be due to the formation of protective scales.

CONCLUSIONS

(i) Corrosion rates of cupronickel and stainless steel 316 L exposed to non condensable gases in vent lines are much higher than those exposed in flash chambers.

(ii) Corrosion rates of carbon steel exposed to non condensable gases in flash chambers are higher than those in the vent lines. Higher corrosion rates are attributed to the severe rusting of carbon steel in presence of moisture and oxygen.

(iii) Stainless steels 316L in general show very little or no corrosion either in vent line or flash chambers.

(iv) Cracking of otherwise compact and adhered scales on mild steel exposed to non-condensable gases in the flash chambers has been attributed to the presence of chloride at the metal matrix/oxide scale interface.

(v) Sulfur is invariably present in the corrosion products of Cu-Ni exposed in vent line.

(vi) High corrosion rates of cupronickel in vent line are attributed to the presence of sulfur in non-condensable gases.

(vii) A maxima in corrosion rates has been generally found in the middle stages # 10-14 in all the alloys. This could be explained on the basis of combined effect of temperature and air in-leakage providing most favorable conditions for corrosion at these stages.

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Table 1: Corrosion of Cu-Ni, Carbon Steel and Stainless Steel 316L in Vent Lines of MSF Desalination Plant, Unit # 100.

S. No.	Distiller Stage # (Cascading)	Exposure Period (h)	Corrosion Rate (MPY)		
			70-30 Cu-Ni	SS 316L	Carbon Steel
1	3 (1-3)	2000	0.41	0.06	3.97
		4000	0.42	0.05	5.55
2	8 (4-8)	2000	0.07	0.06	0.46
		4000	0.05	0.04	1.02
3	13 (9-13)	2000	0.1	0.07	1.95
		4000	0.06	0.07	1.74
4	17 (14-17)	2000	0.04	0.06	0.42
		4000	0.01	0.02	0.99
5	22 (18-22)	2000	0.08	0.04	0.87
		4000	0.12	0.02	0.96

Table 2: Corrosion of Cu-Ni, Stainless Steel 316 L and Carbon Steel in the Flash chambers of MSF Desalination Plant, Unit # 100.

S. No.	Distiller Stage #	Exposure Period (h)	Corrosion Rate (MPY)		
			70-30 Cu-Ni	SS 316L	Carbon Steel
1	1	9000	0.1	0.001	1.80
2	4	9000	0.05	0.001	1.72
3	10	9000	0.02	0.002	2.41
4	14	9000	0.01	0.002	2.40
5	18	9000	0.007	0.001	2.36

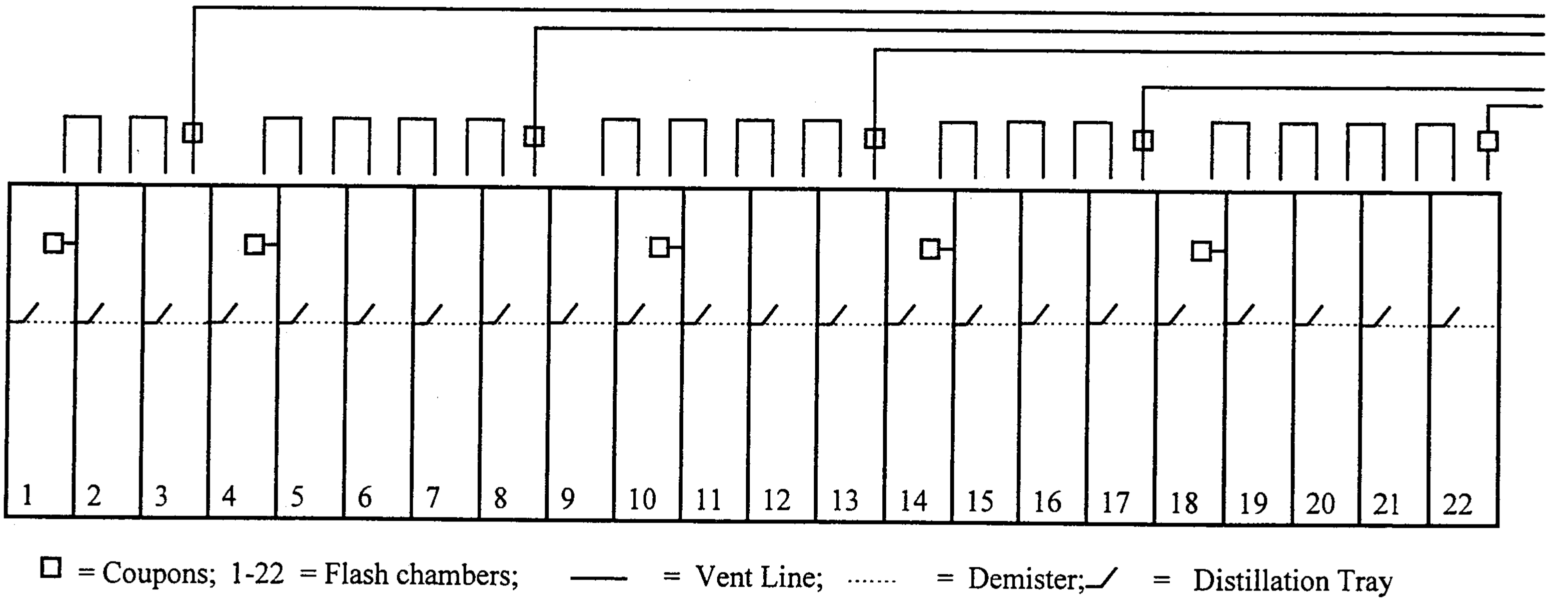

Figure 1: Location of the coupons in vent line and chambers of MSF distiller

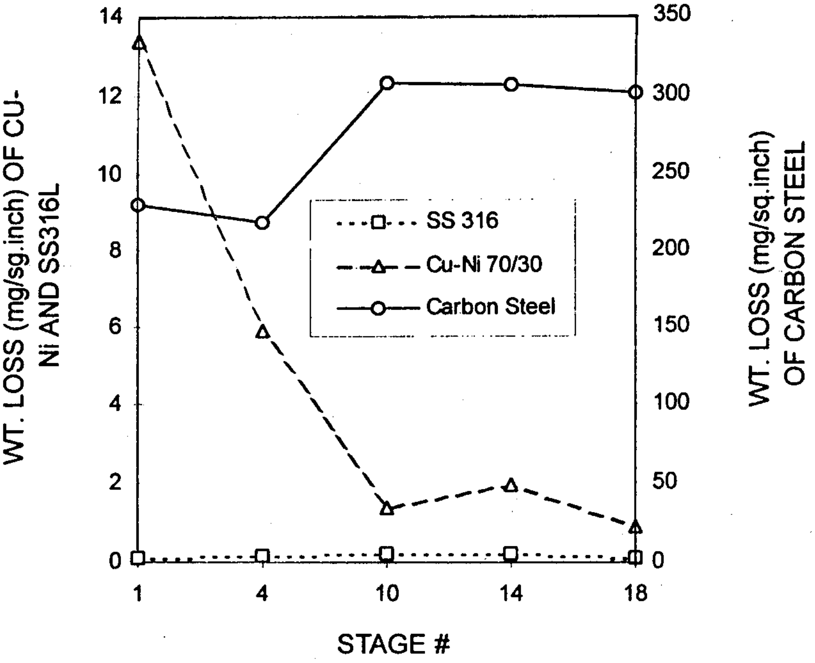

Figure 2: Vapor-side corrosion of different materials exposed for 9000 hrs. in flash chambers of MSF plant

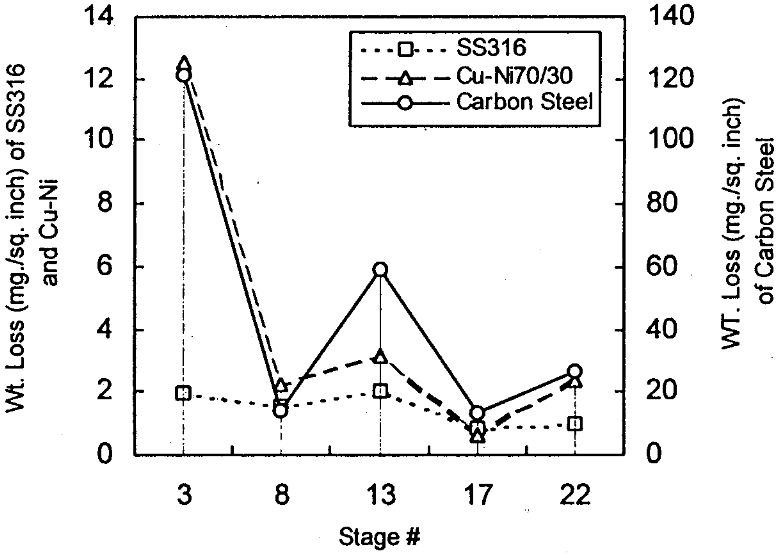

Figure 3: Vapor-side corrosion of different materials exposed for 2000 hrs. in vent line of MSF plant

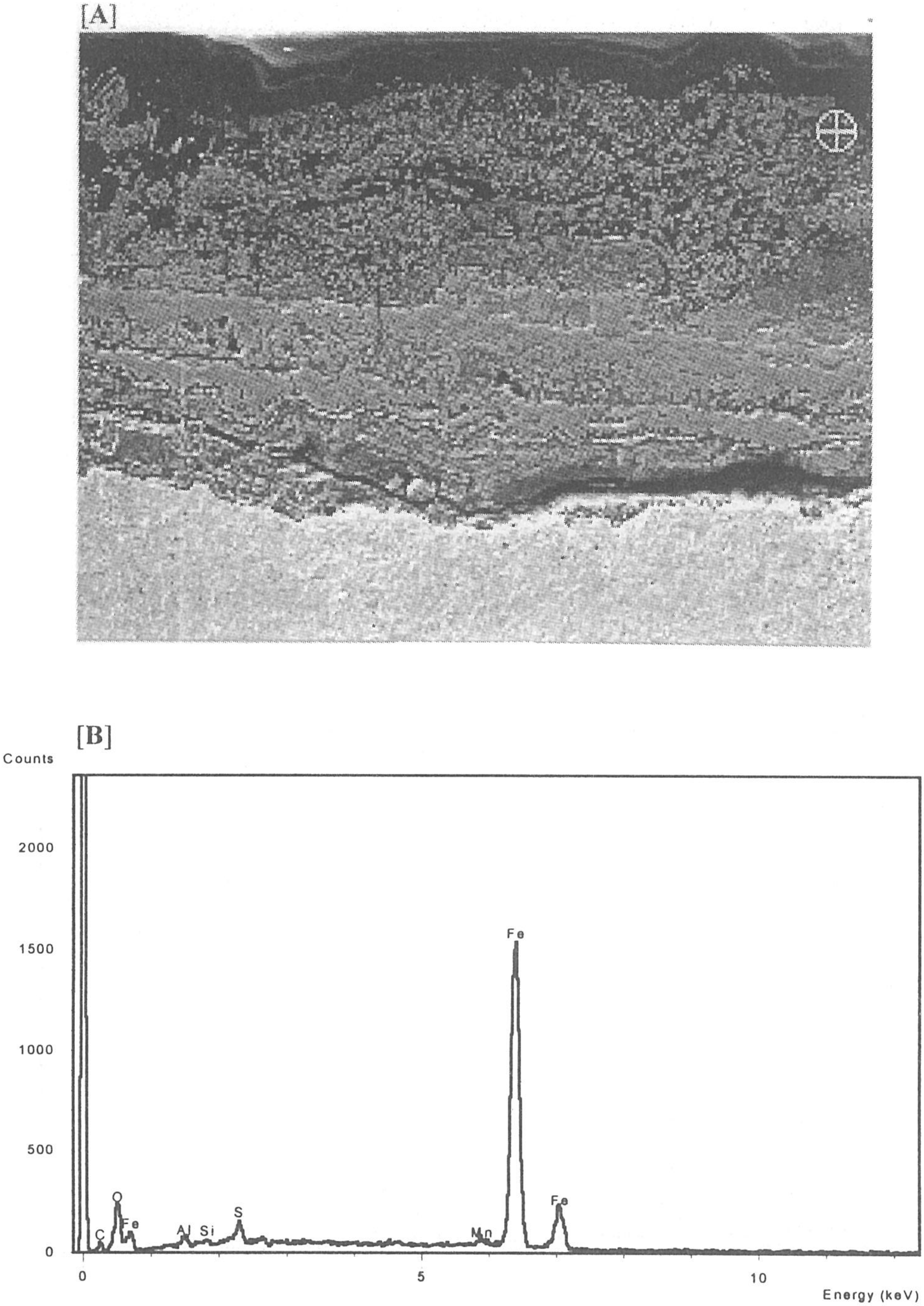

Figure 4. [A] SEM micrograph and [B] EDAX results showing the elemental distribution in the oxide scales formed on carbon steel coupon, exposed for 9000 hrs. in stage #1 of MSF plant

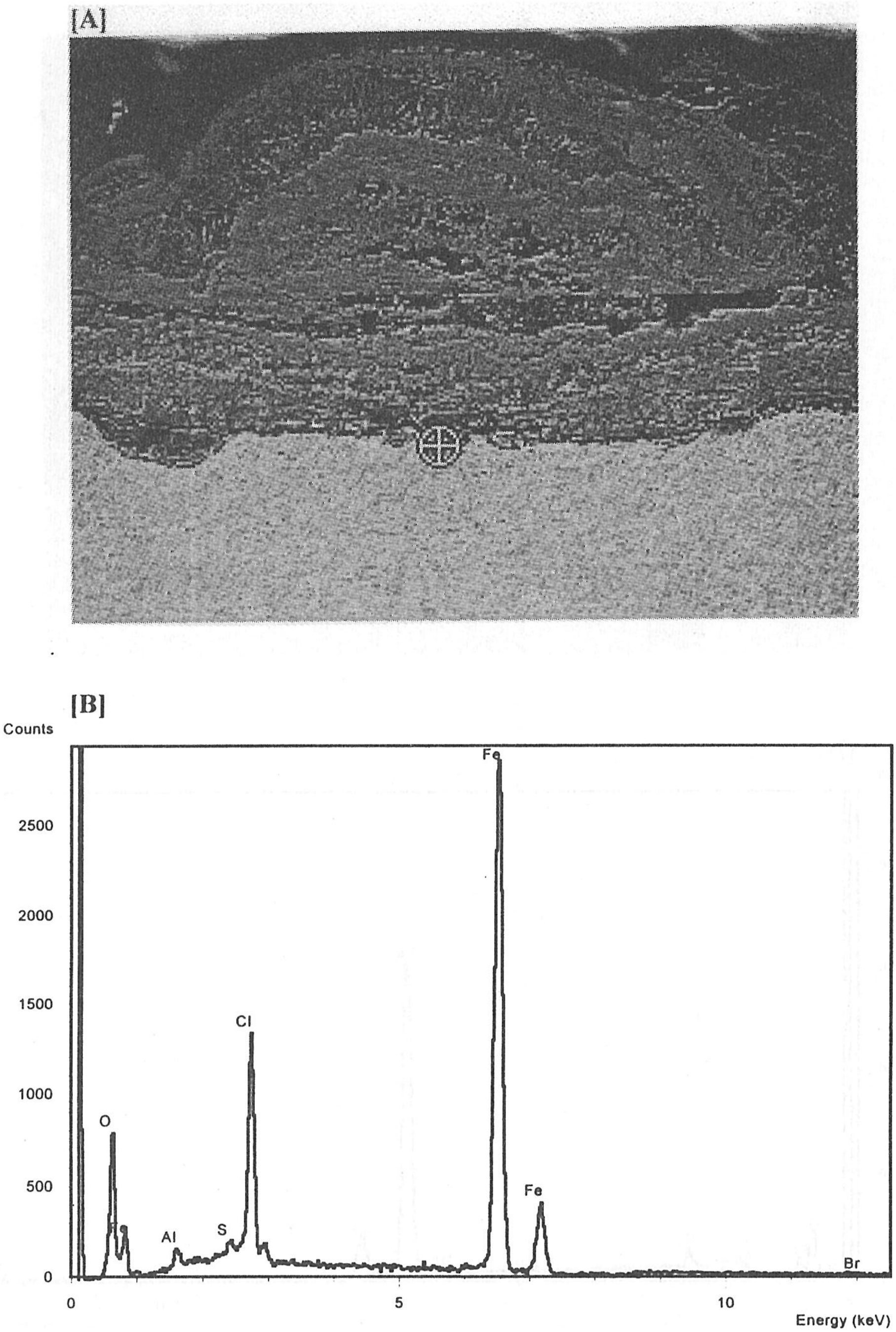

Figure 5. [A] SEM micrograph and [B] EDAX results showing the elemental distribution in the oxide scales formed on carbon steel coupons, exposed in MSF stage #4, for 9000 hrs.

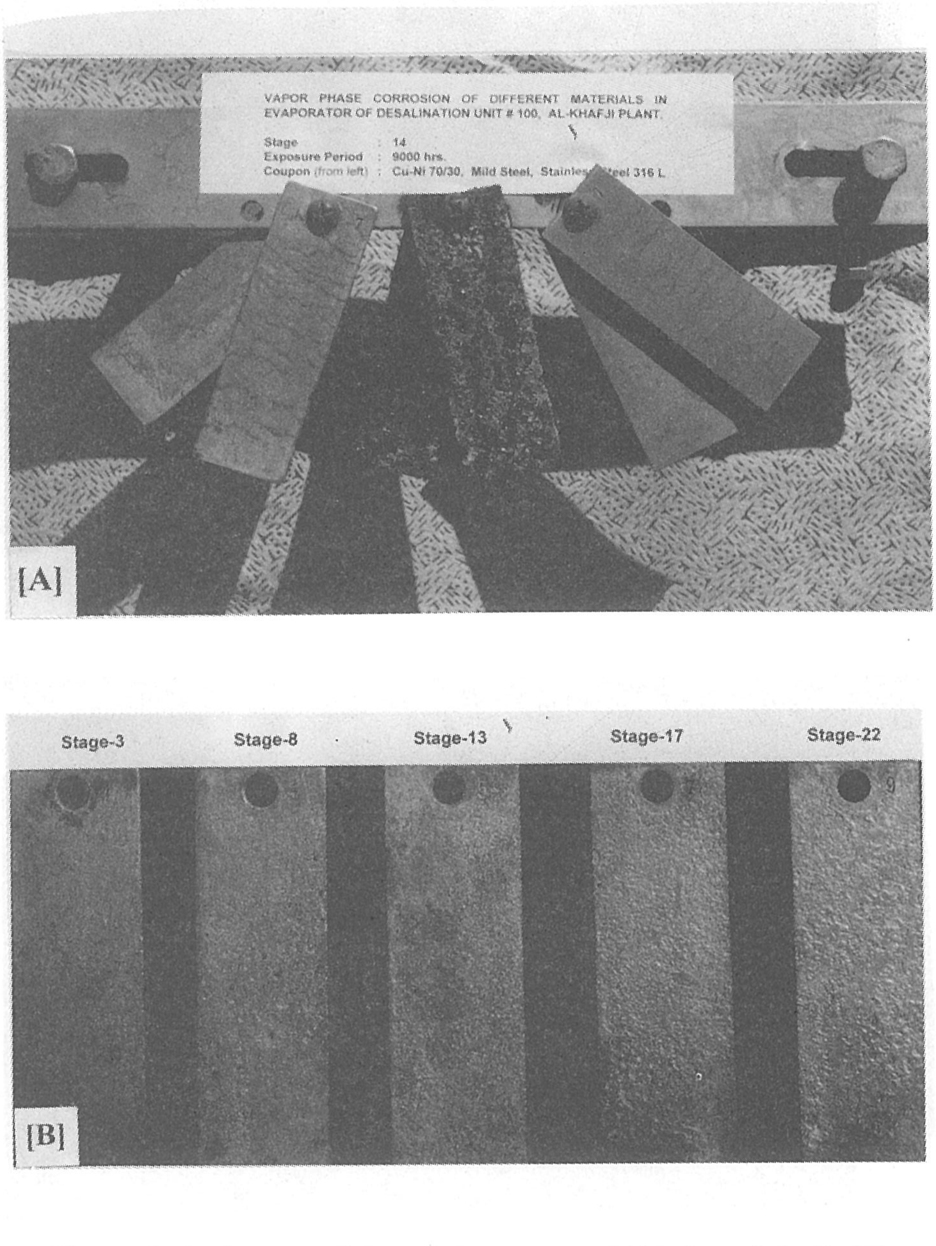

Figure 6. Surface condition of the coupons [A] left to right Cu/Ni alloy, carbon steel and stainless steel 316L, exposed for 9000 hrs. in stage#4 [B] carbon steel coupons exposed for 9000 hrs. in different stages of MSF desalination plant

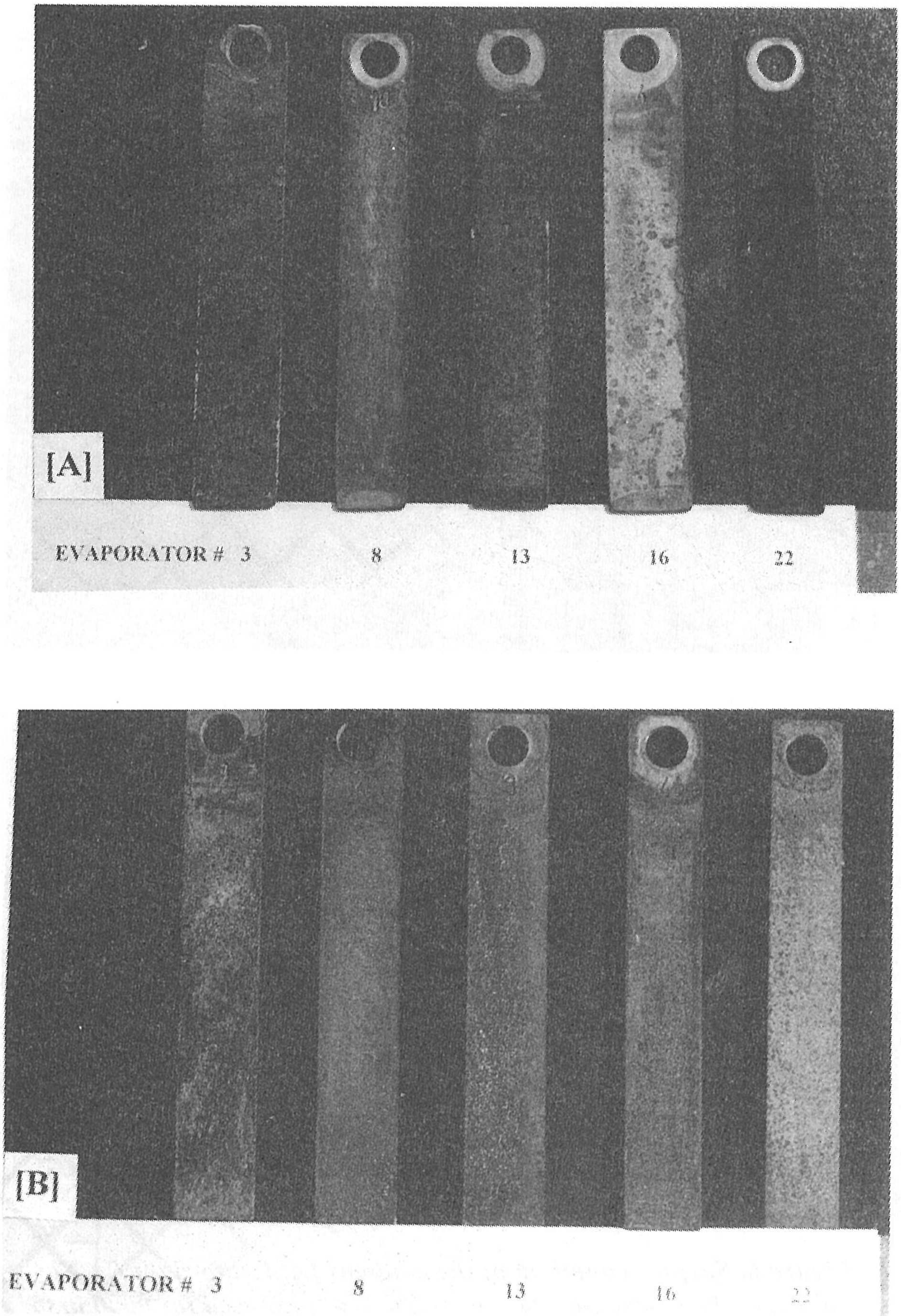

Figure 7 Surface condition of the coupons [A] Cu/Ni 70/30 [B] carbon steel coupons exposed for 4000 hrs. in vent line of different stages of MSF desalination plant

Long Term Electrochemical Noise Measurement for Stainless Steels Used in Desalination Plants

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LONG TERM ELECTROCHEMICAL NOISE MEASUREMENT FOR STAINLESS STEELS USED IN DESALINATION PLANTS

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ABSTRACT

Stainless steel is commonly used in desalination plants due to its resistance to corrosion. However, such material is prone to pitting corrosion in chloride media. Research into the fluctuations in the corrosion potential and current have attracted great interest in the past few years. This is primarily due to the advancement in digital techniques that allows one to easily collect, analyse and interpret a considerable amount of data. Electrochemical noise data are easily collected with the aid of inexpensive microcomputers in the laboratory. In the present work, our research was concerned with the behaviour of noise associated with corrosion potential and current transients. The potential and current transients obtained were analyzed and evaluated for Stainless Steel type 304 and 316 used in desalination plants. Experimental results using electrochemical noise suggest that a better performance of SS316 over SS304 in seawater and rejected brine media.

KEYWORDS: Corrosion, Desalination, Electrochemical Noise, Stainless Steel.

INTRODUCTION

Electrochemical Noise

The investigations of localised corrosion often leads to the analysis of random current fluctuations related to the random birth, growth, and repassivation of pits at an electrode surface. It is generally thought that the analysis of the time series of elementary events occurring at potentials below the pitting potential allows the susceptibility of an alloy or metal to localised corrosion to be assessed. Electrochemical Noise is shown to result from the behaviour of individual micro-chemical events, averaged over time, which when combined, form the potentials and currents normally measured in such systems. Noise has various origins and various models have been proposed to explain its origin. Either deterministic or stochastic technique could be used to develop a model to explain Electrochemical Noise phenomenon. However, semimacroscopic phenomenon related to corrosion are random by nature such as pitting, abrasion, gas evolution. This special property of the processes makes the classical deterministic techniques difficult to perform. Even when the processes seem to be in a steady state regime, non-linearity problems often remain [1].

EXPERIMENTAL SETUP

A major part of the experiments involved the collection of long term electrochemical potential and current noise from stainless steels commonly used in desalination plants in the typical conditions to which they are normally exposed. The practical data acquisition and handling methods used in the practical work were based upon digital procedures controlled by a personal computer. The programming language Pascal was used to write data handling procedures. The time records collected in this work measure fluctuations in corrosion potential between a working electrode and reference electrode and the fluctuations in the galvanic coupling current between two working electrodes. A schematic of the apparatus used is shown in Figure 1. In general terms, the electrochemical cells were set up in a 1 litre Pyrex beaker filled with 500 ml of the appropriate electrolyte.

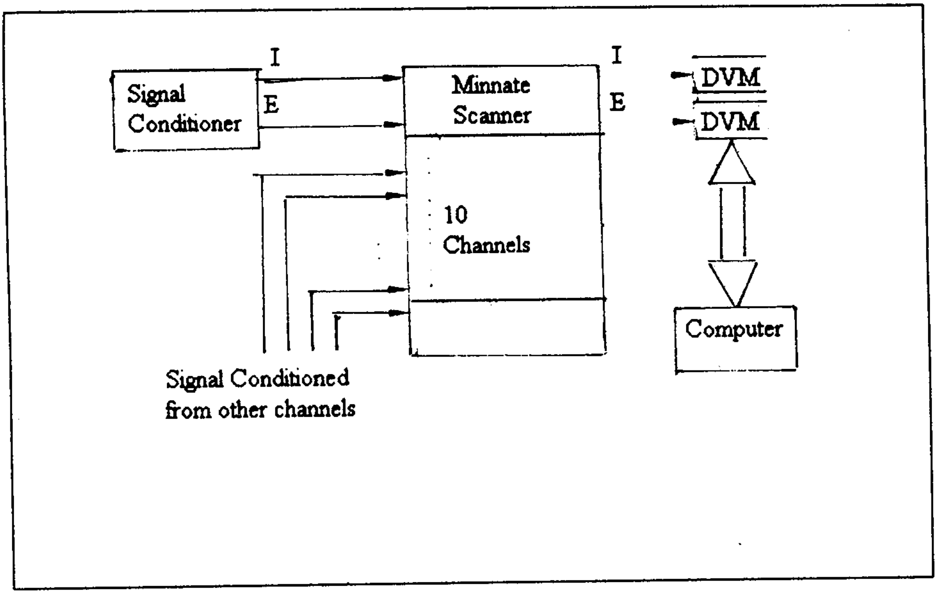

Figure 1: Electrochemical Noise Measurement Set Up

RESULTS

Behavior of SS304 and SS316 in Sea Water

Both SS304 and SS316 were immersed in seawater for a period of 140 hours. The time record for both potential and current are shown in figures 2 and 3. The typical current amplitude recorded for SS304 immersed in seawater is 1.5 μA. The immersion results for SS316 in seawater presented in Figure 3 suggest that the highest current amplitude found is 0.2 μA and the potential difference is 170 mV/SCE.

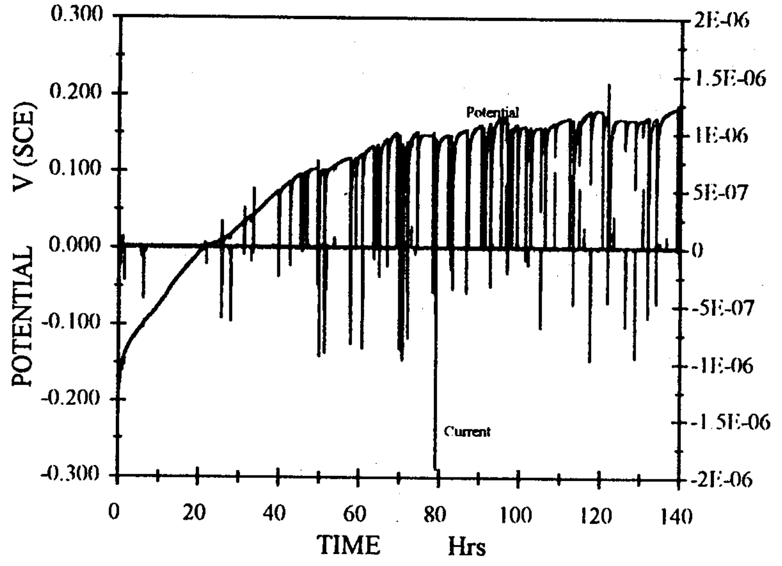

Figure 2: Time Record for SS304 immersed in seawater

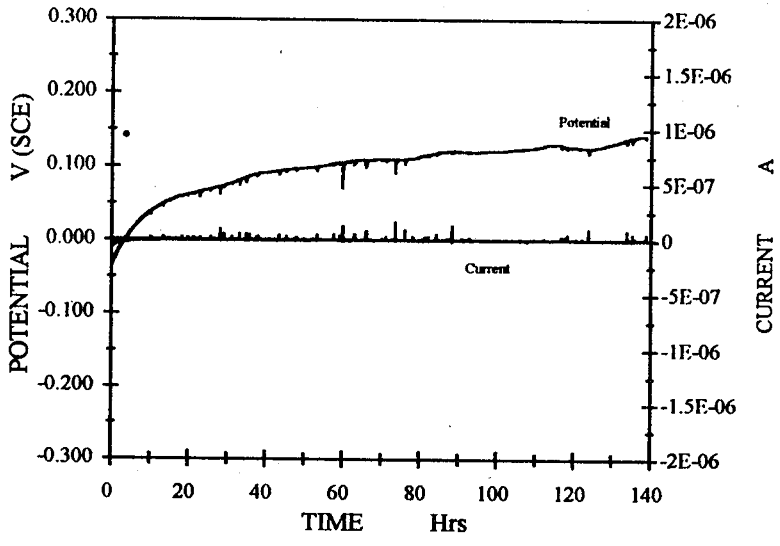

Figure 3: Time Record for SS316 immersed in seawater

Behavior of SS304 and SS316 in Brine

The SS304 and SS316 were also immersed in brine. The SS304 did record current amplitude of 2.0 μA and potential difference of 150 mV/SCE as shown in Figure 4. Figure 5 illustrates the time record for SS316 showing a maximum current amplitude of 1μA and the potential of around 100 mV at the end of exposure.

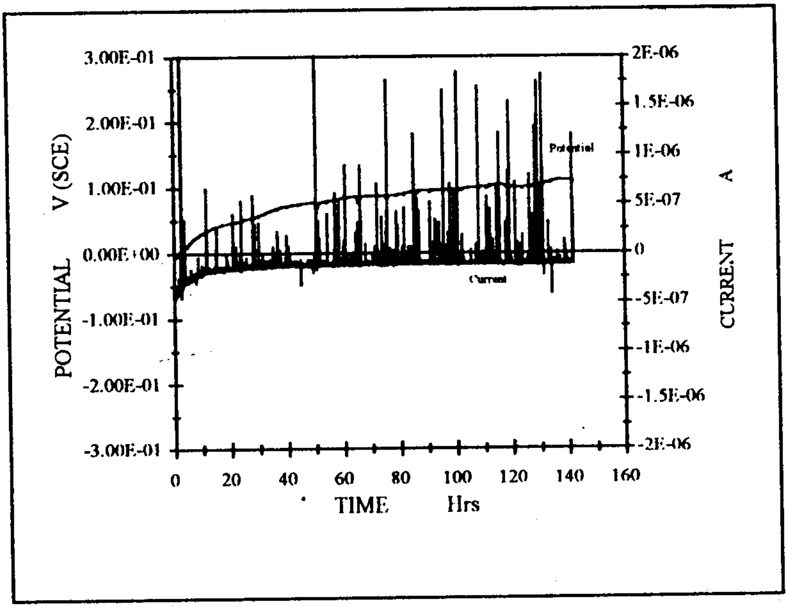

Figure 4: Time Record for SS304 immersed in Brine

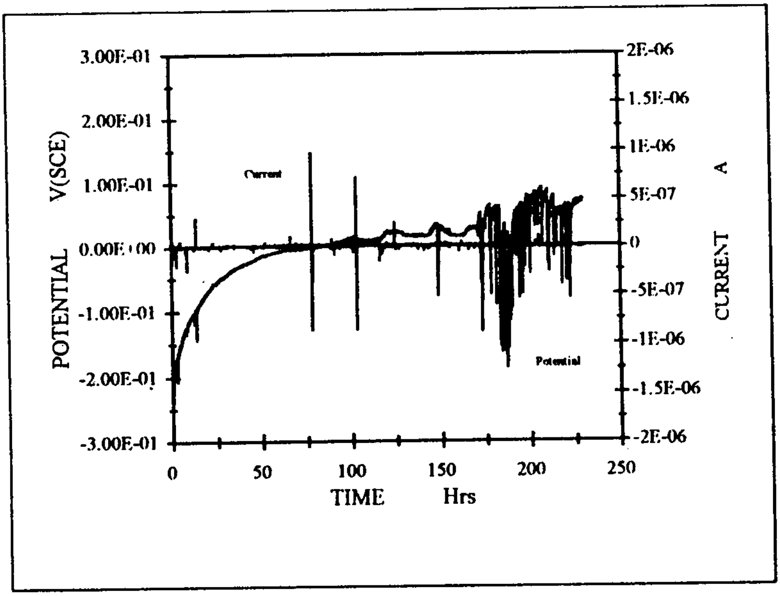

Figure 5: Time Record for SS316 immersed in Brine

DISCUSSION

Shapes of Transients of SS304 in SeaWater

The SS304 in seawater exhibited a mix of transients with stability criteria less than 0.3 Am^{-1}.

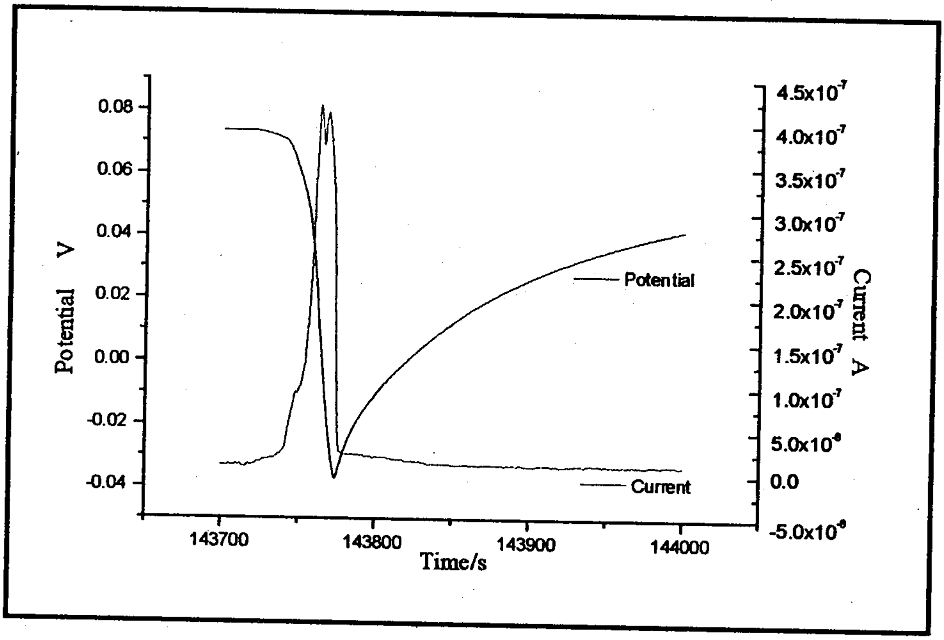

Figure 6: Time Record for SS304 in SeaWater - from 143700 to 144000 seconds

In Figure 6 it can be seen clearly that the stability factor of the metastable pit doesn't exceeded 0.016 Am^{-1}. The slow rise in the current transient is obvious in this situation with probability of two pits formation each have the slow inclination with fast declination, but one is larger than the other. The potential changed by 120 mV in about 30 seconds which responding fast to the current. In Figure 7 which shows slow rise with sharp decline of transient between immersion time 168300 – 168400 seconds.

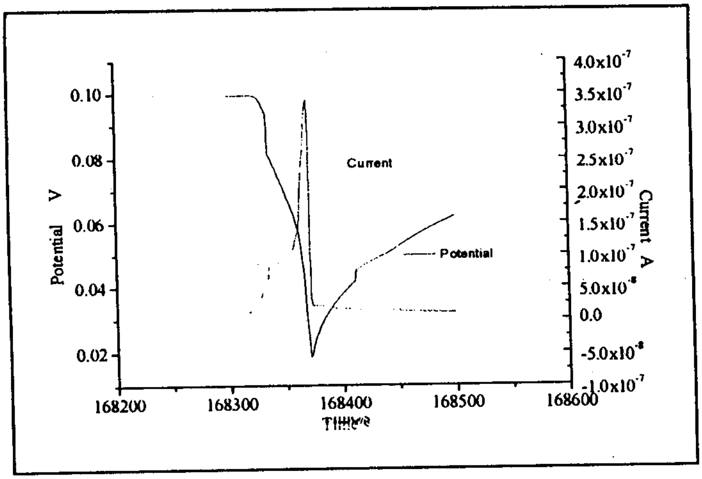

Figure 7: Time Record for SS304 in Sea Water - from 168200 to 168600 seconds.

However both transients in Figures 6 and 7 show similar type of transients, with metastable pit of stability product (ia=0.018 A/m).

Shapes of Transients of SS304 in Brine

After 232500 seconds immersion, the transient in Figure 8 was observed. This had a pit stability factor of 0.012 Am^{-1}, and behaved as if the cover was partially failed and repassivation process was moderate. There is another complimentary explanation of the transient behaviour, that the process of metastable pit went through stages and mainly four stages where at the first stage the cover has partially broken and repassivate then followed next and so on.

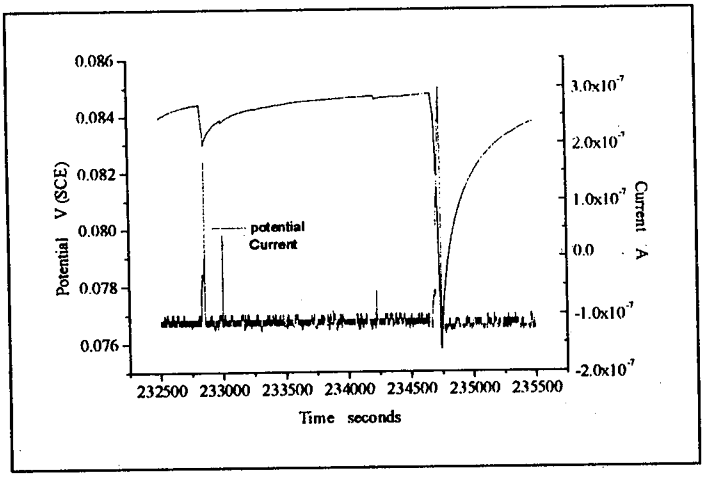

Figure 8: Time Record of SS304 in Brine – from 2323500 to 235500 seconds

Figure 9 demonstrates the same behaviour of the transient reported by Burstein et al[4] but with different current amplitude. Bearing in mind that Burstein et al[4] controlled the potential to hold the metastable event in the region of 0.56 V.

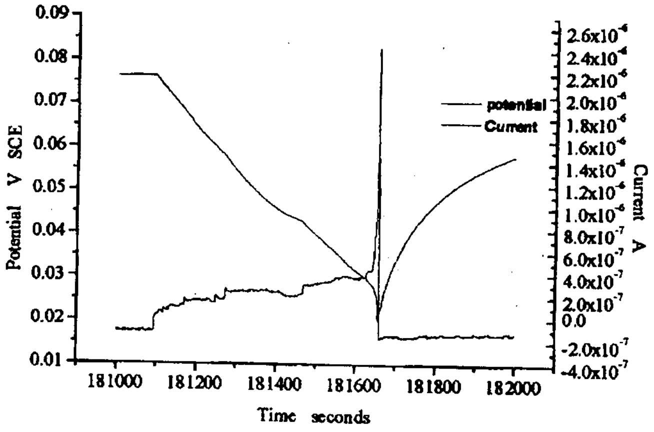

Figure 9: Time Record for SS304 in Brine – from 181000 to 182000 seconds

Shapes of Transients of SS316 in Sea Water

In Figure 10 the transient has stability criteria 0.00341 Am^{-1} and the transient in Figure 11 has stability index 0. 00318 Am^{-1} which suggests a metastable pit is present.

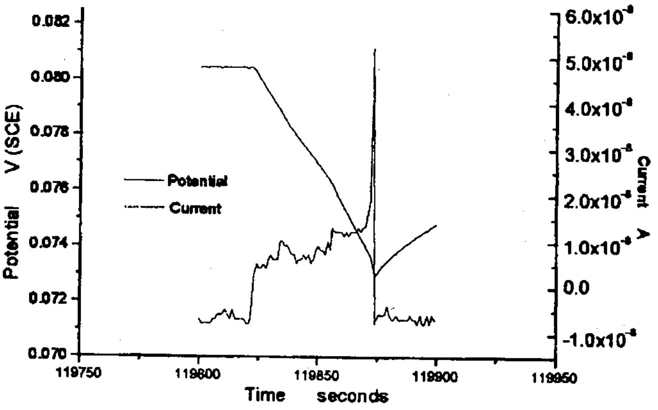

Figure 10: Time Record for SS316 in Sea Water – from 119750 to 119920 seconds

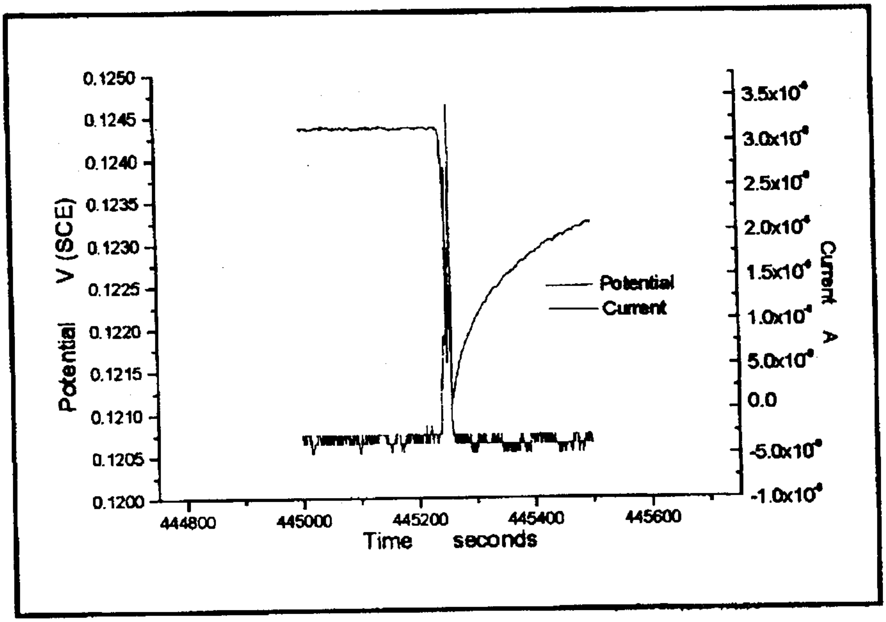

Figure 11: Time Record for SS316 in Sea Water- between 445000 - 445500 seconds

Shapes of Transients of SS316 in Brine

In Figure 12 there were 5 distinct current transients with amplitude varying from 0.16 to 0.35 μA with duration from 30 to 80 seconds. They show a pit stability index from 0.009 to 0.0151 Am^{-1} which suggests that they are still metastable events with low level of pit stability index.

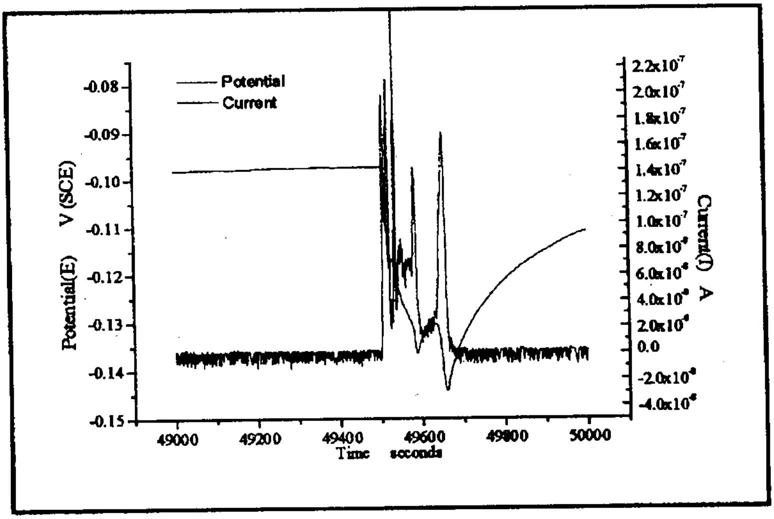

Figure 12: Time Record for SS316 in Brine – between 49000 and 50000 seconds

CONCLUSION

The current and potential transient of both SS316 and SS304 have the behavior of stainless steels current transient that slow increase with sharp decline upon metastable pit. However other forms of transient have been encountered. The general behavior of the time record shows that the SS316 have a better performance than SS304 in seawater and brine environment.

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Studies on the Stress Corrosion Cracking Behavior of Few Alloys Used in the Desalination Plants

T.L. Prakash and A.U. Malik

STUDIES ON THE STRESS CORROSION CRACKING BEHAVIOR OF FEW ALLOYS USED IN THE DESALINATION PLANTS

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ABSTRACT

This paper describes a study on the Stress Corrosion Cracking (SCC) behavior of alloys resulting from the synergistic action of corrodents such as chlorides, oxidants and H_2S. In the this study, the threshold stresses for SCC have been determined for few generic alloys namely, mild steel, 316L, 317L, 430 and Monel 400 used in the desalination plants. The standard Proof Rings and U-Bend samples in NACE and SHELL solutions containing H_2S are used for the purpose. Electrochemical polarization measurements were performed on these alloys in the specified environments to study the effect of electrochemical potential on the intergranular SCC. Structural analysis were conducted by Scanning Electron Microscopy supplemented by Energy Dispersive Spectroscopy. The test results showed that the intergranular and intragranular SCC fracture of mild steel and alloy 430 in H_2S environment occurs only in the limited potential environment, whereas, the alloys viz, 316L and 317L are immune to SCC under the condition of test performed. The alloy Monel 400 was found however susceptible to SCC in presence of H_2S which may be present in some part of desalination plants due to pollution of seawater.

Keywords: Stress Corrosion Cracking; Threshold Stress; Electrochemical Polarization: Sulfide Ion; Passive Film.

INTRODUCTION

Many engineering alloys when in contact with aqueous solutions such as seawater inherently containing number of dissolved ionic species are normally susceptible to corrosion. Chlorides being one of the major constituents of the sea water, its effect on the corrosion of material is of significant concern for the reliability of alloys. Corrosion of alloys under the influence of stresses in service could accelerate and lead to environmentally assisted cracking known as Stress Corrosion Cracking (SCC). Catastrophic failures have occurred due to SCC resulting in loss of life and expensive repairs. There are several types of SCC which include hydrogen embrittlement, sulfide stress cracking and so forth.

In the past, it was thought by several investigators that SCC of a given alloy occurs only in limited range of specific environments [1]. Subsequently, the above notion was diluted when it was found that SCC occurs in wide range of environments [2,3] containing more than one ionic species in solution that act as corrodents. The synergistic action of corrodents namely, Cl_2, CO-CO_2, caustic solutions containing H_2S (elemental Sulfur in solution) etc. have an immense effect on SCC of materials.

The carbon steels are prone to SCC in carbonate, bicarbonate, acetate, phosphate, etc. environment and are identified as main reason of cracking [4]. The low alloy steels are affected by SCC in oxygenated water at high temperature [2], alkaline chloride such as NaCl- $Ca(OH)_2$ [5] and anhydrous ammonia- methanol solutions containing chloride [6]. The J-55 and N-80 steels have shown SCC in chloride solution containing H_2S [7]. Similar observation was also made in AISI 1075 steel [8].

In austenitic stainless steels, SCC was well-known since three decades. The cracking was mainly due to chloride (which were neutral at high temperature, acid at low temperature) and hydroxide solutions [9]. Thiosulphate environments of weld sensitized stainless steels have shown SSC [10]. In the duplex stainless steel SSC is severe at 160 °C in 25% NaCl containing H_2S and also in aerated brine solutions [11].

The nickel base alloys viz., C-276 and alloy 825 were susceptible to SSC in HCl oxidizing solution containing S. In chloride containing solution the SSC has been observed at temperatures above 204 °C [11]. The copper base alloys are subjected to SCC in environments like ammonia, sulfur dioxide, organic complexing solution like acetates and tartrates, copper sulfate solution, etc. [12].

Threshold Stress for SCC

As the name implies the threshold stress is the stress below which no SCC occurs. The main purpose of determining the threshold stress for SCC is to establish a ranking order under given condition of metal environment combination, heat treatment microstructure, type of stressing and its magnitude. An exact threshold stress for a given condition is difficult to define. However, the relative ranking seems quite obvious.

The material which shows highest SCC resistance for a given environment may show susceptibility to SCC when it is heat treated to different microstructure. For example, threshold stress in SCC of carbon and low alloy steels was found to be influenced by heat treatment when it is studied using 5% NaCl - 0.5% Acetic acid solution containing 3000 ppm of H_2S [13]. The heat treatment carried out gave untempered martensite in the structure which is attacked by H_2S and resulted in low threshold stress values for cracking. The result of series of test in Drop Evaporation Test on highly alloyed stainless steel and duplex stainless steels as indicated by their threshold stress values [14], the highly alloyed stainless steels such as 654 SMO (UNS S 32654) was most resistant to SCC than the duplex stainless steels viz., 2205 (UNS S 31803).

The purpose of the present investigation was two fold: 1) To study the SCC behavior of few generic alloys used in the desalination plants resulting from synergistic action of corrodents such as chlorides, oxidants and H_2S and to determine the threshold stresses for SCC. 2) To assess the effect of electrochemical potential on the alloy passivity to corrodent species by performing the electrochemical polarization measurements in the specified environments.

EXPERIMENTAL DETAILS

The materials selected for this study were mild steel (MS), AISI 316L, 317L and 430 stainless steels and Monel 400. The chemical composition of these alloys are shown in Table 1.

Stress Corrosion Cracking Tests

Round and flat tensile samples were machined from the rod / sheet material stocks. All the materials selected were of mill finished commercial grades. The tests were carried out in Cortest Proof Rings (15] with corrosion testing environment chamber. An hour meter and H_2S gas manifold were used to measure the time of failure of specimen and H_2S gas monitoring during test respectively.

The media employed for the tests were i) NACE solution (having composition 5% NaCl + 0.5% CH_3COOH) prepared from distilled water and continuously bubbled with H_2S to maintain H_2S saturation in solution. ii) SHELL solution (having composition 0.5% CH_3COOH) prepared from distilled water and continuously bubbled with H_2S to maintain H_2S saturation as in i).

The samples were tested in ambient temperature with the Cortest Proof Ring at 70, 80, 85 and 90% of their respective 0.2% yield stress (YS) with the help of loading nut and calibration charts. During the tests H_2S was continuously bubbled in the solution. The time to rupture of the samples were recorded. The samples those have crossed 500 hours without rupture were withdrawn from the test. During the test samples were periodically withdrawn for examination of any initiation of cracks or corrosion pit development.

Electrochemical Tests

The electrochemical polarization techniques were performed to measure the absolute corrosion rates. Tafel plots were generated for this purpose on samples by polarizing the specimen about 300 mV anodically (positive-going potential) and cathodically (negative-going potential) from the corrosion potential, E_{corr}. The potential is stepped in staircase waveform. The resulting current is plotted on a logarithmic scale. The corrosion current I_{corr} is obtained from Tafel plot by extrapolating the linear portion of the curve to E_{corr}. The corrosion rate was calculated from the I_{corr}.

Experiments were carried out on the samples using EG&G model 273 Potentiostat with Softcorr Corrosion Software M342. A saturated calomel electrode was used as reference electrode. Graphite electrodes were used as auxiliary electrodes. Button samples of 14 mm dia and 2 mm thick were machined from rod/ sheet stock of the respective material. They were polished at one side to 600 # grade paper. The media employed in the electrochemical tests are i) NACE solution ii) Natural seawater iii) Natural seawater containing varied amounts of sulfide ion concentration obtained by dissolving known quantities of sodium disulfide crystal iv) SHELL solution and v) SHELL solution containing known amount of sulfide ion concentration.

RESULTS AND DISCUSSION

The results of the SCC tests are shown in Fig 1 & 2 . The results revealed that the SCC occurred more readily in MS samples in NACE solution and SHELL solution saturated with H_2S at stress ranges of 70, 80, 85 and 90% YS. The alloys 430 and Monel 400 are also found susceptible to SCC only when they are stressed to 90% YS in NACE solution containing H_2S. The alloys 316L and 317L were found immune to SCC in all the condition of tests in NACE or SHELL solutions containing H_2S.

The fracture of SCC tested samples of MS and alloy 430 were analyzed in Scanning Electron Microscopy (SEM). The fractures revealed intergranular as well as intragranular mode of crack propagation (Fig. 3 & 4). Branching of secondary cracks from the primary cracks, which is typical of SCC failure mode were noticed. The Energy Dispersive Spectrum obtained during SEM fractography at fracture crack tips containing corrosion products showed sulfur rich regions (Fig 3b & 4b). This also confirm the onset of SCC due to sulfide stress cracking.

The threshold stress for MS in NACE and SHELL solution is 75% YS, whereas it is 85% for 430 and Monel 400 alloys in NACE solution. Such threshold stress was not found to exist for the same alloy in SHELL solution at all the stress levels. The alloys 316L and 317L, however, did not show any threshold limits upto 90% YS either in NACE or SHELL solutions containing H_2S. It is possible that threshold stress might be greater than the YS of these materials. MS has shown greater susceptibility to SCC in the tested medium when compared to other alloys.

Sulfides play a dominant role in the structural steels particularly in their resistance to sulfide stress cracking. The water present would obviously assist a corrosion mechanism. The reaction will be of the type

$$H_2S + Fe \rightarrow FeS + 2\,H$$

The nascent hydrogen is then expected to embritle the alloy. The presence of chloride all the more aggravate corrosion leading to early failure of steel. The instances of sulfide stress cracking of stainless steels have been reported [16] wherein failure of specimens have been promoted in high chloride environment (>25% NaCl) at elevated temperature and pressure saturated with H_2S. The chloride content used in some of the test being 5% , it is conceivable that the stainless steel of type 316L and 317L are less likely to be affected by H_2S as evidenced in the experiment. The effect of H_2S in SHELL solution suggest that except MS, other alloys were immune to SCC. MS was found prone to SCC > 75% YS. The synergistic effect of chloride in presence of oxidant (CH_3COOH) and H_2S to promote SCC in alloys 430 and Monel 400 at stresses > 85% YS was clearly demonstrated as seen from the results of NACE solution experiment (Fig. 1).

The data obtained from electrochemical polarization studies are shown in Table 2 & 3. The results obtained from NACE solution and natural seawater containing 0.1M of sulfide indicated that alloys 316L and 317L showed higher passive current densities relative to the Monel 400. The passive current density in natural seawater solution for alloy 430 with 0.1M sulfide was lowest which can be attributed to the development of a stable passive film over the surface of the alloy.

The electrochemical data from the SHELL solution revealed that lowest passive current densities for 317L alloy in 0.1M sulfide solution, while the highest was observed for alloy Monel 400 except MS. In general, for all the alloys studied, high sulfide content moved the corrosion potential to active direction thus enhancing localized corrosion. Lowest passive current densities exhibited by alloys 316L and 317L indicated that they are least susceptible to corrosion in presence of sulfide.

From the results of electrochemical tests it is seen that the synergistic effect of chloride and sulfide on the corrosion behavior were prominent particularly for alloys 316L and 317L. For alloys 430 and Monel 400 such effect were not noticed. However, under the influence of stress as noticed from the SCC test results, the trend was reverse. It is plausible that the passive films formed over the alloy 430 and Monel 400 were less stable and get disrupted easily, leading to SCC.

CONCLUSIONS

The following conclusions are drawn on the basis of investigations carried out.

i) In solutions containing sulfide, the chlorides demonstrated the synergistic effect to promote SCC in alloys 430 and Monel 400 at stress levels > 85% YS.

ii) Alloys 316L and 317L were found SCC resistant under all conditions of the tests performed.

iii) Alloys 316L and 317L had shown higher passive current densities relative to the other alloys in presence of specified oxidant, chloride and sulfide ionic species. Under the influence of stress, they were least susceptible to corrosion.

iv) MS was found prone to SCC at stress levels > 75% YS in solutions containing specified amounts of sulfide, chloride and oxidants.

v) Sulfide ion displaces the corrosion potential in active direction thereby increasing the risk for localized corrosion for all the alloys studied.

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Table 1 Chemical Composition of Alloys

S. No.	Alloy	UNS No.	Composition (%)								Others
			Fe	Cr	Mo	Ni	C	Cu	Mn	Si	
1	Mild Steel	J2503	Bal	0.5 Max	0.2 Max	0.5 Max	0.25 Max	0.3 Max	1.2 Max	0.6	0.04P,0.04S
2	316L	S31603	Bal	16.0	3.0	11.0	0.02	0.2	1.0	1.0	0.04P,0.02S
3	317L	S31703	Bal.	18.5	3.2	13.5	0.02	-	1.0	-	0.08 N
4	430	S43000	Bal.	18.0	-	-	0.12	-	1.0	1.0	0.04P,0.03S
5	Monel 400	N4000	2.5	-	-	66.5	0.3	Bal.	-	0.5	0.024S

Table 2 Potentiostatic Polarization data from NACE Solution and Natural Seawater (NSW).

S.No.	Material	Electrolyte	E corr (mv)	I corr (μA/cm²)	CR (mpy)
1	316L	i) NACE Solution	-59.6	1.53	0.67
		ii) NSW	-216	0.32	0.13
		iii) NSW + 0.064 M Sulfide	-348	0.31	0.137
		iv) NSW + 0.1 M Sulfide	-389	24.64	10.84
2	317L	i) NACE Solution	-212	0.58	0.26
		ii) NSW	-261	0.31	0.13
		iii) NSW + 0.064 M Sulfide	-383	0.39	0.17
		iv) NSW + 0.1 M Sulfide	-392	15.72	8.17
3	430	i) NACE Solution	-332	0.47	0.21
		ii) NSW	-65	0.16	0.06
		iii) NSW + 0.064 M Sulfide	-480	16.72	7.35
		iv) NSW + 0.1 M Sulfide	-503	1.62	7.12
4	Monel 400	i) NACE Solution	-207	9.53	3.7
		ii) NSW	-244	8.85	3.45
		iii) NSW + 0.064 M Sulfide	-315	1.61	0.63
		iv) NSW + 0.1 M Sulfide	-601	7.98	3.11

Table 3 Potentiostatic Polarization data from SHELL Solution

S. No.	Material	Electrolyte	E corr (mv)	I corr ($\mu A/cm^2$)	CR (mpy)
1	Mild Steel	i) SHELL Solution ii) SHELL Solution+0.1 M Sulfide	-705 -715	57.88 37.71	25.46 16.59
2	316L	i) SHELL Solution ii) SHELL Solution+0.1 M Sulfide	-100 -72	1.31 4.81	0.57 2.11
3	317L	i) SHELL Solution ii) SHELL Solution+0.1 M Sulfide	-62 -125	0.34 3.87	0.15 1.7
4	430	i) SHELL Solution ii) SHELL Solution+0.1 M Sulfide	-150 -215	0.93 1.6	0.41 0.7
5	Monel 400	i) SHELL Solution ii) SHELL Solution+0.1 M Sulfide	-492 -450	9.96 11.89	3.88 4.64

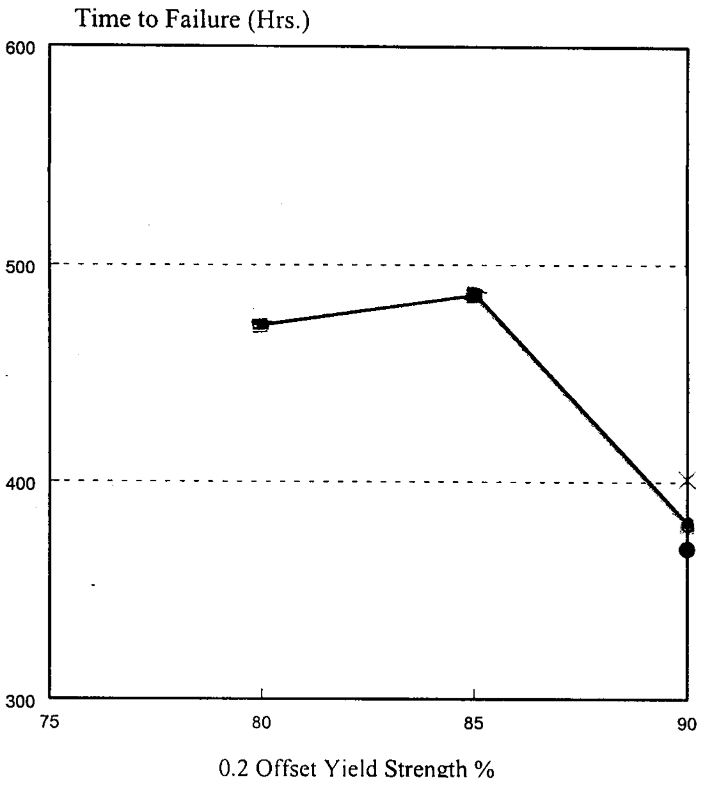

Fig SCC of alloys in NACE Soln. Containing Hydrogren Sulfide

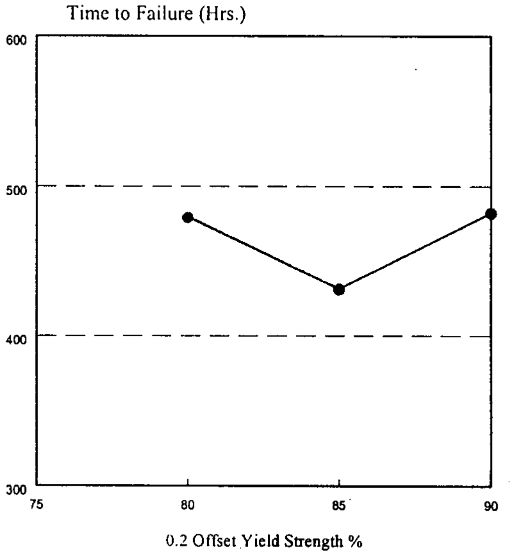

Fig.2 SCC of Alloys in SHELL Solution Containing Hydrogen Sulfide

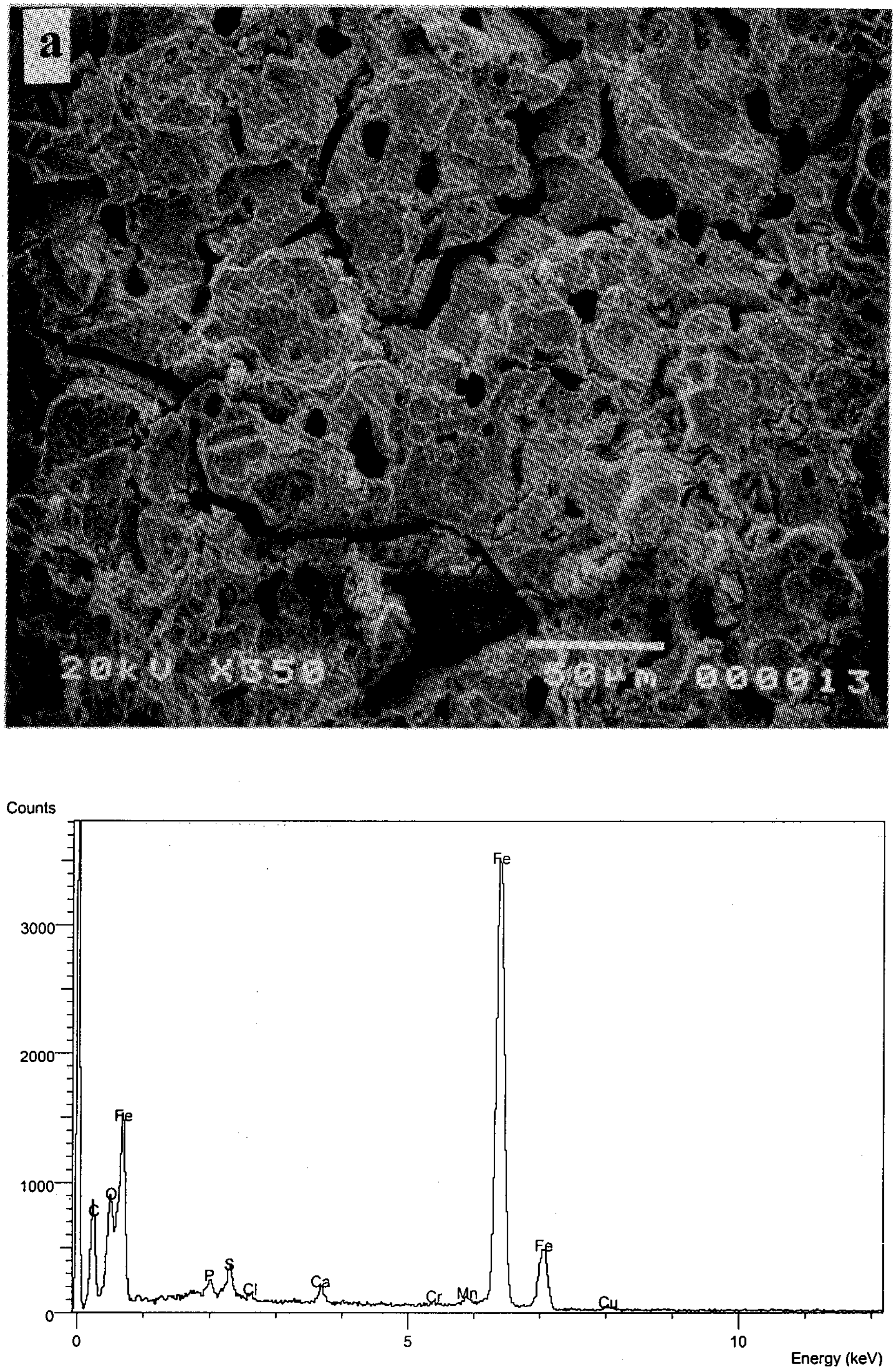

Fig. 3 SEM Fractrograph of SCC tested MS sample.
a) Fractrograph showing intergranular and intragranular fracture modes
b) EDAX spectrum taken at crack tip.

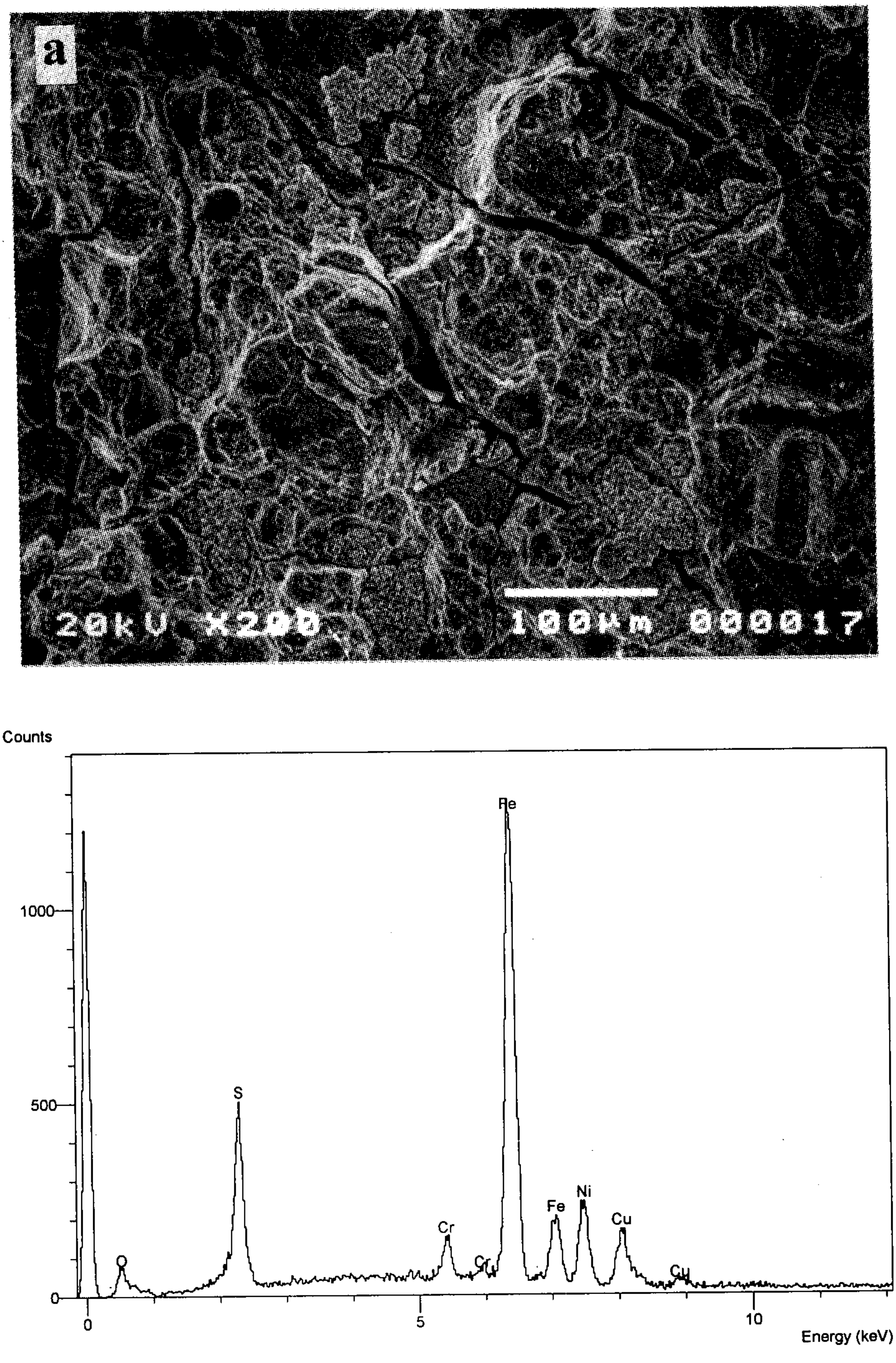

Fig. 4 SEM Fractrograph of SCC tested 430 alloy sample.
a) Fractrograph showing intergranular and intragranular fracture modes.
b) EDAX spectrum taken at a crack tip

Studies on Corrosion of Carbon Steel in Deaerated Saline Solutions in Presence of Scale Inhibitor

Ismaeel Andijani and S. Turgoose

STUDIES ON CORROSION OF CARBON STEEL IN DEAERATED SALINE SOLUTIONS IN PRESENCE OF SCALE INHIBITOR

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ABSTRACT

An investigation has been carried out to evaluate the corrosion behaviour of carbon steel in deaerated 1.0 M NaCl solutions and artificial seawater at pH ranges of 3 to 8.5 and temperatures up to 80°C under static conditions, using the polarization resistance technique. The effect of scale inhibitor addition on the corrosion rates has also been studied.

It has been found that the polarization resistance of carbon steel in deaerated solutions is generally high. With increasing pH, the polarization resistance increases, and is highest at 25°C and lowest at 80°C.

In long-run measurements, for 50°C at pH 8.5 the corrosion rates gave lower values compared to short-run measurements under the same conditions. This reduction in the corrosion rate is due to the formation of a protective hydroxide film on the metal surface, which restrict access of H_2O to the surface.

The results obtained with addition of 20 ppm phosphonobutanetri- carboxylic-acid scale-inhibitor at 50°C at pH 8.5, showed that there was no effect on the R_P values.

Key words: Carbon Steel, Desalination, Dissolve oxygen, Corrosion rate, Inhibitor, Polarization,

INTRODUCTION

Production of water by desalination of seawater is the prime source of drinking or potable water in the Arabian Gulf Countries. Multi Stage Flash evaporation (MSF) is the main process used for production of desalinated water. Carbon steel and wide variety of materials of diverse properties are used in MSF plants. A seawater desalination plant offers numerous corrosion problems due to its process conditions, including factors such as temperature and pH, and operation in relatively aggressive environments consisting of deaerated seawater, seawater-air and salt-air aerosols, corrosive gases, slow moving or stagnant liquids or deposit forming liquids [1]. To control the scale formation, acid treatment and anti-scale additive are used in desalination plants. To reduce the corrosion rates, removal of oxygen from the seawater by physical and chemical methods is used.

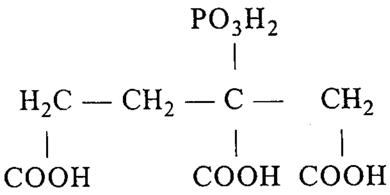
Therefore, mechanistic studies are required, which can provide experimental data and good understanding of the corrosion behaviour of metallic materials in these environments.

Some studies involving the use of a range of materials have been carried out under actual service conditions in desalination plants [2-6]and in laboratory conditions [7,8] using seawater and deaerated seawater at temperatures up to 90°C with different ranges of pH.

This work is concerned with the electrochemical study of carbon steel corrosion in deaerated 1.0 M NaCl solution and artificial seawater with and without addition of 20 ppm phosphonobutanetri- carboxylic-acid scale-inhibitor at different pH and temperatures under static conditions.

EXPERIMENTAL PROCEDURE

Specimens used in the polarization resistance measurements were cut from a mild steel cylindrical rod of 0.785 cm^2 cross section area. The electrodes were approximately 1.5 cm long, covered with adhesive lined hot shrink tube for 50° and 80°C test solution to prevent water entering between electrode and araldite resin, due to expansion of the resin with heat, and consequently to avoid crevice corrosion. A conducting wire was spot welded to each piece of metal; then the wires were insulated with PVC tubes of 6 mm in diameter in order to avoid electrical contact between the wires and the electrolyte, and to hold the electrode during the test. The electrodes were then mounted in araldite setting resin. The specimens were prepared by polishing with a 600 grit SiC abrasive paper, then rinsed with distilled water and dried in a stream of air. The corrosive media are 1.0 M NaCl solution, prepared by dissolved 58.5g of NaCl salt in 1dm^3 distilled water, and artificial seawater. The pH of solutions was adjusted to different range values from pH 8.5 to pH 3.0 with dilute HCl and NaOH solutions. Phosphonobutanetricarboxylic-acid scale-inhibitor, PBTC, with commercial name of BAYHIBIT,



$$\begin{array}{cccc} & & PO_3H_2 & \\ & & | & \\ H_2C\ — & CH_2\ — & C\ — & CH_2 \\ | & & | & | \\ COOH & & COOH & COOH \end{array}$$

was added to the test solutions. Deaeration was achieved by continuous purging N_2 gas (oxygen free) through the sealed test cell.

Polarisation resistance data were obtained using a standard three-electrode arrangement in 1 litre of deaerated test solution, in conjunction with an EG&G model 273 potentiostat/galvanostat and a chart recorder. A saturated calomel electrode and platinum electrode was used as reference and counter electrodes, respectively. The current-potential curve of carbon steel electrode in deaerated solution with and without scale-inhibitor was measured in the region of ± 10 mV around E_{corr} under static condition. The slope, DE/DI, at the corrosion potential, is defined as polarisation resistance (R_P). The corrosion current density is $I_{corr} = B / R_P$ where B is the Tafel slopes determined from the E vs. $\log_{10}$ I plot. The measurements were repeated at three different temperatures, 25, 50 and 80°C. Before starting the experiments, the specimens were left in the test solution for about 1 hour to attain a steady state which was shown by a constant potential. The scan rate of applied potential was 0.25 mV/sec.

The short run measurements were carried out by successively adjusting the pH and measuring corrosion rate on the same electrode. The entire experiment took only a few hours. The long run measurements were carried out by taking readings at different time intervals at fixed temperature over 24 hours. The pH was monitored over this period.

EXPERIMENTAL RESULTS

Figure 1 shows the results of polarization resistance measurements of carbon steel in deaerated 1.0 M NaCl solution at temperatures of 25, 50 and 80°C and over a pH range of 8.5 to 3.0.

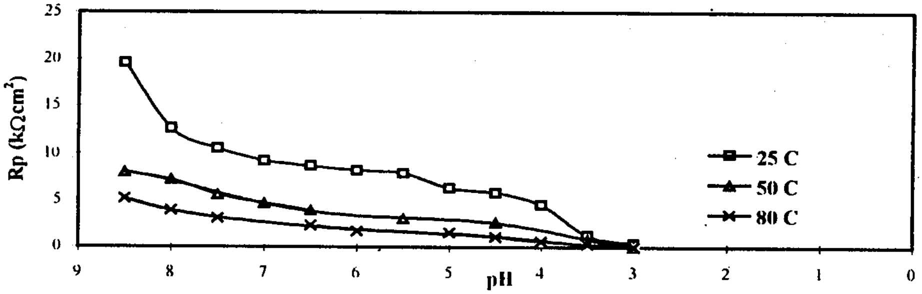


Fig.1 Polarization resistance measurements of carbon steel with pH in deaerated 1 M NaCl solution

At 25°C, the results indicate high R_P values in the neutral pH range and which then decrease with decrease in the pH values, while corrosion potential increases with decrease in the pH. Same phenomena observed at 50 and 80°C as summarised in tables (1,2,3).

Table (1) Short-run Measurements Results of Carbon Steel in Deaerated 1 M NaCl Solution at 25°C

	pH						
	8.5	8.0	7.5	7.0	6.5	6.0	3.5
E_{corr}, (mV)	-785	-780	-777	-775	-772	-770	-750
R_P, (kΩ.cm^2)	19.6	12.6	10.5	9.2	8.7	8.2	1.2
I_{corr}(µA/cm^2)	1.27	1.98	2.38	2.71	2.87	3.05	20.8
CR (mm/y)	0.013	0.02	0.024	0.027	0.029	0.031	0.21

Table (2) Short-run Measurements Results of Carbon Steel in Deaerated 1 M NaCl Solution at 50°C.

	pH						
	8.5	8.0	7.5	7.0	6.5	6.0	3.5
E_{corr}, (mV)	-810	-800	-787	-780	-774	-770	-745
R_P,(kΩ.cm^2)	8.0	7.2	5.6	4.7	3.9	_	0.78
I_{corr}(µA/cm^2)	3.12	3.47	4.46	5.32	5.41	_	31.8
CR (mm/y)	0.031	0.035	0.045	0.053	0.064	_	0.32

Table (3) Short-run Measurements Results of Carbon Steel in Deaerated 1 M NaCl Solution at 80°C.

	pH						
	8.5	8.0	7.5	7.0	6.5	6.0	3.5
$E_{corr,}$(mV)	-830	-810	-803	-798	-794	-790	-740
$R_{P,}$(kΩ.cm^2)	5.2	3.9	3.1	_	2.3	1.7	0.3
I_{corr}(μA/cm^2)	4.81	6.4	8.06	_	10.87	14.71	83.3
CR(mm/y)	0.048	0.064	0.081	_	0.11	0.15	0.83

Table 4 shows the results of polarization resistance measurements with time of carbon steel in deaerated 1 M NaCl solution at 50°C and pH 8.5.

Table (4) Long-run measurements results of carbon steel in deaerated 1.0 M NaCl solution at 50 ^{0} C and pH 8.5.

Time, hr	1	2	3	4	21	22	23	24
I(μA/cm^2)	3.62	3.33	3.2	3	2.58	2.56	2.43	2.4
E(mV)	-797							-818

The tables indicate that the Icorr values decreases with time and the Ecorr shifted to more negative value. Figure 2 shows the Tafel Plot of carbon steel in deaerated 1.0 M NaCl at 50 C and pH 8.5 under static conditions. The experiments were carried out at two different time intervals e.g., 1 hr and 24 hrs.

Fig. 2 Potentiodynamic polarization curves of carbon steel in deaerated 1 M NaCl at different immersion interval

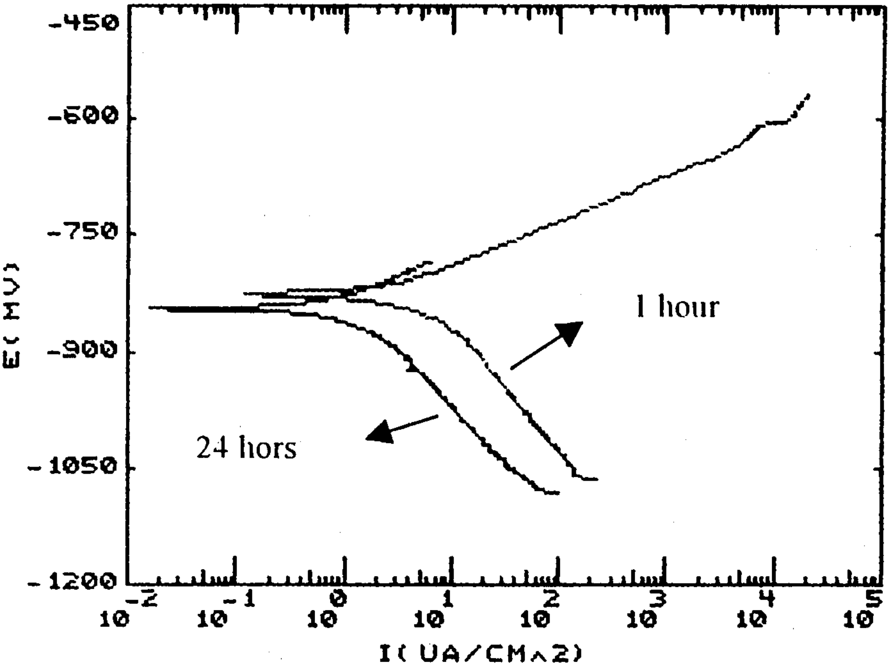


Figure 3 shows the results of polarization resistance measurements with time of carbon steel in deaerated 1.0 M NaCl solution and artificial seawater with and without addition of scale-inhibitor at 50°C and pH 8.5. The objective of this experiment is to distinguish, in terms of corrosion current of steel, between NaCl solution and artificial seawater and between inhibited seawater and uninhibited seawater in long term tests.

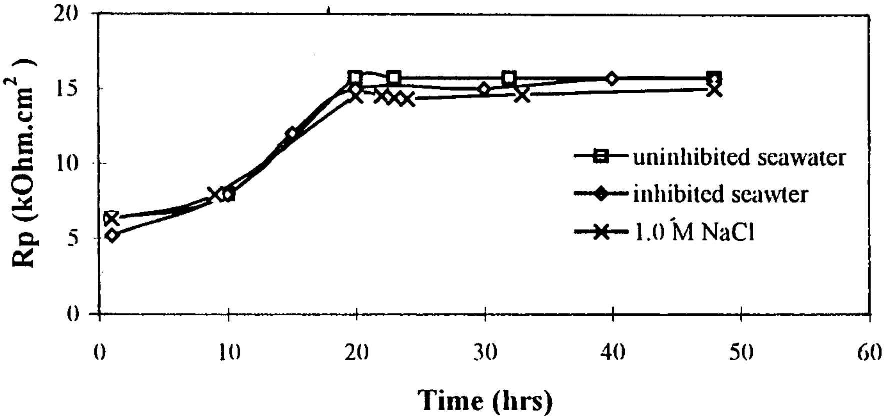


Fig. 3. Polarization resistance measurements of carbon steel with time in deaerated solution

DISCUSSION

The mixed potential theory was derived originally to explain experimental electrochemical laboratory measurements. The useful of E-log I diagrams is that they combine thermodynamics and kinetics to form a whole picture: the potential axis is the thermodynamic factor and the current axis is the kinetic factor. Obviously, the rate of corrosion of steel in deaerated solution is controlled by such factors as:

(i) the anodic polarization line of steel.

(ii) the cathodic polarization line of water reduction in the neutral and slightly alkaline solution.

(iii) the cathodic polarisation line of hydrogen reduction in the acidic solution.

Since water reduction is the only cathodic reaction in the deaerated neutral solution therefore, the corrosion rate depends on how rapidly H_2O diffuses to the metal surface through the barrier of oxide film and on diffusion of Fe^{2+} out through the film. Thus any change in temperature, bulk concentration and disturbance of the oxide film will affect the reaction rate.

Within the acid region of the solution the oxide film is dissolved, the steel surface is in direct contact with the aqueous solution and hence the increased rate of reaction is due to the higher rate of hydrogen evolution. Increasing hydrogen ion concentration in the neutral and slightly alkaline solution i.e., 8.5 - 6.5 the equilibrium potential of the hydrogen electrode becomes more noble, according to this relation $E_{H/H}^{+} = 0.0 - 0.059pH$, although $I_{oH/H2O}$ remain constant, hence $E_{H/H}^{+} - E_{Fe/Fe}^{2+}$ becomes larger and the resulting corrosion current increases. Graphically this may be represented in Evans diagram as shown in Fig. 4

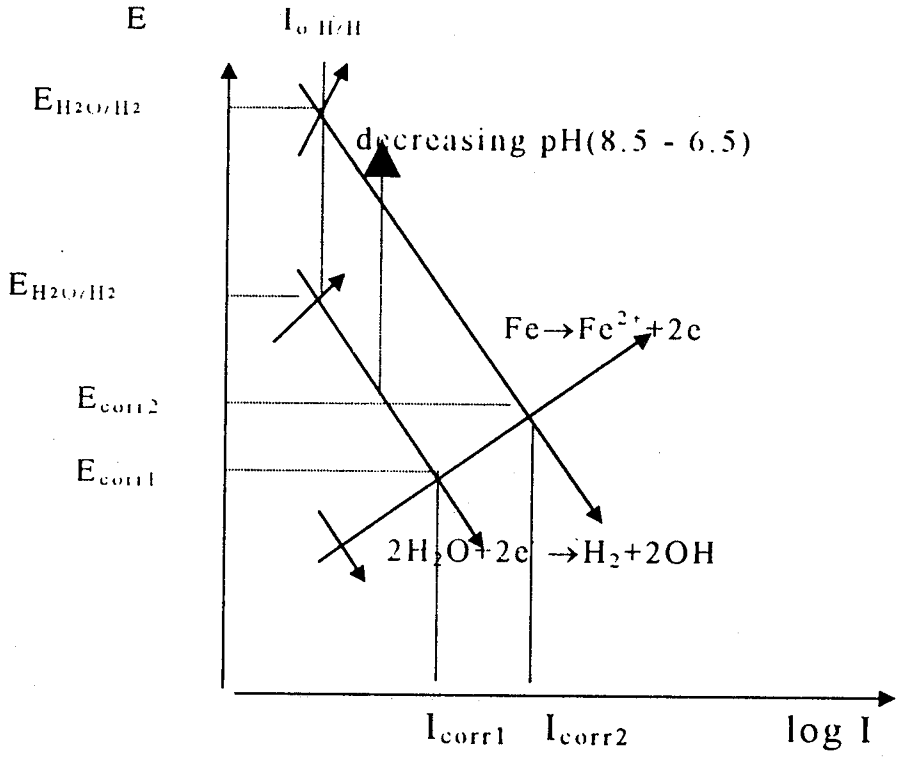


Fig. 4 Evans diagram for pH range of 8.5 and 6.5 for carbon steel in deaerated 1 M NaCl solution at 50°C.

This explains why the corrosion rate of carbon steel in deaerated NaCl solution increases with decreases in pH at neutral and slightly alkaline solution and also E_{corr} shifts to more positive values. When the solution become more acidic, the equilibrium potential of the hydrogen electrode becomes more noble and the exchange current density of hydrogen on steel, $I_{oH/H}^{+}$ on Fe, increases and hence the corrosion rates increases compared to neutral solution. Schematically:

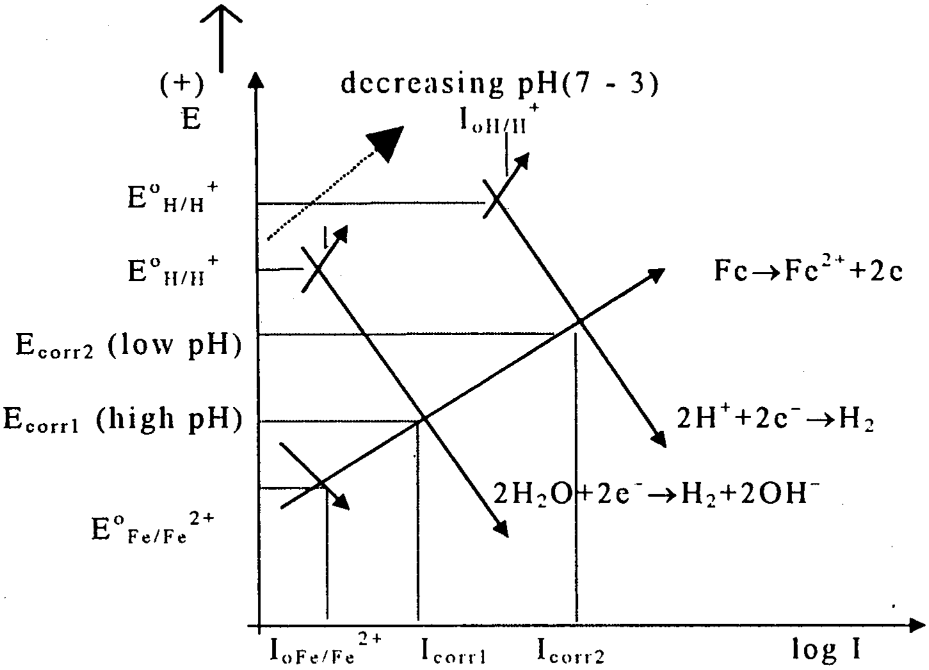


Fig.5 Evans diagram for pH range of 7 and 3 for carbon steel in deaerated 1 M NaCl solution at 50°C.

This explains why the corrosion current of steel in deaerated NaCl solution at 25°C (Table 1) was increased slowly at pH 8.5 to pH 6.0 (1.27 to 3.05 μA/cm^2) where at pH 6.0 to pH 3.5 the corrosion current was increased drastically (3.05 to 20.8 μA/cm^2). At increased temperature, 50 and 80°C, the velocity of corrosion reaction is greater than at 25°C due to $I_{oFe/Fe}{}^{2+}$ increasing at elevated temperatures. However, the relationship between exchange current, $I_{oFe/Fe}{}^{2+}$, and temperature is given by [9]

$$I_{oFe/Fe}{}^{2+} = n.F.f.c^n .\alpha. \frac{N_s}{N_o} \exp\left[\frac{-\Delta G}{RT}\right]$$

where:

n = number of electron.
F = Faraday constant.
f = the frequency of thermal vibrations, ~ 10^{12} Hz, which determine frequency with which atoms exchange thermal energy.
c^n = the concentration of solvating species in the environment and n the number of solvating molecules associated with each cation.

α = the proportion of atoms on a metal surface located at link site, 10^{-3}, it depends on whether the metal is annealed or coldworked.

N_s/N_o = the number of moles of metal per unit area of surface, where N_s is the number of metal atoms per unit area of surface (typically $\sim 2\times10^{13}$ atoms/mm^2) and N_o is the number of atoms/mole (Avogadro,s number, $\sim6\times10^{23}$ atoms/mole).

ΔG = electrochemical activation energy.

R = perfect gas constant.

T = absolute temperature.

Increasing the temperature of the solution will increase $I_{O\ M/M^{n+}}$ of both electrodes, hydrogen reduction reaction and iron oxidation reaction and hence corrosion rate increases.

Increase in temperature also affects the reversible equilibrium potential, $E^O_{M/M^{n+}}$ of both electrodes. The relationship between temperature and reversible equilibrium potential is given by

$$\Delta G^O = n\ F\ E^O$$

$$\Delta G^O = -\ R\ T\ \ln K_{eq}$$

$$E^O = -\ \frac{RT \ln K_{eq}}{nF}$$

With increasing temperature, the reversible equilibrium potential becomes more negative.

Figure (6) shows the temperature influence on the polarisation curves of carbon steel in deaerated 1.0 M NaCl solution at pH 8.5. As the temperature increases from 25 to 80^O C, the E_{CORR} shifts to more negative values and I_{CORR} increases.

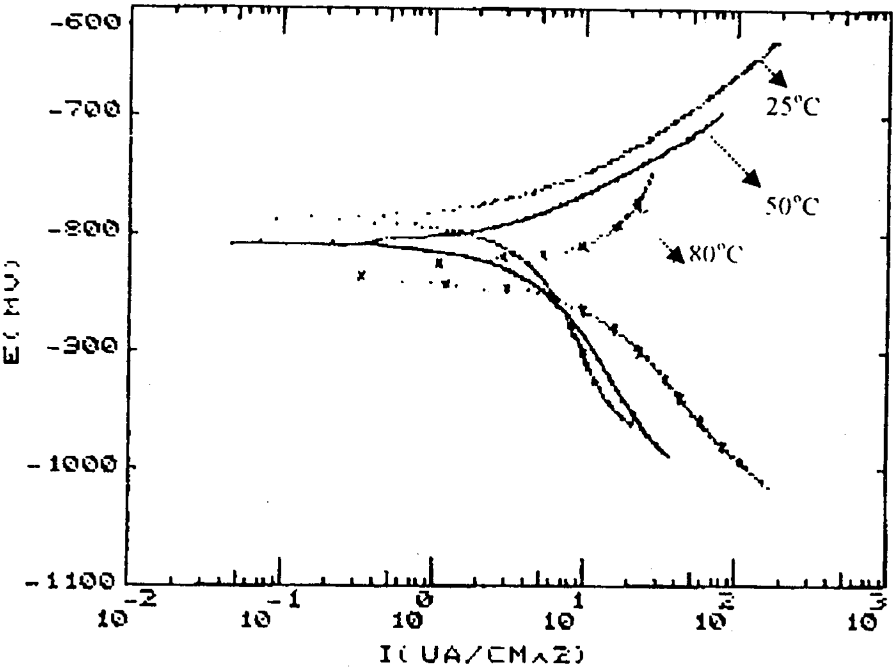


Figure 6. Potentiodynamic polarisation curves of carbonsteel in deaerated 1.0 M NaCl solution at pH 8.5 and different temperature under static condition

Thus increased rate of reaction at increased temperatures from 25 to 80°C is due to the kinetic change of anodic and cathodic reactions. E_{CORR} in general had more negative value at 80°C compare to 50°C and 25°C. at pH 8.5.

Graphically this may be represented in Evans diagram as shown in Fig. 7

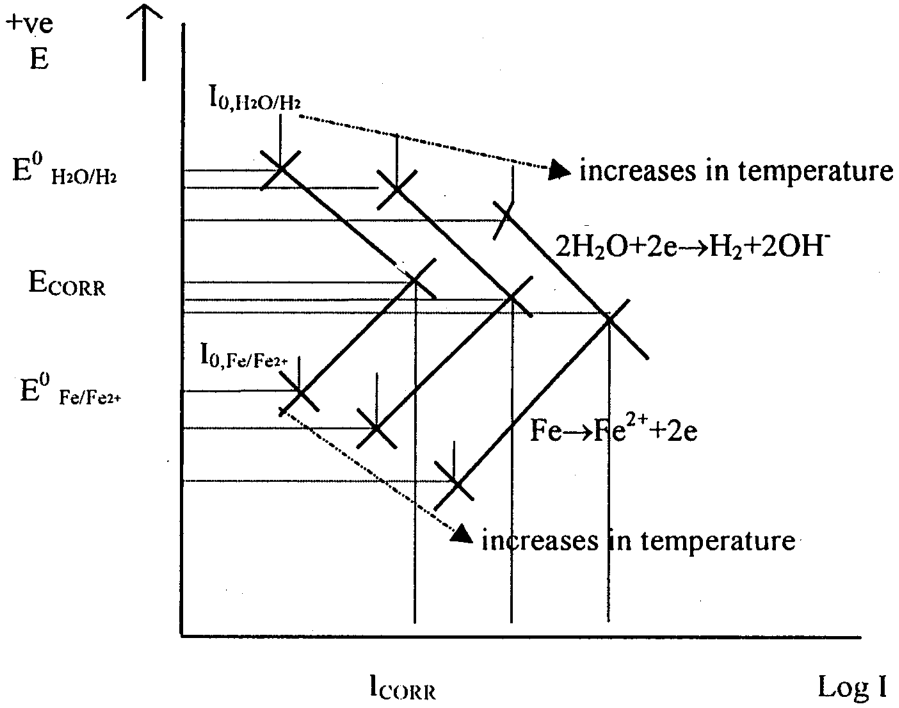


Fig. 7 Evans diagram for carbon steel at temperature 25, 50 and 80°C in deaerated 1.0 M NaCl solution at pH 8.5 under static condition.

This diagram illustrates the effect of increase in temperature on the anodic and cathodic partial reaction. With increase in temperature, $I_{0,Fe/Fe2+}$ increases and $E^0_{Fe/Fe2+}$ becomes more negative, the resulting anodic curve shifts to down right. Similarly, the cathodic curve, with increase in temperature also shifts to down right. Hence, the resulting corrosion current increases and E_{CORR} becomes more negative.

Deaerated Seawater

The results of corrosion behaviour of steel in seawater showed no significance difference compared to NaCl solution in same condition as shown in figure 3. Usually seawater has tendency to form adherent scale deposit on the metal surface at elevated temperature that isolates the metal from the seawater and hence corrosion decreases. But no such case occurred here.

Effect of Addition of Scale-inhibitor

Addition of scale-inhibitor basically is to distort the crystals of the scale-forming material at elevated temperature. The experimental results (figure 8) indicated that the influence of scale-inhibitor on the corrosion rates of the steel was mostly on the cathodic reaction. Decrease of the pH of the solution and consequent corrosion current increase when scale-inhibitor was added to the solution with the increase in E_{corr} means that the corrosion rate is under cathodic control, because when the pH was adjusted to the initial value, the corrosion current was almost the same as without inhibitor. Thus, addition of phosphonobutanetricarboxylic-acid scale-inhibitor at constant pH has almost no effect on the corrosion rates.

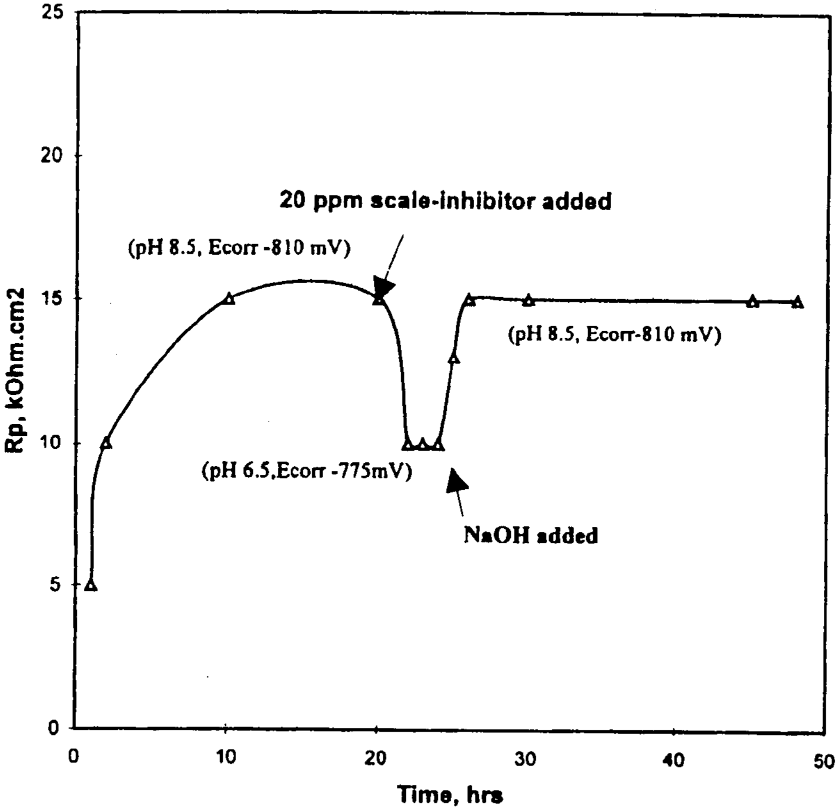


Fig. 8 The addition effect of scale-inhibitor on the corrosion behaviour of carbon steel in deaerated solution at pH 8.5 and 50°C.

Long-run Test

The results of these tests have confirmed that the corrosion rates of carbon steel tend to reduce in seawater and NaCl environments at 50 °C and pH 8.5. This has been explained on the basis of formation of a protective hydroxide film as represented by the following electrochemical reactions: Under deaerated condition, the anodic reaction is represented as follows:

$Fe \rightarrow Fe^{2+} + 2e$ (1)

The cathodic reaction at neutral pH is

$2H_2O + 2e \rightarrow H_2 + 2OH^-$ (2)

Fe^{2+} hydrolysis to form $Fe(OH)_2$

$Fe^{2+} + 2H_2O \rightarrow Fe(OH)_2 + 2H^+$ (3)

Under deaerated condition, the corrosion product, $Fe(OH)_2$ formed on metal surface increases with time but it also impedes the cathodic reaction (2) resulting in lowering of the current density. When sufficient amount of the deposit is formed on the steel surface, the current subsequently becomes constant.

From the Tafel Plot of carbon steel in deaerated 1.0 M NaCl at 50°C and pH 8.5 under static conditions, the study of the cathodic curves indicates that Icorr shifts towards left (to low current value) at 24 hrs and Ecorr is also shifted to less noble value. This behaviour can be represented by Evans diagram:

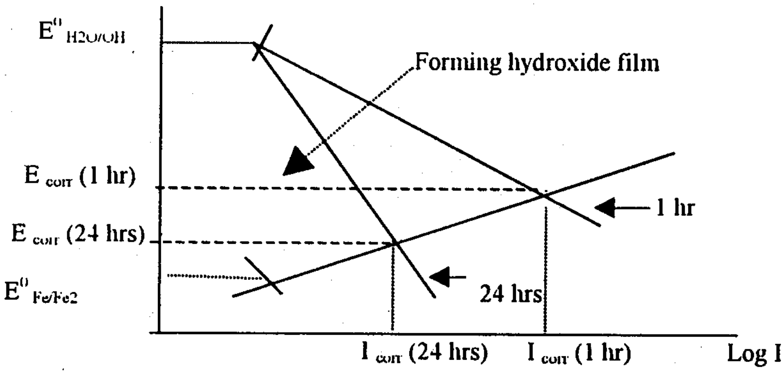


Fig 9 Evans diagram at different immersion interval of carbon steel in deaerated 1 M NaCl solution at 50°C and pH 8.5.

The diagram shows clearly the shifting of I_{corr} to lower value with increase in time. Thus after 24 hrs there will be more formation of $Fe(OH)_2$ than that after 1 hr and this will result in lower value of I_{corr}.

CONCLUSIONS

In the present study, the corrosion behaviour of carbon steel in deaerated 1 M NaCl solution and artificial seawater were studied by polarisation resistance measurements, the following characteristics were found:

- The corrosion rates were very low at 25°C and neutral pH.
- With decrease in pH the corrosion rates increase and the corrosion potential increase [more noble].
- With increase in temperatures [25° to 80°C] the corrosion rates increase [$(0.024\ mm/y)_{pH7.5}$ to $(0.081\ mm/y)_{pH\ 7.5}$] and the corrosion potential decreases (less noble).
- In the neutral and slightly alkaline solution, H_2O diffusion to the metal surface through the barrier of oxide film is the rate determining. While in the acidic solution, the oxide film is dissolved, the rate of hydrogen evolution alone determines the reaction rate.
- The addition of phosphonobutanetricarboxylic-acid scale-inhibitor has no significant effect on the corrosion rates.
- In long-run measurements, at 50°C and pH 8.5, the corrosion rates gave lower values compared to short-run measurements at the same conditions. This reduction in the corrosion rate is due to the formation of a protective hydroxide film on the metal surface which restrict access of H_2O to the surface

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Drinking Water

Comparative Study of Potable and Bottled Mineral Waters Available in the State of Kuwait

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Hamad Al Ajmy

COMPARATIVE STUDY OF POTABLE AND BOTTLED MINERAL WATERS AVAILABLE IN THE STATE OF KUWAIT

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ABSTRACT

Potable water in Kuwait is produced by desalination of seawater in multistage flash desalination (MSF) plants. In this study, the quality of potable water in Kuwait from different sources and stations along the distribution system to the end consumers was examined. Additionally, the quality of bottled mineral waters available in Kuwait was also examined. More than 20 types of bottled mineral water were studied. Both physical properties and chemical constituents were included in this study. Major constituents and trace elements, especially toxic elements, were determined for all water samples, and the results compared with the 1993 World Health Organisation Guideline Values. The results showed that all water samples tested complied with W. H. O. Guidelines. Kuwait's public water supplies showed results in medium ranges for most of the mineral contents, major and minor parameters, compared to bottled mineral water

Key words: potable - ground - mineral - major - trace

INTRODUCTION

Kuwait is located in an arid area, which is characterised by severe weather conditions, very low rainfall and absence of rivers and lakes. It has very limited underground water resources, which are not sufficient in quantity and of variable quality. The State of Kuwait obtains potable water through the desalination of seawater. Five dual-purpose power and desalination plants are located at Shuwaikh, Shuaiba, Doha East, Doha West, and Az-Zour. All these plants are multistage flash (MSF), producing distilled water, which is then blended with 5-10 % brackish water and pumped to the distribution system. The distribution system consists of underground reservoirs, elevated storage towers, pipelines and filling stations for road tankers to distribute potable water to areas of the country not supplied by water.

However, some inhabitants of Kuwait use bottled mineral water. Most bottled mineral waters are groundwater, bottled with or without some treatment process such as filtration and sterilisation. Bottled waters must conform to suitable water quality specifications. The bottled waters studied were supplied from different countries, especially countries of the GCC, as shown in Table 1 & Figure 1.

Naturally, groundwater results from waters that infiltrate from the land-surface and percolate to the underlying strata, (Wanielista, 1990). For groundwater to be safe for human consumption, it must be free from organisms that are capable of causing disease, and also from minerals and organic substances that could produce adverse physiological effects. Drinking water should be aesthetically acceptable; it should be free from apparent turbidity, colour and odour and from any objectionable taste. Drinking water also should be at a reasonable temperature. Water meeting these conditions is termed "potable", meaning that it may be consumed in any desired amount without adverse effects to health (Cotruvo and Vogt, 1990).

Such quality is generally defined by a series of properties related to the physical, chemical and biological properties of the water. These variables have evolved over time and are constantly being modified and updated in line with the expanding uses to which water is put. With the development of analytical capabilities to measure a greater number of substances at ever smaller concentrations (GEMS, 1987), these parameters were collated and published by the World Health Organisation in 1984, and again in 1993, as Guideline Values for Drinking Water Quality (Table 2).

However, water may contain substances, whether natural or anthropogenic, that can affect the quality and existence of life. It is important to distinguish between pure water and safe water. Pure water may be defined as water that is free from all extraneous substances, whether harmless or not, and, (from a practical standpoint), is impossible to produce. On the other hand, Safe water is water that is not likely to cause undesirable or adverse effect, although it may contain certain contaminants. It should be emphasised that drinking water should be clean and safe, and that minute quantities of contaminants present in water should meet the drinking water Guideline Values set by the World Health Organisation, or U.S. Environmental Protection Agency standards for protecting public health (Wang, 1984).

Although the concern for water quality is not new, progress has been made in moving from sensory associations as a means of control, to the application of knowledge and criteria gained from scientific advances in detection and measurement, and from a better understanding of the characteristics of water. From the standpoint of quality, natural water resources, especially groundwater, show wide variations, from low to high concentration of salt constituents, as well as other physical, chemical and biological features (IHD-WHO, 1978).

Many inorganic compounds, especially metal ions, play dual roles in the physiology of the organism; some are essential for life, while most of them are toxic at elevated concentrations. Recent years have brought an increasing concern for the potential toxic effects of metal ions and other inorganic compounds, which constitute part of the products and by-products of recent technologies (Seiler et al, 1988).

There are five behaviour categories for substances in living systems. Some of them are considered as essential nutrients, where deficiency causes impairment of physiological functions that can be relieved by administration of that substance. There are many instances of both essential and nonessential compounds behaving as stimulants, causing stimulation of some metabolic processes, while others are inert or innocuous, such as tantalum, platinum, gold and silver. (These substances are therefore often employed as surgical implants.) There are numerous compounds that serve as therapeutic agents, such as the historical use of arsenic and mercury compounds against parasites, and the current employment of lithium in manic depression. Finally, at high concentrations most substances become toxic, they harmfully affect the activity of living organisms, perhaps irreversibly, in a manner that leads to loss of function, deformity and perhaps death. Depending on concentration and time, a single substance may act in more than one of the five ways in a single organism (Seiler et al, 1988).

Excessive levels of trace elements may occur naturally as a result of normal geological phenomena such as ore-formation, or anthropogenic sources.

These may be released by burning fossil fuel, mining, smelting, discharging industrial, agricultural and domestic waste, and by deliberate application of pesticides. Once made available to the environment, metals are not usually removed rapidly, nor are they readily detoxified by metabolic activity. As a result they accumulate. Thus, their release into the environment must be carefully monitored and controlled (Duffus, 1980).

The aim of this study was to monitor the quality of public potable water supplies, and to compare them with most of the bottwaters available in Kuwait.

MATERIAL AND METHODS

Samples were collected throughout the distribution network, which includes pump stations, reservoirs, towers and tanker filling stations

Water samples were collected according to the WHO recommendations, and the 1992 standard AWWA, APHA and WEF test methods were used

Sampling Sites

A total of 6 water samples were collected from different points in the public distribution system. They were included the main pumping stations of Kuwait's distribution system, Doha, Az-Zour and Shuwaikh, and also main reservoirs and towers, as well as one filling station. 25 brands of bottled water, especially those brands from Gulf countries, were purchased in the local market.

Water Handling

1 litre polyethylene bottles were used for collection of water samples; Water samples in these plastic bottles were used to determine the major constituents and trace elements. All samples were collected from the distribution systems, the taps being run for ten minutes before the sample was collected after rinsing the container two or three times with the water being collected. One bottle contained 1% HNO_3 as a preservative for trace element analyses. pH, salinity and conductivity were measured immediately after sampling. All sample bottles were stored under refrigeration at 4° C until analysis.

Analysis

Chemical analyses were carried out according to the 1992 "Standard methods of Test for Water and Wastewater", as follows:

- pH, salinity and conductivity: determined electronically.
- Total dissolved solids (TDS): determined gravimetrically.
- Alkalinity: determined titrimetrically.
- Chlorides, sulphates, and nitrogen compounds: determined by Auto-Analyser equipment.
- Trace elements: determined by inductive coupled plasma mass spectroscopy or atomic absorption spectrometry.

RESULTS

Table 3 shows the physical and chemical analyses of potable water in Kuwait's distribution system. This table shows that the physical parameters, which include the temperature, pH, electrical conductivity (EC), salinity, total hardness and alkalinity, were all in the acceptable range for aesthetically acceptable potable water. The total dissolved solids and major constituents (mg/l) of the water samples tested were within the appropriate W.H.O. Guideline Values.

Table 4 shows trace element analyses (μg/l) of water from Kuwait's distribution system. From this table, it can be seen that the trace element concentrations were all lower than the W.H.O. Guideline Values.
Table 5 shows the physical and chemical parameters for the bottled potable water samples, which were analysed. This table indicates that the physical parameters, e.g. temperature, pH, electrical conductivity (EC), salinity, total hardness and alkalinity, all were in the acceptable range for palatable potable water. The total dissolved solids and major constituents (mg/l) of the examined water samples were within the appropriate W.H.O. Guideline Values.

Table 6 shows trace elements (μg/l) of the bottled potable waters. As shown, the trace element concentrations were all lower than the W.H.O. Guideline values.

From Figure 2 it is clear that the mean values of physical and chemical analyses of Kuwait's public water supplies fall between the maximum and minimum values of bottled water, except for chloride contents.

Figure 3 compares the differences among the average values of trace metals analyses of Kuwait's distribution system with maximum and minimum values of bottled mineral water. From this figure, the arithmetic mean values of Kuwait's public water supplies were in between the maximum and minimum values for most trace elements, except for copper and nickel, which had higher values (Figs. 3C, 3E).

DISCUSSION

The results of the chemical analyses of the examined water, (Tables 3 & 5), show that the pH of both types of water was approximately neutral and had mid-range values of total dissolved solids and the major constituents. The mean pH values of water from the distribution system and bottled water samples were approximately the same, 7.84 and 7.85 respectively. The pH of the Kuwait distribution system samples was in the range of 7.4 to 8.01, while the pH of the bottled water samples was in the range of 6.58 – 8.41. The W.H.O. Guideline Value for pH is less than 8. About 44% of the bottled water samples had pH values higher than 8, and 8% samples were slightly acidic, less than 7. At the same time, the total dissolved solids (TDS) values were moderate, between 286 mg/l and 364 mg/l in the public water, (mean 323 mg/l), while it varied from 72.5 mg/l to 335 mg/l in the bottled mineral waters.

The alkalinity and bicarbonate contents in the water samples tested, (Tables 3 & 5) were in the range of 25.9 to 54.8 mg/l and 31.6 to 66.9 mg/l respectively, with mean values of 46.9 and 57.2 mg/l in the public distribution system. Tests on bottled water showed a wide range, from 15 to 263.1 mg/l and 18.2 to 319.4 mg/l respectively. No carbonate ions (CO_3^{2-}) were found in samples from the distribution system, but were found to be between zero and 7.1 mg/l in the bottled water samples. Chloride concentrations varied: they were moderate, in the range of (68-78 mg/l) in the municipal water, and showed moderate to low values in bottled water, in the range of 5.7 to 72.2 mg/l . Also, sulphate concentrations of the water samples were in the range of 46-107 mg/l in the municipal one and 5 to 89.2 mg/l in the bottled one.

Cation concentrations – e.g. sodium (Na), potassium (K), calcium (Ca), and magnesium (Mg) in the public water (Table 1A) were respectively 36 - 47.7, 1.38 - 3.2, 37.6 - 46.9, and 5.9 -10.8 mg/l, with means of 44.1, 2.1, 42.8 and 9.3 mg/l respectively. In the bottled water samples, these cation concentrations varied from very low to moderate values (Table 2A). The equivalent ranges were : (Na): 2.5 - 65.1, (K): 0.4 - 10.4, (Ca): 1.8 - 103.3, (Mg): 0.3 - 15.1 mg/l.

Previous studies, (Pocoke et al. (1981), Zielhuis (1981), and Rasmussen et al. (1987), show that there appears to be a relationship between Ischemic Heart Disease (IHD) and Acute Myocardial Infraction (AMI), and magnesium and calcium intake, as the incidence of these diseases are high in areas with low concentration of these elements in drinking water. Zielhuis, (1981) indicated that soft water with calcium content below 40 - 60 mg/l contributed to cardiovascular mortality, while Pocock et al. (1981), suggested that soft water areas (around 0.25 mmol/l, 20 mg/l Ca) had a 10-15% higher cardiovascular mortality than areas of medium hardness (around 1.7 mmol/l). Kuwait's municipal water had mean calcium content of 42.8 mg/l, with 144.8 mg/l mean hardness, while 76% of the bottled water samples had calcium-content less than 40 mg/l.

However, Calabrese and Tuthill (1981), concluded that elevated levels of sodium (Na) in drinking water contributed to elevation of blood pressure (BP). They revealed statistically significant and medically important higher blood pressure distribution among the high sodium areas (107 mg/l), relative to low Na (8 mg/l) areas, for both systolic and diastolic BP, both in boys and girls. Kuwait's public water supplies have median sodium contents in the range of 36 - 47.7 mg/l, while 12 % of the examined bottled water had higher values.

In this study, three nitrogen species, nitrates, nitrites and ammonia, were determined for all water samples (Tables 3 & 5). Ammonia and nitrites were not found in any water samples tested. Nitrate concentrations were low for all samples, in the range of zero to 2.4 mg/l in the municipal water samples, and zero to 6 mg/l in the bottled water samples. Toxic cyanide ions were not found in any water samples tested. A very important consideration is the fact that nitrates can be readily converted *in vivo* to nitrites as a result of bacterial reduction (WHO, 1984).

The silica concentrations in the samples tested showed low values for the public water supplies (Tables 3 & 5), with some high values in bottled water. The mean value in the public supplies water was 1.5 mg/l., but values between zero and 23.6 mg/l wfound in the bottled water samples. About 60% of the bottled water samples had higher values of silica content than were seen in the public water supplies. Generally colloidal silica and their degradation products are poisonous substances, which can react with human erythrocytes. Animal studies showed that high silica contents can cause fibrosis of the liver, enlargement of the spleen, and changes in the kidney resembling interstitial nephritis (Seiler et al., 1988).

The difference between mean values of physical and chemical analyses (Figure 1) of municipal water with maximum and minimum values of bottled water indicate that the mean values of Kuwait's public water supplies were in between the maximum and minimum values for most physical and chemical parameters, except for chloride. However, the maximum values of chloride ions in the bottled water was 72.2 mg/l, while the mean in the public water was 75.5 mg/l.

The trace metals analysed were all below the W.H.O. Guideline Values for Drinking Water (Tables 4 & 6). Lithium (Li), beryllium (Be), boron (B), aluminium (Al), vanadium (V), chromium (Cr), manganese (Mn), cobalt (Co), nickel (Ni), copper (Cu), zinc (Zn), arsenic (As), selenium (Se), molybdenum (Mo), cadmium (Cd), tin (Sn), antimony (Sb), barium (Ba), mercury (Hg), and lead (Pb) were tested in all samples. Mean and maximum values found are shown in the tables. Most of these parameters are toxic and have can severe effects on the human body.

Generally, comparison of the mean values of trace metal analyses for Kuwait public potable water supplies to those of bottled mineral water, (Figure 3), shows that the mean values of municipal Kuwaiti water were less than the Guideline Values for trace elements. The maximum values of copper and nickel contents in bottled water were 2.7 and 3.9 μg/l, while its mean values in the municipal water were 35.7 and 6.4 μg/l.

From a general point of view, some of the bottled water that was analysed had higher contents of trace metals than the public water supplies. However, all these concentrations were less than the 1993 World Health Organisation Guideline Values for drinking water. Some of these are not essential to the human body, but some show accumulative toxic effects and may be carcinogenic. The total contents of trace metals for municipal water were in the range of 36.2 – 164.6 μg/l, with mean of 95.6 μg/l , while that of the bottled mineral waters showed a very wide range of values, from 4 μg/l to 718.2 μg/l.

Pier (1984) classified the human requirements of metals into three groups:

1. metals required in substantial amounts, for which the body has a wide tolerance, e.g., iron

2. metals required in much smaller amounts, and for which the body has a quite narrow tolerance, e.g., the "micro-nutrients" copper and zinc

3. metals for which an essential role in the life processes has not been determined, and which are toxic at low levels, e.g., lead and mercury, Oehme (1978).

Generally, excessive levels of most trace elements have different effects, especially those, which have accumulative effects.

Hara et al. (1987) stated that chronic renal patients on long-term dialysis develop serious conditions due to the toxic effects of aluminium accumulation. Strenuous efforts are made to limit aluminium uptake for such patients from all sources to 30 μg/l. The concentration of aluminium in the public water supply was in the range of 1.4-19.1 μg/l, although sample No. 15 had an aluminium content of 139.7 μg/l, with total mineral content of 205 μg/l , while it had a lower content of total dissolved solids (72.5 mg/l).

CONCLUSIONS

1. The major parameters and trace metals of Kuwait's public water supplies fall within the W.H.O. Guideline values for Drinking Water.

2. The bottled waters showed considerable differences between samples, some had low values for different chemical parameters, while others had high values. However, all parameters were within the W.H.O. Guideline values for Drinking Water.

3. Kuwait public water supplies show medium ranges for most of the mineral contents, major and minor parameters, compared to bottled mineral water

4. In general, the study found that bottled mineral water is not better than the public water supply from a health viewpoint.

5. Careful selection should be followed for the major and minor contents of water used for special purposes – for example, kidney failure and cardiac patients.

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Table 1. Distribution of bottled water samples among countries

Country	No. of Samples	%
Kuwait	3	12
Saudi Arabia	10	40
Emarate	4	16
Bahrain	2	8
Oman	2	4
Qatar	1	8
others	3	12
Total	**25**	**100**

Fig1. Distribution of bottled water samples among countries

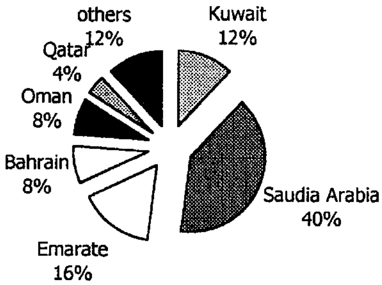

Table 2. World Health Organisation (WHO) Guidelines, 1993

Major constituents	**Guidelines limits (ppm)**	**Trace metals**	**Mam. Permessible (ppb)**
pH	<8.00	Pb *	10.00
EC	---	Cd *	3.00
TDS**	1000.00	Ni *	20.00
T.AlK	---	Cu	2000*
T.Hard.	---	Ba *	700.00
CO3	---	Mn	500*
HCO3	---	Sb *	5.00
CI **	250.00	Zn **	3.000.00
SO4 **	250.00	Sn	---
Ca	---	Se *	10.00
Mg	---	B*	300.00
Na **	200.00	Cr *	50.00
K	---	Hg *	1.00
SiO2	---	Mo *	70.00
CN *	0.07	As *	10.00
NO3 *	50.00	AI **	200.00
NO2 *	3.00	v	---
Br	---	Co	---
NH3 **	1.50	Sr	---
TOC	---	THM	<1

* HEALTH EFFECT ** Consumer complain

Table (3) Physical & Chemical Analysis for Potable Water

S.L./Parameters	Temp.	PH	EC	TDS	T.H.	T.A.	CO_3	HCO_3	SO_4	Cl	Na	K	Ca	Mg	SiO_2	CN	NH_3	NO_2	NO_3
Units	$_o$C		uS/cm	mg/l	mg/l as $CACO_3$	mg/l as $CACO_3$	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
Shwaikh P.S.	23.40	7.40	446.00	286.00	136.00	25.90	0.00	31.60	88.00	72.00	36.00	1.38	37.60	10.20	1.42	0.00	0.00	0.00	0.00
Zoor P.S.	28.60	8.00	468.00	286.00	119.00	51.50	0.00	62.80	46.00	87.00	43.30	3.21	38.00	5.90	0.72	0.00	0.00	0.00	0.00
Doha P.S.	25.20	7.80	543.00	364.00	160.00	54.80	0.00	66.90	107.70	68.00	44.40	1.48	46.40	10.80	2.00	0.00	0.00	0.00	0.00
Reservoir	29.10	7.80	505.00	316.00	138.00	50.40	0.00	61.20	70.40	79.00	45.80	2.83	42.20	8.00	1.55	0.00	0.00	0.00	1.36
Filling station	26.50	8.01	519.00	348.00	155.00	52.00	0.00	63.60	96.30	73.80	47.30	1.81	45.60	10.10	1.90	0.00	0.00	0.00	2.40
Tower	29.90	8.01	552.00	337.00	161.00	46.50	0.00	56.80	102.00	73.40	47.70	1.79	46.90	10.70	1.40	0.00	0.00	0.00	2.40
Average	27.12	7.84	505.50	322.83	144.83	46.85	0.00	57.15	85.07	75.53	44.08	2.08	42.78	9.28	1.50	0.00	0.00	0.00	1.03
Max.	29.90	8.01	552.00	364.00	161.00	54.80	0.00	66.90	107.70	87.00	47.70	3.21	46.90	10.80	2.00	0.00	0.00	0.00	2.40
Min.	23.40	7.40	446.00	286.00	119.00	25.90	0.00	31.60	46.00	68.00	36.00	1.38	37.60	5.90	0.72	0.00	0.00	0.00	0.00

Table (4): Trace Metal Analyses for Potable Water (ug/l)

S.L./Parameters	Li	Be	B	Al	V	Cr	Mn	Co	Ni	Cu	Zn	As	Se	Mo	Cd	Sn	Sb	Ba	Hg	Pb
Shwaikh P.S.	4.31	0.00	7.90	1.38	0.74	2.29	2.72	0.07	7.40	38.08	3.90	0.46	0.04	3.65	0.00	0.00	0.00	3.26	0.00	0.00
Zoor P.S.	0.00	0.08	1.21	1.91	0.06	1.59	1.67	0.04	1.97	31.08	4.57	0.00	1.85	0.17	0.02	0.01	0.00	1.76	0.00	0.13
Doha P.S.	7.03	0.05	16.50	2.80	3.18	3.85	1.41	0.20	11.20	52.10	19.10	1.30	3.87	7.31	0.00	0.00	0.19	2.60	0.00	0.03
Reservoir	5.20	0.05	9.30	19.07	1.50	3.31	1.46	0.07	4.02	22.70	9.40	0.57	4.41	6.80	0.02	0.09	0.01	3.97	0.00	0.00
Filling station	4.50	0.06	10.60	5.00	1.70	4.30	1.34	0.09	6.80	41.10	14.40	0.90	4.04	8.70	0.01	0.09	0.02	5.30	0.00	0.00
Tower	5.80	0.00	14.00	10.80	1.76	2.38	1.46	0.05	6.90	29.40	26.40	0.58	0.31	9.80	0.03	0.08	0.02	5.89	0.00	0.00
Average	4.47	0.04	9.92	6.83	1.49	2.95	1.68	0.09	6.38	35.74	12.96	0.64	2.42	6.07	0.01	0.04	0.04	3.80	0.00	0.04
Max.	7.03	0.08	16.50	19.07	3.18	4.30	2.72	0.20	11.20	52.10	26.40	1.30	4.41	9.80	0.03	0.09	0.19	5.89	0.00	0.13
Min.	0.00	0.00	1.21	1.38	0.06	1.59	1.34	0.04	1.97	22.70	3.90	0.00	0.04	0.17	0.00	0.00	0.00	1.76	0.00	0.00

Table 5. Physical & Chemical Analysis for Mineral Water

Para-meters Units	Temp. $_0$C	PH	EC uS/cm	TDS mg/l	T.H. mg/l as CaCO$_3$	T.A mg/l as CaCO$_3$	CO_3 mg/l	HCO_3 mg/l	SO_4 mg/l	Cl mg/l	Na mg/l	K mg/l	Ca mg/l	Mg mg/l	SiO_2 mg/l	CN mg/l	NH_3 mg/l	NO_2 mg/l	NO_3 mg/l
1	24.56	8.24	370.00	215.67	140.49	130.83	7.06	145.16	38.58	18.90	15.27	4.18	49.90	3.82	23.55	0.00	0.00	0.00	0.03
2	25.10	8.07	205.00	118.90	107.61	74.64	0.00	115.42	11.56	5.68	7.79	0.36	38.00	3.06	2.90	0.00	0.00	0.00	0.07
3	25.00	8.21	523.00	306.40	232.42	146.19	2.48	173.23	51.04	57.42	26.23	2.13	75.30	10.72	13.57	0.00	0.00	0.00	0.02
4	23.40	7.80	296.00	172.80	9.48	66.06	0.00	80.56	10.81	48.08	65.08	0.79	3.00	0.48	2.00	0.00	0.00	0.00	0.00
5	26.40	7.73	214.00	124.10	85.23	49.57	0.00	60.45	24.50	18.59	12.13	0.71	30.30	2.30	3.19	0.00	0.00	0.00	0.00
6	25.20	7.70	246.00	143.16	11.49	25.81	0.00	31.47	6.68	58.35	44.98	2.24	2.70	1.15	3.39	0.00	0.00	0.00	0.05
7	27.50	8.20	290.00	169.33	64.84	78.34	0.00	95.54	29.21	30.00	44.66	0.44	21.60	2.63	4.28	0.00	0.00	0.00	0.00
8	26.50	7.73	273.00	159.40	10.82	53.82	3.55	58.41	5.04	66.60	57.58	1.64	3.80	0.32	1.19	0.00	0.00	0.00	0.00
9	27.30	8.21	251.00	146.06	32.39	49.91	0.00	60.86	15.70	44.86	15.66	0.70	4.70	5.01	8.00	0.00	0.00	0.00	0.00
10	22.20	7.76	190.50	110.49	30.92	51.15	0.00	62.38	35.15	25.37	12.96	0.68	2.30	6.11	5.95	0.00	0.00	0.00	0.07
11	21.86	8.41	395.00	230.24	66.84	95.12	0.00	116.01	35.15	58.26	27.56	1.72	1.80	15.13	0.92	0.00	0.00	0.00	0.03
12	23.45	8.04	516.00	302.00	220.60	263.06	0.68	319.43	48.43	63.16	22.78	1.91	78.60	5.85	10.50	0.00	0.00	0.00	0.00
13	24.60	7.58	241.00	140.24	36.81	46.85	0.00	57.13	36.07	27.22	37.51	1.71	13.80	0.56	1.20	0.00	0.00	0.00	0.02
14	27.40	7.72	570.00	334.84	263.69	196.98	0.00	240.21	28.89	50.30	30.45	0.72	103.30	1.32	4.57	0.00	0.00	0.00	0.00
15	24.90	7.99	284.00	165.82	20.55	23.73	2.12	24.61	5.89	72.16	52.49	2.25	5.40	1.71	0.92	0.00	0.00	0.00	0.03
16	23.56	8.32	248.00	144.32	24.91	63.97	1.39	75.18	15.07	35.39	15.06	0.89	3.70	3.80	4.35	0.00	0.00	0.00	0.00
17	22.87	6.76	125.00	72.46	8.38	14.95	0.00	18.23	13.28	17.78	13.38	0.58	2.20	0.70	11.88	0.00	0.00	0.00	0.00
18	23.80	8.09	355.00	207.10	137.51	147.75	4.20	171.63	29.62	14.56	15.16	3.79	39.20	9.59	9.96	0.00	0.00	0.00	0.00
19	23.40	7.78	266.00	155.29	32.88	51.67	0.00	63.01	9.02	33.18	40.13	1.30	5.80	4.46	1.19	0.00	0.00	0.00	0.00
20	23.40	7.50	190.60	110.10	14.29	18.48	0.00	22.54	26.91	34.61	33.02	1.23	4.30	0.86	1.25	0.00	0.00	0.00	0.00
21	25.70	6.58	365.00	243.00	116.94	51.33	0.00	62.60	89.20	30.00	26.90	10.40	24.20	13.70	1.60	0.00	0.00	0.00	0.00
22	24.20	8.24	377.00	220.00	173.40	141.98	6.19	160.53	42.00	21.79	15.86	3.67	65.90	2.10	17.00	0.00	0.00	0.00	0.00
23	22.70	8.24	354.00	206.00	157.99	126.22	2.74	148.35	40.15	23.12	20.88	1.62	58.40	2.91	0.00	0.00	0.00	0.00	0.00
24	22.99	7.97	258.00	150.00	46.65	89.54	0.00	109.20	10.25	26.48	21.99	2.70	15.10	2.16	11.49	0.00	0.00	0.00	0.00
25	21.98	7.40	139.44	85.59	70.60	36.98	0.00	45.10	12.50	6.10	2.50	1.69	22.80	3.30	0.54	0.00	0.00	0.00	6.00
Average	24.40	7.85	308.46	177.33	84.71	84.60	1.22	100.69	26.83	35.52	27.12	2.00	27.04	4.15	5.81	0.00	0.00	0.00	0.25
Max	27.50	8.41	570.00	334.84	263.69	263.06	7.06	319.43	89.20	72.16	65.08	10.40	103.30	15.13	23.55	0.00	0.00	0.00	6.00
Min	21.86	6.58	125.00	72.46	8.38	14.95	0.00	18.23	5.04	5.68	2.50	0.36	1.80	0.32	0.00	0.00	0.00	0.00	0.00

Table 6: Trace Metals analyses for Mineral Water (ug/l)

Parameters/S.No.	Li	Be	B	Al	V	Cr	Mn	Co	Ni	Cu	Zn	As	Se	Mo	Cd	Sn	Sb	Ba	Hg	Pb
1	7.81	0.11	8.02	0.15	45.06	8.06	0.02	0.08	1.34	1.14	4.82	4.51	0.00	0.32	0.11	0.73	0.10	139.00	0.00	0.06
2	0.60	0.02	0.57	0.14	0.09	0.11	0.00	0.06	0.36	0.18	4.14	0.72	1.78	0.00	0.04	0.06	0.03	4.64	0.00	0.03
3	0.77	0.21	45.41	0.47	0.35	0.14	0.20	0.03	0.16	0.59	6.50	0.50	6.08	0.00	0.18	0.17	0.08	4.39	0.00	0.12
4	0.80	0.18	45.42	0.37	0.37	3.30	0.21	0.04	0.15	0.52	6.40	0.52	3.24	0.00	0.15	0.12	0.19	4.20	0.00	0.10
5	2.90	0.10	8.11	0.14	0.93	2.56	0.01	0.04	0.60	0.06	5.37	0.22	1.03	0.00	0.08	0.06	0.05	15.30	0.00	0.03
6	5.01	0.15	37.84	0.21	0.31	1.51	0.02	0.02	0.31	0.32	4.94	0.46	2.86	0.00	0.21	0.07	0.07	5.76	0.02	0.06
7	2.59	0.18	23.15	0.11	0.23	4.63	0.04	0.04	0.48	0.43	6.28	0.31	4.27	0.00	0.17	0.52	0.06	13.17	0.00	0.09
8	6.45	0.06	60.38	0.50	0.10	2.71	0.13	0.03	0.40	0.76	1.60	0.19	2.87	0.00	0.05	0.00	0.11	4.86	0.00	0.21
9	2.72	0.01	10.06	0.18	0.03	3.79	0.06	0.03	0.43	0.00	4.31	0.13	3.76	0.00	0.03	0.08	0.02	9.88	0.05	0.01
10	1.59	0.07	9.59	0.34	0.02	3.05	0.01	0.01	0.21	0.09	3.55	0.07	1.07	0.00	0.04	0.00	0.02	6.43	0.00	0.03
11	4.08	0.00	10.23	0.25	0.13	7.34	0.01	0.01	0.06	0.06	6.23	0.15	2.65	0.00	0.00	0.07	0.00	15.40	0.06	0.00
12	5.80	0.00	9.85	0.54	1.65	9.25	0.13	0.14	3.83	2.74	3.31	0.54	3.50	0.00	0.00	0.00	0.13	31.29	0.00	0.15
13	1.57	0.00	26.07	2.90	1.78	2.16	0.13	0.00	0.82	0.36	3.44	0.38	0.00	12.70	0.00	0.53	0.00	13.70	0.00	0.00
14	10.37	0.00	8.57	4.82	1.27	4.87	0.33	0.12	3.89	0.69	0.79	0.32	0.82	3.97	0.00	0.00	0.61	64.11	0.00	0.10
15	7.64	0.00	40.94	139.68	0.57	1.28	0.15	0.06	0.85	0.18	1.43	0.35	0.00	0.98	0.00	0.00	0.00	10.90	0.03	0.00
16	2.42	0.00	9.57	4.47	0.11	2.38	0.07	0.00	0.63	0.14	5.19	0.11	0.00	0.61	0.00	0.00	0.00	17.51	0.17	0.00
17	0.30	0.00	1.44	28.24	0.00	0.46	3.03	0.00	0.75	1.02	224.76	0.00	0.00	0.24	0.16	0.00	0.58	14.20	0.01	0.02
18	7.20	0.00	6.83	26.03	42.05	3.62	0.08	0.00	1.30	1.08	4.16	3.52	0.00	0.79	0.00	0.00	0.04	136.09	0.01	0.00
19	8.57	0.00	19.83	13.96	0.20	1.19	0.13	0.00	0.75	1.30	2.65	0.21	0.00	0.48	0.00	0.00	0.00	13.97	0.03	0.08
20	0.12	0.00	37.47	11.26	0.27	0.56	0.09	0.00	0.46	0.11	3.29	0.13	0.00	0.00	0.00	0.00	0.00	9.55	0.28	0.00
21	15.42	0.00	100.98	10.42	8.10	1.46	0.24	0.00	0.40	2.73	2.80	0.09	0.00	15.89	0.00	0.00	0.00	4.86	0.00	0.00
22	8.17	0.02	8.22	0.27	45.40	8.21	0.02	0.07	1.36	1.11	4.32	3.76	2.63	0.00	0.03	0.00	0.04	144.00	0.00	0.05
23	5.44	0.09	8.99	0.34	1.35	5.47	0.08	0.08	1.52	1.40	15.00	0.29	1.56	0.00	0.10	0.09	0.04	28.60	0.11	0.11
24	0.78	0.00	12.39	12.52	10.92	4.21	0.18	0.03	0.63	0.78	6.93	0.59	6.09	6.26	0.07	0.25	0.03	15.61	0.00	0.40
25	0.03	0.00	18.41	0.66	0.05	4.30	0.17	0.05	0.15	0.40	0.51	0.50	0.00	0.78	0.05	0.05	0.05	2.66	0.10	0.05
Average	4.37	0.05	22.73	10.36	6.45	3.46	0.22	0.04	0.87	0.73	13.31	0.74	1.77	1.72	0.06	0.11	0.09	29.20	0.03	0.07
Max.	15.42	0.21	100.98	139.68	45.40	9.25	3.03	0.14	3.89	2.74	224.76	4.51	6.09	15.89	0.21	0.73	0.61	144.00	0.28	0.40
Min.	0.03	0.00	0.57	0.11	0.00	0.11	0.00	0.00	0.06	0.00	0.51	0.00	0.00	0.00	0.00	0.00	0.00	2.66	0.00	0.00

Figure 2: The difference among Municipal water's mean values with Bottled water's maximum& minimum values for physical & chemical parameters

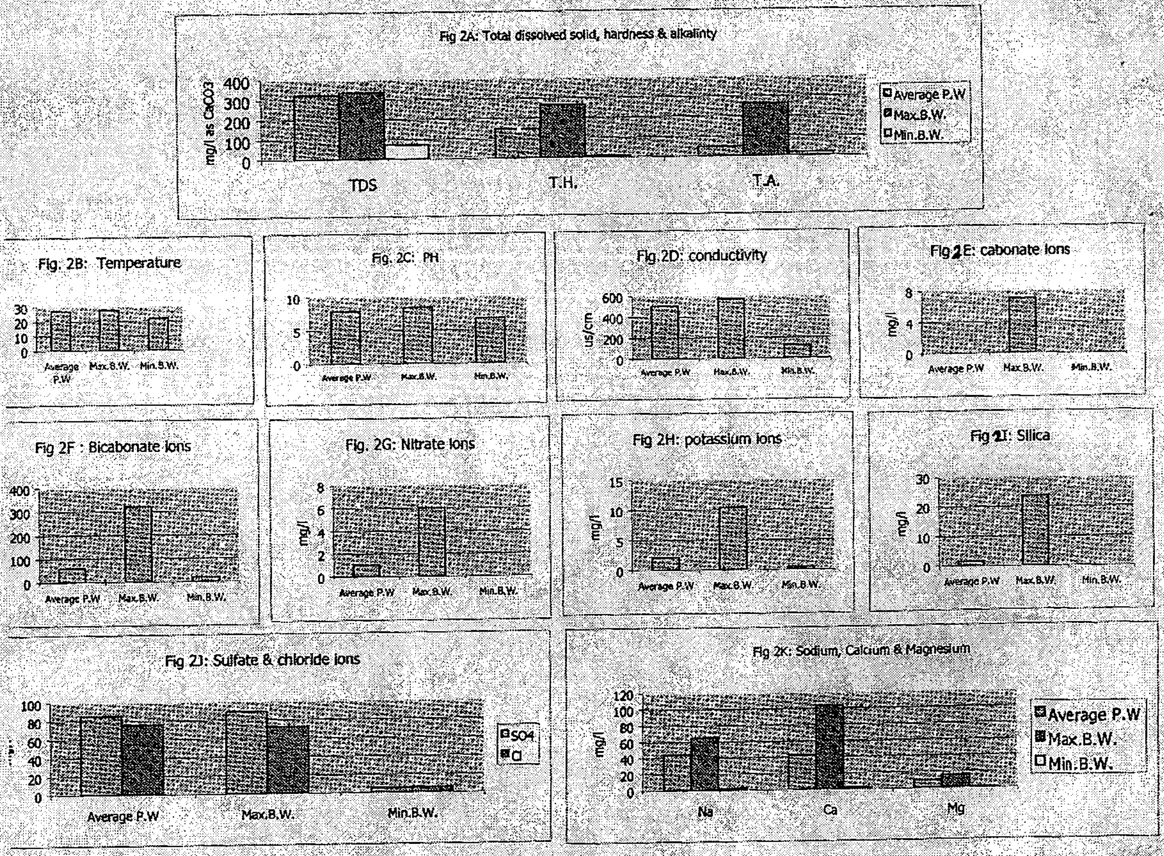
Fig 2A: Total dissolved solid, hardness & alkalinty

Fig. 2B: Temperature

Fig. 2C: PH

Fig.2D: conductivity

Fig 2E: cabonate ions

Fig 2F : Bicabonate ions

Fig. 2G: Nitrate ions

Fig 2H: potassium ions

Fig 2I: Silica

Fig 2J: Sulfate & chloride ions

Fig 2K: Sodium, Calcium & Magnesium

Figure3: The difference among Municipal water's mean values with Bottled water's maximum& minimum values for trace elements

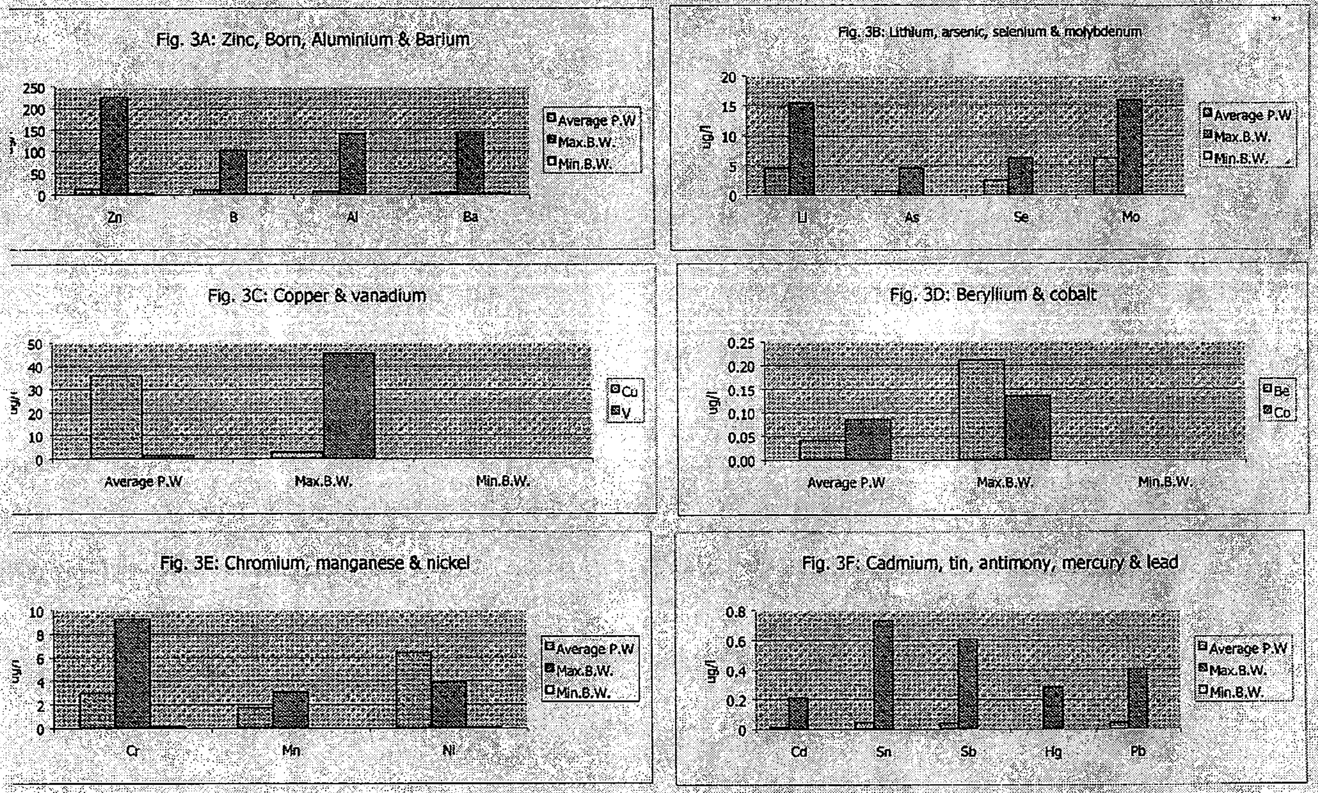

Fig. 3A: Zinc, Born, Aluminium & Barium

Fig. 3B: Lithium, arsenic, selenium & molybdenum

Fig. 3C: Copper & vanadium

Fig. 3D: Beryllium & cobalt

Fig. 3E: Chromium, manganese & nickel

Fig. 3F: Cadmium, tin, antimony, mercury & lead

Wastewater

Technologies of Wastewater Treatment in West Asia

Dr. Fouad Kanbour

TECHNOLOGIES OF WASTEWATER TREATMENT IN WEST ASIA

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ABSTRACT

The ever increasing demands for water resources in the region of West Asia necessitates the search for non-conventional water resources. One alternative source of water in the region is the utilization of reclaimed wastewater. This paper gives an overview of the various technologies of wastewater treatment utilized in the Arab countries of the West Asia region. The overview shows that the most common technology used in the region is the stabilization ponds. This technology constitutes the backbone of wastewater treatment in at least three countries. The most common wastewater treatment preferred by urban municipalities is the Activated Sludge Technology.

The Application of Flowsheeting Approach in Wastewater Treatment and Water Reuse

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THE APPLICATION OF FLOWSHEETING APPROACH IN WASTEWATER TREATMENT AND WATER REUSE

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ABSTRACT

The shortage of water in the Arabian Gulf countries with limited and high cost of water sources make the full recycling of wastewater an attractive alternative. Many wastewater treatment processes are available but most of them are applied to certain application and certain waste. No general strategy of treatment appeared in literature of wastewater.

Flowsheeting is widely used in petrochemical industries for process synthesis and selection through economic analysis of various processes with the help of computer programs. This flowsheeting approach is applied in the present work to achieve the suitable treatment strategy for any wastewater influent "either industrial or domestic" not only for its disposal but also to produce useful water qualities for better utilization and reuse.

A computer program was developed to achieve this target. Various application runs were conducted with the developed program and useful water products like agricultural water, potable water and bolier water were obtained from wastewater influents.

KEYWORDS : Water reuse, Wastewater treatment, Flowsheeting, Process selection

INTRODUCTION

In Saudi Arabia, the request for water is increasing annually to fulfill the developments requirements in the field of drinking, agricultural and industrial sectors.

The only water sources available for these requirements are desalted sea water which is costly and underground water with its renewal limitation. Hence, more attention has to be given to recycling and reuse for both domestic and industrial wastewaters.

The major contaminants present in both domestic and industrial wastewaters are: suspended solids, organics "usually expressed as BOD, COD or TOC", dissolved salts "expressed as TDS", nitrogen and phosphorous compounds, oils and grease, bacteria and viruses, heavy metals and toxic compounds.

The various types of treatment processes available in literature are usually classified according to the level of treatment, i.e. primary "physical treatment", secondary "biological treatment", the tertiary and the quaternary treatments "advanced treatment technologies". But, the selection of the appropriate treatment strategy for specific wastewater to be reused or recycled, is not an easy job especially from both economic considerations and wide spectrum of expected pollutants in both industrial and domestic influents. Flowsheeting "or process selection" is the key for this problem.

FLOWSHEETING

Chemical process design, usually starts with the selection of a series of processing steps and their interconnection into a flowsheet. Thus, a flowsheet is the diagrammatic representation of the process steps with their interconnections.

It is worth mentioning that chemical engineers have traditionally synthesized process flowsheets using experience, insight invention and sequential evaluation of alternatives. But recently, there has been some analysis and computerization of this process [1,2].

In fact, flowsheeting technology started recently to find its extensive use in chemical and petrochemical processing and some computer programs are now available to help in these fields like Chemcad and Aspin Plus [2].

Although tremendous work has been carried out in the past on water treatment, water reuse and wastewater treatment processes, none of these previous studies encountered the flowsheeting approach in the selection of the proper treatment process for various water feeds [3].

Only two studies appeared in literature dealing with the process selection for wastewater treatment [4,5]. The first study [4] outlined a simplified and preliminary procedure for developing a wastewater treatment process consisting only of two steps. The first step is the characterization of all wastewater streams to evaluate the treatment needs for each stream. Then streams requiring the same type of treatment are combined together and treated as a composite stream [4]. But no flowsheeting analysis was conducted in that work.

In the second paper published recently by Belhateche [5], the various treatment processes were reviewed and evaluated. This work gives some guide lines for the selection of the suitable treatment step for a certain pollutant present in the wastewater stream, but it lacks flowsheeting methodology and economic analysis [5].

In the present work, the flowsheeting approach will be applied on various wastewaters to produce specific water qualities suitable for reuse in certain fields like agriculture, industry and even drinking or ultra pure water.

MAIN POLLUTANTS OF FEED STREAMS

Various contaminants present in industrial and domestic wastewaters are diverse and complex; but for practical and engineering purposes, those pollutants can be characterized by both organic and physical parameters as shown in Table (1).

The organic parameters measure the organic impurities present in wastewater like BOD (biochemical oxygen demand), COD (chemical oxygen demand) and TOC (total organic carbon). Physical characteristics of wastewater include TSS (total suspended solids), TDS (total dissolved salts), pH, color and odour [5]. Other significant pollutants like nitrogen compounds, phosphorous and heavy metals are also considered separately by their specific parameters [5] as shown in Table (1).

In the present work, thirty nine significant pollutants are considered during the design of the computer program by which the proper treatment strategy will be obtained.

VARIOUS TECHNOLOGIES OF WASTEWATER TREATMENT PROCESSES

There are different types of treatment processes available in literature [3, 6]. Those are usually classified into four levels; i.e., primary, secondary, tertiary and quaternary treatment [3,5]. Primary treatment involves physical processes like suspended solids removal and entrained oil separation while secondary treatment is usually applied to remove organic pollutants by biological oxidation. Tertiary and quaternary treatment are advanced levels of processing by which specific contaminants like nitrogen, phosphorous compounds, TDS and heavy metals are finally removed from wastewaters [5, 6].

In addition to these four levels of treatment, there are four classes of treatment technologies, i.e. physical, chemical, biological and thermal processes as shown in Fig. (1); [7]. Table (2) indicates some of these treatment technologies with their expected effect on the considered pollutants of wastewater [3].

In the present work, a wide spectrum of treatment processes was considered in order to give a wide chance for the designed computer program to achieve the proper treatment strategy for each type of considered wastewater feed. These treatment processes considered in the present work are as follows:

1. Screening "for large suspended solids removal".
2. Neutralization "for pH adjustment".
3. Coagulation, Flocculation and Sedimentation "for fine suspended particles".
4. Flotation "for fine oil particles".
5. Biological treatment "for BOD and COD".
6. Filtration "for residual fine suspended particles".
7. Air stripping "for NH_3, phenol and CN".
8. Activated carbon "for refractory TOC".
9. Membrane separation "for TDS".
10. Demineralization "by ion exchange or electrodialysis for heavy metal separation and hardness".
11. Disinfection "for taste, odor and bacteria".

It is worth mentioning that Fig. (2) indicates these treatment steps in a form of generalized flowsheet which will be applied in the following computer program.

THE COMPUTER PROGRAM

The main target of the present work is to establish an engineering strategy by which the suitable treatment flowsheet can be achieved for each considered wastewater and each desired water product. In order to reach this goal a computer program was constructed based on the following assumptions:

(a) Thirty nine pollutants have been selected as previously mentioned. Those pollutants are widely found in various domestic and industrial wastewaters [3, 8, 9].

(b) Major treatment methods have been identified (as shown in Fig. 2).

(c) Each treatment unit will treat only one principle pollutant "this assumption is only considered at the first phase of the present program to avoid complexity and interference between the function of these units".

(d) A definite percentage of pollutant removal has been chosen for each treatment method "based on data available in literature and also practical experience" [8, 9].

(e) The specific characteristics of the desired water products are identified according to the regulation rules and given to the computer program. Four qualities of treated water are considered in the present phase of work, i.e. potable water, agricultural water, industrial water and treated wastewater for disposal purposes [3, 8, 9].

Table (4) indicates the characterization of these four water qualities

THE PROGRAM ALGORITHM

This program starts with reading the wastewater characteristics and the demand water "product specifications". Then the program compares the level of pollutants in the wastewater feed with that required in the product water. Based on this comparison, the first treatment step is selected or bypasssed. After exiting from this unit, another comparison is conducted to decide for the next treatment step and so on until the desired water quality is achieved. It is worth mentioning that this numerical program is written in Fortran language. Then the program will stop and print the details of required treatment steps and properties of wastewater after each treatment unit. Fig. (3) indicates the flow chart of this program.

RESULTS OF THE COMPUTER PROGRAM

Preliminary runs of the computer program were successfully carried out using various types of wastewater feeds and specified water quality products "suitable for actual reuse purpose like drinking water, irrigation, industrial or even high purity water". Promising results were achieved for each case in the printout of each program run where a definite number of treatment processes "flowsheet" were finally given by the program. Moreover, an actual case "from actual treatment plants in Jubail city" was used in some runs and the obtained treatment sequence "flowsheet" by the program was close to the actual designed one "or even more improved than it".

Two samples of these results are given here below; the first one is for producing agricultural water from domestic wastewater while the second one is for the production of boiler water "for low pressure boilers" from the same waste.

SAMPLES RESULTS OF THE COMPUTER RUNS

(1) The Production of Agricultural Water from Domestic Wastewater: (see attached sample of computer output)

(i) Properties of wastewater feed:

1. Suspended solids = 200 ppm
2. pH = 7.8
3. BOD = 240 ppm
4. COD = 485 ppm
5. N. compounds = 40 ppm
6. TDS = 1240 ppm
7. Odour comp. = 25
8. Colour comp. = 25

Other pollutants are Nil.

(ii) Properties of required agricultural water product:

1. Suspended solids = 10 ppm
2. pH = 7.0
3. BOD = 10 ppm
4. COD = 150 ppm
5. N. compounds = 40 ppm
6. TDS = 4500 ppm

7. Odour comp. = 20
8. Colour comp. = 20

Other pollutants are Nil.

(iii) The selected treatment steps as given by the program are "see also Fig. (4)":

1. Screening
2. Neutralization
3. Sedimentation
4. Biological treatment
5. Filtration
6. Disinfection

(2) The production of low pressure boiler water from domestic wastewater:

(i) Properties of wastewater feed:
As previously given in the first sample of results.

(ii) Properties of the required low pressure boiler water:

1. Suspended solids = 10 ppm
2. pH = 7.0
3. BOD = 0.0 ppm
4. COD = 5.0 ppm
5. N. compounds = 0.1 ppm
6. Silica = 30 ppm
7. Aluminium = 5.0 ppm
8. Manganeeze = 0.3 ppm
9. Bicarbonate = 170 ppm
10. Hardness = 350 ppm
11. Alkalinity = 350 ppm

Other pollutants are Nil.

(iii) The selected treatment steps as given by the program are: "see also Fig. (5)":

1. Screening
2. Neutralization
3. Sedimentation
4. Biological treatment
5. Filtration
6. Air stripping
7. Activated carbon
8. Reverse osmosis
9. Disinfection

It is worth mentioning that these obtained preliminary results encouraged us to make more improvements on the program and also to start the second phase of the project concerning detailed design and economic analysis of the obtained flowsheets.

CONCLUSIONS

(1) There is a potential need for wastewater reuse in the Kingdom of Saudi Arabia; either domestic or industrial wastewater.

(2) The flowsheeting approach is utilized in the present work to achieve the suitable treatment strategy for various wastewaters.

(3) The main characteristics of various wastewaters "either industrial or domestic" were defined in a clear method suitable for computer calculations.

(4) A computer program was designed and executed successfully; by which the suitable treatment flowsheet was obtained for each polluted water to give the desired treated water quality.

(5) Promising results were obtained from the preliminary runs of the computer program even for actual cases considered like production of boiler's water from domestic wastewater and also the production of agricultural water from the same waste.

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Table (1). Typical wastewater contaminants [5]

Parameter	Concern
Bulk Organic Parameters	
TOC	Can be toxic; depletes oxygen
COD	Can be toxic; depletes oxygen
BOD	Depletes oxygen in receiving waters
Oil and Grease/TPH	Damages vegetation and wildlife
Physical Parameters	
TSS	Turbidity; toxic to aquatic life
pH	Acidity or alkalinity is toxic to aquatic life
Temperature	Toxic to aquatic life
Color	Aesthetic; destroys algae
Odor	Can be toxic to aquatic life and humans; aesthetic
Redox Potential	Can be toxic to aquatic life
Contaminant-Specific Parameters	
NH_3/NO_3	Toxic to aquatic life; eutrophication
Phosphates	Eutrophication
Heavy Metals	Toxic to aquatic life and humans
Surfactants	Toxic to aquatic life and humans; aesthetic
Sulfides	Toxic to aquatic life and humans; aesthetic
Phenol	Toxic to aquatic life and humans
Toxic Organics	Toxic to aquatic life and humans
Cyanide	Toxic to aquatic life and humans

Key:
TOC = Total Organic Carbon
COD = Chemical Oxygen Demand
BOD = Biochemical oxygen Demand
TPH = Total Petroleum Hydrocarbons
TSS = Total Suspended Solids

Table (2). Wastewater treatment technology [3]

	Primary Treatment	**Secondary Treatment**	**Advanced Treatment**
pH		Neutralization	
SS	Screening Sedimentation	Chemical Precipitation Filtration	
BOD	Sedimentation Chemical Precipitation Methane Fermentation	Activated Sludge Trickling Filter	Activated Carbon Chemical Precipitation Membrane Filtration
COD	Sedimentation Chemical Precipitation Methane Fermentation	Activated Sludge Trickling Filter	Activated Carbon Ozonation Membrane Filtration
Oil	Floatation	Chemical Floatation Filtration	
N compound	Activated Sludge Trickling Filer	Biological Denitrification	Activated Carbon
Phenol		Activated Sludge	Activated Carbon
CN		Chlorination, AS	Catalytic Wet Oxidation
Cr		Reduction Sedimentation	Ion Exchange, ED
Fe	Sedimentation Chemical Precipitation	(Hydroxide) Precipitation Filtration	Ion Exchange Electrodialysis
Heavy Metal		(Hydroxide) Precipitation Filtration	Ion Exchange Electrodialysis
Cl		Neutralization by Sodium Thiosulfate	Activated Carbon
Sulfide		Activated Sludge Chemical Oxidation	Reverse Osmosis
Odor		Activated Sludge Chemical Oxidation	Activated Carbon
Color		Chemical Precipitation Oxidation & Reduction	Activated Carbon Reverse Osmosis

Table (3). Characteristics of different qualities of the desired water, in ppm. [3, 8, 9]

Pollutants	**Potable**	**Agriculture**	**Sewage For disposal**	**Industrial**
BOD	0 - 3	10	100 - 400	100 - 300
COD	0 - 10	150	200 - 1000	300 - 2000
Oil & Grease / TPH	0	< 8	< 120	15 - 55
TSS	0 - 0.25	15	2000	15 - 45
pH	7 - 8.5	6.5 - 8.5	5 - 10	6 - 9
N-compound		40 (NH_3)	10 - 50 (NH_3)	1 - 10
Phosphate		1.0	5 - 20	0.2
Phenol	0.001-0.002	< 0.1	< 150	0.1 - 1.0
Cyanide	< 0.02	< 0.05	1.0	0.1
Cd	< 0.001	10	0.5	-
As	< 0.05	0.1	1.0	-
Cr	0.05	100	200	-
Cu	0.05 - 1.5	200	1.0	-
Pb	< 0.05	5	1.0	-
Hg	< 0.001	0.001	0.001	-
Zn	5 - 15	200	10	-

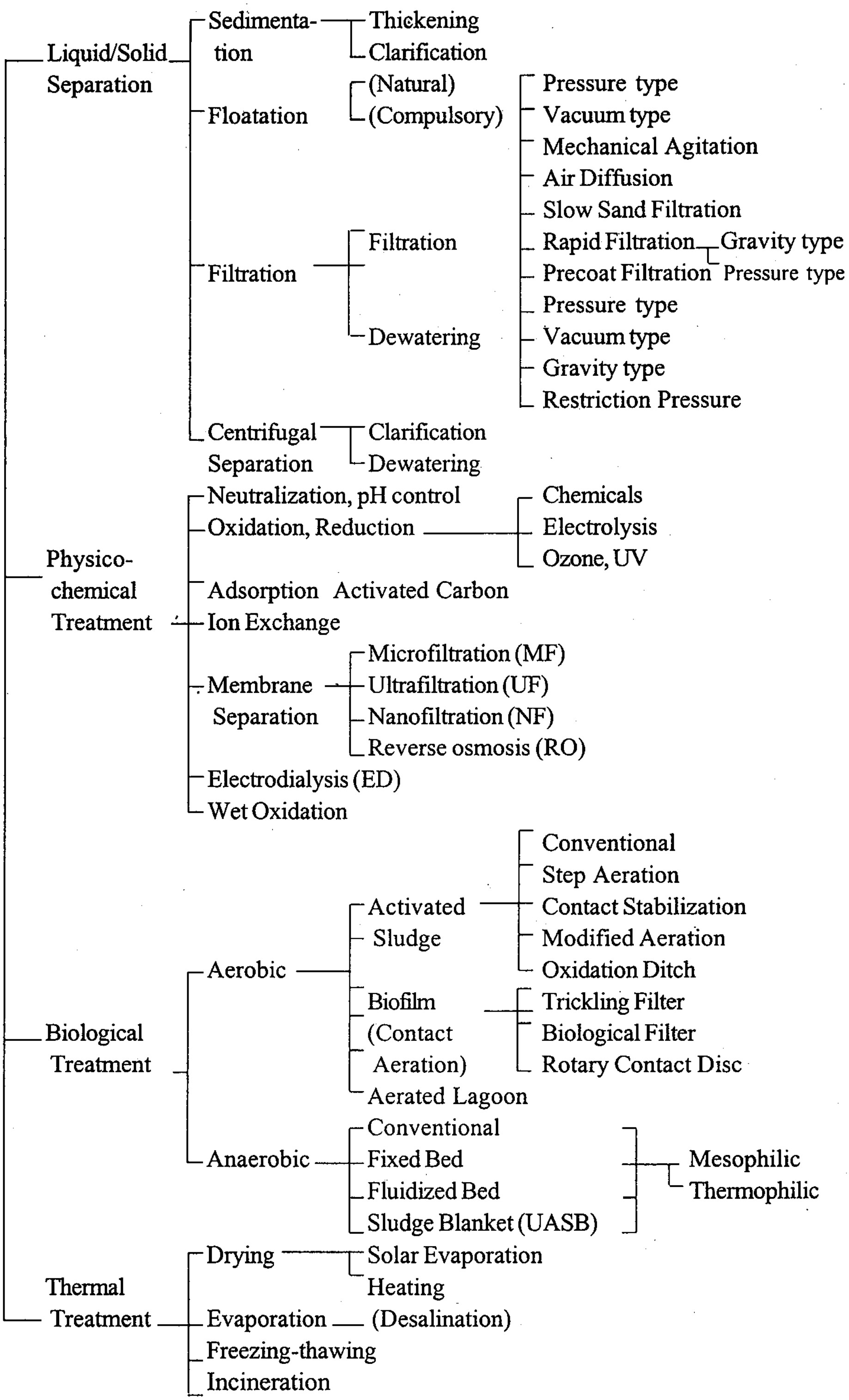


Fig. (1) : Classification of Treatment Processes. [7]

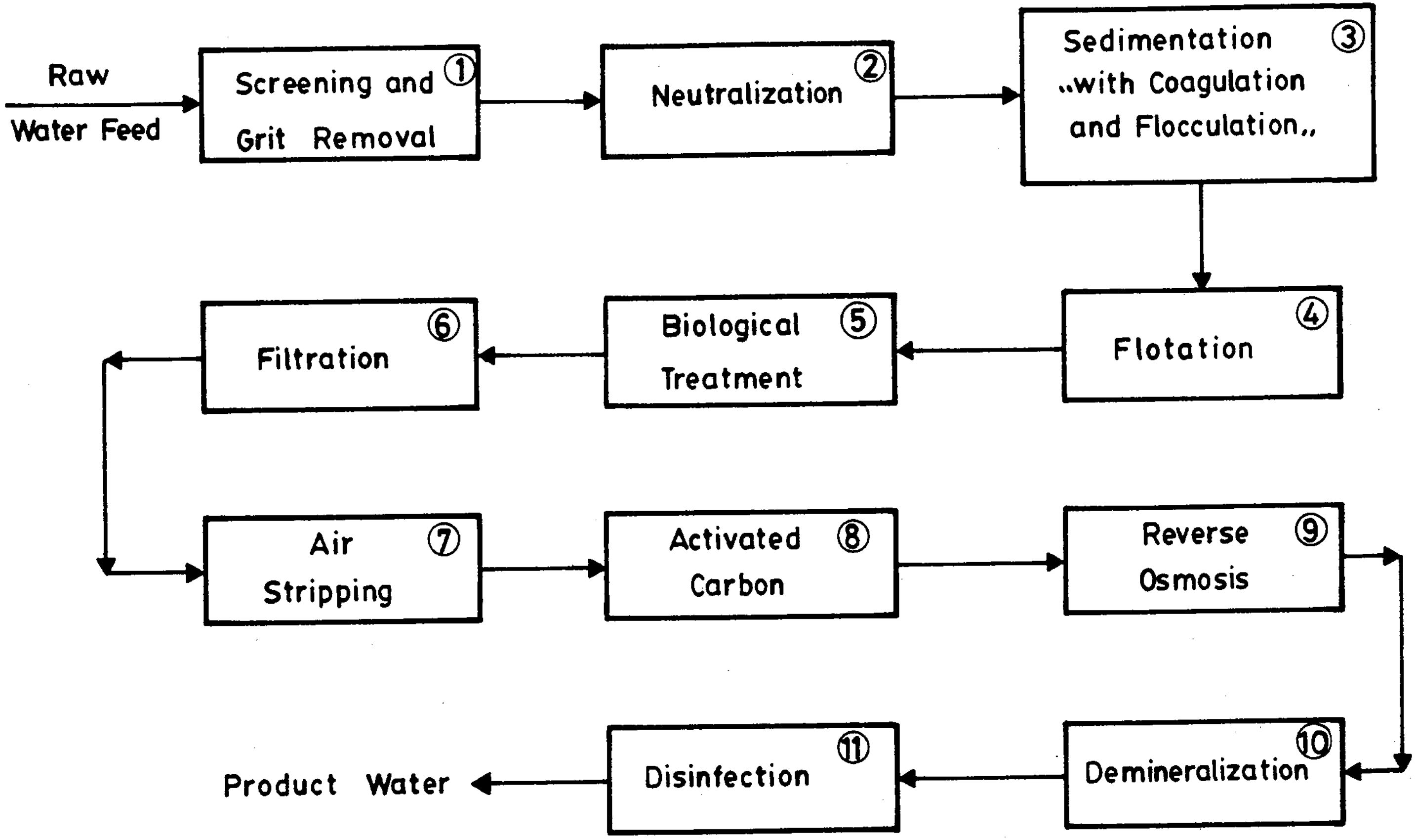


Figure 2: Schematic Flowsheet of the Major Treatment Processes

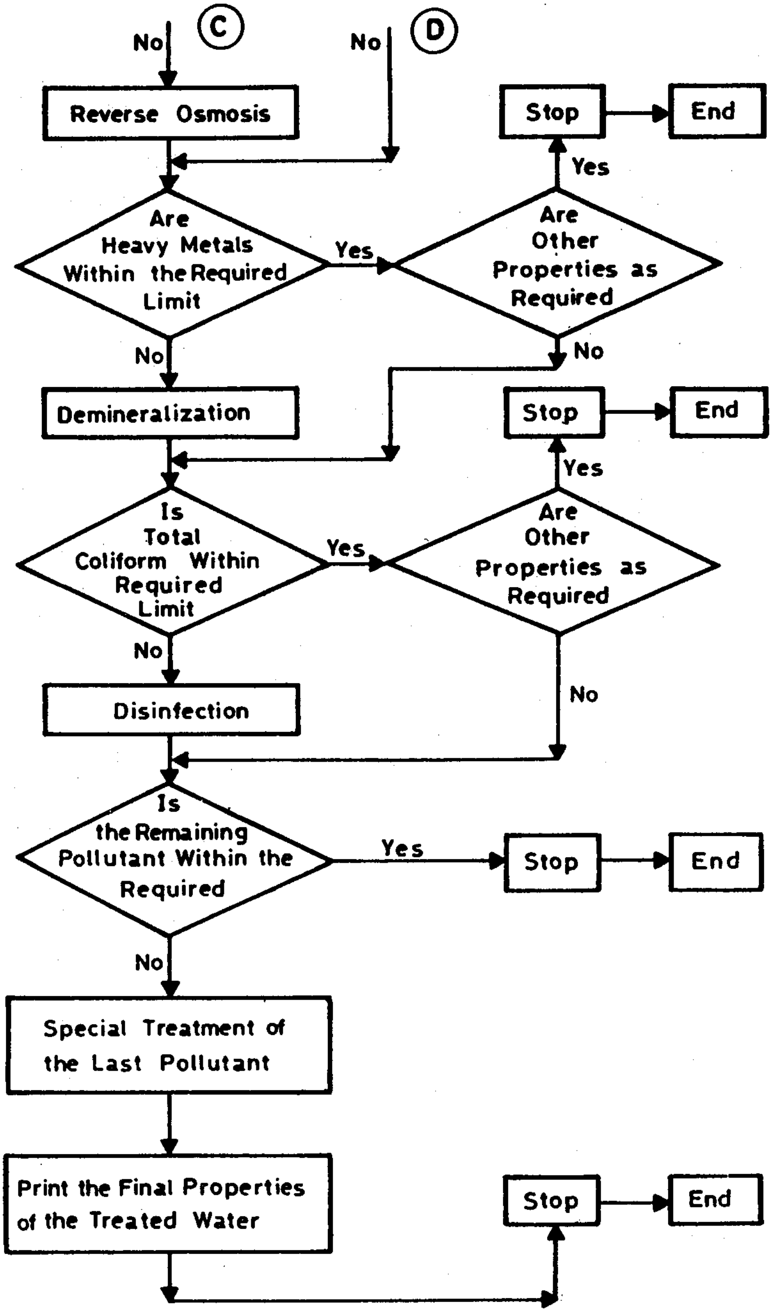
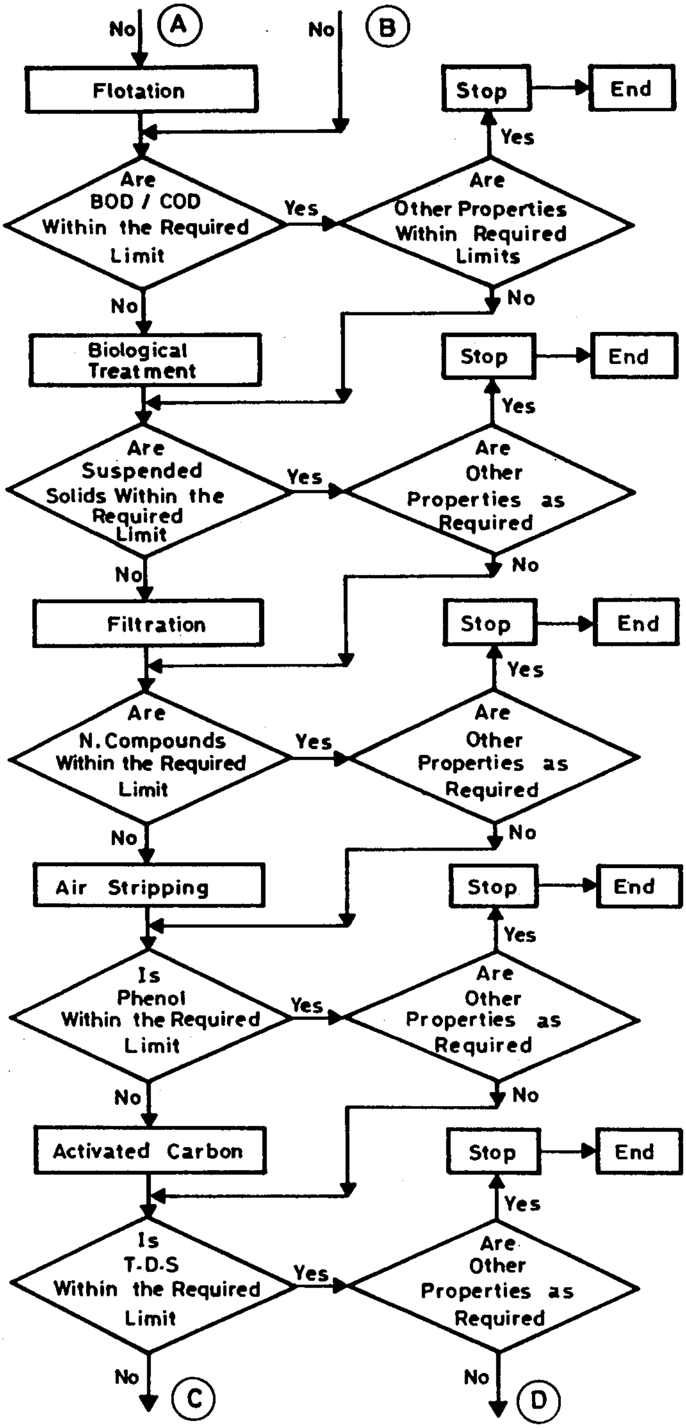
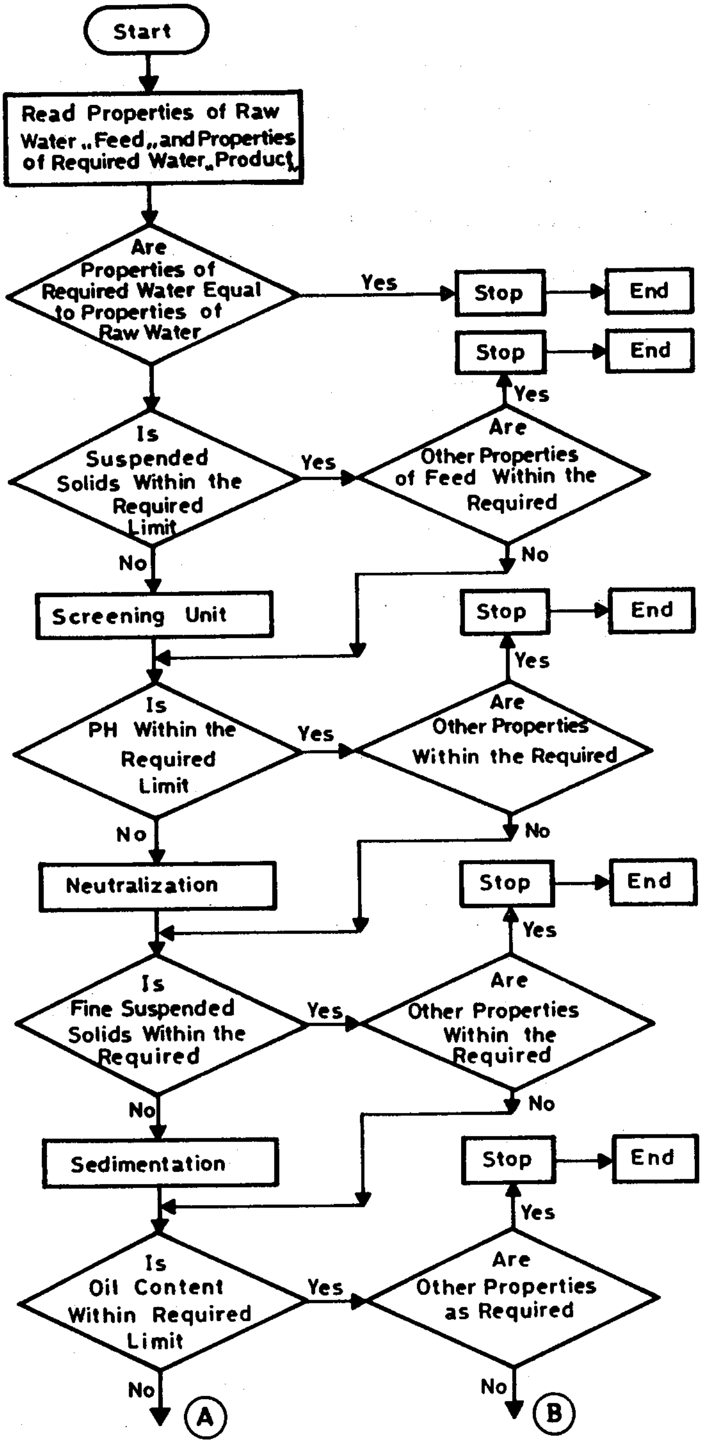


Figure 3: The Flow-Chart of the Computer Program


No
A
No
B
Flotation
Stop
End
Yes
Are BOD / COD Within the Required Limit
Yes
Are Other Properties Within Required Limits
No
No
Biological Treatment
Stop
End
Yes
Are Suspended Solids Within the Required Limit
Yes
Are Other Properties as Required
No
No
Filtration
Stop
End
Yes
Are N. Compounds Within the Required Limit
Yes
Are Other Properties as Required
No
No
Air Stripping
Stop
End
Yes
Is Phenol Within the Required Limit
Yes
Are Other Properties as Required
No
No
Activated Carbon
Stop
End
Yes
Is T.D.S Within the Required Limit
Yes
Are Other Properties as Required
No
C
No
D


Start
Read Properties of Raw Water „Feed„ and Properties of Required Water„Product„
Are Properties of Required Water Equal to Properties of Raw Water
Yes
Stop
End
Is Suspended Solids Within the Required Limit
Yes
Are Other Properties of Feed Within the Required
Yes
Stop
End
No
No
Screening Unit
Is PH Within the Required Limit
Yes
Are Other Properties Within the Required
Yes
Stop
End
No
No
Neutralization
Is Fine Suspended Solids Within the Required
Yes
Are Other Properties Within the Required
Yes
Stop
End
No
No
Sedimentation
Is Oil Content Within Required Limit
Yes
Are Other Properties as Required
Yes
Stop
End
No
A
No
B

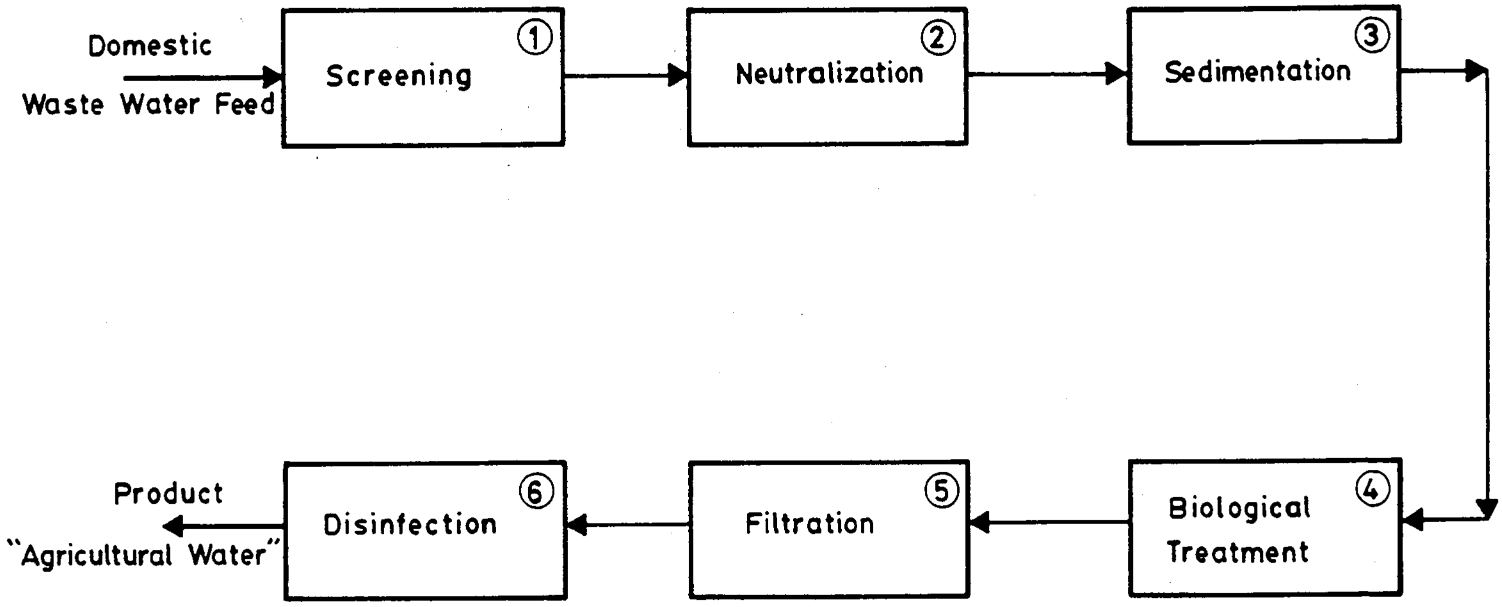


Figure 4: The Selected Treatment Sequence for Producing Agricultural Water from Domestic Waste Water

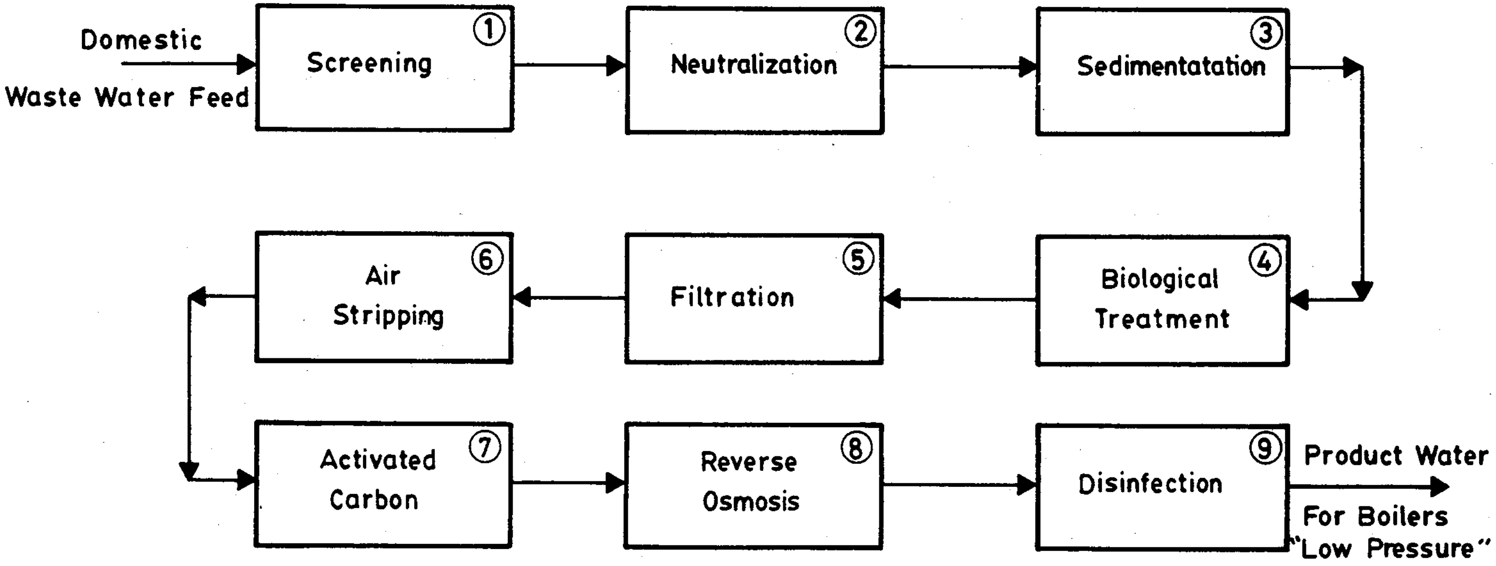


Figure 5: The Selected Treatment Sequence for the Production of Low Pressure Boiler Water from Domestic Waste Water

Compact Technologies for Urban Wastewater Recycling

Slim Zeghal, Gerard Velter and Abdelkader Gaid

COMPACT TECHNOLOGIES FOR URBAN WASTEWATER RECYCLING

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ABSTRACT

Wastewater treatment regulations are getting much more stringent. The need for recycling is increasing. Treated water quality and reliability of operation requirements lead to the use of new treatment technologies. On the other hand, site constraints, a sensitive neighborhood and a limited available construction space urge for technologies that can be easily integrated in their environment.

Primary sedimentation has seen several developments in recent years. The Actiflo process, combining the benefits of lamella sedimentation with enhanced coagulation around microsand particles, allows upflow velocities higher than 100 m/h, to be compared to a few m/h in conventional sedimentation. Several biofilter reactors with fixed biomass, used at an industrial scale, are presented to illustrate the most recent developments in wastewater treatment. Down-flow aerated biofilters consist of a single compact process associating a biological reactor and a liquid-solids separator. A floating bed biofilter has also been developed. It offers the possibility to completely eliminate nitrogen in a single reactor cell. Another biofilter design has been developed adding a further simplification of the treatment train by eliminating the need for primary sedimentation. In all cases, the system offers a reliable secondary treatment that does not suffer bulking or mixed liquor losses, lending itself to a good further chlorine or UV disinfection, or being an excellent pre-treatment to a tertiary process unit. This tertiary process unit can rely on membranes microfiltration. Another compact design was developed by merging the secondary biological step with the membranes separation. This compact system can therefore easily perform nutrients removal and deliver a treated water with turbidities of less than 0.5 NTU.

In order to achieve various treatment goals for wastewater, various innovative solutions are proposed, allowing the optimization of treatment plants. The use of compact treatment techniques allows bringing the plants close to urban centers where the pollution is produced and recycled water is needed, thus cutting on infrastructure costs related to collection systems. This has also the advantage of allowing an easier integration of the plants in their environment, by landscaping them and providing a better control of their odor and noise emissions.

KEY WORDS: Wastewater reuse; reclamation; Sedimentation; Actiflo; Biofilters; Nutrients removal; Biocarbone; Biostyr; B2A; Membranes; BioSep

INTRODUCTION

The concept of deriving beneficial uses from treated municipal and industrial wastewater coupled with increasing pressures on water resources has prompted the emergence of wastewater reclamation, recycling, and reuse as integral components of water resource management. The inherent benefits associated with reclaiming treated wastewater for supplemental applications prior to discharge or disposal include preservation of higher quality water resources, environmental protection, and economic advantages. (Asano and Levine, 1995). Several technologies are now available to suit various reuse needs and standards. They need to be reliable and designed for the various applications. Compactness is a favorable characteristic as well. Indeed, such technologies would allow to treat the wastewater close to its production and potential application sites. Treatment systems can be landscaped and perfectly integrated into the environment.

ENHANCED SEDIMENTATION

Mexico City's population is exceeding 20 millions. The water authorities have been considering wastewater reuse as a potential new water resource for agriculture as well as an improvement of the environment conditions by lowering the discharges of organic pollution and nutrients. Due to various constraints and after a significant analysis process, it has been concluded that advanced primary treatment is the best suited solution to treat the considerable amount of wastewater Mexico city produces (74.5 m^3/s - 1700 MGD).

The novelty about the approach adopted here is to consider that there is benefit in leaving the organics (BOD) and nutrients (Nitrogen and Phosphorus) into the treated water in order to preserve its fertilizing capacity and to focus mainly on helminth eggs removal (recommendation of below 1 egg per liter) and pathogenic bacteria indicated by the fecal coliforms count (below 1000 coliforms in 100 ml).

Starting with a raw sewage with concentrations higher than 10^7 Coliforms /100 ml, an activated sludge or a biofiltration system alone will not be sufficient to cope with the bacterial standard, leading to only 2 to 3 Log units removal. Disinfection would then be required, but neither chlorination nor UV would have any effect on helminth eggs, and ozone would not be sufficient either. The best option for the helminth eggs is to physically remove them from the system through sedimentation, filtration at a sufficient threshold or other physico-chemical processes.

The treatment line that allows to deal with the objectives of microbiological decontamination and low BOD and nutrient removals at an acceptable cost is composed of :

- Advanced primary physico-chemical treatment to remove most of helminth eggs and suspended solids.
- Rapid sand filtration if the removal of helminth eggs is not sufficient on the first step
- Disinfection with Chlorine or UV radiation.

Conventional primary treatment is effective for removal of particulate matter larger than about 50 µm from wastewater. In general, about 50 % of the suspended solids and 25 to 50 % of the BOD_5 are removed from the untreated wastewater by primary treatment processes (Metcalf & Eddy, 1991). The Actiflo® process is an optimized physico-chemical treatment particularly suited for the particular application described above. Initially developed for drinking water treatment, it has been progressively adapted to various wastewater applications, ranging from stormwater treatment in combined sewer overflows to tertiary treatment and filters backwash water reject and eventually to high rate primary sedimentation.

The principle is that microsand is added to the influent at the coagulation-flocculation stage. This has two major advantages :

- The initiation of the reaction is improved by the presence of microsand, which increases the number of free particles. The retention time can be shortened and the size of flocculation tank minimized;
- The resulting floc is heavier, and therefore settles more easily so that higher upflow velocities are allowed.

An Actiflo® prototype plant (130 m^3/h - 0.8 MGD) has been used during 6 months in order to demonstrate the ability of the process to treat Mexico city wastewater. The results on SS and helminth eggs show the efficiency and reliability of this process even at high upflow rates up to 200 m/h (Le Poder et al., 1998). The helminth eggs concentration in the treated water is almost always belows the 1 egg/l threshold and certainly below 5 eggs/l at all times. These results obtained together with an average 25 mg/l SS in the treated effluent could be constantly monitored at high upflow rates and low chemical dosing (< 60 ppm of Alum).

The combination of the Actiflo system and a rapid sand filtration followed by disinfection allows to comply with World Health Organization recommendations (< 1 eggs/l and < 1 000 Fecal Coliforms / 100 ml).

BIOFILTRATION

Down-flow filtration

The Biocarbone filter is a down flow process using expanded clay media for biomass attachment and particle retention. During operation, the filtering media is not fluidised because filtration is carried out in the direction of the media compression. This feature allows very low suspended solids residuals. A periodic counter current backwash removes the accumulated excess biomass and the retained solids. The backwash gives priority to the highly loaded part of the bed in contact with the incoming solids. This process has been widely used in various sensitive locations where it has allowed a good integration of the wastewater treatment plant into the urban environment. Examples from the French Riviera include the Antibes facility, the first in the world to obtain an ISO 14001 certification and the Monaco plant. The Biofilters, loaded between 5 and 7 kg COD/m3.d, provide excellent carbon removal after primary sedimentation using lamella settlers. The plant is conveniently located downtown, close to the pollution production center. Designing the complete treatment system including the plant and the collection pipes leads therefore to an optimized solution.

Floating-bed Biofilter

The Biostyr filter is an up flow process using a floating polystyrene media for biomass attachment and particle retention. The floating media allows an aeration within the filter, therefore leading to the creation of an aerated zone above an anoxic zone, which enables both nitrification and denitrification to be carried out within a single reactor. Denitrification, being the first stage in the process, lowers the carbonaceous load of the water prior to entering the nitrifying stage, an essential requirement for the development of autotrophic nitrifiers. This also ensures that water entering the denitrification stage has the highest possible carbon content. A certain percentage of the nitrified water exiting the reactor is recycled back to the denitrifying stage to achieve the required nitrate removal. Filtration and backwash advantages from the Biocarbone are further enhanced here by the gravity backwash and lightweight media in use.

For a carbon and secondary nitrification application, the Cergy treatment plant (200 000 p.e.) combines deep tank high load activated sludge with lamella clarification and 15 nitrifying Biostyr cells of 63 m2 each. COD is lowered from an inlet value of 800 mg/l on average to 350 mg/l out of the lamella settlers and down to 50 to 60 mg/l in the effluent. Despite high ammonia loads which can reach up to 1.8 kg/m3.d on peak, the average daily concentration of ammonia is 5 mg/l and total TKN below 7 mg/l. The average outlet SS is about 18 mg/l.As a teriary nitrification application, in Evreux (France, 70 000 p.e.), the initial medium load activated sludge has been upgraded by the addition of Biostyr cells treating an average flow of 10 250 m3/d in 1995. This tertiary treatment refines COD from 86 mg/l down to less than 50 mg/l. N-NH4 is oxidised from 34 mg/l to 1 mg/l in the outlet. Excellent nitrification performance is achieved due to the low inlet organic matter to the biofilters. The Saint-Jean d'Illac treatment plant is composed mainly of a lamellae settler followed by 5 Biostyr cells of 16 m2 each used for nitrification/denitrification. It serves a population equivalent of 14000, corresponding to a maximum flow rate of 2100 m3/d. Mean water characteristics are shown on table 1.

Parameters (mg/l)	SS	COD	N-NH4	TKN	NO3	TN	P
Inlet	250	750	65	80	0	80	15-20
Outlet settler	120	500	65	75	0	75	14-18
Outlet Plant	8	55	1-2	4-6	12	15-20	10-15

Table 1: Average characteristics of waters at St Jean d'Illiac plant

An effluent standard of 20 mg/l TN, 30 mg/l SS and 90 mg/l COD is respected as can be seen from the previous table.

Pre-precipitation using ferric chloride is currently carried out at the Nyborg treatment plant which serves 60000 p.e. in Denmark. The biological stage is composed of 8 Biostyr filters of 63 m2 for nitrification-denitrification. The Biofilter building covers the same surface area as the existing conventional primary settlers. Table 2 shows the average desing loads and the effluent requirements, Up to now, 3 years of experience has been gained in running that plant. The mean effluent quality values, which meet the discharge standards, appear from Table 3.

Parameters		Inlet Design data	Outlet requirement
COD	mg/l	230	
BOD5	mg/l	90	15
TN	mg/l	46	8
NH4-N	mg/l		3
P	mg/l	2	15
SS	mg/l	50	10

Table 2: characteristics of waters at Nyborg plant

Total-P effluent of less than 1.5 mg/l could be continuously achieved with an average inlet around 8 mg/l. Despite an increasing inlet concentration, the effluent ammonia is around 2 mg/l and the nitrate concentration less than 6 mg/l. This could be achieved by methanol addition and more recycle. The results remain steady and meet the Danish standards. It should be pointed out however that high molar ratios of Metal ion/P were used for the chemical dosage, and precipitation degrades the C/N ratio of the biological stage by eliminating more carbon in the primary settler, which then requires carbon addition to denitrify.

Parameters		Mean value	Standard deviation
BOD5	mg/l	7	(2)
TN	mg/l	6,5	(1,0)
NH4-N	mg/l	1,8	(1,3)
P	mg/l	0,8	(0,4)
SS	mg/l	11*	(5)

Table 3: Mean effluent values at Nyborg plant (official analyses 1994)
**** SS value in influent 150% of the design basis***
(76 mg/l SS instead 50mg/l)

Co-precipitation of phosphorus has been demonstrated at an industrial scale demonstration filter with a surface area of 22 m^2 and an average flow rate of 800 m^3/d that was constructed at the Rambouillet treatment plant. The waste water treated by the reactor comes from a combined urban sewer which has undergone preliminary treatment and primary settling. Chemical precipitation was carried out using Alton P1, an alkaline minêral reagent. Sammut et al. (1992) report removal rates of 70 to 80% on PO_4-P for a water with 10 to 12 mg/l PO4-P at the inlet. The addition of Alton P1 does not appear to have any effect on nitrification rate (0.9 kg NH4N/m^3 aerated zone/d), which would agree with the findings of Kraft and Seyfried (1990). It only slightly reduces denitrification rates, unlike the case of pre-precipitation, as colloidal organic matter seems to remain available to the denitrifying bacteria as a carbon source. Headloss build-up is slightly faster and of a more linear nature during the addition of chemical.

Nitrification, denitrification and biological phosphorus removal has been achieved on fixed film reactors. As described by Goncalves et al. (1993), a pilot unit consisting of 5 Biostyr columns (0.2 m diameter, 1.5 to 2.6 m material height) was used for these experiments. It was operated in the following conditions: of the 5 biofilters, one will be temporarily under anaerobic conditions. The anaerobic phase moves from one biofilter to the following, once maximum P release is achieved. Aeration then starts up again in the biofilter previously under anaerobic conditions. The total influent settled waste water flow to the pilot (rich in organic substrate) is forwarded to the filter under anaerobic conditions. After the anaerobic filter, it is then equally divided among the aerobic filters placed in parallel downstream. At the end of a filter's anaerobic phase, a short aeration period (30 min) takes place with the objective of purifying the volume of phosphorus-rich water remaining in the reactor and minimizing its impact on the effluent quality. Filter backwash takes place at the end of an aerobic phase when the biomass is saturated in phosphorus.

Parameters (mg/l)	SS	COD	NH4-N	TKN	NOx-N	tot-P	PO4-P
Inlet	126	359	44.1	52.6	0	11.3	7.2
Outlet	7.3	39	2.6	4.4	10.5	3	1.9

Table 4: Influent/effluent characteristics for nitrification/denitrification+BPR

Between 8 and 9 mg P total/l were eliminated (Table 4), reaching P removal efficiencies of 60 to 70 %. The efficiency of the anaerobic phase was strongly dependent on the influent characteristics and the carbon substrate present. Nitrification is not affected by this operational mode. Denitrification is improved by introducing an anaerobic contact time to the system, confirming observations made by other authors when working with an activated sludge system (Clayton et al., 1991).

Using a 300 mm diameter, 2.3 m bed height pilot column, Zeghal et al. (1995) researched the maximum removal capacities of the floating biofilter for post denitrification. The effects of different inlet nitrate concentrations and temperatures were investigated. It was found that the Biostyr filter could totally remove loads from 1.2 kg of N-NO3 per cubic meter of reactor and per day (kg N-NO3/m3.d) at 10 mg/l N-NO3 inlet concentration and 16 °C to 2.5 kg/m3.d at 20 mg/l N-NO3 inlet and 20°C. The carbon addition needs were around 5 to 7 mg COD per mg N-NO3, including the consumption due to dissolved oxygen. The suspended solids removal allowed an average outlet of 6 mg/l SS with an average inlet of 55 mg/l. In situations of unfavorable inlet COD/N ratios where low outlet residuals are desired on a consistent basis, it is wise to use a treatment train combining pre and post denitrification. Pre-denitrification ensures that the waste water carbon is used as much as possible. Post-denitrification on part of the flow relieves the hydraulics on the previous stages of treatment by lowering the needs for recycle.

Another up-flow Biofilter

The B2A process is another up-flow filter. It has been developed using variable granulometry media, which enables it to take pre-treated raw sewage without primary sedimentation. It is particularly suited for small-medium wastewater treatment plants, providing long cycle time, minimal maintenance, simplified process train and high reliability. As with the Biostyr, the B2A offers the possibility to nitrify and denitrify in a single cell.

Some results are presented here from the Gratentour (France) treatment plant, designed for a pollution of 3000 population equivalent. The daily average COD in the effluent is below 40 mg/l for an inlet concentration of around 500 mg/l in average and a load of 5 kg COD/m3.d. Under a loading of 0.8 kg-NH4/m3.d, total nitrogen is consistently below 20 mg/l which was the target. Co-precipitation of phosphorus was also trialled at this plant, eventough not required. Residuals as low as 1 mg/l were achievable without disturbing the other biological processes within the system.

Membrane Bioreactor

Microfiltration membranes can be used as a tertiary treatment in lieu of a sand filtration step for effluent polishing. In this case, they are fed with secondary treated water of suspended solids generally below 20 to 30 mg/l and act as a polishing step. They have also been used as a solids separation step in the activated sludge process. Initial developments used the membranes into an external recirculation loop. The process delivered high quality water with however a significant energy consumption.

The use of immersed membranes as biomass separators in secondary treatment systems is an innovative approach which holds interesting promises in the context of wastewater reclamation or plants rehabilitation (Buisson et al., 1998). The membranes used in such configurations replace the secondary clarifier in conventional systems. They allow to work with high biomass concentrations (up to 25 g/l), high sludge retention times (higher than 20 days), and to effectively decouple mass loading and volumetric loading. The resulting systems are compact and lead to a good nitrification and a low sludge production. The following table gives a results summary of two pilot studies conducted in the USA and France:

Parameter	***Units***	***Feed***		***Permeate***		***Reduction***	
		USA	**FR**	**USA**	**FR**	**USA**	**FR**
COD	mg/l	237	482	16	10	93 %	98 %
TSS	mg/l	96	120	ND	ND	> 99 %	> 99 %
Turbidity	NTU	—	148	0.12	0.24	—	99.8 %
Ammonia-N	mg/l	25	39	0.8	0.4	97 %	99 %
TkN-N	mg/l	38	54	1.8	2	95 %	99 %
Nitrate-N	mg/l	—	—	26	9.9	—	—
Total-N	mg/l	38	54	28		26 %	80 %
Total Coliform	MPN/ 100 ml	5.6x10^{7}	5.9x10^{7}	20	43	6.4 log	6.1 log
Bacteriophages	N/ml	3.7x10^{4}	1.48x10^{3}	—	—	> 4.5 log	> 3.8 log
Temperature	°C	26-33	17 - 21	—	—	—	—

Table 4 – Biosep process performance (immersed membranes)

The immersed membrane activated sludge process offers the advantages of simplicity and compactness. The treatment does not require chemicals and leads to a low sludge production with a high effluent quality (COD, SS, N). The membranes provide an efficient barrier for microbes and an excellent pretreatment to reverse osmosis, should the latter be necessary to remove salinity.

CONCLUSIONS

The successful implementation of wastewater reuse options into a water resources management program requires careful planning; economic and financial analyses; and effective design, operation and management of wastewater reclamation, storage, and distribution facilities. Technologies for wastewater reclamation and purification have developed to the point where it is technically feasible to produce water of almost any quality and advances continue to be made. Current water reclamation strategies incorporate multiple measures to minimize health and environmental risks associated with various reuse applications. A combination of source control, advanced treatment process flowschemes, and other engineering controls provides a sound basis for increased implementation of water reuse applications. The feasibility of producing reclaimed water of a specified quality to fulfill multiple water use objectives is now a reality due to the progressive evolution of technologies and risk assessment procedures. However, the ultimate decision to harvest reclaimed wastewater is dependent on economic, regulatory, and public policy factors reflecting the demand and need for dependable water supply.

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On-Line Monitors for Wastewater Treatment

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ON-LINE MONITORS FOR WASTEWATER TREATMENT

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ABSTRACT

Wastewater reuse is gaining interest worldwide as planners have to deal with water scarcity as well as sewage disposal in environmentally sensitive areas. Reusing the wastewater provides therefore an excellent alternative by offering a new water resource and reducing the pollution load.

Even though the most important parameters relevant to reuse relate to microbiological aspects, nutrients need to be lowered to acceptable levels and monitored consistently. This paper will present a set of on-line sensors that have been reliably used on settled and secondary treated wastewater. Turbidity, ammonia, nitrates and phosphates were continuously measured in and out of a biological nutrients removing activated sludge system.

They provided an excellent visualization of these parameters trends and had relative deviations of 4% to 9% for the high concentrations range (settled wastewater) and 20 to 30% for the low range (treated water) while being calibrated for the high range and commuted from inlet to outlet. The absolute deviation is still less than 0.5 mg/l at the outlet as shown in the table below. It is however advised to use separate monitors for the inlet and outlet of the plant when very detailed and precise information is needed.

On-line analyzers have been reliably used to monitor turbidity, ammonia, phosphates and nitrates. They allow a continuous tracking of the various influent effluent characteristics at a reasonable cost, regarding the amount of data gathered. The plant is therefore easier to operate. Operating conditions (such as DO set-points, recycle ratios etc.) can be optimized confidently. Costing of capital and operating expenses is made and can be used in assessing the benefit of implementing such monitors as part of the complete control system

KEY WORDS: Wastewater reuse; reclamation; Nutrients removal; Analyzers; Process Control;

INTRODUCTION

One of the major concerns when dealing with wastewater reuse is related to public health and the impact of the reused wastewater on the receiving bodies such as aquifers, lakes and storages. The water has to comply with various standards depending on its final use (and the local legislation) and the system has to be designed with significant level of flexibility and reliability. Contingency plans need to be developed and monitoring tools are used to ensure of the proper operation of the plant.

On-line monitors have made significant progress in wastewater applications. They can now be used reliably at acceptable capital and operating cost. The benefit they provide for a better operation of the plant and an improved analytical program at a lower cost gives them a payback period of less than 2 years. This paper gives an example of monitoring equipment set-up. The particular equipment selection will depend not only on the technical characteristics sought but also on the service quality and availability that can be obtained from the supplier.

MATERIAL AND METHODS

The monitoring and control system includes an ultra-filtration unit, on-line analyzers, and data acquisition system as detailed in table 1 below. A schematic showing the way the different elements of the on-line control system are connected is shown in figure 1.

Parameter	**Brand**	**Model**	**Method**
Turbidity	ABB-Kent-	4670/200	Nephelonometry
	Taylor	System	
Nitrate	Datalink	NT 100	U.V absorbance (2 wavelengths)
Ammonia	Contronic	BE462	Specific gas electrode with pH adjustment.
Phosphate	HACH	Series 5000 Model 60001	Colorimetry (Molybdovanadate)
Data acquisition	Eurotherm-Chessel	4250 M	Chart + PCMCIA card

Table 1: On-line analyzers

TURBIDITY-METER: *ABB-Kent-Taylor 4670/200 System*

The sample flows continuously through the sample cell in front of the light source. The scattered light at 90° is measured by the detector. The turbidity-meter is provided with a cleaning module: a wiper cleans both the lamp and the detector regularly.

NITRATE ANALYZER: *Datalink NT 100*

The nitrate meter measurement is based on the ability of the nitrate to absorb U.V light for a particular wavelength: 220 nm). The sample circulates continuously into a quartz cell which is illuminated with a U.V lamp. A detector measures the light absorbed by the sample. The analyzer is equipped with an automatic back-washing system : an acid solution can be pumped into the cell on a regular basis (programmable) to prevent or stop biofilm growth. An indicator shows the cleanness of the cell and the reliability of the measurement.

AMMONIA MONITOR: *Contronic BE462*

The ammonia monitor uses a specific gas electrode. The sample to be analyzed is continuously pumped with caustic soda to maintain a basic pH (11 to 13) which allows the measurement of the ammonia as N-NH3+. The analyzer has an automatic calibration which can be programmed from 0 to 7 times a week.

PHOSPHATES ANALYZER: *Hach series 5000 Model 60001*

The phosphate analyzer works on a batch method (one measurement every 10 minutes) making use of a colorimetric method: Molybdovanadate reagent is added and reacts with orthophosphate to form a yellow-colored complex which color intensity is proportional to the orthophosphate concentration of the sample. The light absorbance of the solution is measured at 480 nm and the PO4- concentration calculated accordingly. When the analysis is complete, a new sample flow purges the sample cell and the analysis cycle repeats automatically. It is possible to program a weekly automatic calibration.

ULTRA-FILTRATION UNIT: *Process Styming AB's*

An ultra-filtration unit has been installed to pre-treat the settled wastewater before ammonia and phosphate analyses. It is composed of two tangential flow membrane filters mounted on a PVC panel. While one filter is in use, the other is cleaned. A flow-meter is also incorporated.

DATA ACQUISITION: *Euroterm Chessel 4250M*

The signals from the analyzers are recorded by a data acquisition system on two different formats: a chart recorder for a visual display of what is going on in real time for the plant operator and a PCMCIA card which allows data to be transferred on a computer as Excel spreadsheets. The recorder is completely programmable and each channel can be set-up independently.

COMMISSIONING AND CALIBRATION

The different items forming the control system were connected together as shown on figure 1, then commissioned and calibrated individually. The commissioning included each analyzer set-up and programming of the recorder. The output of each analyzer has been calibrated and the signal received by the recorder checked. The ammonia, phosphate and nitrate analyzers calibration were checked with standard solutions. The results of these calibrations tests are:

	Standard solution	**Standard value**	**Unit**	**Analyzer value**	**Unit**
Monitor					
Ammonia	NH4Cl	40	mg N/l	41.2	mg N/l
Nitrate	KNO3	25	mg N/l	24.7	mg N/l
Phosphate	KH2PO4	6	mg P/l	5.9	mg P/l

Table 2: Analyzers Calibration

This shows a good measurement of standard solutions. The turbidity-meter has been calibrated using a "dry calibration standard" provided with the monitor.

RESULTS

After commissioning, the precision of the ammonia, phosphate and nitrate analyzers was estimated by taking samples to be analyzed in the lab. The average deviation between the monitor readings and the laboratory analysis of the sample was calculated. Plots showing the correlation between monitors readings and laboratory analysis results are given in figure 2 for ammonia, figure 3 for phosphates, 4 for turbidity with SS and 5 for turbidity with COD.

	N° Sample	Monitor reading	Analysis results	Average deviation	
Monitor		mg/l	mg/l	Absolute mg/l	Relative %
Phosphate (SW)	20	5.1	4.6	0.47	10.2
Phosphate (Effluent)	7	0.8	0.6	0.23	39.3
Ammonia (SW)	35	39.6	40.8	2.42	5.9
Ammonia (Effluent)	23	0.7	1.2	0.51	42.7
Nitrate (Effluent)	8	9.6	9.7	0.30	3.1

Table 3: On-line analyzers deviation

The analyzers give therefore a continuous and reliable reading as can be seen from the above table 3. The average deviation is less than 10% for the high concentration range for which the sensors have been calibrated (5 to 60 mg/l for ammonia and 10 mg/l for phosphorus). This average relative deviation is increased up to some high values in the lower concentrations range, where one is also getting close to the monitors detection limit and accuracy. The absolute deviation is still less than 0.5 mg/l and trends can be clearly observed on treated water. It is however advised to use separate monitors for the inlet and outlet of plant when very detailed and precise information is needed. This leads however to the doubling of the monitoring cost.

For turbidity, correlations with other variables not accessible on-line have been developed. These variables are SS and COD. The calculations have been made for the 24 hr. average samples (same volume sampled every 15 min for 24 hr) and curves are shown in figures 4 and 5.

For COD (N =16), $y = 3.64x+88.4$ ($R^2 = 0.64$)
For SS(N=14), $y = 1.171x-103.75$ ($R^2 = 0.76$)

These correlations are acceptable to relate the on-line turbidity measurement to trends in COD and SS in the system. They are not universal and need to be derived for each water quality under consideration. They need to be revised if the water quality varies significantly.

MAINTENANCE REQUIREMENTS

The analyzers described in this paper required a quite simple basic maintenance. An example of maintenance schedule is shown in table 7. The approximate monthly time consumption for the analyzers maintenance is given in table 4:

Monitor	**Time consumption** *h/month*
Turbidity-meter	1.5
Nitrate	0.5
Ammonia	4
Phosphate	0.5
Ultra-filtration unit	1.5
Total	**8**

Table 4: Monthly maintenance time consumption

The Nitrate and Phosphate analyzers require small attention as they are highly automated. The turbidity-meter is more time consuming because the calibration has to be executed manually. The Ammonia analyzer requires more care and must be followed closely to know when the ammonia membrane needs to be replaced.

The total monthly time required by the analyzers is one working day. This is little compared to the time requirement for similar laboratory analysis. In the particular case of a pilot plant, the amount of data collected by the analyzers is very important to allow close monitoring and optimize testing time and design adjustment. Data is available for each parameter every 5 to 30 minutes for weeks. It would be very expensive to try to collect this data with laboratory analyses: In the case of a full scale plant, its operation can be improved, as the data is available in real time and the plant operation can be immediately adjusted.

TURBIDITY-METER

The turbididy-meter operation is very easy. It needs to be drained every day. The lamp, receptor and cleaning wiper are cleaned once a week: unscrew the different items flush them with water as well as the flow chamber, dry the optical windows with a soft paper before putting them back in place. The calibration has to be checked once a month.

Operating Cost

The average operating cost for each analyzer on a yearly basis including the reagents and miscellaneous is the following :

Nitrate analyzer:	$100
Ammonia analyzer:	$300
Phosphate analyzer:	$3,000
Turbidity-meter:	$0
Total:	**$3,400**

Table 6: Operating Cost for on-line Monitors

The operating cost for the phosphate analyzer is the highest. It could be cut down by nearly two by on site preparation of the acid and phosphate standard solutions. Nevertheless, the total operational cost is low regarding the cost of a laboratory program.

As the initial investment for analyzers can be recovered within 2 years, the savings for the following years are more than $30,000/ year, without taking into account the potential plant operational cost savings.

CONCLUSIONS

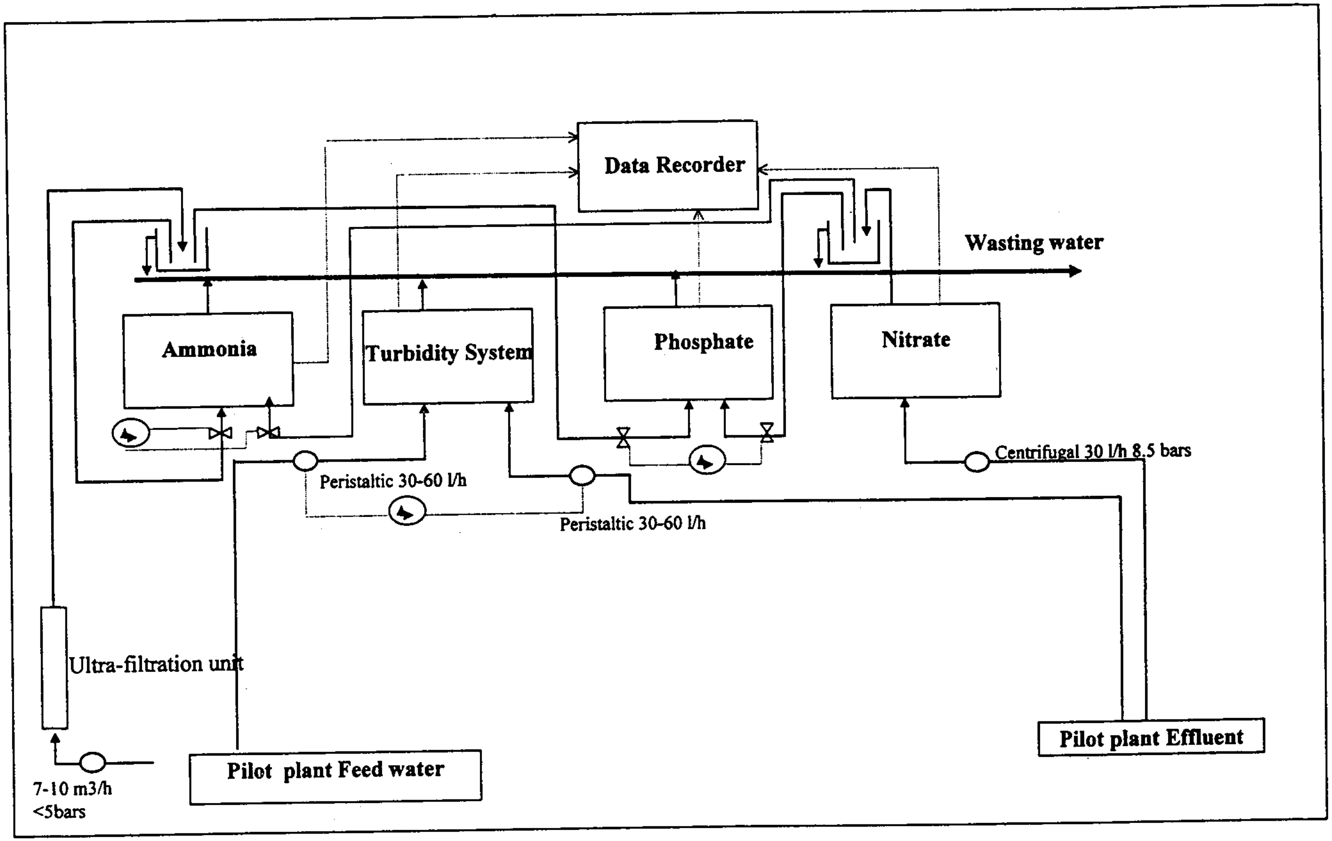
On-line analyzers have been reliably used in wastewater treatment to monitor turbidity, ammonia, phosphates and nitrates. They allow a continuous tracking of the various influent and effluent characteristics at a reasonable cost, regarding the amount of data gathered. The plant is therefore easier to operate as any trends in the various components concentrations can be detected and corrected for. Operating conditions (such as DO set-points, recycle ratios etc...) can be optimized confidently on a full scale plant. A detailed costing of capital and operating costs has been presented and can be used in assessing the benefit of implementing such monitors as part of the complete control system. The investment and operating costs are paid for in less than two years, not taking into account the benefits of optimized plant operation.

As quality requirements and plant reliability are crucial for reuse applications, using on-line monitors offers an excellent tool for enhanced plant operation. On the one hand, it allows for a more efficient use of the existing assets. On the other hand, it enables to get a good insight into performance data and to analyse the risks.

Table 7 : Maintenance Schedule Sample

Day	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
Turbidimeter:																														
Draining	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■
Cleaning		■							■						■							■							■	
Calibration	■																													
Nitrate analyser:																														
Zero checking			■							■						■							■							■
Washing solution preparation	■																													
Ammonia analyser:																														
Electrodes rinsing	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■	■
Vessel cleaning			■							■						■							■							■
Alkali preparation	■				■				■				■				■				■				■				■	
Standard preparation	■																													
pH calibration				■													■													
Phosphate analyser:																														
Measurement cell cleaning				■							■						■							■						
Ultrafiltration unit:																														
Filter washing.					■							■						■							■					

Figure 1: On-line monitoring installation



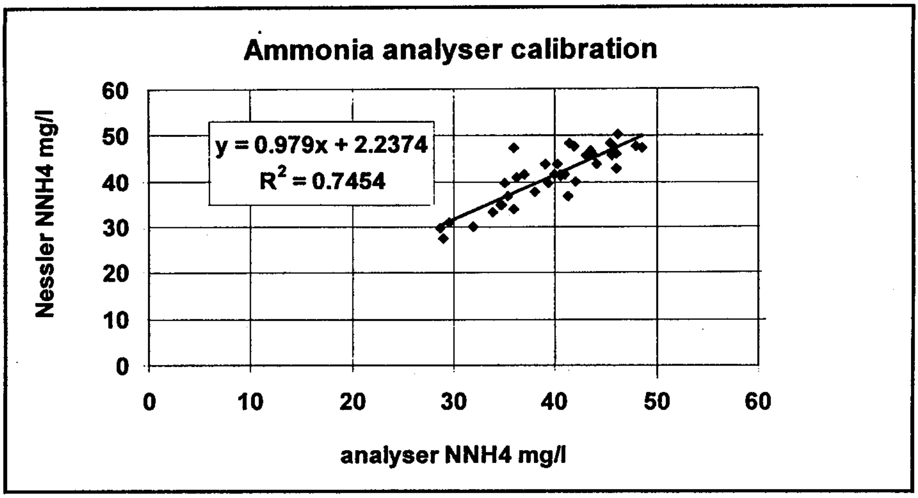


Figure 2 : Ammonia analyser calibration

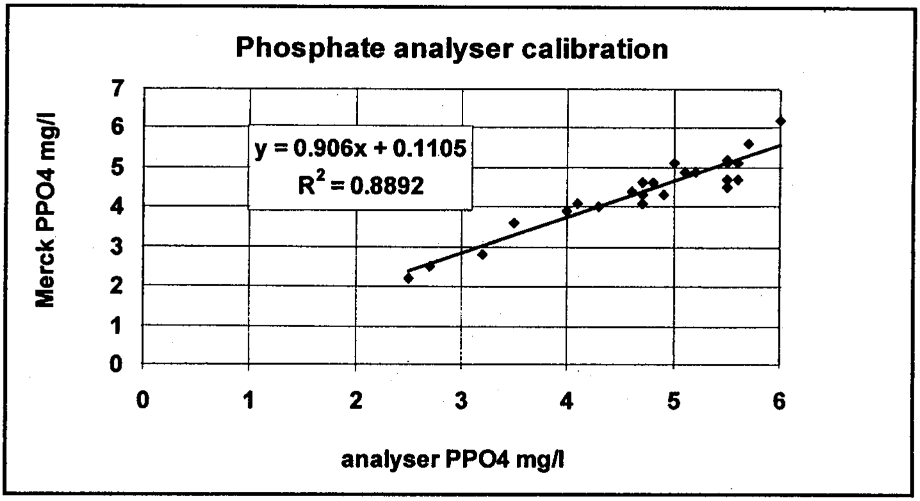


Figure 3 : Phosphate analyser calibration

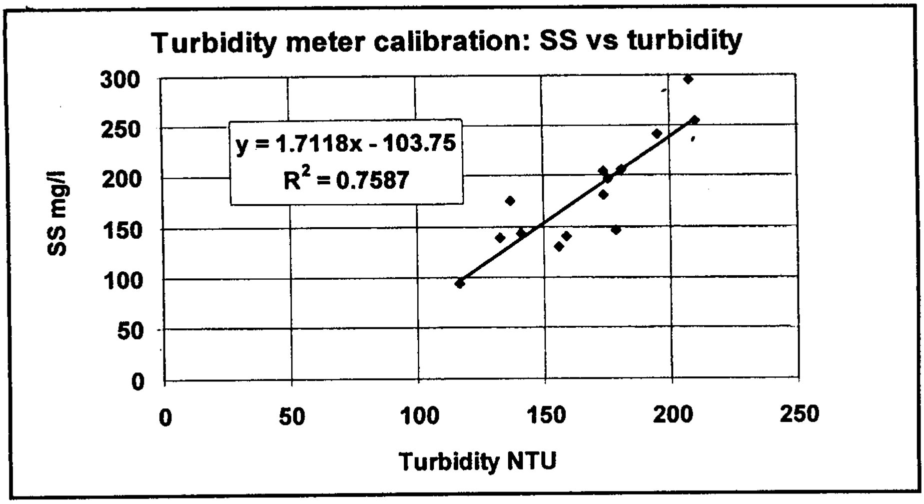


Figure 4 : Turbidity meter calibration : SS vs turbidity

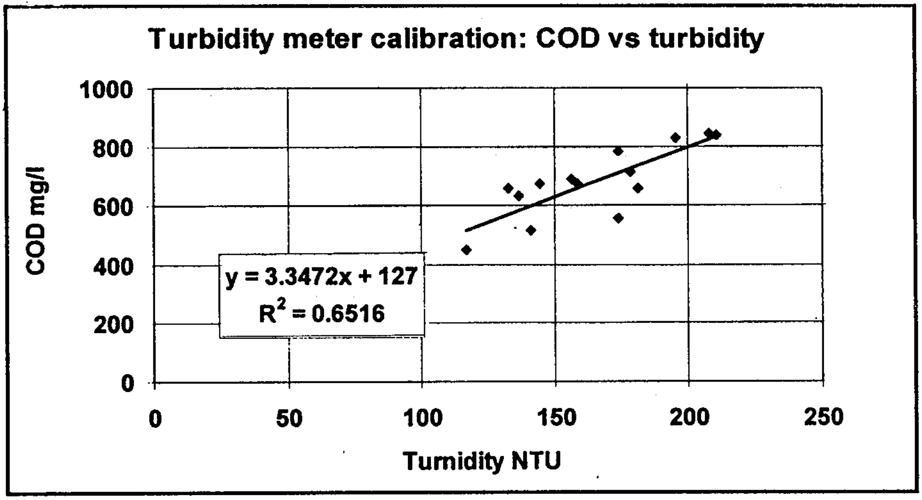


Figure 5 : Turbidity meter calibration: COD vs turbidity

Optimization of Sand Filtration Process for Tertiary Treatment

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OPTIMIZATION OF SAND FILTRATION PROCESS FOR TERTIARY TREATMENT

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ABSTRACT

Tertiary treatment using sand filtration process is considered reliable, simple and efficient process for the purification of secondary treated wastewater and produces water qualities which is within the recommended standards for agricultural reuse. However there seems to be some controversy over the suitable range of major design parameters, namely filtration rate and filter depth, to be used in designing such process. The main objectives of this study were to evaluate the efficiency and suitability of local sand for tertiary filtration and to determine the optimum values for flow rates and media depth relevant to the local conditions. A pilot plant, which consisted of three filtration units filled with local sand, was constructed at Riyadh wastewater treatment plant (RWTP). The pilot plant was fed with Riyadh South Plant secondary effluent and operated at different filtration rates and media depths. Filtration rates ranging from 4 to 30 m / hr. and media depths of 100 cm and 150 cm were used. Samples were collected at different depths and analyzed for BOD, SS and turbidity to study the effect of depth and flow rate on removal efficiencies and the head loss through the filtration depth. The results showed that local sand with grain size range of 2 - 3.35 mm and effective size of 2.4 mm with uniformity coefficient of 1.2 is a suitable media for tertiary filtration of secondary effluent. Using flow rates in the range of 8 - 12 m / hr. produce very high quality effluent with BOD, and SS less than 10 mg/L. These rates may be used in designing sand filtration process as an average and peak filtration rates respectively. Also it was found that increasing media depth from 100 cm to 150 cm had little effect on the filtration efficiency, therefore, using media depth more than 100 cm is not recommended.

KEY WORDS: Tertiary Filtration - Sand Filtration - Tertiary Treatment-Filtration of Secondary Effluent.

INTRODUCTION

Treated wastewaters are an asset that must be fully developed and integrated into the total water resources especially in arid regions. Saudi Arabia has put into operation a very ambitious program for wastewater reuse for irrigation, industrial, and recreational purposes. Most new wastewater treatment plants provides tertiary treatment level and many old plants were upgraded with tertiary treatment stage to produce high quality effluent in conformity with the water quality standards for unrestricted agricultural reuse.

The Riyadh wastewater treatment plants (RWTP) treat at present more than 400,000 m^3/day. The Petromin oil refinery uses some 20,000 m^3/day of the effluent and the balance is available for irrigation at Dirab, Dariyah and Ammariah.

Due to the increasing demand for reuse of treated wastewater effluent in the agricultural and industrial sectors and the stringent criteria imposed by the Ministry of Agriculture and Water for such uses, filtration of secondary effluent became a must. Riyadh wastewater treatment facility consist of two separate plant known as South plant (SP) and North plant (NP). The SP average design capacity is 200,000 m^3/d. The plant is a high-rate trickling filter which was upgraded recently with sand filtration units. The NP average design capacity is 200,000 m3/day. The plant is a single-stage activated sludge incorporating the nitrification-denitrification process followed by sand filtration.

The Riyadh Plants Filtration systems were designed by different consultants and each system was designed with different parameters. The designed parameters were chosen based on American and European experiences which might not be suitable for local conditions and could result in over or under design for such system. The SP Filtration units were designed with Filtration rates of 8 and 12 m/hr for average and peak flow respectively and media depth of 150 cm while the Filtration rates for the NP Filtration units were 3.12 and 5 for average and peak flow respectively and media depth of 90 cm.

Due to these discrepancies in the design parameters and in order to establish some optimum design parameters relevant to Saudi Arabia which will help in producing better and more economical design a Filtration pilot scale study was conducted for two years. The objectives of the pilot study were to evaluate the suitability and efficiency of local sand for tertiary filtration and to determine the optimum values for flow rate and media depth relevant to the local conditions.

BACKGROUND

The water quality standards for irrigation with treated wastewater effluents are based on health aspects, agronomic factors and type of irrigation practice [1]. It is granted that the most important consideration is potential hazard to public health, however, quality requirements have to be determined by the possible adverse effects on the irrigated crop by chemical constituents in the water and also by the potential problems of clogging the irrigation system [1].

The proposed criteria for Treated Wastewater Effluents to be used for unrestricted irrigation and the International (FAO) and American (EPA) quality guidelines are shown in Table 1. It should be noted that the proposed Saudi Arabian Standards for BOD and SS agree with those presently adopted by the Gulf States [2].

In order to maintain high quality treated effluent acceptable for unrestricted irrigation and meet the requirements of Riyadh Irrigation Master Plan, the Riyadh Region Water and Wastewater Authority (RRWWA) and ARRIYADH Development Authority (ADA) specify the use of rapid granular filtration and disinfection as a final stage of treatment [3, 4, 5, 6, 7, 8]. There is almost a unanimous agreement among researchers and designers that rapid granular filters are required before reclaimed water can be utilized for unrestricted irrigation [1, 9, 10, 4, 12].

Proper media selection can save money and improve performance. Single, dual and multimedia as shown in Table 2 have been used in wastewater filtration including fine and coarse media options. However, there is no general agreement as to which is better or more effective [13].

The current trend is to use single-medium, deep filters with large grain size having an effective size of 1.5 to 2.0 mm. Many studies had shown that the single medium deep filter performance is competitive with dual and multi media filters [14, 15, 16]. Sarah [17] concluded that media size has dramatic effects on removal efficiency and headloss. Increasing media size (1.85 mm) allows use of the entire bed for removal and reduces headloss and decreases in particle removal may be offset by increasing the filter depth.

Young [18] reported that the physical nature of biological wastewater solids is such that media size up to 2.5 mm effective size will provide satisfactory removal.

Filters were usually designed for a standard rate of 5 m/hr which was the accepted rate for water filtration plants. Numerous studies on tertiary filtration conducted over the years indicated that high rates are possible to attain without sacrificing effluent quality [15, 19, 20, 21, 22, 23, 24, 25]. The filtration rates studied by different researchers are summarized in Table 3. Other investigators had concluded that the optimum filtration rates for tertiary filtration of secondary effluent are in the range of 7 to 12 m/hr [16, 18, 20]

PILOT PLANT

Three filtration units filled with local sand were constructed at RWTP. The pilot plant was fed with SP secondary effluent and operated at filtration rates ranging from 4 to 30 m/hr and media depths of 100 cm and 150 cm. The size of the sand media used was in the range of 2 - 3.35 mm, which is similar to that used in the SP, brought from a site at Al-Saffaneyyah in the eastern province of Kingdom of Saudi Arabia. The details of filtration unare shown in Fig. (1).

Testing of filters media samples was carried out, in the College of Engineering Laboratory [26], according to the methods specified or recommended by American Water Work Association (AWWA) standards for filtering materials [27], and all tests were performed in duplicate to ensure accuracy. The test results of Filter media materials are shown in Table 4.

After two hours of starting operation, the first water samples were collected form the filters influent and at media depths of 100 cm and 150 cm. Sampling was repeated every 4 hours until the end of the run. These samples were analyzed for BOD, SS and turbidity according to the standards method [34]. Also headloss measurements were taken after, two hours of starting the filtration cycle at seven levels starting from the top of the media, at 12.5 cm depth and at five levels with 25 cm intervals of media depth.

Filters operation was continued until reaching a termination headloss of 170 cm after which the backwash cycle was started. Backwashing continue for 15 minutes with water and air and chlorine was added to the wash water at concentration of 20 mg/L to prevent algal growth in the media.

Table 1. Summary of Water Quality Criteria for Unrestricted Irrigation [1,28,29]

Maximum Contaminant Level (MCL) in mg/L except where noted				
Parameter	Proposed (a) Saudi Standard[28]	F A O [29]	E P A [1]	
			Agricultural	Landscaping
BOD	10 (b)	-	-	20
TSS	10 (b)	-	-	15
Aluminum	5	5	5.0	5.0
Arsenic	0.1	0.1	0.10	0.1
Beryllium	0.1	0.1	0.1	0.1
Boron	0.7	0.75	-	-
Cadmium	0.01	0.01	0.010	0.01
Chromium	0.1	0.1	0.10	0.1
Cobalt	0.05	0.05	0.050	0.05
Copper	0.4	0.2	0.20	0.2
Cyanide	0.05	-	-	-
Fluoride	2.0	1.0	1.0	2.0
Iron	5.0	5	5.0	5.0
Lead	0.1	5	5.0	5.0
Lithium	2.5 (c)	2.5	2.5 (c)	2.5
Manganese	0.2	0.2	0.20	0.2
Mercury	0.001	-	-	-
Molybdenum	0.01	0.01	0.010	0.01
Nickel	0.2	0.2	0.20	0.2
Selenium	0.02	0.02	0.02	0.02
Vanadium	0.1	0.2	0.1(	0.1
Zinc	4.0	2	2.0	2.0
pH in units	6.0 to 8.4	-	-	6.0 to 9.0
Oil & Grease	Absent	-	-	Nil
Phenol	0.002	-	-	50
Intestinal Nematodes	1 Viable egg/L	-	-	-
Fecal Coliform MPN	2.2/100ml(d)	-	1000/100mL	2.2/100 mL.
Turbidity	2.0 NTU	-	-	-
Chloride (e)	280	-	-	100-200
Sulfate (e)	-	-	-	200-400
Nitrogen as N (f)	-	-	-	-
Sodium adsorption ratio	-	-	-	8-18

a. Wastewater quality must meet or be better than the MCL listed 80% of the time, based on the last 20 consecutive samples.

b. Monthly average BOD and TSS concentrations shall not exceed 10 mg/L, and weekly average values shall not exceed 15 mg/L.

c. Value is 0.07 (Saudi Standards), 0.075 (others) for citrus crops.

d. MPN shall not exceed 2.2/100 mL for seven consecutive days. MPN shall not exceed 23/100 mg/L, severe problems.

e. According to WPCF Manual (2), for total salt content below 500 mg/L, no problems; and above 2000 mg/L, severe problems.

f. According to WPCF Manual (2), Nitrogen in irrigation water causes no problems if its concentration is below 5 mg/L. increasing problems if it is between 5 and 30 mg/L, and severe problems if it exceeds 30 mg/L.

Table 2. Media Options [15]

Type		Material	Size (mm.)	Depth cm
Single medium	Fine	Sand	0.35 - 0.60	25-50
	Course	Anthracite coal	1.3 - 1.7	90-150
Dual media		Sand	0.45 - 0.60	25-30
		Anthracite coal	1.0 - 1.1	50-75
Multimedia		Garnet *	0.25 - 0.40	5-10
		Sand	0.45 - 0.55	20-30
		Anthracite coal	1.0 - 1.1	45-60

* Other types of material such as metal oxides are also used. (31)

Table 3. Some Filtration Rates Reported in the Literature

Author's and Researchers	Filtration Rates (m / hr)
Brown [30]	5.04, 7.2, 9.72
Baylis [31]	5.04, 9.72, 11.16, 12.24
Robeck et al [32]	5.04, 9.72, 14.76
Culp [19]	10 – 40
Tchobanoglous [20]	10 – 40
Conley [22]	20 – 25
W.I.Rechwal et al. [14]	5-8(Average) and 8-21 (Maximum)
Tebbutt [23]	4.17 – 25
Darby et al. [33]	6.48 - 13.32
Dawda et al. [24]	7.32, 9.78, 12, 14.64
M.J.Hammer [25]	7.2 - 9.7

Table 4. Test Results for Filter Media [26]

No.	Test	Sample No. 1	Sample No. 2	AWWA Standards (B100-80)	Remarks
1	Acid solubility	.15 %	0.12 %	Not more than 5%	
2	Specific gravity	2.66	2.66	Not less than 2.50	
3	Sieve analysis				
	a. Effective size, E.S. (mm.)	2.4	2.4		
	b. Uniformity coefficient	1.17	1.21		Tolerance
	c. Size range (mm.)	2-3.35	2-3.35		± 1 %
4	Organic impurities	Lighter than std. color	Lighter than std. color	Not darker than std. color	
5	Fine particles	0.5%	0.6%	Not more than 2%	

RESULTS AND DISCUSSION

The pilot plant results show that the single-media filters comprised of unstratified local coarse sand with grain size range of 2 - 3.85 mm and effective size of 2.4 mm with uniformity coefficient of 1.2 is a suitable media for tertiary filtration of secondary effluent and produce very high quality effluent with BOD and SS less than the recommended standards for unrestricted reuse of 10 mg/L.

The pilot filters were operated at filtration rate ranging from 4 to 30 m/hr. The results show that operating at filtration rates of 15 m/hr or higher produces effluent with BOD and SS concentrations higher than the recommended level of 10 mg/L and the removal efficiency decreases as the filtration rates increases as shown in Fig. (2). Also the filtration cycles at these rates were shorter than the recommended 24-hr cycle due to high headlosses as shown in Fig. (3) and Fig. (4).

For typical secondary effluent (SS £ 30 , BOD £ 30) the filtration rates of 8 - 12 m/hr produce very high quality effluent with BOD and SS less than recommended level of 10 mg/L (most of the time) as shown in Fig. (2) and Fig. (5) through Fig. (9). However, in cases where BOD and SS

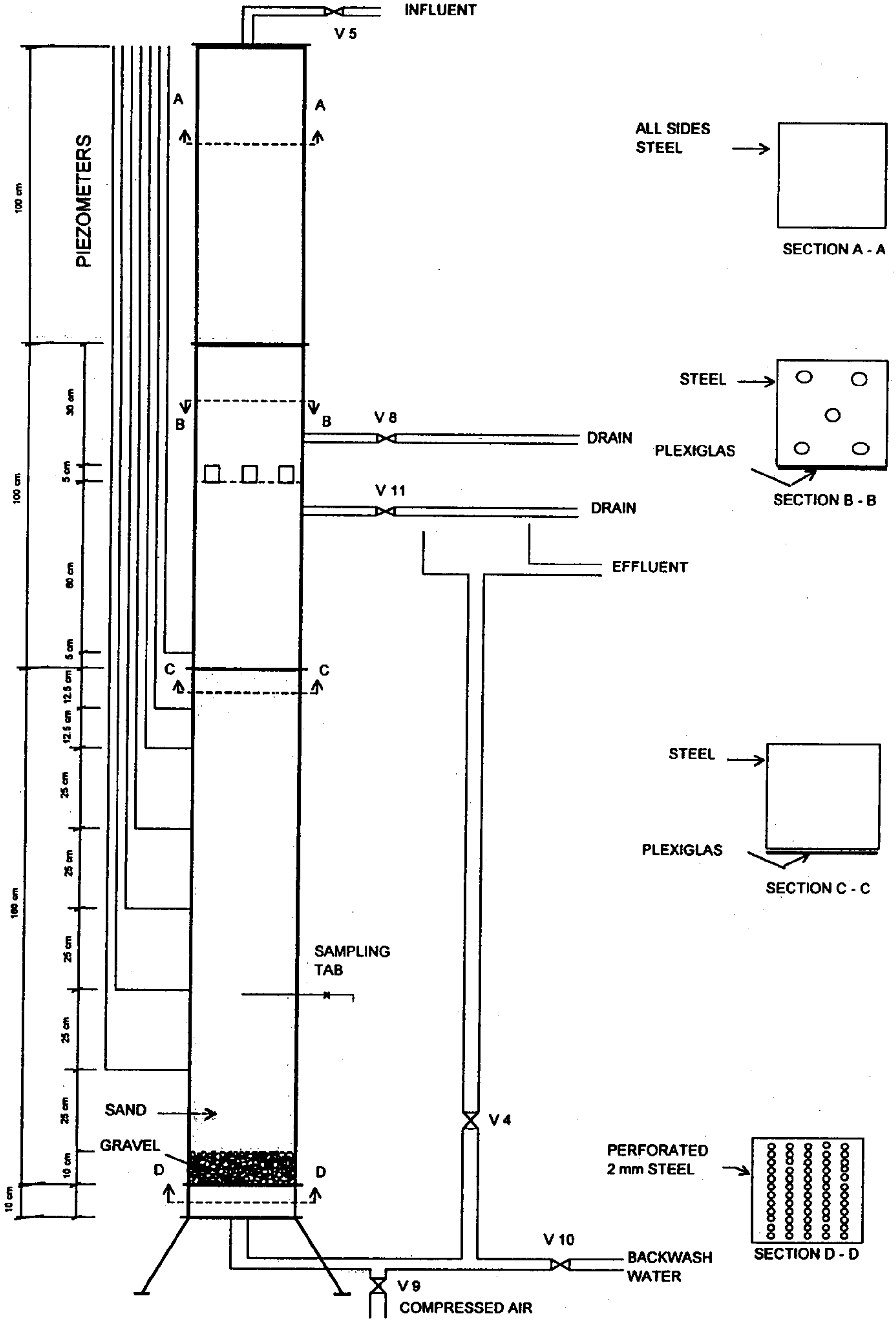

Figure 1: Detailed Sections of Pilot Filters

concentrations was higher than 30 mg/L, the filtration plant failed to produce water quality within the specified limits for unrestricted agricultural reuse.

Media depth of 100 cm and 150 cm had been evaluated for all different filtration rates. The results showed that there was no significant differences in the removal efficiency between the two depths as demonstrated in Fig. (2) and Fig. (5) through Fig. (8). Therefore increasing media depth beyond 100 cm is not justifiable from both operational and economical point of view.

Headloss development in a granular-media filter is a function of hydraulic and solid retention capacity. Since the filters were operated with terminal headless of 1.7 m, all filters that running at filtration rates higher than 12 m/hr were failed to complete the 24-hr cycle as illustrated in Fig. (4). The headloss development in the filters at different filtration rates are presented in Fig. (3) which also showed that it is not possible to achieve the required 24-hr cycle for filtration rates higher than 12 m/hr.

As shown in Fig. (10) more than 90 percent of the headloss occurred in the top 25 cm of the 150 cm for the optimum filtration rate of 8 m/hr and 12 m/hr. This means that most of the suspended particulates in the secondary effluent were removed in the upper 25 cm of the filter media and increasing the depth of filter media beyond a certain depth would not improve the solids removal efficiency for the single media filter.

If the design of the existing filtration plants at the SP and NP is to be revised based on the results of this study the filter area of the NP would be reduced by 60% while the filter media depth in the SP would be reduced by more than 30%. This would result in a great saving in the capital cost as well as in the operational cost.

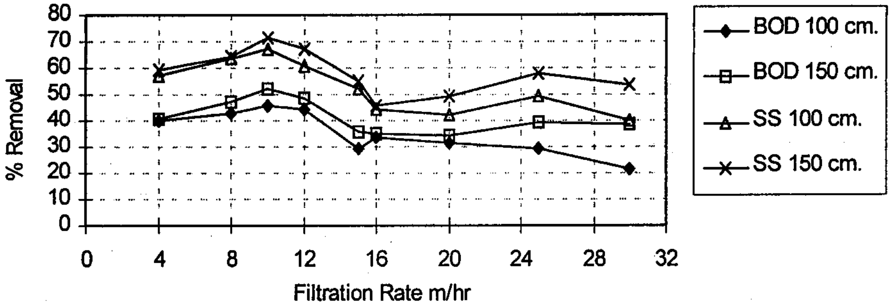

Figure 2. SS and BOD Removal Effeciency and Filtration Rate Relationship

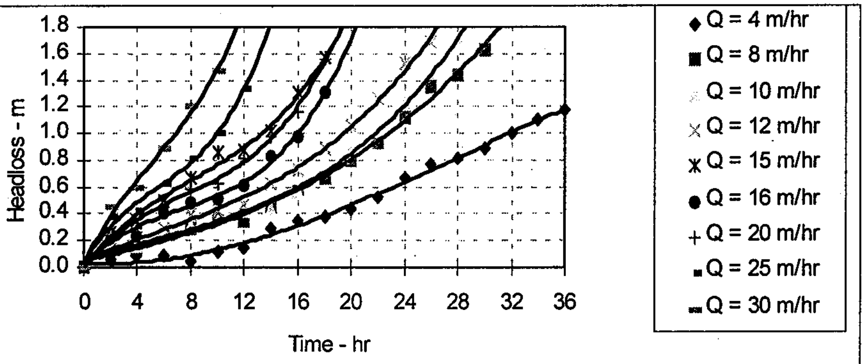

Figure 3 Total Head Loss for Different Filtration Rates

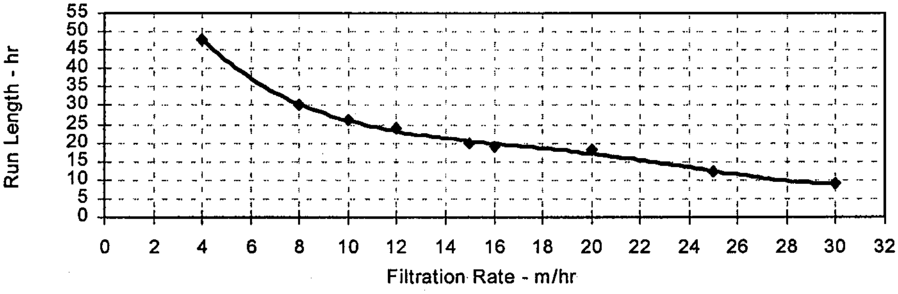

Figure 4. Filtration Rate and Run Length Relationship

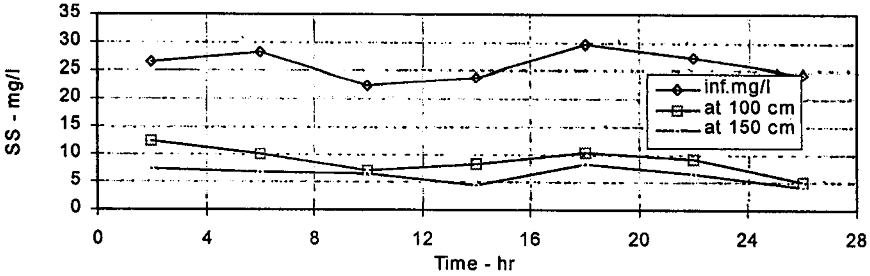

Figure 5. SS Concentration for Filtration Rate of 12 m/hr

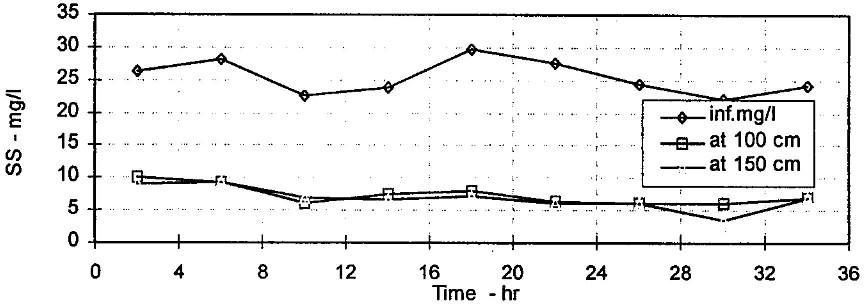

Figure 6. SS Concentration for Filtration Rate of 8 m/hr

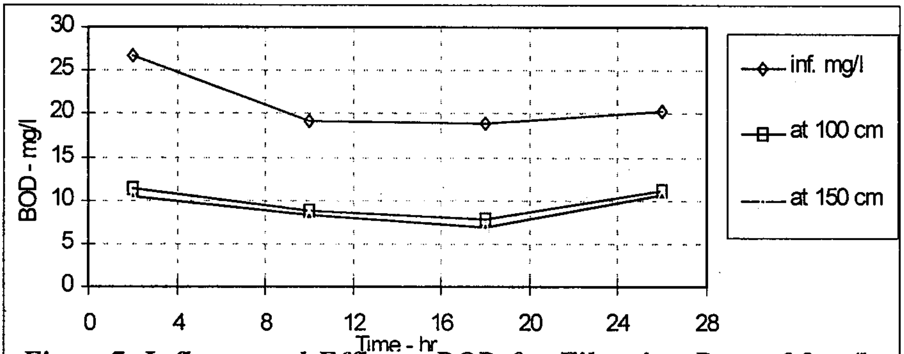

Figure7. Influent and Effluent BOD for Filtration Rate of 8 m/hr

Figure 7. Influent and Effluent BOD for Filtration Rate of 8 m/hr

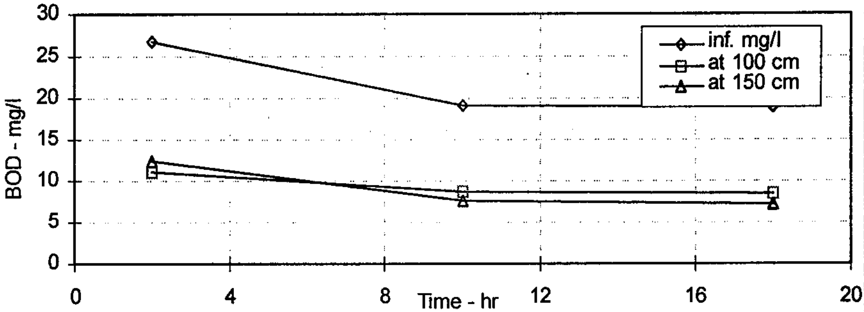

Figure 8. Influent and Effluent BOD for Filtration Rate of 12 m/hr

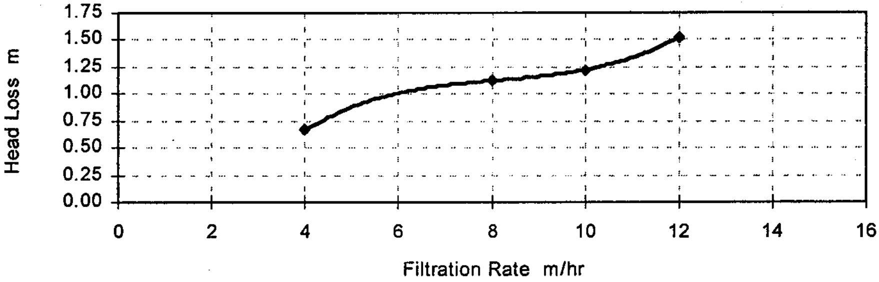

Figure 9. Total Head Loss Development Curve

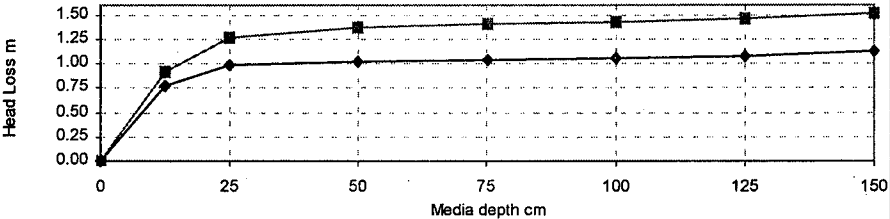

Figure 10. Head Loss Distribution in Filter Media for the Rates of 8 and 12 m/hr

CONCLUSION AND RECOMMENDATIONS

It is very important to establish and approve local sources for filtration media materials because importing such material from abroad, which is the current practice for most filtration plants in the Kingdom is very costly. Single media filters comprised of local course sand with grain size range of 2 - 3.35 mm and effective size of 2.4 mm with uniformity coefficient of 1.2 is a suitable medium to be used for treating secondary effluent and producing effluent with SS and BOD concentrations satisfying the standard criteria for reuse of < 10 mg./L. Increasing the media depth beyond 75 cm would have little effect on filtration efficiency, therefore a maximum media depth of not more than 100 cm is recommended. Filtration rates of 8 m/hr and 12 m/hr for average and peak flow rates respectively, are recommended to be adopted in designing future tertiary filters for conventional secondary treatment plants in the Kingdom.

Filtration rates of higher than 12 m/hr are not recommended because they produced effluent with high BOD and SS concentrations and very high headlosses with short filtering cycles. Achieving turbidity level of NTU specified by Saudi Standards for unrestricted irrigation was not possible. The turbidity level of filter effluent was in the range of 4 - 6 NTU. Other standards such as FAO and EPA did not specify limits for turbidity therefore turbidity level specified by Saudi Standards must be revised because the 2 NTU level is very restricted.

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Comparative Performance Analysis of Two Techniques to Renovate Tertiary Treated Wastewater

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COMPARATIVE PERFORMANCE ANALYSIS OF TWO TECHNIQUES TO RENOVATE TERTIARY TREATED WASTEWATER

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ABSTRACT

In Kuwait, wastewater effluent is treated to secondary/tertiary levels. The relatively low salinity of the treated wastewater, i.e., 1,500 parts per million (ppm), compared to that of brackish water, i.e., 4000 to 5000 ppm, makes it a potentially excellent source for good quality water to be used for general purposes, except for direct human consumption. Different methods can be employed to renovate/polish this treated wastewater effluent for agricultural, industrial and as feed water for reverse osmosis plants. The most commonly used methods are conventional treatment and membrane separation processes. Conventional treatment is cumbersome and costly. Microfiltration (MF) is a simple, viable and possibly cost-effective separation process for wastewater reclamation. In Kuwait, both techniques are evaluated under comparable conditions at Ardiya wastewater treatment plant. Characterization of the feed and filtrates of both treatment methods and problems are investigated. This paper discusses the performance evaluation of the two techniques in polishing tertiary treated wastewater as pretreatment for reverse osmosis plant, which were operated for one year.

KEYWORDS: Dual-media filtration, conventional, microfiltration, flocculation, coagulation, cost

INTRODUCTION

Water Resources in Kuwait are limited to desalinated fresh water which constitute 95% produced by seawater desalination processes, mainly multi-stage flash (MSF) and the remaining,5% is added from ground water resources (MEW statistical year book 1996). The rapid increase in demand for more water for needed industrial, agriculture and domestic necessitates the search for other available water resources.

In Kuwait wastewater generation is usually estimated to be about (70-80%) of the fresh water consumption for domestic (Al-Awadhi. et. al, 1992). This indicates that wastewater generation is directly proportional to water consumption and time. Hence this huge source of water of low salinity (<1500 mg/l) that is partially discharged to the sea, could be treated directly by advanced process such as MF technology to produce high quality water suitable for agricultural, industrial and non domestic uses at a reasonable cost.

The wastewater treatment process in Kuwait consists mainly of screening, grit removal and biological treatment by activated sludge. Sludge is thickened and digested before being further thickened and air-dried on open sludge beds where the sewage gas from the digestion process is burned off. All effluent from the treatment plant receives tertiary treatment. The flow passes of contact tanks where it is chlorinated and then passed through sand filters to remove the last remaining impurities and suspended materials.

Advanced wastewater treatment is designed to remove those constituents that are not adequately removed in the secondary treatment stage. These include nitrogen, phosphorus, and other soluble organic and inorganic compounds. Nitrogen and phosphorus are nutrients that accelerate the plant growth in the receiving water. Ammonia is toxic, exerts nitrogenous oxygen demand, and increases chlorine demand. Heavy metals, hydrogenated hydrocarbons, phenolic compounds, etc., are also toxic to aquatic life, concentrate in the food chain, and may create taste and odor problems in water supplies. Many of these constituents must be removed to meet more stringent water quality standards and also to be used for industrial, irrigation, recreational, and other non-potable water use.

The most commonly used advanced wastewater treatments are chemical precipitation of phosphorus, nitrification, denitrification, ammonia stripping, breakpoint chlorination, filtration, carbon adsorption, ion exchange, reverse osmosis (RO), Ultrafiltration (UF) and microfiltration (MF).

Membrane separation is used to polish wastewater effluent. MF is a successful filtration technique that is already applied to surface seawater, brackish water and industrial and urban wastewater. It relies on a membrane separation technique that requires low pressure to force good quality filtrate water through the membrane that rejects harmful contaminants including bacteria (kolega,et el,1991).

MF technique is applied in the range of 0.1 to 2 µm and can, theoretically, reject most of the harmful constituents of the treated wastewater. The MF technique requires very low pressures (low energy consumption) and low chemical consumption (for membrane cleaning). The novel MF process has been successfully used in the recovery of valuable products from effluents, sterilization of beverages and pharmaceuticals. MF has found application in analytical and quality assurance tests in industrial wastewater (Fane et al., 1995). Applying the MF technique to polish treated wastewater is expected to produce a technically feasible and cost effective method of treatment.

A good quality feed water is required for reliable operation of a reverse osmosis plant. Membranes easily fouled by colloidal particle in the wastewater feed stream, further treatment or polishing the tertiary effluent is necessary to ensure a successful operation of reverse osmosis plant. The present work covers the performance evaluation of two pretreatment technique as pretreatment for reverse osmosis unit.

DESCRIPTION OF TREATMENT SYSTEMS

1. Conventional treatment system

In-line coagulation and flocculation of the feed water are the basic unit of the conventional treatment system. During normal operation, the feed is treated and filtered through a pressurized filtration system.

A schematic diagram of the conventional treatment system is shown in Fig. 1.The filtration system consists of two 1-m^3 glass reinforced plastic (GRP) filter tanks (multi-media sand filter), feed and backwash pumps, filtrate tank (pipes and valves), two chemical dosing pumps and agitators for Ferric Cloroulphate ($FeClSO_4$) and Polyelectrolyte.

The buffer tank is fed with tertiary treated wastewater from the outfall of the Ardyia wastewater plant. The $FeClSO_4$ and Polyelectrolyte are dosed into the feed stream before entering the buffer tank. The feed pump delivers the effluent to the sand filters, where the water is filtered and fed to the filtrate tank. The Silt Density Index (SDI) is continuously monitored. The system is cabable to produce 1 m^3/h. filtrate water.

2. Micofiltration system

The tertiary effluent is pumped by pool pumps to a break tank after passing through a coarse pump strainer and injected with 2.5 ppm of sodium bisulphite ($NaHSO_3$) to remove any traces of chlorine. The feed stream is fed to a continuous microfiltration (CMF) unit (20M10C Memcor) to remove fine solids and micro-organisms. The CMF unit comprises of two filtration modules, each containing hollow fiber polypropylene membranes having an average pore size of 0.2 μ. Table 1 presents technical specifications for a MF unit. Water enters the modules at one end, flowing along the outside of the membrane and through the wall, removing virtually all suspended solids, fecal and other bacteria and parasites. Fig. 2 shows a schematic diagram of a MF system.

A key element of the CMF system is the patented air backwash. When pressure across the fiber membrane reaches a pre-set level, filtration stops and high-pressure air is injected into the fiber tubes. The air blasts through the fiber wall dislodging accumulated sediments sitting on the outer surface. A sweep of feed flushes the dislodged sediment to a sewer.

Over time, the driving pressure climbs to a level (around 120 kPa) where chemical cleaning of the membrane is required. This is achieved by a fully automatic clean-in-place (CIP) system using a 2% caustic solution and surfactant.

A filtrate tank of 250 *l* is installed at the product side of the system to maintain a steady flow of filtrate water to the silt density index (SDI) system during the measuring time.

Table 1. Average Chemical Characteristics for the Feed and the Filtrat of the conventional and Microfiltration systems

Parameter	**Feed (mg/*l*)**	**Filtrate (mg/*l*)**	
		Conventional	**Microfiltration**
TDS	957	958	972.6
Total Alkalinity	130.8	132.8	133
p-Value	0	0	0
HCO_3	137	126.6	126.6
CO_2	9.5	7.8	6.6
Na^+	157.8	168	168.6
K^+	11.36	18.56	22.96
Mg^{+2}	20.6	20.9	20.9
Ca^{+2}	87.8	85.6	85.8
Al^{+3}	< 0.1	< 0.01	< 0.01
6th-	0.31	0.287	0.276
CL^-	275.8	258.3	258.4
SO_4	260	263	265.8
NO_2	0.322	0.342	0.35
NO_3	0.36	0.358	0.361
Ammonia	21.6	22.0	22.0
Total Fe	< 0.05	0.34	< 0.05
TOC	< 1	< 1	< 1
COD	27.8	20	11.25
BOD	6.3	3.8	2.6
Total Count	40×10^7	31×10^7	21×10^7
Fecal-Coli	0	0	0
Salmonella	0	0	0

Table 2. Operational Parameters for the Feed of the conventional and Microfiltration Units

Parameter	**Average**	
	Conventional	**Microfiltration**
Feed Flow (m³/h)	1.0	1.37
Filtrate Flow (m³/h)	1.0	'1.37
Feed SDI	> 6.00	> 6.00
Filtrate SDI	5.2	2.98
Feed Temperature (°C)	25.7	25.7
Filtrate Velocity (m/s)	0.18	0.22
Feed Conductivity (μS/cm)	1642	1642
Filtrate Conductivity (μS/cm)	1643	1645

RESULTS AND DISCUSSION

Both Microfiltration and conventional treatments were tested at Ardyia wastewater treatment plant under same conditions. In both cases, the units were fed with tertiary treated wastewater effluents. Table 1 shows the chemical characteristics for both systems and table 2 shows the operational parameters for both units, whereas Fig. 3 and 4 show the average monthly temperature of the feed water and average monthly conductivity of the feed stream respectively. It can be seen from these figures that the temperature of the feed was in the range of 22-32 °C during winter and summer and showed no significant changes on day to day basis. Therefore, the effect of temperature in the flux is almost negligible in both units. The conductivity of the feed was 1500 μS/cm and kept steady most of the time, except when diluted with rain water in the open outfall of Ardyia plant. The silt density index (SDI) for the two systems where measured and plotted as shown in Fig. 5 and 6. It can be seen that the SDI values of the filtrate from the conventional system were not steady as they vary between (4-6 %), whereas the SDI values of MF unit remained almost consistent with an average value 2.89 %. Fig. 7 and 8 show the filtrate flow rate for both units. It can be seen that the productivity of MF is kept steady with flow rate of 1.4 m³/hr whereas the filtrate of the conventional treatment was varying from 0.98 – 1.05 m³/hr.

Table 2, shows the quality of the feed and filtrate streams for both systems. It can be seen that no significant reduction in chemical oxygen demand (COD) and biological oxygen demand (BOD) values in the filtrate of the conventional treatment system, whereas, these values were reduced in the filtrate at the MF system.

During the operation of the conventional treatment (multi media filters) there was a need to backwash the system using the same filtrate of the unit. This process resulted in loss of production, whereas in the case of MF system, it uses the feed water as backwash media.

The techno-economic study indicates that the total water unit cost by MF is 12.264 fils/m^3 compared to 28.153 fils/m^3 for the conventional system (56% reduction in unit water cost) as indicated in Table 3.

Table 3. Comparison of Unit Cost of Waste water treatment by Alternative Systems

Cost Components	Unit Cost (fils/m^3)	
	Conventional	**Microfiltration**
A. Capital Cost (Depreciation)	10.365	3.537
Feed Pumps	0.119	0.119
Chlorine Dosing Plants	0.015	0.119
Structure-Intake	0.387	-
Structure-Outfall	0.387	0.387
Treatment Plant	9.547	-
Microfiltration Plant	-	2.644
B. Operating Cost	17.788	8.727
Electricity	9.648	7.090
Chemicals	4.662	1.637
Filters	3.478	-
Total unit Cost (A+B) (fils/$m^{3)}$	28.153	12.264

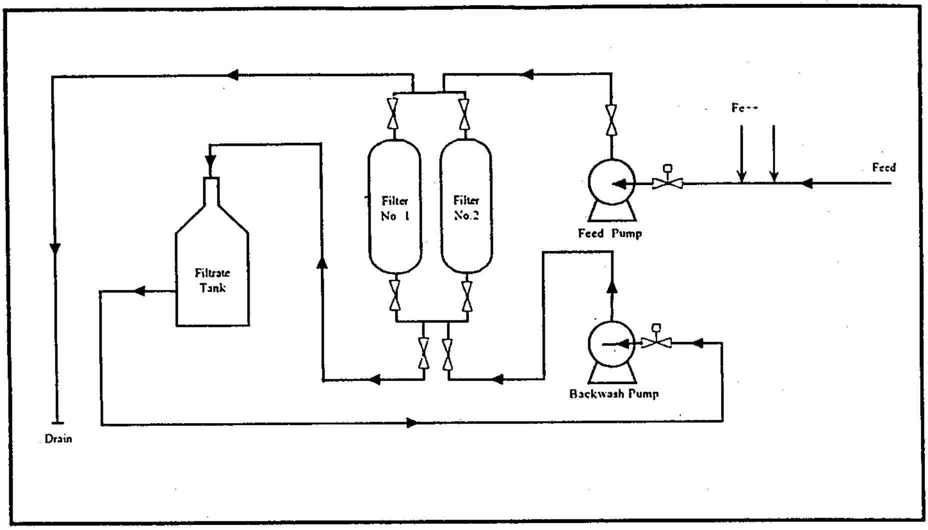

Figure 1: Schematic diagram of a conventional wastewater treatment system

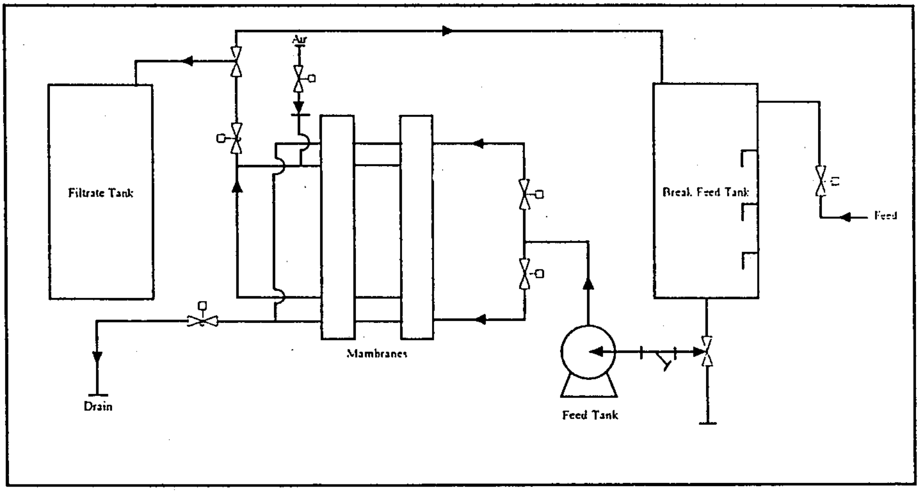

Figure 2: Schematic diagram of microfiltration system

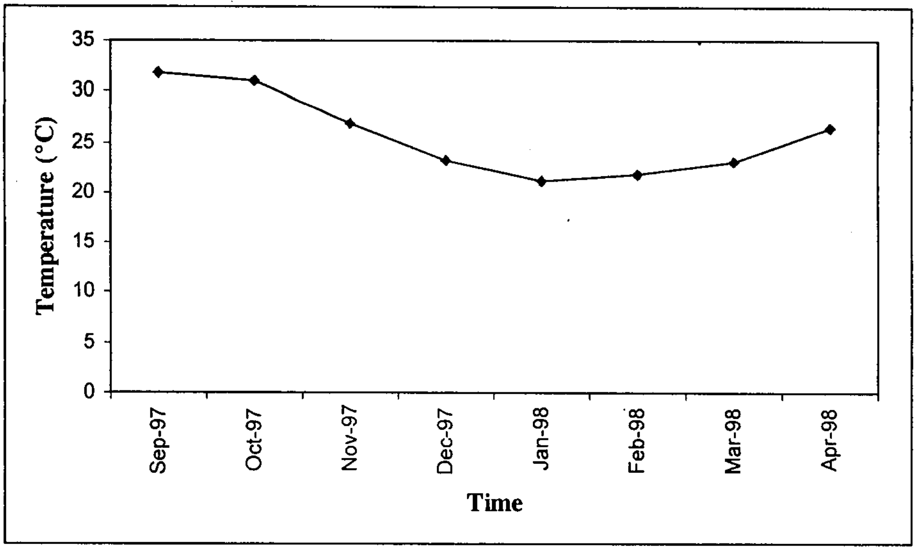

Figure 3: Average monthly temperature of the feedwater versus time

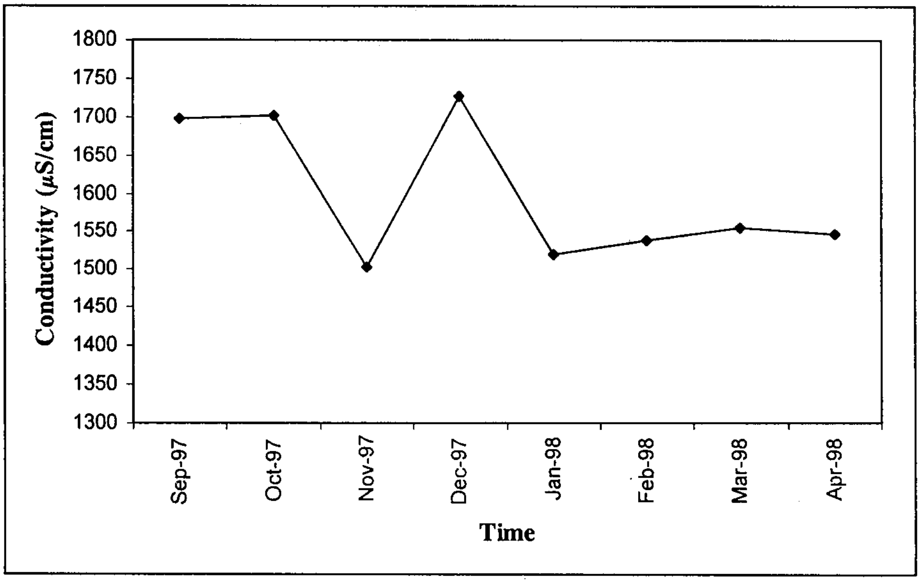

Figure 4: Average monthly conductivity of the feedwater versus time

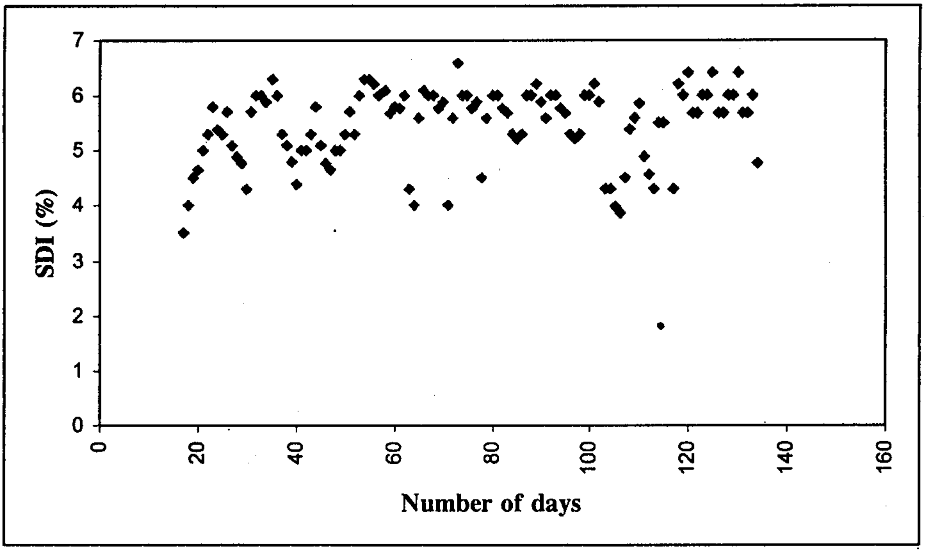

Figure 5: Silt density index (SDI) of conventional treatment versus number of days

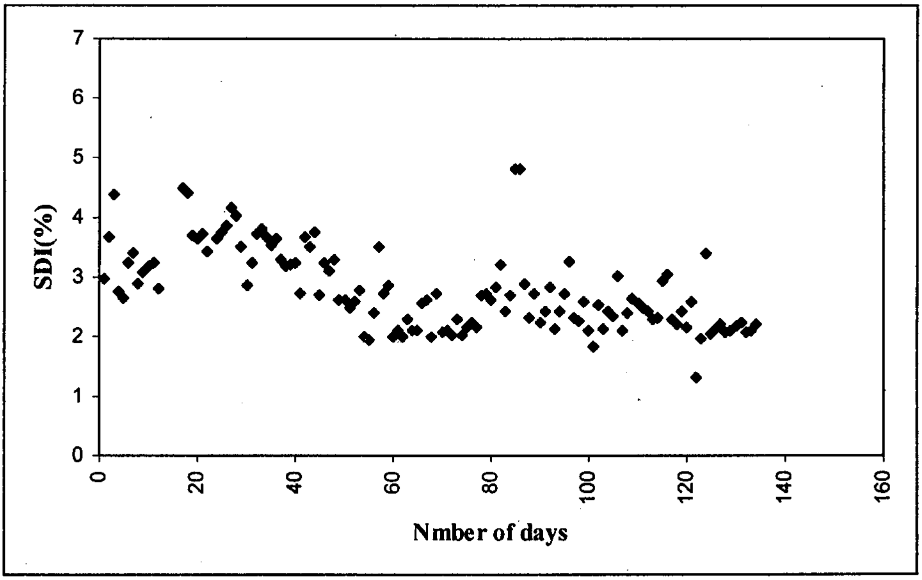

Figure 6: Silt density index (SDI) MF treatment versus number of days

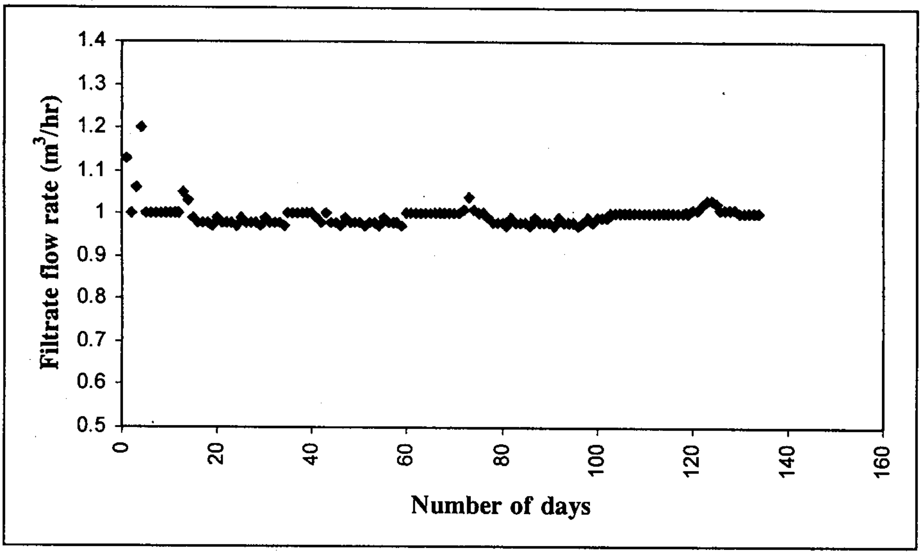

Figure 7: Filtrate flow rate of conventional treatment versus number of days

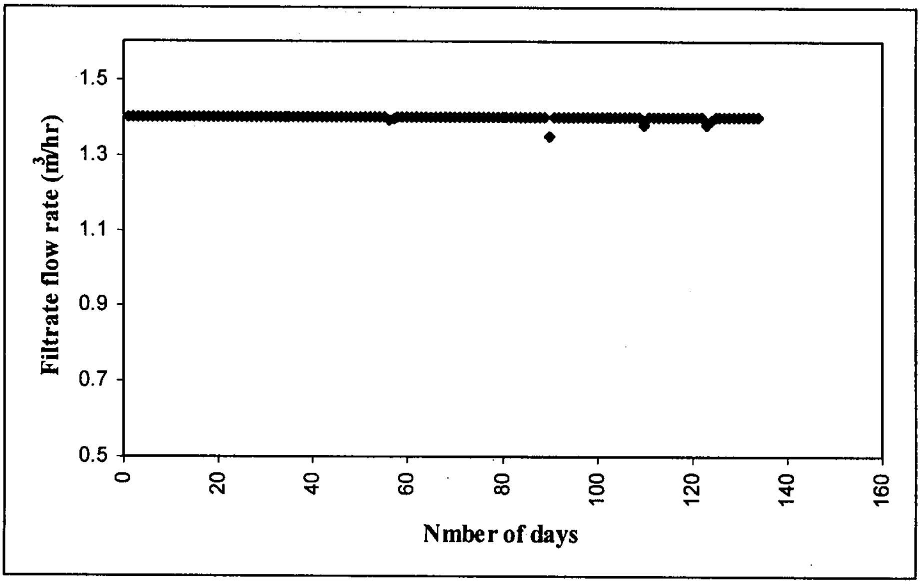

Figure 8: Filtrate flow rate of MF treatment versus number of days

CONCLUSIONS

The study shows that the microfiltration system was very effective in removing naturally occurring wastewater impurities.

Chemical analysis results revealed that the MF system has significantly improved the quality of the effluents with an average SDI (2.98%). There was consistent reduction in BOD, COD and bacteria-total count. Hence the MF filtrate can be considered safe for agricultural, industrial and as feedwater for RO plant.

The techno-economic study indicates that MF system is the most effective system for the treatment of wastewater effluent as compared with the conventional system. The MF has also the added advantages over conventional treatment as shown below:

1. Reduction in capital cost due to:
    - No standby capacity.
    - Elimination of chemical dosing systems.
2. Reduction in operating costs due to:
    - Less membrane replacements costs.
    - Less chemical consumption costs (no chemical are needed for disinfection and Coagulation)
    - Less maintenance costs for pumps and instruments.
    - Less labor costs.

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Lime Recovery form Riyadh Water Treatment Plants

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LIME RECOVERY FROM RIYADH WATER TREATMENT PLANTS

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ABSTRACT

There are two sources that provide Riyadh city with potable water. These include desalinated sea water from Al Jubail plants which accounted for about 60%. The remaining 40% comes from the treated brackish water from deep and shallow wells. Riyadh Water Treatment plants are operating at close to design capacity and presently operated and maintained by Riyadh Water and Sewage Authority. Massive amount of chemicals are used in these plants. The type and amounts of chemicals often used in water treatment plants basically depend on the ground water quality and chemical characteristics. However, large quantities of lime, $Ca(OH)_2$, are consumed for softening. The objective of the softening process is to prevent scaling of membranes in the RO units. The produced sludge from the softening is drained to the sludge thickener on a daily basis where it should be disposed. Proper sludge disposal is required for safe environment. So sludge must be dewatered before discarding. The dewatered softening sludge contains about 85% calcium carbonate ($CaCO_3$) and 12% magnesium hydroxide, $Mg(OH)_2$. The insoluble magnesium hydroxide can be converted into soluble magnesium bicarbonate by carbonation of the sludge with carbon dioxide, $C0_2$. The remaining solid will be rich in calcium carbonate which can be easily separated. After recalcining of the carbonated sludge calcium, oxide is produced which can be hydrated readily into lime.

This paper will focus on the feasibility of the recovery of lime from Riyadh water treatment plants. Brief description of softening process in Riyadh water treatment plants will be summarized as well as the characteristics of the softening sludge produced from these plants. Processes used for lime

production from softening sludge will be reviewed including sludge conditioning and dewatering. The technical description of these processes will be given.

KEYWORDS: Lime, lime recovery, sludge, water treatment.

INTRODUCTION

Riyadh has five large water treatment plants. These are Shemessy Water Treatment Plant, Malaz Water Treatment Plant, Sulbukh Water Treatment Plant, Manfouha Water Treatment Plant and Buwayb Water Treatment Plant. These plants were built by Degremont of France and use reverse osmosis system with hollow fine fiber polyamid membranes of B-9 (0840) permeators manufactured by DuPont except Buwayb plant which has a spiral wound cellulose acetate membranes of CAROGA 81SOS manufactured by UOP[1].

Shallow wells in Wadi Nisah, Wadi Namar and Wadi Hayir and the deep wells of Minjur formation are the principal sources of raw water in Riyadh. Table 1 shows the typical analysis of raw water after cooling and net water produced by the Riyadh Water Treatment Plants. Clearly large amount of hardness and alkalinity had to be removed as well as the disolved soilds[2].

Water is pumped from deep wells to cooling towers in the water treatment plants. The bulk of cooled water (the mainstream) is treated chemically, while a smaller portion (the slipstream) is sent directly to filters and then, to a mixing chamber for blending with the reverse osmosis permeate. The chemicals treatment of the mainstream is mainly lime and soda softening, coagulation, and pH adjustment. Acidified softened water is diverted to coarse and fine media of two stages open sand gravity filters. Filtered water is then stored in reverse osmosis feed storage tanks. SDI value of about 1.5 is achieved at this point. In order to minimize water waste, filter effluent (upon back washing and rinsing) is circulated back to clarifiers inlet for repeated chemical conditioning. Sulfuric acid is added to reduce RO feed pH to 6.5. Sodium hexametaphospahte ($(NaPO_3)_6$) is also dosed about 4 mg/l. Also AMF Cuno (SS 316) micro filters with 5 micron cartridges are installed for micro filtration of reverse osmosis feed water. Reverse osmosis product is post treated with sodium carbonate and transferred to a mixing chamber where it is blended with undesalinated but treated (and cooled only raw water). Post chlorinating is carried out in the mixing chamber prior to distribution.

*Table 1: **Typical Water Analysis***

Item		Raw Water	Net
Temp.	°C	33	32
pH		8.0	7.5
T. Hardness	ppm as $CaCO_3$	670	180
Ca. Hardness	ppm as $CaCO_3$	470	100
Mg. Hardness	ppm as $CaCO_3$	200	80
Total alk.	ppm as $CaCO_3$	160	55
Cond.	µS/cm	2000	770
TDS	mg/l. as ion	1350	490
Ammonia	mg/l. as ion	<0.24	0
Nitrite	mg/l.as ion	<0.016	0
Nitrate	mg/l. as ion	10	8.5
Chloride	mg/l.as ion	265	100
Sulfate	mg/l.as ion	500	170
Iron	mg/l. as ion	0.4	0.06
Alu.	mg/l. as ion	0	0.009
Chlorine	mg/l. as ion	0	0.36
Silica	mg/l. as ion	<26	-

In the softening process, large amounts of sludge are produced which must be disposed. The environmental legislation call for proper disposal of sludge. The problem of sludge disposal can be significantly reduced if lime recovery facilities are utilized. During the year 1417H, the average total amounts of sludge produced from the Riyadh Water Treatment plants were about 97080 m^3 /month, with maximum amount of 142700 m^3/month as presented on table 2. The produced sludge have a minimum monthly mass rate of 2852 tons of solids after drying which corresponds to a maximum monthly mass rate of about 4350 tons. Calcium carbonate ($CaCO_3$) and magnesium hydroxide ($Mg(OH)_2$) contents of the dry solids are 2166 and 384 tons, respectively as illustrated on table 3.

Table 2 Amount of sludge and net water production per month on the year 1417H

Plant	Sludge Prod.(m^3)		Net Prod.(m^3)		Brine
	Avg.	Max.	Avg.	Max.	(m^3)
Manfouha	30520	42740	1351490	1924010	170090
Buwayb	19560	35240	849410	1523150	102640
Sulbukh	28860	35110	1245990	1522380	166730
Malaz	4640	6830	206820	303330	26250
Shemessy	13500	22780	622760	1055170	38690
Total	171390	226050	10278250	12815870	504410

Table 3 Content of sludge per month (ton)

ITEM	Malaz	Sulbukh	Buwayb	Manfouha	Shemessy	Total
Density	1	1	0.9	1	1	
Amount of Solid after drying(min)	123	577	781	926	445	2904
Calcium carbonate (as $CaCO_3$)	97	308	544	834	383	2166
Magnesium carbonate (as $CaCO_3$)	11	214	103	24	32	384
Magnesium sulfate ($MgSO_4$)	0.3	17	23	9	0.3	50
Calcium sulfate ($CaSO_4$)	2.4	12	17	15	8	54.4
Calcium chloride ($CaCl_2$)	5.7	3	15	0.46	4.4	29
Magnesium chloride ($MgCl_2$)	0.6	3	7.8	0.18	3	14.6
Iron mg/l. as ion	0.2	0.6	2.5	4.5	1.6	9.4
Silica mg/l. as ion	2.4	7	19	11.8	6.3	46.5
Alu. mg/l. as ion	0.1	0.8	2	0.5	1	4.4

Lime-softening sludge can be dewatered by various methods such as Sludge thickeners, drying beds, lagoons, vacuum filters, centrifuges and dual cell gravity concentrator. The dewatered softening sludge contains on average 75-85% calcium carbonate ($CaCO_3$) and about 12% magnesium hydroxide ($Mg(OH)_2$). Table 4 shows the analysis of sludge after drying in various Water Treatment Plants in Riyadh. The insoluble magnesium hydroxide can be converted into soluble magnesium bicarbonate by carbonation of the sludge with carbon dioxide, CO_2. The remaining solids will be rich in calcium carbonate which can be easily separated. After recalcining of the carbonated sludge calcium oxide is produced which can be hydrated readily into lime.

Table 4 Analysis of sludge after drying

Content of sludge	Malaz	Sulbukh	Buwayb	Manfouha	Shemessy
Total amount of solid (gm/lit)(min)	26	20	40	30	32
Calcium carbonate (as $CaCO_3$)%	78.68	53.4	69.7	90.12	86.02
Magnesium carbonate (as $CaCO_3$)%	8.80	37.05	13.16	2.67	7.28
Magnesium sulfate ($MgSO_4$)%	0.30	2.90	3.00	1.00	0.08
Calcium sulfate ($CaSO_4$)%	2.01	2.10	2.14	1.66	1.00
Calcium chloride ($CaCl_2$)%	4.70	0.56	1.96	0.05	1.00
Magnesium chloride ($MgCl_2$)%	0.47	0.52	1.00	0.02	0.68
Iron mg/l. as ion %	0.16	0.11	0.33	0.49	0.37
Silica mg/l. as ion %	2.00	1.20	2.44	1.28	1.43
Alu. mg/l. as ion %	0.095	0.15	0.26	0.06	0.23

SOFTENING CHEMISTRY

The lime softening process was patented in 1841. The first municipal soften plant which used lime only was install in 1854 in England. In 1903, the first lime soda ash water softening plant was established. Lime - soda softening reactions can be summarized by the following categories:

a) removal of carbon dioxide: This reaction is completed at pH of 8.3.

$$CO_2 + Ca(OH)_2 \rightarrow CaCO_3\downarrow + H_2O$$

b) removal of carbonate hardness: Carbonate hardness caused by calcium and magnesium bicarbonates are precipitated as calcium carbonate and magnesium hydroxide by the addition of lime, calcium hydroxide $Ca(OH)_2$. Magnesium bicarbonate is first converted to magnesium carbonate at a pH of 9.4. Since magnesium carbonate is soluble, excess lime is added to precipitate it as magnesium hydroxide. The pH is raised to 10.6.

$$Ca(HCO_3)_2 + Ca(OH)_2 \rightarrow 2CaCO_3\downarrow + 2H_2O$$

$$Mg(HCO_3)_2 + Ca(OH)_2 \rightarrow CaCO_3\downarrow + MgCO_3 + 2H_2O$$

$$MgCO_3 + Ca(OH)_2 \rightarrow CaCO_3\downarrow + Mg(OH)_2\downarrow$$

c) removal of calcium noncarbonate hardness: Noncarbonate hardness caused by calcium sulfale and calcium chloride are precipitated as calcium carbonate by the addition of soda ash, sodium carbonate (Na_2CO_3).

$$CaSO_4 + Na_2CO_3 \rightarrow CaCO_3\downarrow + Na_2SO_4$$

$$CaCl_2 + Na_2CO_3 \rightarrow CaCO_3\downarrow + 2NaCl$$

d) removal of magnesium noncarbonate hardness: Noncarbonate hardness caused by magnesium sulfale and magnesium chloride are precipitated as magnesium hydroxide by the addition of lime, calcium hydroxide $Ca(OH)_2$. These reactions produce calcium noncarbonate hardness, $CaCl_2$ and $CaSO_4$. So soda ash, sodium carbonate (Na_2CO_3), must be added to precipitate these calcium noncarbonate hardness, $CaCl_2$ and $CaSO_4$, to calcium carbonate.

$$MgCl_2 + Ca(OH)_2 \rightarrow Mg(OH)_2\downarrow + CaCl_2$$

$$CaCl_2 + Na_2CO_3 \rightarrow CaCO_3\downarrow + 2NaCl$$

$$MgSO_4 + Ca(OH)_2 \rightarrow Mg(OH)_2\downarrow + CaSO_4$$

$$CaSO_4 + Na_2CO_3 \rightarrow CaCO_3\downarrow + Na_2SO_4$$

In Riyadh Water Treatment Plants, lime and soda ash softening process is utilized. Lime and soda ash are added into the precipitatos. Sodium aluminate($NaAlO_2$) and ferric chloride ($FeCl_3$) are dosed into the presoftened water as coagulants. The main objective of sodium aluminate and ferric chloride addition is to remove silica and colloids to prevent their precipitation on membrane surfaces. In addition, an anionic polyarcrylamid is dosed as a flocculation aid. Each precipitator consists of five zones (primary mixing and reaction, secondary reaction, sludge blanket, clarified water zone, sludge concentrators).

Raw water and coagulant enter each precipitator in the primary mixing and reaction zone. Lime and soda ash are added into the primary mixing zone for softening. Mixing and reaction in this zone is promoted by the lower blades of the central impeller, which also induces recirculation. Accumulated sludge below the primary mixing zone is drained to the sludge thickener on a daily basis. The clearance of the sludge is regulated by pneumatically operated valves in the sludge pipes. Each valve, in turn, opens automatically for a variable period and intervals. Sludge can be drained from the bottom of the primary reaction zone by manual opening a pneumatically operated sluice valve which remains open until accumulated sludge is cleared.

The sludge is allowed to settle in the central conical tank. Water from this tank is returned to the distribution chamber for retreatment. Thickened sludge from the conical tank is pumped out to the sludge lagoons. Thickened sludge is discharged to one of three sludge lagoons. This can be achieved either by gravity flow or by pumping. The sludge from the bottom of the thickener is discharged via a pneumatically operated valve.

The pH of the softened water is about 11. Carbon dioxide and/or sulfuric acid are then used to lower the pH to about 8.6 in order to prevent any deposition of the remained hardness. Calcium and magnesium are converted back into the soluble bicarbonate forms. This practice is know as recarbonation.

SLUDGE DEWATERING OPTIONS

Large amount of sludge produced daily from the softening process. Softening sludge volume generally ranges from 0.3 to 5% of the volume of treated raw water. About 2.5 lb (1.1 kg) of sludge will be formed for each pound of lime used. The solid content of settled sludge varies from 2 to 30%. The typical composition of softening sludge is shown in table 5 for raw water contain 14 ppm CO_2, carbonate hardness(CH) of 265 ppm, noncarbonate hardness(NCH) of 20 ppm and lime impurities of 8% of insoluble materials[3]. The exact composition depends on raw water hardness and degree of treatment.

Table 5 Estimated Typical Softening Sludge Composition [3]

Components	ppm	wt.%.
$CaCO_3$		
CO_2base	32	4.6
Carbonate hard.	530	76.9
non-Carbon. hard	20	2.9
total $CaCO_3$	582	84.4
$Mg(OH)_2$	81	11.7
SiO_2	4	0.6
Insolubles		3.3
Total		100.0

Proper sludge disposal is required for safe environment. So sludge must be dewatered before discarding to reduce its water content. The selection of sludge disposal process depends on the type of sludge to be disposed and the available disposal sites. The relative ease of dewatering softening sludge depends on the ratio of calcium to magnesium and the amount of gelatinous solids present. The higher the ratio of calcium to magnesium, the easier the dewatering [4]. Sludge thickeners, drying beds, lagoons, vacuum filters, centrifuges and dual cell gravity concentrator are considered possible alternatives in water treatment plants for sludge dewatering processes. Thickening is a preliminary step to final dewatering by other methods. Sludge thickeners are circular basins where sludge solids are allowed to settle by gravity.

In drying beds, a layer of sand is placed over graded aggregate and an underdrain system. Sands of 100 to 200 mm deep are usually placed on top of 200 to 460 mm graded aggregate (gravel or stone). Sludge tends to dewater slowly into the underdrain system. Large areas are required. Underdrained water can bc recycled to the plant, discharged to a waterway or used for land irrigation.

Like sand-bed drying, lagoons can be economical if land is readily available and relatively inexpensive. Dewatering sludge by lagoons has been widely practiced. Lagoons are usually about 3 in deep with 4000 m^2 or more size.

Vacuum filtration can be used for dewatering lime sludge that are low in magnesium content. Sludge dewatering by vacuum filtration yield a crumbly cake with solid content from 45 to 70%. The continuous flow vacuum filter consists of a horizontal cylindrical drum that is 20-35% submerged into the sludge. The cylindrical drum is covered with a porous fabric woven from fine metal or natural or synthetic fibers that act as a filter medium. As the cylindrical drum rotates the medium is rolled on and off the drum. Sludge is pulled onto the filter medium by a vacuum with the filter drum. The cake is cleaned from the medium by roll discharge or by a scraping blade.

Centrifuges are sedimentation tanks that use centrifugal force to speed up the separation of sludge solids from water at a force of 3000 to 10,000 times the gravity force. They require small space and can be fully automated. They have the ability to handle dilute or thickened sludge. Softening sludge enters through a feed pipe into the rotating tank (bowl). The typical rotation of this bowl is between 200 and 8000 rpm. A screw-type conveyor scrapes solids that settle at the wall of the bowl and moves them to the solids discharge end. Sludge dewatering by centrifuges yield sludge cake with solid content from 50 to 60%.

Dual cell gravity (DCG) concentrator is a continuous mechanical gravity sludge dewatering device that has found wide application in dewatering of metal hydroxide as well as sanitary sewage sludge[5]. Sludge is introduced on the dewatering cell where initial liquid-solids separation takes place. Detailed description of Dual cell gravity concentrator is given by Schwoyer and Luttinger[5].

General considerations for selection of sludge dewatering processes are as follows[6]:

1. Sludge characteristics.
2. Volume reduction attainable.
3. Energy inputs and other operation and maintenance costs. Skills required to operate the process must also be considered.
4. Chemical input.
5. Other operations in the sludge processing train.
6. Ultimate disposal.
7. Concentration of solids in the reject liquid.
8. Space requirements. Mechanical processes require much smaller areas than land dewatering.

Recalcining offers the potential for complete disposal of sludge. It is considered as a possible method for lime recovery from softening sludge. Recalcining is carried out at several locations in the United States as show in table 6[3]. Thre are four principal types of furnaces, calciners or klins employed for recalcining lime sludge. These types are rotary klin(RK), flash calciner(FC), fluidized bed calciner(FB) and multiple hearth calciner(MH)[3].

Table 6 US Major Sludge Recalcinning Plants

City	Water Source	Capacity mgd	Prod. tons/d	Type	Const. Date
Marshalltown, Iowa	wells	3.5	10	FC	1943
Miami, Florida	wells	180	80	RK	1948
Lansing, Michigan	wells	20	30	FB	1954
Saliva, Kansas	wells	-	24	FC	1956
Geensville, Florida	wells	-	6	FB	1957
Dayton, Ohio	wells	96	150	RK	1957
San Diego, California	river	-	25	RK	1961
Ann Arbor, Michigan	-	-	24	FB	1968
Lake Taboe, Nevada	w.water	7.5	10.8	MIH	1968
S. Paul, Minn.	lakes	120	50	FB	1969

RECALCINING PROCESS DESCRIPTION

The main purpose of calcining is to convert calcium carbonate ($CaCO_3$) produced from treatment plant process into quicklime (CaO) to be hydrated into lime for reuse. Before recalcining, sludge had to be dewatered and carbonated. Dewatering of sludge accomplished by one of the prementioned processes; sludge thickeners, drying beds, lagoons, vacuum filters, centrifuges and dual cell gravity concentrator. Carbonation is an effective method of selectively dissolving magnesium from the sludge. The highly insoluble megnesium hydroxide is converted into highly soluble magensium bicarbonate by inroducing gaseous carbon dioxide into the sludge. Carbonation reactions are as follows:

$$Mg(OH)_2\downarrow + CO_2 \rightarrow Mg(HCO_3)_2 + MgCO_3$$

$$MgCO_3 + CO_2 + H_2O \rightarrow Mg(HCO_3)_2$$

It is important to avoid the formation of saturated solution of the bicarbonate to prevent the formation of the solid magnesium carbonate trihydrate. This is formed according to the following reaction:

$$Mg(OH)_2\downarrow + Mg(HCO_3)_2 + H_2O \rightarrow 2MgCO_3.3H_2O\downarrow$$

Carbonation of calcium carbonate according to the following reaction:

$$CaCO_3\downarrow + CO_2 + H_2O \rightarrow Ca(HCO_3)_2$$

does not occur to any significant degree. The saturated solution of calcium bicarbonate has a lower pH than the saturated solution of magnesium bicarbonate[3]. Carbon dioxide for carbonation can be availabe from the lime klin in far grater quantities than the amount required to dissolve all of the magnesium hydroxide present in the sludge[7].

In the recalcining process, sludge is dried then the temperature is raised up to a temperature of 870-1200 °C where calcination reaction starts. This high temperature has to be maintained in order to keep the following endothermic reactions going:

$$CaCO_3 \rightarrow CaO + CO_2$$

$$Mg(OH)_2 \rightarrow MgO + H_2O$$

$$MgCO_3 \rightarrow MgO + CO_2$$

$$Mg(HCO_3)_2 \rightarrow MgO + 2CO_2 + H_2O$$

In these reactions quicklime (CaO) and magnesium oxide are produced. Magnesium oxide is precipitated into the addition of water:

$$MgO + H_2O \rightarrow Mg(OH)_2\downarrow$$

Calcining process may take place in a lime rotary klin(RK) which is a cylindrical rotating oven with a length between 40 and 120 m having a diameter between 2 and 4 m. Wet lime mud is fed into the high end of the kiln by conveyor belt with solid content of 60-70%. The solid phase moves downhill inside the kiln and hot gases flow in the opposite direction.

Description of the lime recovery process at Dayton, Ohio can be briefly given as follows[5]. Softening sludge is pumped to a sludge recarbonation basin where it is intimately mixed by diffusion with scrubbed kiln gas containing about 20% carbon dioxide, CO_2. Most, but not all, of $Mg(OH)_2$ is selectively dissolved from the calcium carbonate during the 50 minutes retention period provided. The recarbonated slurry passes to a 130 ft thickener from which the clear supernatant containing the magnesium, in the form of soluble magnesium carbonate, overflows and passes to the sewer. The thickened sludge mainly $CaCO_3$ with some undissolved $Mg(OH)_2$ and some insoluble, passes to a sludge storage tank. From there it passes to three Bird centrifuges that concentrate the thickened slurry to about 60% solids. This thick paste is then conveyed to the kiln where it is calcined into high quality quicklime. The CaO content is usually ranges between 92-93%. The magnesium concentration of the recalcined lime is about 3 to 4%. Silica (SiO_2) content is about 1.2% while iron and aluminium are about 2.3%.

CONCLUSIONS

In the softening process, large amounts of sludge were produced which had to treated for proper disposal. The problem of sludge disposal can be significantly reduced if lime recovery facilities are utilized in the water treatment plants. Lime recovery from softening sludge is technically feasible process. Calcium carbonate in the sludge is about 85%. Quiklime with about 90% CaO content can be produced in lime recovery process. For every ton of lime fed to the softening process about 1.3 tons can be recovered, the excess is due to calcium content present in the raw water.

In lime recovery process, lime-softening sludge is first dewatered. Various methods such as sludge thickeners, drying beds, lagoons, vacuum filters, centrifuges and dual cell gravity concentrator can be used for sludge dewatering. The dewatered softening sludge contains on average 75-85% calcium carbonate and about 12% magnesium hydroxide. The insoluble magnesium hydroxide can be converted into soluble magnesium bicarbonate by carbonation of the sludge with carbon dioxide. The remaining solids will be rich in calcium carbonate which can be easily separated. After recalcining of the carbonated sludge calcium oxide is produced which can be hydrated readily into lime.

Several benefits can be obtained from lime recovery. The cost of softening will be reduced substantially. It eliminate the problem of sludge disposal. It also provides amble source of carbon dioxide to be used for recarbonation.

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Agricultural Water

Wastewater Quality and its Reuse in Agriculture in Saudi Arabia

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WASTEWATER QUALITY AND ITS REUSE IN AGRICULTURE IN SAUDI ARABIA

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ABSTRACT

The water supply in Saudi Arabia, to meet the growing agricultural requirements, increased from 1.75 billion m^3 in 1975 to 22.93 billion m^3 in 1992. The total amount of wastewater available according to the references given is around 1.32 million m^3 per day. The total water salinity ranges between 1000-7273 mg L^{-1} in different areas. The quality of the wastewater ranges between low salinity-low sodium to very high salinity-very high sodium hazards which could create substantial soil and crop management problems when used for irrigation. Application of certain mathematical models have shown that the use of these wastewaters can affect the exchangeable-sodium-percentage (ESP) and soil salinity to a create substantial soil and crop management problems. However, wastewater quality demands careful assessment if it was to be reused as a supplemental source of crop irrigation for agricultural expansion. Experiments have shown that use of wastewaters in Saudi Arabia as a supplemental irrigation has not only increased crop production, water use and nitrogen use efficiencies but also served as a source of plant nutrients. It was also found that the use of wastewater can save up to 50 % application of inorganic nitrogen fertilizer if the wastewater contains 40 mg N L^{-1}. The short and long term uses of different types of wastewaters for irrigation did not show any significant increase in the bioaccumulation of heavy metals in crops and soils. However, there exists a lot of potential for reuse of wastewater in agriculture after appropriate treatment and by following certain management practices such as leaching requirements, proper crop selection, estimation of crop water requirements, adoption of improved irrigation methods and application of right amount of fertilizer.

Keywords: wastewater, water quality, reuse, agriculture, Saudi Arabia

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INTRODUCTION:

Saudi Arabia is an arid country with a total land area of 2.253 x 10^6 km^2. The total cropped area in the Kingdom has increased from 1.25 million hectares in 1988 to 1.59 million hectares by 1994 (MAW, 1994). As a result, the demand for irrigation water increased from 1.75 billon m^3 in 1975 to 22.93 billion m^3 in 1992 (Dabbagh and Abderrahman, 1996). Currently, more than 80 % of water demand in agriculture sector is met from non-renewable groundwater sources (MAW, 1994). In order to meet the growing demand for water in domestic, industrial and agriculture sectors, alternative source of water, especially for agriculture, has to be developed. As a result of urban and rural expansion in the last two decades, a large quantity of wastewater has become available in different regions of Saudi Arabia. For example, from Riyadh about 100 million m^3 is discharged in Wadi Hanifah annually (Al-Degaither, 1992). Another 11.64 million m^3 per annum of drainage water in Al-Ahsa Oasis (HIDA 1984) can be used in agriculture.

Generally, the quality of Riyadh wastewater and Al-Ahsa drainage water is marginal for irrigation. The total water salinity ranges between 1000 to 7273 mg L^{-1} in different areas (Hussain and Sadiq 1991). The quality of water for irrigation is determined by its chemical composition and the conditions of use. Presently, there is a lot of awareness regarding water pollution and environmental health hazards that may result from the use of wastewater. Thus there is a growing concern regarding recycling and reuse of wastewater. Presence of certain toxic elements such as Pb, Ni, Cd, Co, Cu, Mo, Hg etc in the wastewaters may create some environmental problems and needs evaluation prior to its use in agriculture. The main objective of this study is to review wastewater availability, its quality and potential use in agriculture in Saudi Arabia.

WASTEWATER AVAILABILITY

Main sources of wastewaters include agricultural drainage water, run-off losses from irrigation systems and fields, sewage water from cities, leakage from water supply systems, high percolation losses from the irrigated fields, wastewater from water treatment plants and drainage water from aquaculture facilities in the Kingdom of Saudi Arabia (Table 1).

Table 1. Quantity of Wastewater Available in the Kingdom

Sr.No.	City	Quantity (m^3 day^{-1})	Source
1.	Al-Riyadh	420,000	(Al-Degaither (1992).
2.	Jeddah	118,000	
3.	Buraidha	10,000	
4.	Unaizah	7,000	
5.	Al-Ahsa	293,782	WRD, HIDA (1984).
6.	Dammam	187,143	Alkhuzayium et al.(1994).
7.	Al-Khobar	96,000	Alkhuzayium et al.(1994).
8.	Al-Qatif	29,571	Alkhuzayium et al.(1994).
9.	Aqua.Effluent	20,000	Al-Jaloud et al.(1993).
10.	Saihat	15,717	Bader (1994).
11.	Safwa	8,021	Bader (1994).
12	Jubail	12,639	Bader (1994).
13	Hofuf Areas	1,04,907	Bader (1994).
	Grand Total =	**1,322,780**	

INDICATORS OF WATER QUALITY FOR IRRIGATION

The criteria used to evaluate quality of wastewater for use in agriculture are 1) **Salinity** of irrigation water for salt build up in soils and its adverse effects on plant growth (Gallatin et al., 1963, 2) **Sodium Adsorption Ratio (SAR)** for its deleterious effect on soil physical properties (Bower, 1972), 3) **Residual Sodium Carbonate (RSC)** for its effects on final soil water SAR value with the loss or gain in Ca and Mg concentration due to the precipitation or dissolution of alkaline earth carbonates (Bower et al., 1965) and the **Toxic effects of Specific Ions** in irrigation water such as Na, Cl, SO_4 and B on plant growth and yield (Eaton, 1942).

OTHER WATER QUALITY INDICATORS

In addition to the above indicators, some other mathematical equations and models were applied to evaluate the data on water quality for its possible use in agriculture. Soil salinity development (SSD), adjusted sodium adsorption ratio (adj.SAR), Adjusted Na Ratio (adj. R_{Na}), and Exchangeable Sodium Percentage (ESP) were calculated from the analytical data.

(a) The SSD was calculated according to FAO (1985). The equation used was:

$$SSD = EC_{dw} = EC_{iw}/LF \qquad (1)$$

where

EC_{dw} Salinity of the drainage water percolating below the root zone.
EC_{iw} Salinity of the wastewater to be used for irrigation.
LF Leaching fraction is the amount of irrigation water that leaves the root zone as drainage water.

(b) The adj. SAR was calculated as (Ayers and Westcot, 1985):

$$adj.\ SAR = SAR\ [1 + (8.4 - pH_c) \qquad (2)$$

$$pH_c = [pK' + pK_{calcite}] + p[Ca + Mg] + pHCO_3$$

where p refers to the negative logarithm, K is the second dissociation equilibrium constant of the carbonic acid, K_c is the solubility equilibrium constant for calcite. Concentration of Ca, Mg, and HCO_3 in meq l^{-1}.

(c) The adj.R_{Na} was determined according to Suarez (1981) using the following equation:

$$adj.R_{Na} = Na/\ [Ca_x + Mg/2]^{0.5} \ (3)$$

where all concentrations in meq L^{-1}, Ca_x represents concentration after counting for HCO_3 of wastewaters.

(d) The ESP was predicted as:

$$ESP = \frac{100\ (-0.0126 = 0.01475 \text{ x } SAR)}{1\ (-0.0126 + 0.01475 \text{ x } SAR)} \qquad (4)$$

where SAR is the SAR of the soil solution resulting from irrigation with wastewaters.

The salinity and sodicity hazards of the wastewaters were determined according to the classification given by USDA Handbook No. 60, 1954.

The wastewaters were also categorized for crop production according to the standards of FAO (1985) and for health protection according to WHO (1989). This provided a highly reliable tool for safe evaluation of quality of the existing wastewaters in the Kingdom of Saudi Arabia.

CHEMISTRY OF AVAILABLE WASTEWATERS

The chemical composition of wastewaters is determined for evaluation and classification with special reference to its reuse in agriculture. The chemistry of some waters from different sources is presented in Tables 2-4.

DRAINAGE WATERS

The chemical composition of drainage waters from Al-Ahsa Oasis is shown in Table 2. The salinity ranges between 3500 to over 7200 mg L^{-1} in different wastewaters. Sodium (Na) and Cl were the most abundant cation and anion in the drainage waters, respectively.. The drainage waters contain more salts than other wastewaters. This was because salts from the soil profile were being washed into the drainage canals during reclamation and cropping seasons.. The salinity and sodicity hazards of the drainage waters were classified as C_4S_2 (very high salinity with medium sodicity problems) according to USDA (1954).

C. AQUACULTURE EFFLUENTS

The ranges for different water quality parameters are: pH 6.6-7.2, EC between 1.1-12.1 dS m^{-1}, TDS 550 - 6050 mg L^1 and SAR from 2.9-10.3 (Table 3). Sodium and Cl were the dominant cation and anion, respectively. These were followed by Ca, Mg, K and HCO_3 in descending order according to the concentration levels. Additionally, the aquaculture effluent also contained Fe in the range of 0.1-3.8 mg L^{-1}. The main source of Fe in these effluent waters was the fish feeds.

ANTICIPATED EFFECT ON SOILS

IMPACT ON ESP OF SOILS

The ESP of soils, which would be irrigated with wastewater, was predicted for its hazards on physical properties of soils. The SAR, adj. R_{Na} and adj. SAR were all significantly correlated to ESP (r= 0.999, 0.981 and 0.939, respectively). The ESP values predicted from adj. SAR were higher than those calculated from adj. R_{Na} and SAR. The ESP values greater than 15 indicate Na hazards (USDA Handbook No. 60, 1954). It was, therefore, anticipated that there would not be a serious Na problem from the use of these waters for crop irrigation. Similarly, ESP of soils could be predicted for various quality irrigation waters intended for use in agriculture. This would help proper management of marginal quality waters especially containing high level of sodium. Because irrigation with high Na waters will disperse the soils by replacing Ca and partly Mg from soil exchange complex.

DEVELOPMENT OF SOIL SALINITY

Irrigation water is the main source of addition of salts to soils and its magnitude depends on the total water salinity, water composition, leaching requirements considered, soil type, drainage conditions, climate, management practices, and the consumptive use of crops to be cultivated. Salt build up showed an increase with a decrease in leaching fractions. For example: If the leaching requirement (LR) is 10 %, the salt build up will be 10 times to that of a given irrigation water salinity and 5 times against 20 % leaching requirements. The data further infers that salinity development will be faster in soils irrigated with high total salinity irrigation waters than less saline waters (Table 4). It also implies that only medium to high salt tolerant crops would be possible with proper agricultural practices. In addition to the above, highly saline waters could successfully be used for the reclamation of highly salt affected soils at its initial stage of development. Because the final soil salinity at equilibrium will not exceed that of irrigation water salinity. A similar exercise was done taking into consideration the total salinity of Al-Ahsa drainage waters (Table 4)..

Table 2. Chemical Composition of Drainage Waters

Sample	pH	EC $dS\ m^{-1}$	TDS $mg\ L^{-1}$	Ca^{+2}	Mg^{+2}	Na^{+}	K^{+}	CO_3^{-2}
				$meq\ L^{-1}$				
Al-Mazroohiya	7.43	11.30	7273	18.2	24.2	76.1	3.2	0.6
D-1 (Al-Auba0	7.55	9.01	5766	21.0	18.6	58.7	2.5	1.8
D-2 (Outfall)	7.64	8.87	5677	23.8	20.4	52.2	2.2	0.6
P-21/D-1.6	7.42	5.83	3731	18.8	11.8	32.9	1.8	2.5
HARC	7.35	7.37	4717	24.2	11.4	50.1	2.2	2.0
Al-Uyan	8.08	6.57	4205	18.9	12.3	36.9	2.0	2.0
KFU	8.00	7.38	4723	29.2	11.8	45.6	1.0	1.4
Al-Qarah	7.58	8.01	5126	22.0	17.2	43.5	1.7	3.0
D-1.5	7.72	6.20	3968	18.2	11.8	39.1	0.9	2.8
Al-Mansoora	7.57	8.04	5146	23.1	19.1	45.6	0.5	2.4
HARC	7.18	7.23	4627	29.6	12.2	34.8	0.5	1.8
F-1.11/D-1	7.56	5.50	3540	16.6	11.2	26.5	0.8	3.0
D-1	7..97	5.52	3533	16.2	10.8	26.8	1.0	2.6

	HCO_3^-	Cl^-	SO_4^{-2}	SAR	Adj. SAR	Adj. R_{Na}	Class
	meq L^{-1}						
Al-Mazroohiya	6.1	88.3	26.7	16.5	44.8	19.7	C_4S_2
D-1 (Al-Auba)	6.1	65.4	27.5	27.5	36.7	16.7	C_4S_2
D-2 (Outfall)	7.1	59.3	31.6	11.1	30.8	14.3	C_4S_2
P-21/D-1.6	5.8	40.2	16.6	8.4	23.6	11.0	C_4S_2
HARC	5.8	48.2	31.9	11.9	32.9	16.6	C_4S_2
Al-Uyan	4.4	43.9	19.8	9.3	25.2	11.7	C_4S_2
KFU	1.4	48.0	36.7	10.1	23.5	11.5	C_4S_2
Al-Qarah	3.9	53.4	24.1	9.8	26.7	12.1	C_4S_2
D-1.5	4.3	40.2	22.7	10.1	27.8	12.4	C_4S_2
Al-mansoora	5.9	55.2	24.8	9.9	27.8	12.7	C_4S_2
HARC	3.5	44.4	27.4	7.6	19.8	10.1	C_4S_2
F-1.11/D-1	4.7	34.9	12.5	7.3	20.3	8.9	C_4S_2
D-1 (ARAMCO)	3.7	38.8	9.7	7.3	20.0	8.9	C_4S_2

HARC = Fofuf Agriculture Research Station, D = Drainage Canal
KFU = King Faisal University, Al-Ahsa.
(Source: Hussain and Sadiq, 1991).

Table 3. Chemical Composition of Aquaculture Effluent

Source	Area	pH	EC dS m^{-1}	TDS	Ca^{+2}	Mg^{+2}	Na^+	K^+
					mg L^{-1}			
Al-Ballah	Al-Qassim	7.20	1.08	691	72	34	118	9.8
Al-Rashid	Al-Ahsa	6.96	4.41	2832	193	97	613	28.5
KACST	Dirab	7.87	2.96	1894	269	81	263	17.0
Al-Dhahir	Al-Riyadh	6.82	12.1	7744	1010	382	1288	32.5
Bin-Ammar	Al-Riyadh	7.22	8.80	5632	770	276	950	25.8
Karnak	Al-Qatif	6.70	6.34	4058	321	140	813	35.5
Al-Dakhil	Al-Qassim	7.33	2.22	1421	152	68	243	16.3
Al-Emmar	Al-Kharj	6.65	2.39	1530	173	95	200	12.7
KACST	Al-Qassim	7.15	3.79	2426	152	58	513	10.0
Essam	Al-Qatif	7.05	4.86	3110	297	116	525	30.0
Al-Kabir	Al-Qassim	6.75	3.68	2355	112	73	575	9.8
Al-Rajhi	Al-Kharj	7.00	1.83	1171	200	53	145	8.2

		HCO_3^-	Cl^-	SO_4^{-2}	N_{Total}	Fe^{+2}	SAR	Class
		mg L^{-1}						
Al-Ballah	Al-Qassim	210	189	145	15.6	0.43	2.9	C_3S_1
Al-Rashid	Al-Ahsa	210	2045	615	8.9	0.28	8.9	C_4S_3
KACST	Dirab	240	375	368	40.8	0.14	3.6	C_4S_2
Al-Dhahir	Al-Riyadh	377	8625	2913	20.5	3.40	8.7	C_4S_4
Bin-Emmar	AL-Riyadh	185	1190	595	73.0	0.13	7.5	C_4S_4
Karnak	Al-Qatif	210	633	1300	11.5	0.23	9.5	C_4S_3
Al-Dakhil	Al-Qassim	44	316	165	12.1	0.37	4.1	C_3S_2
Al-Emmar	Al-Kharj	93	1272	550	11.2	0.17	3.0	C_4S_1
KACST	Al-Qassim	244	866	598	12.3	0.32	8.9	C_4S_3
Essam	Al-Qatif	337	2804	795	11.2	0.08	6.5	C_4S_3
Al-Kabir	Al-Qassim	151	871	579	5.6	0.67	10.3	C_4S_3
Al-Rajhi	Al-Kharj	229	456	498	16.8	0.21	2.3	C_3S_1

(Source: Al-Jaloud et al. 1993), N = nitrogen

Table 4. Prediction of Soil Salinity (EC_e) from Irrigation with Drainage Waters.

		Leaching Fraction					
Water Source	Water EC	0.10	0.15	0.20	0.25	0.30	0.40
		dS m^{-1}					
Al-Mazroohiya	11.3	113	75.3	56.5	45.2	37.7	28.3
D-1 (Al-Auba)	9.01	90.1	60.1	45.1	36.0	30.0	22.5
D-2 (Outfall)	8.87	88.7	59.1	44.4	35.5	29.6	22.2
P-21/D-1.6	5.83	58.3	38.9	29.2	23.3	19.4	14.6
HARC	7.37	73.7	49.1	36.9	29.5	24.6	18.4
Al-Uyan	6.57	65.7	43.8	32.6	26.3	21.9	16.4
KFU	7.38	73.8	49.2	36.9	29.5	24.6	18.4
Al-Qarah	6.57	65.7	53.4	40.1	32.0	26.7	20.0
D-1.5	6..20	62.0	41.3	31.0	24.8	20.7	15.5
Al-Mansoora	8.04	80.4	53.6	40.2	32.2	26.8	20.1
HARC	7.23	72.3	48.2	36.2	28.9	24.1	18.1
F-1.11/D-1	5.50	55.0	36.7	27.5	22.0	18.3	13.8
D-1 (ARAMCO)	5.52	55.2	36.8	27.5	22.1	18.4	13.8

PROBLEMS OF EFFICIENT WATER USE IN IRRIGATION

WATER LOSSES

One of the major problems faced by water users is the loss of irrigation water through evaporation from irrigation systems (concrete canals in Al-Ahsa Oasis) carrying water and deep percolation due to over irrigation to field crops by flood system of irrigation. These water losses has considerable impact on the amount of agricultural drainage water. For example, the water losses from irrigation canals in Al-Ahsa Oasis, estimated using Class A-pan evaporation data, were 3.75-4.53 x 10^6 m^3 per year. If the total surface area for all the drainage canals is also considered then the total evaporation losses of water will be twice as much. Correspondingly, keeping in view the consumptive use of different crops for the same area, the total crop water use for the period from 1974-83 was calculated. The total crop water requirements varied between 26.65-137.82 x 10^6 m^3.

Recently, Al-Ajaji (1985) found in a field study that high percolation losses are the main cause for excessive drainage flow in Al-Ahsa drainage canals. He found that actual deep percolation (ADP) losses averaged upto 58 % of the actual irrigation water applied. This was also very well supported by the high negative correlation between actual deep percolation (ADP) and application efficiency of low quarter (AELQ r = 0.86). He further concluded that ADP is related mainly to basin length (L), average depth applied (ADA) and the ratio between soil moisture deficit (SMD) and management allowed deficit (MAD) by the following equation.

$$\mathbf{ADP = 69.20 - 0.77\ L = 7.15\ ADA - 78.13\ SMD/MAD}$$

Because this equation has high statistical significance and a high coefficient of determination value of 99.9 and 0.91, respectively.

SALINIZATION OF AGRICULTURAL LANDS

Salinization of irrigated agricultural lands is a serious problem in an arid environment. The salinization could be the end result of many factors such as salt in irrigation water, climate, geology and configuration of the terrain which determine soil properties, land drainage, soil, water and crop management practices adopted (Khatib 1977).The problem of resalinization is world-wide (United Nations 1977). The countries of the Middle and the Near East suffer from this problem to a greater extent.

The middle east countries do show signs of soil resalinization but no statistics are available. However, according to Vidal (1951), the arable land in Al-Ahsa Oasis was estimated to be about 16,000 hectares. But during the last 15 years, this area decreased to around 8,000 hectares because of resalinization due to under/over-irrigation and poor field drainage conditions.

INADEQUATE LAND DRAINAGE

Restricted drainage is a key factor for the use of marginal-quality irrigation waters for agricultural expansion. Because it will reduce soil permeability and create temporary water ponds (a perched water table condition) below the soil surface. This will not only restrict salt leaching but could be the main source of soil resalinization by capillary action.

UNSCIENTIFIC USE OF IRRIGATION WATER

Inadequate information and lack of proper education of the farming community on the use of water have resulted in over or under-irrigation of crops in some cases. This has caused soil degradation especially when the depth to hard layer or the claypan is shallow. Such typical situation of shallow hard layer coupled with soil salinization is clearly evidenced in the Al-Ahsa Oasis, Kingdom of Saudi Arabia (Vidal 1951).

EXPERIENCE ON WASTEWATER RE-USE

SALINE IRRIGATION

Increased use of marginal-quality waters, as a supplemental source of irrigation, plays a significant role in the area of agricultural expansion. Hussain et al. (1994) conducted a pot experiment to determine the effect of saline water on the establishment of windbreak trees and soil properties. Survival period of trees decreased significantly with increase in soil salinity resulting from the irrigation water salinity. The survival period of Prosopis juliflora was significantly more than Casuarina equisetifolia and Eucalyptus camaldulensis. The decrease in total biomass yield was significant with increase in soil salinity. Soil salinity and sodicity increased significantly with an increase in irrigation water salinity and sodicity. Prosopis juliflora tolerated soil salinity (EC_e) up to 38.3 dS m^{-1} with irrigation water salinity of 13.5 dS m^{-1}, Casuarina equisetifolia up to 27.6 dS m^{-1} with irrigation water salinity of 6.6 dS m^{-1} and Eucalyptus camaldulensis up to 15.2 dS m^{-1} with irrigation water salinity of 2.12 dS m^{-1} for proper establishment provided 15 % extra water is applied as leaching requirement to control soil salinity. The experiment proved the sequence in salt tolerance for different trees as prosopis > casuarina > eucalyptus. The results suggested that Prosopis juliflora should be cultivated as windbreak trees in landscape and sand stabilization projects.

Recently, Hussain et al. (1995) found a mean fresh weight yield of alfalfa as 56.72 g (Hassawi), 54.29 g (Supreme) and 59.91 g (CUF-101) per pot with water having total salinity of 7.8 dS m^{-1}. The study indicated that a reasonable alfalfa production can be obtained using saline irrigation in a pot experiment with proper management practices. Nabulsi et al. (1994) observed in a field trial on rice crop that irrigation by drainage water reduced straw yield from 666 to 460 g m^2, plant height from 0.88 to 0.73 m, tillers per plant from 8.2 to 5.3, kernel yield from 546 to 201 g m^{-2}, spikes from 336 to 251 m^{-2} and 100 kernel weight from 4.2 to 2.5 g. Generally, the application of 150 and 200 kg N ha^{-1} significantly improved crop performance and yield under salinity stress. More recently, Helalia et al. (1996) found that alfalfa produced good growth and dry matter yield under both canal (3.2-3.4 dS m^{-1}) and mixed canal and drainage (6.4-8.6 dS m^{-1}) irrigation waters. The irrigation with drainage water (10.2-15.6 dS m^{-1}) reduced the yield significantly especially at bi-weekly irrigation frequency and at low nitrogen and phosphorus levels.

AQUACULTURE EFFLUENT

A field experiment was conducted on sandy-clay-loam soil using aquaculture effluent as a supplemental source of nitrogen fertilizer to wheat crop (Al-Jaloud et al. 1993). They found that results obtained with 25 to 50 % nitrogen application under aquaculture effluent irrigation were comparable with those obtained with 75 and 100 % nitrogen application under well water irrigation. In conclusion, a 50 % saving in nitrogen application as an inorganic fertilizer can easily be achieved if crops are irrigated with aquaculture effluent containing around 40 mg N L^{-1}. Hussain and Al-Jaloud (1995) concluded that application of 150-225 kg N ha^{-1} for well water irrigation and 75 and 150 kg N ha^{-1} for aquaculture effluent containing 40 mg L^{-1} would be sufficient to obtain higher WUE of wheat in Saudi Arabia. Recently, Al-Jaloud et al. (1996) conducted a field experiment on rapeseed cultivars. In an other experiment, Hussain et al. (1996) stated that yield of wheat and its nitrogen use efficiency were greatly affected by nitrogen application and treated effluent irrigation. The wheat grain yield ranged between 5.20-6.87 Mg ha^{-1} for well water and 5.20-6.54 Mg ha^{-1} receiving aquaculture effluent. The nitrogen use efficiency (NUE) ranged between 10.49-50.23 kg grain kg^{-1} N (well water) and 20.65-91.56 kg grain kg^{-1} N (aquaculture effluent). Overall, it was found that a higher grain yield and NUE of wheat crop can be achieved with low application rate of nitrogen if the crop is irrigated with treated effluent containing nitrogen in the range of 20 mg L-1 and above.

PLANT GROWTH AND SOIL PROPERTIES

Effects of wastewaters on plant growth and soil properties were studied in a pot experiment (Al-Jaloud et al. 1993). Mean biomass ranged between 159 and 210 g per pot for maize and between 165 and 212 g per pot for sorghum in different water salinity treatments. Mean drymatter yield ranged between 28.9 and 38.3 g per pot for maize and between 34.9 and 50.4 g per pot for sorghum. The crop yield showed significant increase with an increase in water salinity probably due to the nutrients present in these wastewaters, especially the nitrogen. Plant yield decreased slightly at water salinity level of 2330 mg L^{-1} (TDS), indicating that higher water salinity can neutralize the beneficial effects of nutrients in wastewaters. Soil salinity and sodicity increased significantly with corresponding increase in water salinity and sodicity (r values of 0.98 for maize and 0.98 for sorghum with respect to soil salinity, and 0.96 for maize and for 0.95 sorghum with respect to SAR of soils). The interaction between crop and water treatments was significant for soil salinity ($LSD_{0.05}$ = 0.48) and SAR of soil ($LSD_{0.05}$ = 2.55). Overall, the soil salinity and sodicity were significantly more in sorghum than maize. The results showed that wastewaters can successfully be used to grow corn ad sorghum as forage crops, provided 15 to 20 % excess water is applied to meet leaching requirements to maintain soil salinity within acceptable limits for optimal agricultural production. In conclusion, wastewater irrigation did not increase the concentrations of plant nutrient elements and the heavy metals in corn and sorghum plants to hazardous limits according to the established standards and could safely be used for crop irrigation according to the findings of this limited experimental results (Al-Jaloud et al. 1995). However, intensive investigations over a long period of time are needed to establish the use of such wastewaters for crop irrigation without health hazards.

WATER REQUIREMENTS FOR LAND RECLAMATION

Estimation of water requirement for reclamation of salt affected soils is important for adequate planning, management and economical use of water resources.. A number of studies conducted on sandy salt affected soils showed that a 20-30 cm depth of water is required to reclaim the surface 0-30 cm depth of soil, a 60-90 cm depth of water is required to reclaim the subsurface 30-60 cm depth of soil, and a 50-60 cm depth of water is required to reclaim the whole profile 0-60 cm depth of soil for a given salt affected soil and method of water application (Hussain et al. 1989). The study also provided a useful tool in the form of a leaching curve to determine water requirement for the reclamation of some salt affected soils in Al-Ahsa, Kingdom of Saudi Arabia. Similarly, Abdelhadi and Hussain (1987) showed that soil salinity was reduced from 75.50-16.75 dS m^{-1}, SAR from 20.50-12.50 with 20 cm depth of water application. However, soil salinity and SAR (sodicity) of soil reached safe limits by applying 60 cm depth of water in a three month period. The gypsum contents decreased significantly with increase in irrigation water application. The study indicated that a highly salt-affected soil can be reclaimed with 60 cm depth of water for 0 - 30 cm surface soil and with 100 cm depth of water for 0 - 90 cm profile in a 3 to 5 month period.

CROP IRRIGATION

Some research in Saudi Arabia has shown a lot of potential for the reuse of saline water for crop irrigation without significant yield losses. Fallatah and Hussain (1988) and Al-Tahir et al. (1989) also observed that highly saline drainage waters could be used efficiently by mixing it with normal irrigation water (Tables 5 and 6). Similarly, Al-Mashhady et al (1983) investigated on the use of highly saline water for wheat and barley irrigation. They found that yields of both wheat and barley crops decreased significantly with increase in irrigation water salinity (Table 7).

Table 5. Effect of Water Salinity on Greenmatter and Drymatter Yield of Faba Bean Cultivars.

Water Salinity (TDS)	Greenmatter (g/pot) F-32	F-21	FSLH	Drymatter (g/pot) F-32	F-21	FSLH
1200	81.5	78.5	84.0	13.2	12.0	13.6
2225	69.5	64.7	71.9	11.3	10.5	11.7
3350	59.5	57.5	57.2	9.6	9.3	9..3
4180	54.2	53.8	51.2	8.8	8.4	8.3
5250	43.9	44.9	37.1	7.1	7.6	6.0
$LSD_{0.05}$	4.96	8.16	9.40	0.8	0 .22	1.52

Table 6. Effect of Drainage and Normal Irrigation Waters on Dryweight (mg) of Different Parts of Baba Bean

Plant	**Normal**		**1:1 Mixture**		**Drainage**	
Parts	Hassa-1	F-32	Hassa-1	F-32	Hassa-1	F-32
Root	137.6	132.4	123.7	120.4	95.6	78.1
Stem	98.3	94.7	84.8	81.3	65.2	56.7
Leaves	201.8	207.8	183.7	154.6	110.7	86.9

Table 7. Effect of Highly Saline Water Irrigation on Wheat and Barley (g m^2)

Cultivar	1300	12000	16000	20000	24000
		mgL^{-1}			
Wheat					
Chenab 7	44	275	133	153	99
Madani	573	277	170	109	66
Barley					
Beecher	864	281	210	197	113
BCO Mr. Godiva	363	72	87	65	6

CONCLUSIONS

The comprehensive review has indicated a lot of potential for recycle and reuse of wastewaters in agriculture with an appropriate water treatment provided certain soil, water and crop management practiced such as crop selection, estimation of crop water requirements, proper fertigation, leaching requirements, adoption of improved irrigation methods, time and amount of fertilizer application are followed. Wastewater utilization as a supplemental source of irrigation to meet ever growing needs of water for agriculture expansion is gaining momentum in many arid and semi-arid countries of the world. Since Saudi Arabia falls under the category of arid climate, the use of wastewater for irrigation has a special significance due to the country's non-renewable water resources as well as due to the presence of appreciable amount of crop nutrients.

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