STUDYING THE EFFECT OF OPERATING CONDITIONS ON WATER TREATMENT USING RO

A Thesis
Submitted to the College of Engineering
Of Nahrain University in Partial Fulfillment
Of the Requirements for the Degree of
Master of Science
in
Chemical Engineering

by

Marwa Asad Saleh

B.Sc.in Chemical Engineer 2006

Thu Al-Hija 1430

December 2009

Abstract

The present study is to analyze and model concentration polarization in spiral wound brackish water membrane elements. The aim of the present work is to study the effect of different operating parameters on the reverse osmosis system which are feed flow rate, feed concentration, and feed temperature on permeate concentration and mass transfer coefficient for the brackish water system. The range of these parameters were; feed concentration (100,200,300,400and500ppm), feed flow rate (0.666, 1, 1.333 and 1.666-2 cm^3/s), and feed temperature (15, 25 and 35 °C), salts (NaCl, $CaCl_2$ and KCl).

Membrane efficiency is effected by the phenomena of concentration polarization (accumulation of solute) and fouling at the membrane surface. Results show that the polymer membrane is very sensitive to changes in the feed temperature. There was up to a 6 % average difference in the permeate concentration between feed temperatures of 15 and 35 ^{o}C. Doubling of the feed flow rate decreased the permeate concentration by to 13-36 %, but only at a high solute concentration. Membrane parameters were estimated using an analytical osmotic pressure model for high salinity applications. The modeling studies showed that the membrane transport parameters were influenced by the feed salt concentration and temperature.

List of Contents

Appendices.

Nomenclature

Symbol	Meaning	Unit
A	Area	cm^2
A	The permeability constant for membrane	$g.mole\ H_2O^{-1}$ $cm^{-2}.sec^{-1}.atom^{-1}$
C	The solute concentration in the boundary layer	g/cm^3
c	The molarity density for solution	$g.mole/cm^3$
C_b	Feed Concentration	g/cm^3
C_m	Membrane Surface Concentration	g/cm^3
C_p	Permeate Concentration	g/cm^3
D	The solute diffusion coefficient	cm^2/s
dx	Membrane thickness	cm
k	Mass transfer coefficient	$cm\ s^{-1}$
NA	the average of solute flux	$gm.mole/cm^2.$ sec
NB	the average of water flux	$gm.mole/cm^2.$ sec
J_v	Volumetric flux	$cm^3\ cm^{-2}\ s^{-1}$
XA2	The mole fraction for solute at the surface between the solution and membrane	
XA3	The mole fraction for solute in the product water	

	Greek Letters	Unit
A	solvent permeability through the membrane	s^{-1}
σ	The reflection coefficient	-
δ	Film thickness	cm

ΔP	The applied pressure	bar
$\Delta \pi$	The osmotic pressure difference	bar

Abbreviations

AC	Activated Carbon
CP	Concentration Polarization
ED	Electrodialysis
ME	Multi effect distillation
MF	Microfiltration
MSF	Multistage flash
NF	Nanofiltration
RO	Reverse Osmosis
TDS	Total Dissolve Solid
UF	Ultrafiltration
VC	Vapor Compression
PPM	Part Per Million
mgd	Mill Gram per Day
gpd	Gallon per Day
FO	Forward osmosis

List of Tables

List of Figures

CHAPTER ONE

Introduction

1.1 Introduction:

The world's water continually circulates among the oceans, atmosphere, and land masses. The quality of water is governed by various chemical and physical interactions within and among these domains. The efficiency and economy of reliable process become a dominate aspect when solving water treatment problems (Kirk and Othmer, 1981).

A desalination process essentially separates saline water into two parts. One that has a low concentration of salt (treated water or product water) and the other with a much higher concentration than the original feed water, usually referred to as brine concentrated or simply as concentration (Mario, 2005).

There are basically two types of desalination technologies used throughout the world today. These include thermal technologies and membrane technologies. Thermal technologies, as the name implies, involve the heating of saline water and collecting the condensed vapor (distillate) to produce pure water. Membrane technologies use thin semi permeable membranes to separate the feed flow into two flows of different concentrations, a product and concentrate flow. The feed is either brackish water or sea water (Krishna, 2003).

The concepts of "osmosis" and "reverse osmosis" have been known for many years. In fact, studies on osmosis were carried out as early as 1748 by the French scientist Nollet, and many researchers investigated these phenomena over the next three centuries.

However, the use of reverse osmosis (RO) as a feasible separation process is a relatively young technology. In fact, only in the late 1950's did the work of Reid show that cellulose acetate RO membranes were capable of separating salt from water, even though the water fluxes obtained were too small to be practical. Then, in the early 1960's, Loeb and Sourirajan developed a method for making asymmetric cellulose acetate membranes with relatively high water fluxes and separations, thus making RO separations both possible and practical (Williams, 2003).

Since then, the development of new-generation membranes such as the thin-film, composite membrane that can tolerate wide pH ranges, higher temperatures, and harsh chemical environments and that have highly improved water flux and solute separation characteristics has resulted in many RO applications. In addition to the traditional seawater and brackish water desalination processes, RO membranes have found uses in wastewater treatment, production of ultrapure water, water softening, and food processing as well as many others (Williams, 2003).

The driving force for the development and use of RO membranes is the advantages that these have over traditional separation processes such as distillation, extraction, ion exchange, and adsorption. Reverse osmosis is a pressure-driven process so no energy-intensive phase changes or potentially expensive solvents or adsorbents are needed for RO separations. Reverse osmosis is a process that is inherently simple to design and operate compared to many traditional separation processes. Also, simultaneous separation and concentration of both inorganic and organic compounds is possible with the RO process. In addition, with nanofiltration membranes selective solute separations based on charge and molecular weight/size differences are possible. Finally, reverse osmosis technology can also be combined with ultrafiltration, pervaporation, distillation, and other separation techniques to produce hybrid processes that result in highly efficient and selective separations(Williams, 2003).

Reverse osmosis (RO) systems can often improve the quality of water. The reverse osmosis water treatment method has been used extensively to convert brackish or seawater to drinking water, to clean up wastewater, and to recover dissolved salts from industrial processes. It is becoming more popular in the home market as homeowners are increasingly concerned about contaminants that affect their health, as well as about non-hazardous chemicals that affect the taste, odor, or color of their drinking water. People considering the installation of a water treatment system to reduce toxic chemicals should first have their water tested to determine how much if any hazardous compounds are in the water (Barbara et al., 2005).

Aims of the present work:

1. Study the effect of feed flow rate, temperature and different types of salts on the process of reverse osmosis (NaCl, $CaCl_2$ and KCl).
2. Determine the mass transfer coefficient for the above parameters with reverse osmosis.
3. Apply the empirical of the experiment work in the simulink to calculate the mass transfer coefficient and compare it with experiment data.

CHAPTER TWO

Literature Review

2.1 Literature Review:

- In 1967, Kimura & Sourirajan analyzed the practical results for the Reverse Osmosis from cellulose membrane type occurring on the diffusion coefficient in the membrane, and all one relationship of the operating variables by other variables. They depended in analyzing them on the theoretical analysis, where they said, due to the mechanical pressure effect, both of the dissolved and low value for solute diffusion coefficient will be running out in the membrane will accumulate a big part of the solute on the membrane surface constituting boundary layer. This phenomenon is called "concentration polarization". They depended in the theoretical analysis on the transport equations of (Merten & Lonsdal), where the average of water flux is: (Kimura & Sourirajan, 1967)

NB = A [P – {Л (XA2) – Л (XA3)}] 2.1

Where:

NB = the average of water flux (gm.mole/cm^2.sec.)

A = the permeability constant for membrane (gm.mole H2O/ cm^2.sec.atom.)

P = the operating pressure (atom.)

XA2= the mole fraction for solute at the surface between the solution and membrane.

XA3 = the mole fraction for solute in the product water.

While the average of solute flux is:

$$NA = \frac{c * DAM}{K * \delta} (XA2 - XA3) \qquad 2.2$$

Where:

NA = the average of solute flux (gm.mole/cm^2.sec.)

c = the molarity density for solution (gm.mole/cm^3)

DAM= the coefficient of solute diffusion in the water (cm^2/sec)

K = the ratio of solute distribution between the solution and membrane.

δ = the membrane thickness (cm)

XA2= the mole fraction for solute at the surface between the solution and membrane.

XA3= the mole fraction for solute in the product water.

- In 1975, Ohya & Taniguchi analyzed the obtained results from the practical experiments of the Reverse Osmosis from spiral wound type (Roag – 4100) occurring on properties and specifications of the system, depending on the transport equations of Kimura & Sourirajan.(Ohya & Taniguchi, 1975).

- In 1978, Taniguchi analyzed the empirical information of Reverse Osmosis system from spiral wound type (Roag – 4100 & Roag 4160), developed after (Roag – 4000) to find properties and specifications of the system by depending on the transport equations of Reverse Osmosis process and by the same manner found by Kimura & Sourirajan. Taniguchi found the solute transport coefficient value (K) for two types as (0.8×10^{-3} – 1.4×10^{-3}) cm/sec. (Taniguchi, 1978).

- In 1981, Tweddle determined properties and specifications of the Reverse Osmosis system for commercial shapes by analyzing empirical results of three shapes of membranes (Spiral wound, Hollow fiber, Tubular). (Tweddle, 1981).

- In 1986, Mohammed presented a designing study and an economical comparison for the methods of desalination with Reverse Osmosis and Electrodialysis and he concluded the following: (Mohammed Al-hewayzi, 1986).

  1. Limited polarization value is increased with feed water concentration increment, and the required removal percentage.
  2. Electrical energy is increased with the increment of the entered feed water concentration and removal percentage.
  3. The requisite membrane area for the transport in the Reverse Osmosis method is increased with feed water concentration increment and removal percentage.

- There is no desalination process which can be called a clear winner. However, some processes are viable in certain situations. The range of applicability is determined primarily by the salinity and feed water composition. In general, the two processes best suited for seawater desalination are distillation and Reverse Osmosis. For brackish water desalination, Reverse Osmosis and Electrodialysis are currently viable. In principle, freezing can be used for seawater desalting. However, it is not commercially available. Conclusions regarding the most suitable technology for desalination at a particular location cannot be based on the state-of-the-art alone. Costs for desalination plants are quite site specific. Hence, local costs are very important. The optimum process at a particular location depends on the specifics of the application. The optimum type of plant for one application may not be the optimum for another. (Khan, 1986).

- In 1997, Xiaohua & Menachem compared the fouling behavior of a thin- film composite and cellulose acetate Reverse Osmosis membranes, and concluded the fouling results presented so far clearly showing that the cellulose acetate membranes have a lower fouling tendency than the thin-film composite membranes. Possible explanations for the differences in the fouling behavior may include lower permeate flux of the cellulose acetate membranes at the trans membrane pressures investigated, differences in the surface chemistry of the membranes, and differences in the surface roughness of the membranes. (Xiaohua & Menachem, 1997).

- In 2001 Ibrahim S. Al-Mutaz the performance of reverse osmosis system was studied and analysis theoretically and by running some known computer software. Operating data of Riyadh water treatment plants were also used. The effect of temperature on membrane performance is the most important parameter. It was found that higher feed water temperature will result in a better RO operation. When temperature of feed water is increased for constant product flow the required applied feed pressure decreases and the product water salinity increases. Energy consumption is decreased as the applied pressure decreases. If the permeate flow is let to increase as the temperature increase fewer membrane elements will be required. This leads to a considerable saving in the water production cost. The rate of change in permeate flux is about 3 % per degree Celsius increase in water temperature.

- (Abdul Sattar et al. 2003) in their study concluded the following:

1. Percent salinity is inversely proportional to the percent recovery. Higher percent recovery increases the dilution of salt ions that passed through the membrane, and therefore lower permeate salinity.
2. The fouling process affects membrane performance leading to the reduction of permeate flow rate, and accordingly a higher feed pressure is required to attain the design flow. Usually there is a parallel increase in salt passage resulting in a higher salinity of permeates.

- (Mark Wilf et al.2004) in their study concluded the following:

1. The polymer membrane is very sensitive to changes in the feed temperature. There was up to a 60 % increase in the permeate flux when the feed temperature was increased from 20 to 40°C.
2. When the feed flow rate was decreased from 18 to 9 L/min, at constant temperature, the permeate flux decreased, the permeate flux decreased, the permeate flux decreased, but only at the higher feed pressures and higher salt concentrations. This suggests increasing resistance to flux due to enhanced solute build-up at the membrane surface. The flux-pressure relationship was linear for all our results, indicating that only concentration polarization was occurring (i.e. absence of gel layer formation).
3. The permeate flux decreased significantly with increase in feed salinity. At 5% NaCl feed solution the permeate flux reduced to 16 L/m2h presumably due to concentration polarization of solute at the membrane surface.

4. The spiral wound polymer membrane is very sensitive to the feed temperature and to a lesser extent the feed flow rate.
5. The modeling studies showed that the mass transfer coefficient was very sensitive to the feed salt concentration as well as the feed temperature. The mass transfer coefficient, k, and the membrane permeability, PM, decreased with an increase in feed salt concentration, and a decrease in feed temperature.

2.2 Desalination Processes:

Desalination of seawater or brackish groundwater or river water for a public potable water supply is increasingly being considered or adopted around the world in areas where demand has been increasing beyond sustainable supply, where water sources are fragile or overdrawn and climate change is making previously reliable sources unreliable.

Rapidly increasing populations are also placing pressure on existing water sources, forcing governments to turn the desalination to provide additional water supplies when existing sources are fully extended. There is also potential for desalination to process already treated wastewater, normally returned to the environment, to a higher quality level for use in industry or mining processing, thus taking industrial demand away from public water supplies. An additional benefit of this would be reducing the volume of treated wastewater disposed to the environment. This option can produce water at a lower cost of treating sea water in many cases (Oakland, 2006).

Desalination for major public supply is already employed fairly extensively in areas such as the Middle East and North Africa, the Caribbean, Southern Europe and now in China, Singapore and the USA.

A desalination plant essentially separates saline water into two streams: one with a low concentration of dissolved salts (the fresh water stream) and the other containing the remaining dissolved salts (the concentrate or brine stream).The plant requires energy to operate and can use a number of different technologies for the separation of the saline water. The amount of the feed water discharged to waste in the brine stream varies from 20 to 70 percent of the feed flow, depending on the technology employed and the salt content of the feed water.

Desalination is becoming more economically viable as the technology improves. Desalination plants can be provided in a wide range of outputs to cater for small isolated communities or to contribute substantially to water supplies for large cities and even for irrigation.

One of the processes for desalination is reverse osmosis (RO). It is a membrane separation process in which the water from a pressurized saline solution is separated from the dissolved material by flowing through a membrane. No heating is necessary for this separation. In principle, the saline feed water is pumped into a closed vessel where it is pressurized to overcome the osmotic pressure of the solution before diffusing through the membrane. As a portion of the water passes through the membrane, the remaining feed water increases in salt content. This portion of this feed water is then discharged without passing through the membrane (Oakland, 2006).

2.3 Classification of Desalination Processes:

Many methods have been proposed for desalting saline water, but few were commercially used. The two most popular methods for classifying the well-known desalination processes are as follows:

1. Processes in which desalination taking place involves phase change. There are three main methods:

- Vapor-compression (VC) distillation
- Multi-effect (ME) distillation
- Multistage flash (MSF) distillation

2. Processes in which desalination takes place without any phase change. These include the following two main methods:

- Reverse osmosis (RO)
- Electrodialysis (ED)

There are three other membrane processes which are not considered desalination processes but are relevant.

- Microfiltration (MF)
- Ultrafiltration (UF)
- Nanofiltration (NF)

Ion exchange is not regarded as a desalination process. It is generally used to improve water quality for some specific purpose. E.g. boiler feed water (Patrick, 1976).

2.3.1 Thermal Processes:

Distillation is one of the oldest and most commonly used desalting techniques. In this process, the water evaporation and vapor condensation occur to obtain distillate at the end (Judson, 1971).

2.3.1.1 Vapor compression (VC):

Vapor-compression distillation uses mechanical energy rather than thermal energy. It is based on a simple principle. Saline water is sprayed over an evaporator tube bundle. The vapor formed at some temperature and pressure is then compressed either thermally in a steam ejector, or mechanically (high and low pressure) in a compressor, causing the condensation temperature and pressure to increase and the volume to decrease. Compressed vapor is passed through the evaporator bundle, where it condenses and forms distilled water. The heat of condensation could be recycled to evaporate more brine (Degremont, 1979).

The vapor compression distillation (VCD) process is used either in combination with other processes such as the MED, or by itself. The heat for evaporating the water comes from the compression of vapor, rather than the direct exchange of heat from steam produced in a boiler (Buros, 2000). Vapor compression (VC) units have been built in a variety of configurations. Usually, a mechanical compressor is used to generate the heat for evaporation. The VC units are generally small in capacity, and are often used at hotels, resorts and in industrial applications.

2.3.1.2 Multi-Effect Distillation (ME):

Multi-Effect (ME) distillation was the first process used to produce a significant amount of water from the sea. This process takes place in a series of effects (vessels) and uses the principle of reducing the ambient pressure in the various effects in order of their arrangement. This causes the feed water to undergo boiling in a series of effects without supplying additional heat after the first effect. Vapor generated in the first effect gives up heat to the second effect for evaporation and is condensed inside the tubes. This continues for several effects (Shaffer and Mintz, 1980).

The seawater is either sprayed, or otherwise distributed onto the surface of evaporator tubes in a thin film to promote rapid boiling and evaporation.

The condensate from the boiler steam is recycled to the boiler for reuse. The larger the number of effects, the less heat that is required as heat sources. There are vertical- and horizontal-tube evaporation effects. The vertical tubes could be of the rising or the falling-film type. The formation of falling films of water on the inner surfaces of the heating tubes affects evaporation in the vertical-tube evaporators, so the falling films are heated by the steam passing outside the tubes.

However, with horizontal effects, evaporation takes place on the outer surfaces of the heating tubes, steam for heating being condensed inside the tubes.

Multi-Effect distillation plants tend to come in a much greater variety of plant designs than do Multi-Stage Flash (MSF) Distillation processes. The designer could select a number of heat-transfer surface configurations and a number of flow sheet variations, thus leading to a larger number of possible combinations (George, 1997).

2.3.1.3 Multi-Stage Flash (MSF) Distillation:

This process involves the use of distillation through several (multi-stage) chambers. In the MSF process, each successive stage of the plant operates at progressively lower pressures. The feed water is first heated under high pressure, and is led into the first ‘flash chamber’, where the pressure is released, causing the water to boil rapidly resulting in sudden evaporation or ‘flashing’. This ‘flashing’ of a portion of the feed continues in each successive stage, because the pressure at each stage is lower than in the previous stage. The vapor generated by the flashing is converted into fresh water by being condensed on heat exchanger tubing that run through each stage. The tubes are cooled by the incoming cooler feed water. Generally, only a small percentage of the feed water is converted into vapor and condensed.

Multi-Stage flash distillation plants have been built since the late 1950s. Some MSF plants can contain from 15 to 25 stages, but are usually no larger than 15 mgd in capacity. MSF distillation plants can have either a ‘once-through’ or ‘recycled’ process. In the ‘once-through’ design, the feed water is passed through the heater and flash chambers just once and disposed of, while in the recycled design, the feed water for cooling is recycled.

MSF plants are subject to corrosion unless stainless steel is used extensively. In addition to corrosion, MSF plants are also subject to erosion and impingement attack (Planners, 2003). Erosion is caused by the turbulence of the feed water in the flash chamber, when the feed water passes from one stage to another.

Distillation processes produce about 3.4 billion gpd globally, which is about 50 percent of the worldwide desalination capacity. MSF plants provide about 84 percent of that capacity. Most of those plants have been built, primarily in the Middle East, where energy resources have been plentiful and inexpensive.

2.3.2 Membrane Separation Technology:

The membrane can be defined essentially as a barrier, which separates two phases and restricts transport of various chemicals in a selective manner. A membrane can be homogenous or heterogeneous, symmetric or asymmetric in structure, solid or liquid can carry a positive or negative charge or be neutral or bipolar. Transport through a membrane can be affected by convection or by diffusion of individual molecules, induced by an electric field or concentration, pressure or temperature gradient. The membrane thickness may vary from as small as 100 micron to several mms.

A membrane separation system separates an influent stream into two effluent streams known as the permeate and the concentrate. The permeate is the portion of the fluid that has passed through the semi-permeable membrane. Whereas the concentrate stream contains the constituents that have been rejected by the membrane (Srikanth, 2000).

Various types of membrane separation have been developed for specific industrial applications. Some of the widely used processes are discussed hereunder:

2.3.2.1 Microfiltration (MF):

Microfiltration is a membrane separation process with membrane pore sizes between 0.05 and 5 µm, operating at pressures up to 0.1 MPa. The membrane rejects particles and dissolved macromolecules larger than 0.1 µm. Microfiltration membranes have a relatively large average pore size and consequently can retain only relatively large impurities (i.e. suspended fine particles, but not colloidal matter), but operate at high flux rates. Microfiltration is used industrially for the removal of particulate material and has been used extensively in drinking water treatment and for the treatment of domestic sewage (Vassilis, 1988 and Al-Malack and Anderson, 1997).

The process can be used in conjunction with precipitation processes, provided the precipitated particles are suitably coarse. Organic as well as inorganic microfiltration membranes can be used, depending on the characteristics of the feed water. Ceramic microfilters have been used in the nuclear industry for high activity wastewaters because of that material’s radiation stability (Peter, 1985 and John, 2003).

2.3.2.2 Ultrafiltration (UF):

The ultrafiltration membrane is considerably more porous (i.e. its nominal pore size is larger) than the reverse osmosis membrane. As a result, most soluble species, including inorganic salts, pass through the membrane with the water. Colloids, suspended solids and high molecular weight organic molecules do not pass through the membrane with the water. They are rejected and remain in the concentrate stream. The porous nature of the ultrafiltration membrane allows the process to be operated with high fluxes at relatively low pressures (e.g. 0.2–1.4 MPa). This is possible because the osmotic pressure of colloids and high molecular weight organics is extremely low. The degree and quantity of the separation are a result of the pore size of the membrane and the molecular structure, size, shape and flexibility of the colloids and organic molecules. Pore sizes ranging from 0.001–0.01 µm allow separation from solution of molecules with a molecular weight between 500 and 300 000. Ultrafiltration is used industrially for the removal of macromolecules and colloids from wastewater and has been used extensively in the food and dairy industries (Sablani et al., 2001 and Goosen et al., 2004).

In some applications the primary function of ultrafiltration systems is to remove colloids and other particulate foulants from feed streams that are to be further treated by reverse osmosis systems (Jarusutthirak et al., 2002).

2.3.2.3 Nanofiltration (NF):

Nanofiltration (NF) membranes have been largely developed for this past decade. These membranes have a porous structure with pores less than 2 nm in diameter (between ultrafiltration and reverse osmosis membranes). NF membranes can be used for the removal of salts in water, the fractionation of salts and small molecules in industrial waste water, drinking and industrial water production. A great advantage of NF compared with RO is operating pressure that can be much smaller while keeping relatively high permeate flux. Various mechanisms are likely to be involved in the separation process of NF membranes such as sieving, electrostatic interactions between the membrane and the charged species, differences in diffusivity and solubility, differences in born self-energy and dielectric exclusion (Labbez, 2003).

Nanofiltration is a form of filtration that uses membranes to separate different fluids or ions. NF is typically referred to as "loose" RO due to its larger membrane pore structure as compared to the membranes used in RO, and allows more salt passage through the membrane. Because it can operate at much lower pressures, and passes some of the inorganic salts, NF is used in applications where high organic removal and moderate inorganic removals are desired. An advantage of NF over RO is that NF can typically operate at higher recoveries, thereby conserving total water usage due to a lower concentrate stream flow rate. NF is not effective on small molecular weight organics, such as methanol (Srikanth, 2000).

2.3.2.4 Reverse Osmosis (RO):

Reverse osmosis (RO) is a membrane separation process that recovers water from a saline solution pressurized to a point greater than the osmotic pressure of the solution. The membrane filters out the salt ions from the pressurized solution, allowing only the water to pass. RO post-treatment includes removing dissolved gasses (CO_2), and stabilizing the pH via the addition of Ca or Na salts.

Pressurizing the saline water accounts for most of the energy consumed by RO. Since the osmotic pressure, and hence the pressure required to perform the separation is directly related to the salt concentration, RO is often the method of choice for brackish water, where only low to intermediate pressures are required. The operating pressure for brackish water systems ranges from 15 – 25 bar and for seawater systems from 54 to 80 bar (the osmotic pressure of seawater is about 25 bar) (Buros, 2000). Since the pressure required for recovering additional water increases as the brine stream is concentrated, the water recovery rate of RO systems tends to be low. A typical recovery value for a seawater RO system is only 40% (Spiegler, and El-Sayed, 1994).

Since most of energy losses for RO result from releasing the pressure of the concentrated brine, large scale RO systems are now equipped with devices to recover the mechanical compression energy from the discharged concentrated brine stream with claimed efficiencies of up to 95% (Hauge, 1995). In these plants, the energy required for seawater desalination has now been reported to be as low as 9 kJ/kg product (Pique, 2000).

This low value however is more typical of a system treating brackish water. RO membranes are sensitive to pH, oxidizers, a wide range of organics, algae, and bacteria and of course particulates and other foulants (R. Semiat, 2000). Therefore, pretreatment of the feed water is an important consideration and can a significant impact on the cost of RO (El-Dessouky et al., 1999), especially since all the feed water, even the 60% that will eventually be discharged, must be pretreated before being passed to the membrane.

2.3.2.5 Electrodialysis (ED):

Electrodialysis (ED) utilizes a direct current source and a number of flow channels separated by alternating anion and cation selective membranes to achieve the separation of water and dissolved salts (Figure 2-1) (Buros, 2000).

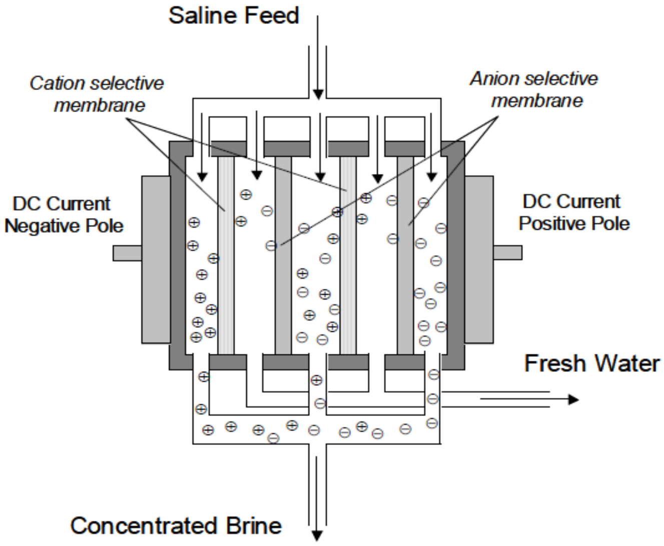


Figure 2-1 Schematic diagram of Electrodialysis desalination process (Buros, 2000).

In the ED process, saline water is fed in parallel to each of the separate channels. Cations and anions then migrate in opposite directions in response to the applied voltage. Due to the charge selectivity of the membranes, the ion concentration increases and decreases in alternating channels of the apparatus. A single membrane stack may consist of hundreds of these alternating channels. Since the resistance in the stack changes from top to bottom, the separation is typically carried out is a series of small steps. This makes the process more economical and easier to control. Like RO, the energy required to separate the ions from solution increases with concentration, thus ED is generally limited to brackish waters containing only a few thousand ppm of dissolved solids.

The membrane of ED units are subject to fouling, and thus some pretreatment of the feed water is usually necessary. Precipitation of scale can be facilitated in the ED process by changes on pH that occur near the membranes as a result of the transport of H+ and OH- ions (Spiegler, and El-Sayed, 1994).However, since there is not a flux of water through the membranes, ED can treat water with a higher level of suspended solids than RO. Also, since nonionic solids, e.g. silica, are not concentrated by the process, these components are of less concern (Buros, 2000). The electrodialysis reversal (EDR) process was developed to help eliminate membrane fouling. In the EDR process, the membrane polarity is reversed several times an hour. This has the effect of switching the brine channels to freshwater channels, and the freshwater channels to brine channels, and breaks up and flushes out deposits (Spiegler et al., 1994 and Buros, 2000).

2. 4 Advantages and Disadvantages of Desalination Processes:

The desalination processes have many aspects and common problems, such as energy consumption, scale formation, fouling, corrosion, heat transfer, and others. Desalting processes are often compared on the basis of the product water quality. In any specific case, the main factors before a particular desalting process is selected may include product water quantity and quality, feed water quality, characteristics, temperature, and availability of energy,

location, process limitations and process economics, as indicated in table (2-1). (Mousa, 1999).

Table (2-1) Characteristics of desalination processes (Mousa, 1999).

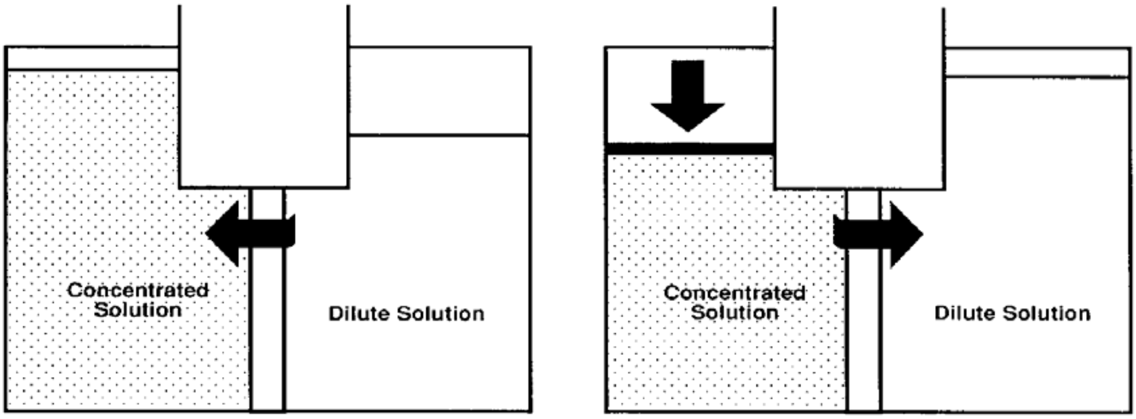
Method of desalination	Advantages	Disadvantages
• Multi Effect (ME)	•High production capacity. •High purity production (Less than 30 mg/l TDS). • Minimal skilled operate.	•Dependence of output on power availability. •Expensive to build & operate. • Long construction period. •Difficult to control water quality. • Large space and material requirements. • The recovery ratio is low.
•Multi Stage Flash (MSF)	•The salinity of the feed water does not have much impact on the process or costs. • High purity production (Less than 30 mg/l TDS). • High production capacity. • Low skill requirement. • Low energy input.	• Requires pretreatment of feed water. • High operating costs. •High construction requirements. • Limited potential for improvement. • The recovery ratio is Low.

•Vapor Compression (VC)	• High water quality. • Short construction period. • Small space requirement. • Operation flexibility.	• High operational costs. • High energy consumption. •Difficult of water quality control.
• Reverse Osmosis (RO)	•Suitable for both sea and brackish water. •Low power requirement compared with (Multi Effect Desalination and Vapor compression). •Simple Operation and build cheap. • Flexibility in site location • The use of chemicals for cleaning purposes is low.	• Requires high quality feed water. • High pressure requirements. •Long construction time for large plants. • Relatively high capital and operating costs. •Reverse Osmosis membranes are expensive and have a life expectancy of (2-5) years.
• Electrodialysis (ED)	•Low operation and capital costs. • The recovery ratio is high. • Low energy consumption. • Low space and material requirements. • Can be operated at low, to moderate pressure.	•Low to medium brackish water (3,000 mg/l). •Requires careful pretreatment of feed water. • Low production capacity. • Purity affected by quality of feed water. • Leaks sometimes occur in membrane stacks.

2.5 Mechanism and Principles of Reverse Osmosis:

The phenomenon of osmosis occurs when pure water flows from a dilute saline solution through a membrane into a higher concentrated saline solution.

The phenomenon of osmosis is illustrated in Figure 2-2. A semi-permeable membrane is placed between two compartments. “Semi-permeable” means that the membrane is permeable to some species, and not permeable to others. Assume that this membrane is permeable to water, but not to salt. Then, place a salt solution in one compartment and pure water in the other compartment. The membrane will allow water to permeate through it to either side. But salt cannot pass through the membrane.



Water diffuses through a semi permeable membrane toward region of higher concentration to equalize solution strength. Ultimate height difference between columns is “osmotic” pressure.

Applied pressure in excess of osmotic pressure reverses water flow direction. Hence the term “reverses osmosis”

Fig. 2-2 The phenomenon of osmosis (Dow Company, 2005).

As a fundamental rule of nature, this system will try to reach equilibrium. That is, it will try to reach the same concentration on both sides of the membrane. The only possible way to reach equilibrium is for water to pass from the pure water compartment to the salt-containing compartment, to dilute the salt solution.

Figure 2-2 also shows that osmosis can cause a rise in the height of the salt solution. This height will increase until the pressure of the column of water (salt solution) is so high that the force of this water column stops the water flow. The equilibrium point of this water column height in terms of water pressure against the membrane is called osmotic pressure.

If a force is applied to this column of water, the direction of water flow through the membrane can be reversed. This is the basis of the term reverse osmosis. Note that this reversed flow produces pure water from the salt solution, since the membrane is not permeable to salt (Dow Company, 2005).

When tried to separate pure water and a salt solution through a semipermeable membrane, the pure water diffuses through the membrane and dilutes the salt solution. The membrane rejects most of the dissolved salts, while allowing the water to permeate. This phenomenon is known as natural osmosis (Figure 2-3-a).

As water passes through the membrane, the pressure on the dilute side drops, and the pressure of the concentrated solution rises. The osmotic flux continues until equilibrium is reached, where the net water flux through the membrane becomes zero (Figure 2-3-b).

At equilibrium, the liquid level in the saline water will be higher than that on the waterside. The amount of water passing in either direction will be equal. The hydrostatic pressure difference achieved is equal to the effective driving force causing the flow, called osmotic pressure. This pressure is a strong function of the solute concentration and the temperature, and depends on the type of ionic species present.

Applying a pressure in excess of the osmotic pressure to the saline water section slows down the osmotic flow, and forces the water to flow from the salt solution into the waterside. Therefore, the direction of flow is reversed, and that is why this separation process is called reverse osmosis (Figure 2-3-c) (Khawla, 1998).

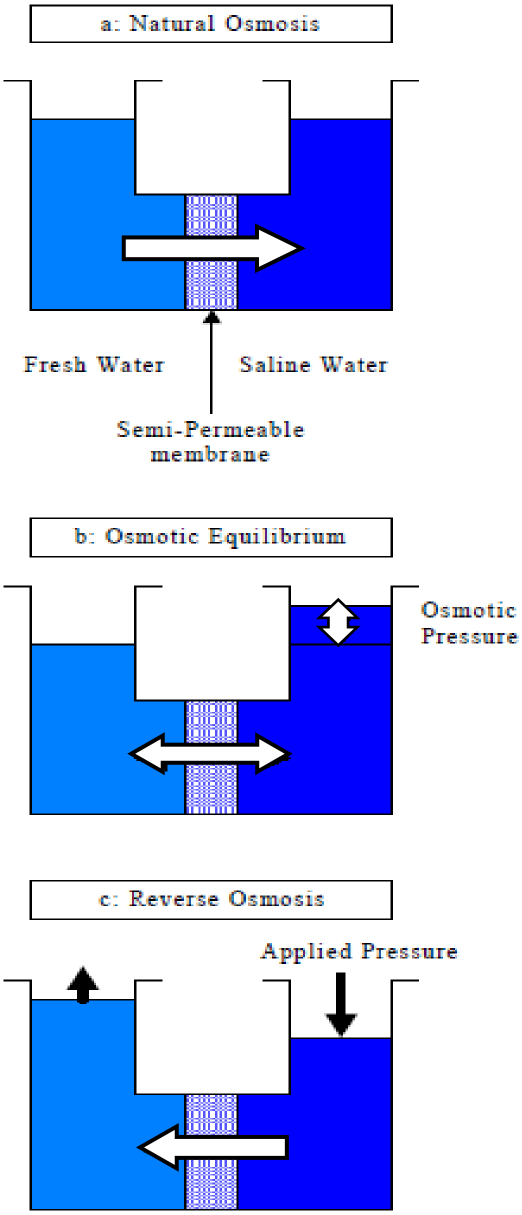


Figure 2-3 Principle of reverse osmosis (Khawla, 1998).

2.6 Membrane Module Types and Their Characteristics:

The various membrane materials described in the previous section are available in one or more of the following modules: plate and frame, spiral-wound, tubular, and hollow fiber.

2.6.1 Plate and Frame:

Process Description: (In Figure 2-4), the arrows show the upstream and permeate paths. The upstream leaves as the retentate and is enriched in non-permeate. Permeates is collected from channels in support plates and leaves enriched in the most permeable component.

Advantages: Easy to clean and replace membranes.

Disadvantages: Low membrane area per volume (a problem in high pressure application where pressure vessel costs are significant) (Syed Ali et al., 2000).

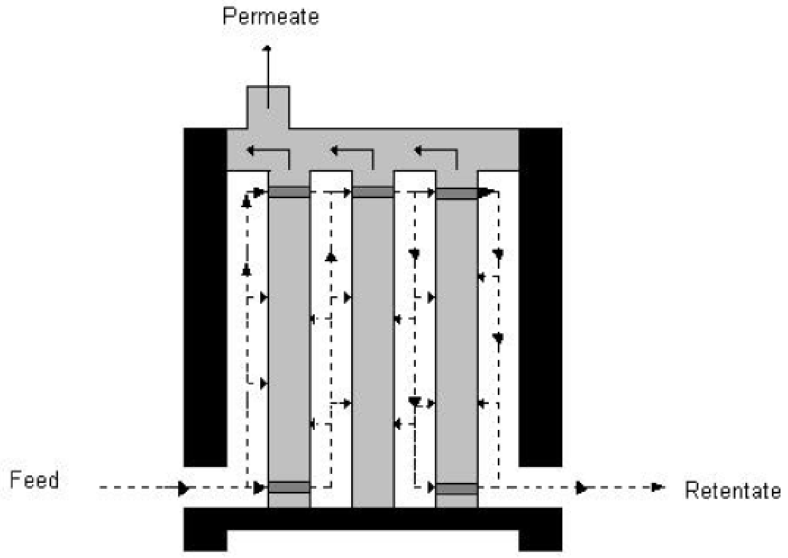


Figure2-4 Plate and Frame Schematic (Syed Ali et al., 2000).

2.6.2 Spiral-Wound Module:

Process Description: The feed passes through membrane that is spirally wound around the porous tube (refer to Figure 2-5). The membrane, feed spacer, and permeate spacer are glued on three sides and terminates at its fourth side into the porous pipe that collects the permeate. This module is wrapped into a spiral and placed in a cylinder shell.

Advantages: Easy and inexpensive to adjust hydronomics by changing feed spacer thickness to overcome concentration polarization and fouling.

Disadvantages: Low membrane area per volume. May become expensive for high-pressures because extra high-pressure shells must be purchased. By passing of feed may occur due to nonuniform wrapping of module spiral (Syed Ali et al., 2000).

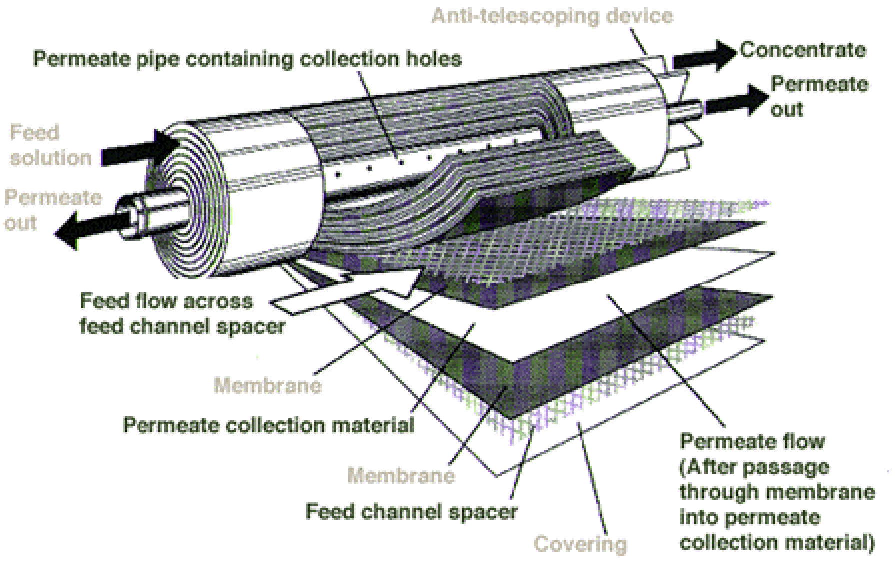


Figure 2-5 Spiral Wound Schematic (Syed Ali et al., 2000).

2.6.3 Hollow-Fiber, Capillary, and Tubular:

Process Description: Hollow fibers (<0.5 mm diameter (D)), capillaries (0.5-5 mm D), and tubes (5-15 mm D) can be configured in bore or shell feed modules (refer to Figures 2-6 and 2-7). Packing density can be as high as

50%. Positions of feed and permeate ports affect flow in the module. High-pressure applications tend to use shell feed, and low pressures favor bore feeds.

Advantages: Eliminates bypassing when used in bore feed mode. Huge membrane area per volume possible with fibers.

Disadvantages: Membrane formation is more complex because the support and the selecting layer are formed as an integral cylindrical unit during spinning for fibers and capillaries. To avoid excessive pressure drop in bore for bore feed mode, large diameter fibers may be required in some cases. Bypassing of feed in shell feed mode due to any nonuniformities in fiber packing. Fouling due to high surface areas if feed contains particulates or other foulants (Syed Ali et al., 2000).

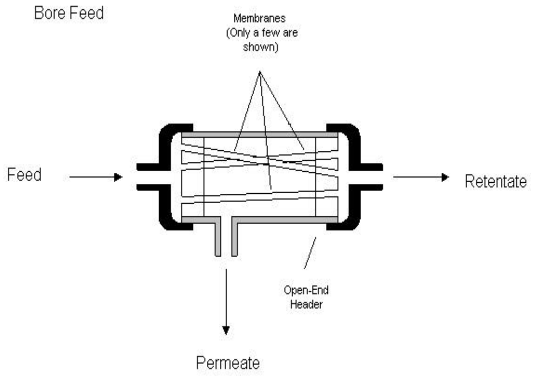


Figure 2-6 Bore Feed Schematic

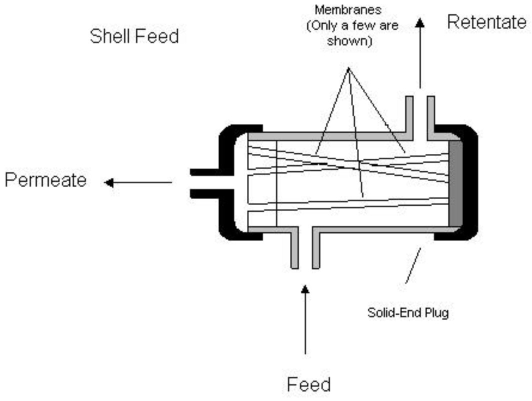


Figure 2-7 Shell Feed Schematic

Table 2-2 lists the packing density (m^2/m^3), resistance to fouling, ease of cleaning, relative cost, and main applications for each of the modules (Syed Ali et al., 2000).

	Plate and Frame	Spiral Wound	Tubular	Hollow-Fiber
Packing density, m^2/m^3	30 to 500	200 to 800	30 to 200	500 to 9,000
Resistance to fouling	Good	Moderate	Very Good	Poor
Ease of cleaning	Good	Fair	Excellent	Poor
Relative cost	High	Low	High	Low
Main application	D, RO, PV, UF, MF	D, RO, GP, UF, MF	RO, UF	D, RO, GP, UF

2.7 Membrane Performance:

The overall performance of a membrane based separation process depends on the characteristics of the membrane, the feed solution being treated and the general operating practices that are employed. During the operation of membrane systems, membrane performance typically will decrease over time.

2.7.1 Membrane Compaction:

The compressive force applied to the membrane under the system hydraulic pressure causes membrane compaction. It is a plastic creep process in which the thin membrane skin grows in thickness as it is compressed into the underlying porous substratum that supports the membrane. As the membrane thickens, its permeability decreases (Lacey and Leob, 1972).

When pure or clean water feed is processed at a constant pressure, the flux through asymmetric reverse osmosis membranes will gradually decrease. The effect of compaction on the maximum attainable flux for the applied pressure can be predicted. The flux losses are irreversible (Dale and Okos, 1983).

Compaction of the membrane structure usually takes place during the initial period when feed is introduced to the membrane system. After pressure is continuously applied to the membranes, the loss in membrane flux should not exceed 10% (Hussein and James, 1989 and Taha, 2000).

During operation of RO systems, membrane material is exposed to very high pressure that may increase in the density of membrane material (compaction).

The compaction decreases the diffusion rate of water and dissolved constituents through the membrane. Therefore, higher pressure has to be applied to maintain the design permeate flow. In parallel, a lower rate of salt diffusion will result in lower permeate salinity. The effect of compaction is more significant in asymmetric cellulose membranes than in composite polyamide membranes. In seawater RO systems, where the feed pressure is much higher than that in brackish applications, the compaction process is more significant. Higher feed water temperature will also result in a higher compaction rate. Usually membrane compaction results in few percent flux decline, and has strongest effect during the initial operating period (Ibrahim Khaled, 2007).

2.7.2 Concentration Polarization:

Concentration polarization can be defined as an increase in solute concentration at the membrane wall (relative to the bulk concentration) caused by the convective flow of solute to the wall being greater than the solute's rate of diffusion away from the wall.

The negative effects of concentration polarization include:

- A decrease in permeate flow.
- An increase in solute passage through the membrane.
- A contributor to fouling (Bhattacharyya and Williams, 1992).

In pressure-driven membrane processes, convective permeate flow causes a buildup of solute at the membrane active layer surface. Referred to as concentration polarization (CP), this phenomenon reduces permeate water flux due to increased osmotic pressure that must be overcome with hydraulic pressure (Tzahi and Amy, 2006).

CP due to water permeation is not limited to pressure-driven membrane processes and also occurs during osmotic-driven membrane processes, on both the feed and permeates sides of the membrane. When the feed solution flows on the active layer of the membrane (like in RO), solutes build up at the active layer. This may be called concentrative external CP and is similar to CP in pressure-driven membrane processes. Simultaneously, the draw solution in contact with the permeate side of the membrane is being diluted at the permeate–membrane interface by the permeating water. This is called dilutive external CP. Both concentrative and dilutive external CP phenomena reduce the effective osmotic driving force. The adverse effect of external CP on osmotic-driven membrane processes can be minimized by increasing flow velocity and turbulence at the membrane surface or by manipulating the water flux (Mulder, 1997).

However, because water flux in FO is already low, the ability to diminish external CP by reducing flux is limited. For modeling external CP phenomena in FO, equations similar to those developed for CP of pressure-driven membranes can be used (Sablani et al., 2001).

Due to the low hydraulic pressure used in FO, membrane fouling induced by external CP has milder effects on water flux compared to the effects in pressure-driven membrane processes.

It has been shown that external CP plays a minor role in osmoticdriven membrane processes and is not the main cause for the lower-than-expected water flux in such processes (McCutcheon et al., 2006).

2.7.3 Membranes Fouling:

In general, fouling is defined as the accumulation of material on the surface or in the pores of a membrane that decreases the water flux through the membrane. The loss of membrane efficiency due to fouling is one of the main impediments to the development of membrane processes for use in drinking water treatment. Membrane fouling is dependent on the water quality as well as the membranes properties and construction. The consequences of fouling can be severe; fouling can reduce the water flux through a membrane up to 90 percent.

There are five broad fouling categories, microorganisms, colloidal or particulate matter, and dissolved organic, sparingly soluble inorganic and chemical reactants.

The continuous accumulation of the components that are rejected by the membrane in the dead-end cells maximizes precipitation and fouling. Dissolved organic and colloidal matters are considered to be serious foulants due to the difficulty in removing them with pretreatment processes. Inorganic fouling can often be controlled by acid addition. Inorganic fouling are typically removed with acid solutions, where as organic and biological fouling are typically removed with alkaline (Ahmed Al-Rubaie, 2006).

Fouling is an accumulation of particulate matter on the membrane surface which results in the plugging of the membrane and therefore a decrease in the permeate flow (Bhattacharyya and Williams, 1992).

Foulants can be classified in three broad categories:

1. Sparingly soluble inorganic compounds,

2. Particulate matter, and

3. Dissolved organic compounds (Potts et al., 1981).

To reduce concentration polarization and fouling, SWMs are designed with a plastic netting inserted between the membrane leafs to increase flow turbulence and thus increase mixing. The drawback to the increased turbulence is an increase in hydraulic pressure losses through the length of the membrane.

Membrane fouling has a negative effect on membrane performance, and in extreme cases may result in non-reversible membrane degradation. The membrane fouling process is usually referred to as deposition of inorganic or organic substances on the membrane surface and/or blockage of feed channels. In the initial stages of membrane fouling, performance changes are similar to those caused by the compaction process. The fouling process is usually associated with an increase in pressure drop. An uncontrolled fouling process may lead to very severe performance degradation and even to complete destruction of membrane elements. The most effective way to control membrane fouling is to identify the origin of the fouling process and eliminate it by the modifying pretreatment process or operating conditions. Foulant deposits can be removed from the membrane surface by chemical cleaning. However, success of the cleaning procedure depends on the age of the foulant deposit, and on proper selection of the cleaning solution (Ibrahim Khaled, 2007).

2.7.4 Membrane Cleaning:

Even if all possible measures have been taken to prevent membrane fouling, changes in the composition of the feed, breakthrough in the sand filter, production upsets and a host of other irregular operating conditions may result in membrane fouling. The fouling will, in most cases , be detectable as a gradual decrease in plant productivity and later as a gradual increase in the salt content of the permeate. A detailed log of the plant operation and analysis of the production parameters helps to identify a problem situation at an early stage before the problems become severe. If fouling is allowed to develop, cleaning becomes much more difficult.

In the case of mineral fouling, chemical cleaning agents are used. Acid generally will remove inorganic salts. Cellulose acetate membranes do not tolerate low or high pH values, whereas polyamide membranes generally can be cleaned down to pH 2. Acetic acid, which also is a complexing agent for some metals, often is used with good results. Complexing agents like quaternary ammonium compound operate better at elevated pH values, which renders them of little use for cellulose acetate membranes.

2.8 Factors Effecting Reverse Osmosis System:

Permeate flux and salt rejections are the key performance parameters of a reverse osmosis or a nanofiltration process. Under specific reference conditions, flux and rejection are intrinsic properties of membrane performance. The flux and rejection of a membrane system are mainly influenced by variable parameters including:

2.8.1 Effect of Pressure:

Feed water pressure affects both the water flux and salt rejection of RO membranes. Osmosis is the flow of water across a membrane from the dilute side toward the concentrated solution side. Reverse osmosis technology involves application of pressure to the feed water stream to overcome the natural osmotic pressure. Pressure in excess of the osmotic pressure is applied to the concentrated solution and the flow of water is reversed. A portion of the feed water (concentrated solution) is forced through the membrane to emerge as purified product water of the dilute solution side (Figure2- 8).

With increasing effective feed pressure, the permeate TDS will decrease while the permeate flux will increase as shown in Figure (Dow Company, 2005).

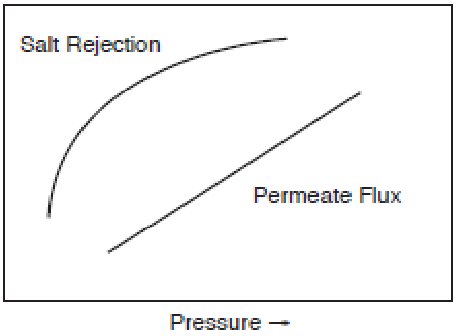


Fig 2- 8 Effect of Feed Water Pressure on Flux and Salt Rejection

(Dow Company, 2005).

2.8.2 Effect of Temperature:

As figure (2-9) demonstrates, membrane productivity is very sensitive to changes in feed water temperature. As water temperature increases, water flux increases almost linearly, due primarily to the higher diffusion rate of water through the membrane. Increased feed water temperature also results in lower salt rejection or higher salt passage. This is due to a higher diffusion rate for salt through the membrane. The optimum water temperature for most Reverse Osmosis membrane is 25 ºC. As the temperature drops to 25 ºC the capacity of the Reverse Osmosis unit will be reduced to less than one half. (Dow Company, 2005).

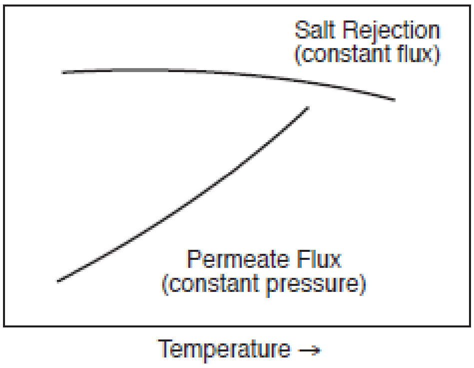


Fig. 2-9 Effect of Feed Water Temperature on Flux and Salt Rejection

(Dow Company, 2005)

2.8.3 Effect of Salt Concentration:

Osmotic pressure is a function of the type and concentration of salts or organics contained in feed water. As salt concentration increases, so does osmotic pressure. The amount of feed water driving pressure necessary to reverse the natural direction of osmotic flow is, therefore, largely determined by the level of salts in the feed water. Figure (2-10) demonstrates that, if feed pressure remains constant, higher salt concentration results in lower membrane water flux. The increasing osmotic pressure offsets the feed water driving pressure, also illustrated in figure (2-10) is the increase in salt passage through the membrane (decrease in rejection) as the water flux declines. (Dow Company, 2005).

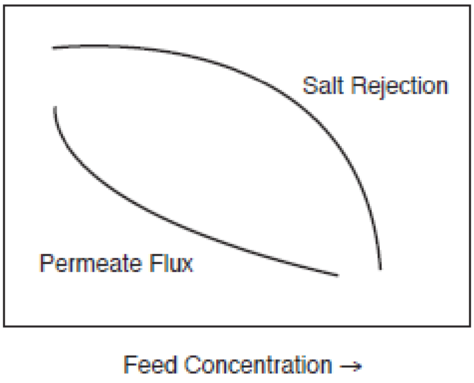


Figure 2-10 Effect of increasing salt concentration on flux and salt rejection

(Dow Company, 2005)

2.8.4 Effect of Recovery:

As shown in figure (2-11), Reverse Osmosis occurs when the natural osmotic flow between a dilute solution and a concentrated solution is reversed through application of feed water pressure. If percentage recovery is increased and feed water pressure remains constant, the salts in the residual feed become more concentrated and the natural osmotic pressure will increase until it is as high as the applied feed pressure. This driving effect of feed pressure, slowing or halting the Reverse Osmosis process and even stop as shown in figure (2-11).

The maximum percent recovery possible in any Reverse Osmosis system usually depends not on a limiting osmotic pressure, but on the concentration of salts present in the feed water and their tendency to precipitate on the membrane surface as mineral scale. Chemical treatment of feed water can be used to inhibit mineral scaling. (Dow Company, 2005).

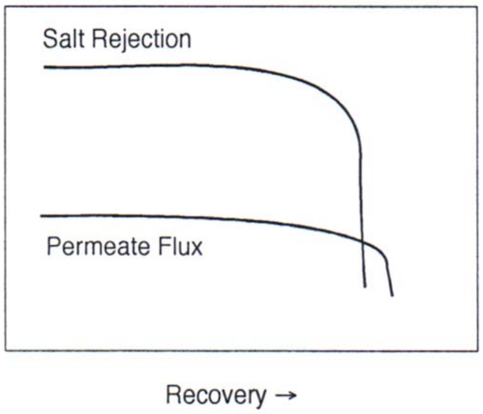


Figure 2-11 Effect of increased recovery on flux and salt rejection

(Dow Company, 2005)

2.8.5 Effect of pH:

The (pH) tolerance of various types of Reverse Osmosis membranes can vary widely. Membrane salt rejection performance depends on (pH). Water flux may also be affected. Figure (2-12) shows that water flux and salt rejection for membranes are essentially stable over a broad (pH) range. (Dow Company, 2005).

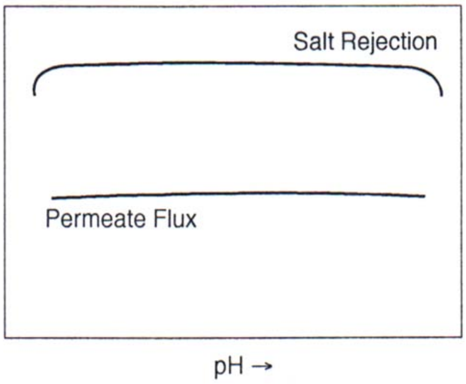


Figure 2-12 Effect of feed water pH on water flux and salt rejection

(Dow Company, 2005)

2.9 Reverse Osmosis Model:

Membrane separation such as reverse osmosis, ultrafiltaration and others are recently developed separetion techniques. As compared with other separation methods, membrane processes possess many advantages such as no phase change, simple equipment, low energy consumption and ease of operation. As a result they have a great tendency to be used in industry as well as in scientific and technological circles.

The main task in designing reverse osmosis devices is selecting the optimum hydraulic parameters to reduce the power consumption of the system and extend the life of the membrane, thus lowering production cost. For the optimization of hydraulical parameters, concentration polarization which seriously affects the performance of the separation system is one of the important control factors influencing the system design. It could be reduced to a certain extend by some appropriate measures, but complete elimination of this phenomenon is impossible in any practical operating system (Wang Xuesong, 1987).

2.9.1 Concentration polarization:

Figure 2-13 illustrates the mass transfer system in reverse osmosis for a membrane under steady state conditions (Vineet, 2007). When the solution flows through the system parallel to the membrane surface at a given rate, both the solute the solvent are forced to pass through the membrane owing to the action of the pressure difference. The solvent can pass through the membrane completely, but most of the solute accumulates at the surface due to the rejection caused by the membrane. Thus a concentration gradient is built between the menbrane surface and bulk solution, which makes the solute diffuse back towards the bulk solution. The higher the concentration of solute at the membrane surface, the lower the permeation rate of the solvent. This unfavorable phenomenon is called "concentration polarization" (Wang Xuesong, 1978 and van et al., 2001).

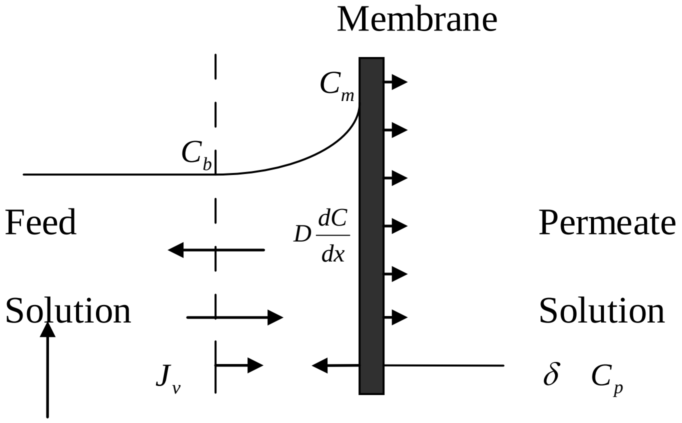


Fig 2-13 Schematic of concentration profile of the solute in the feed and the permeate solution near the membrane surface.

The starting point for our mathematical description of RO/NF separations is the solution–diffusion model. The model assumes that the permeation driving force is the gradient in chemical potential of the solute (Wijmans and Baker, 1995). When the transport equation is expressed in terms of solvent flux (*J*), it is given as

$$J_v = A(\Delta P - \sigma\Delta\pi) \qquad 2.3$$

Where *A* is the solvent permeability through the membrane, ΔP is the applied pressure, $\Delta\pi$ is the osmotic pressure difference between the membrane surface and the permeate, and σ is the reflection coefficient. The reflection coefficient represents the intrinsic salt rejection by the membrane, but when intrinsic salt rejection is over a 0.98, which is typical for reverse osmosis separations, σ may be assumed equal to unity (Bhattacharjee *et al*., 2001**).** When intrinsic salt rejection is significantly less than 0.98 (i.e., nanofiltration), the reflection coefficient should be used to more accurately predict the resultant *trans*-membrane osmotic pressure (Murthy and Gupta, 1997, Eric *et al.*, 2002).

Equation (2.3) serves as the starting point for the design of most modern RO/NF separations. The rejection of ionic species results in an elevated salt concentration near the membrane surface creating a local concentrated layer (CP), this layer quickly reaches a steady state, and the transverse solute flux through the CP layer is constant(Eric *et al*., 2002, Jeffrey *et al*., 2007). The solvent flux (*J*) may then be determined by the following one-dimensional, steady-state mass balance across the CP layer:

$$J_v.C_p = J_v.C - \left[D(\frac{dC}{dx}) \right] \qquad 2.4$$

Where, J_v is the permeate flux through the membrane, C_p is the permeate solute concentration, C is the solute concentration in the boundary layer and D is the solute diffusion coefficient in water (Dharmesh, 2002).

Integrating the one dimensional (transverse) convection- diffusion mass balance from the membrane surface out to a finite mass boundary (film) layer thickness, δ, yields the relationship between concentration polarization and permeate flux. The result is,

$$CP = \frac{C_m - C_p}{C_b - C_p} = \exp\left(\frac{J_v \delta}{D}\right) \qquad 2.5$$

or

$$CP = \frac{C_m - C_p}{C_b - C_p} = \exp\left(\frac{J_v}{k}\right) \qquad 2.6$$

where C_m is concentration at the membrane surface, or channel wall, for the rejected salt, C_b and J_v are the bulk solute concentration and the permeate water flux through the membrane respectively, and (D/δ) = mass transfer coefficient(k)(Suhan Kim, Eric M.V. Hoek, 2005).

It provides an interactive graphical environment and a customizable set of block libraries that let you design, simulate, implement, and test a variety of time-varying systems, including communications, controls, signal processing, video processing, and image processing.

Calculate J_v from:

$$J_v = \frac{Q}{A} \tag{2.7}$$

Where

$J_v =$ volumetric flux (cm^3 cm^{-2} s^{-1})

Q $=$ Feed flow rate (cm^3/s)

A $=$ Area of membrane (cm^2)

Chapter Three

Experimental Work

3.1 purpose of the present work:

The aim of the present work is to study the effect of different operating parameters on the reverse osmosis system which is feed flow rate, feed concentration, and feed temperature on permeate concentration and mass transfer coefficient for the brackish water system. These parameters were; feed concentration (100,200,300,400and500ppm), feed flow rate (0.666, 1, 1.333 and 1.666-2 cm^3/s), and feed temperature (15,25 and 35 °C), salts (NaCl, $CaCl_2$ and KCl).

3.2 Equipment and apparatus:

A schematic diagram of the equipment is presented in fig. (3.1), which consists of the following part:

1. Storage tank (capacity 20 Liter).
2. High Pressure Pump used for pumping the water under the high pressure to a semi-permeable membrane.

The specifications of the pump are as follow:

- Santoprene and polypropylene materials
- High safety (driver 24 VDC)
- Soft mounting for less noise and vibration
- Maximum pressure : 80 psi
- Power : 220 – 240 V
- Current : 1.2 A

3. Rotameter:

Calibrated rotameters used to measure the volumetric flow rate of permeate and concentrated water.

4. Pre-Filtration and Purification:

Three filters of cartridge-polypropylene type, 5 microns opening, (250 psi) pressure limits, 50 °C temperature limits, are used to filtrate the thick and solid impurities from water before the feed water is fed through the pump pressure to the second stage, to rid of organic chemicals, chlorine, to remove the unpleasant taste and odors from water.

5. Heater:

In order to maintain the temperature at a certain value, two submersible electrical coils (220 Volt, 1000 Watt) and thermostat of range from 0 to 80 ^{o}C were used as a heating media.

6. Total Dissolved Solid (TDS) Meter:

A digital TDS hand – held meter is used to measure the concentration of the brine feed solution (Oakton instruments).

The specifications of the TDS meter are as follow:

Type = Waterproof TDSTestr High+
Range = 2,000 – 20,000 ppm
Operating Temperature = 0 – 50 ^{o}C
Accuracy = ± 3%
Power supply = 6 Volt

7. Activated carbon (AC):

Activated carbon is similar to ion exchange resin in density and porosity. It adsorbs many dissolved organics and eliminates chlorine or other halogens in water. It does not remove salts. AC filters are one of the only low-cost methods available to remove low-molecular weight (<100 MW) organics and chlorine. AC filters may become a breeding site for bacteria and pyrogenic materials. The carbon must be sanitized or changed periodically to avoid bacterial growth, and when all adsorption sites are used it must be reactivated by a controlled heat process. This is not easily reactivated in the field. The suspended solids accumulated in the bed from most water sources require frequent backwashing of the filter unless installed after reverse osmosis or ultrafiltration.

7. Separation device:

Table 3-1 Separation device for Reverse Osmosis Membrane.

Element Configuration	Spiral Wound
Membrane Type	Polymer
Type No.	Ro-130 JB
Diameter	9 cm
Length	15 cm

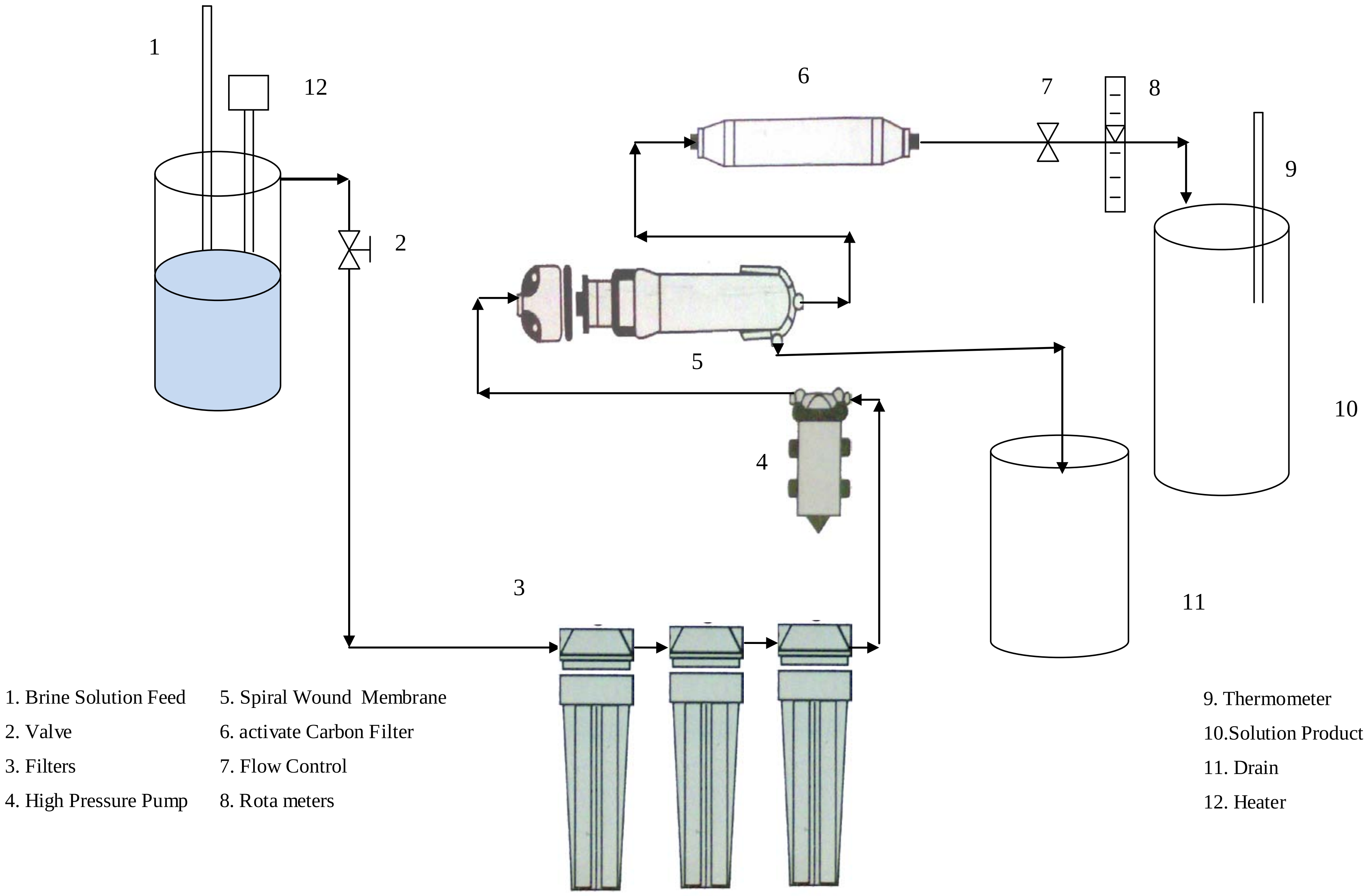


Figure 3.1 Schematic Diagram of Reverse Osmosis System.

3.3 Description of Water Treatment Process:

Brine feed solutions was prepared in the glass vessels by dissolving the solid salts in tap water, then the solution are fed to the reverse osmosis system by gravity of 9 m height through a storage tank (20 liter and 88.2 Kpa); this gives uniform feed of flow to the system by regulated valve. The brain water passes through three filters (5 microns) by the head of the building to remove macromolecules, colloids and suspended solid.

Then the filtered brine water fed to reverse osmosis system (membrane support) by means of a high pressure pump then the permeate water go to the activated carbon filter then go out to another tank as treated water.

3.4 Experimental Procedure

Measurement for each run:

1. Measuring the total dissolved solid for input tap water.
2. Adding salts to the tap water (NaCl, KCl and $CaCl_2$) individually in deferent concentrations (100, 200, 300, 400 and 500ppm) to get brine feed.
3. Measuring the total dissolved solid for brine feed.
4. Measuring the total dissolved solid for feed of permeate water.
5. Repeating step 3 and 4 for deferent flow rates (0.6, 1, 1.3, 1.6 and 2 cm^3/s) for three deferent temperatures 15, 25, 35 ºC (where the maximum designed working temperature for system is 45 ºC).

3.5 Simulation procedure:

A full simulation model was designed by sing Matlab Simulink® version 7.6.0.324 (R2008a), this model depends on the results obtained from the experimental work, and by using the relationships between feed data and resulted data we designed this simulation.

Chapter Four

Results and Discussion

4.1 Introduction:

The present work of reverse osmosis system was carried out to determine the mass transfer coefficient through membrane for various operating conditions of feed flow rate and temperature. This chapter will analyze and discusses the parameters which affect the efficiency of the reverse osmosis (RO) unit.

The membrane separation system utilized in this work was polymer membranes constructed as a spiral - wound module (which is considered as a commercial membranes).

4.2 Effect of Feed Flow Rate:

4.2.1 Effect of Feed Flow Rate on Permeates Concentration:

The decrease in permeate concentration can be obtained at a higher flow rate, as shown in figures (4-1 to 4-3) for sodium chlorine, calcium chlorine and potassium chlorine respectively.

From these figures, was noted that increasing the feed flow rate prevents the concentration buildup in the solution at the vicinity of the membrane surface, and results in increasing of driving force ($\Delta\pi$). The greater shear generated at the surface due to a higher turbulence in membrane enhanced the rate of back-transport of polarized solute into the bulk of the solution, this could be a major reason for the decrease of permeate concentration.

The increase in the feed flow rate reduces concentration polarization value due to increase in turbulence near the membrane resulting in decrease in the boundary layer thickness and solute concentration.

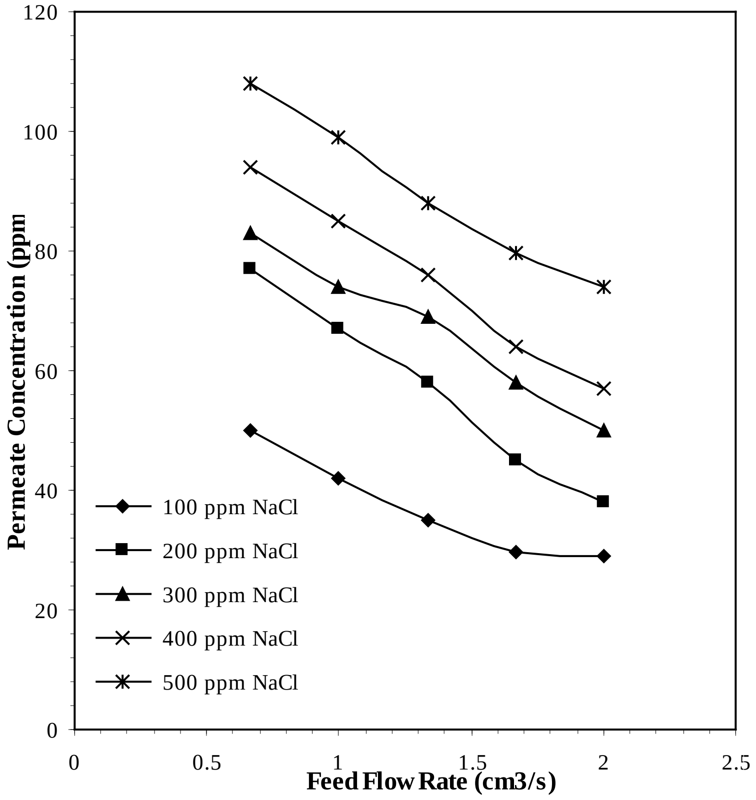


Fig. 4-1 Effect of Feed Flow Rate on Permeate Concentration NaCl

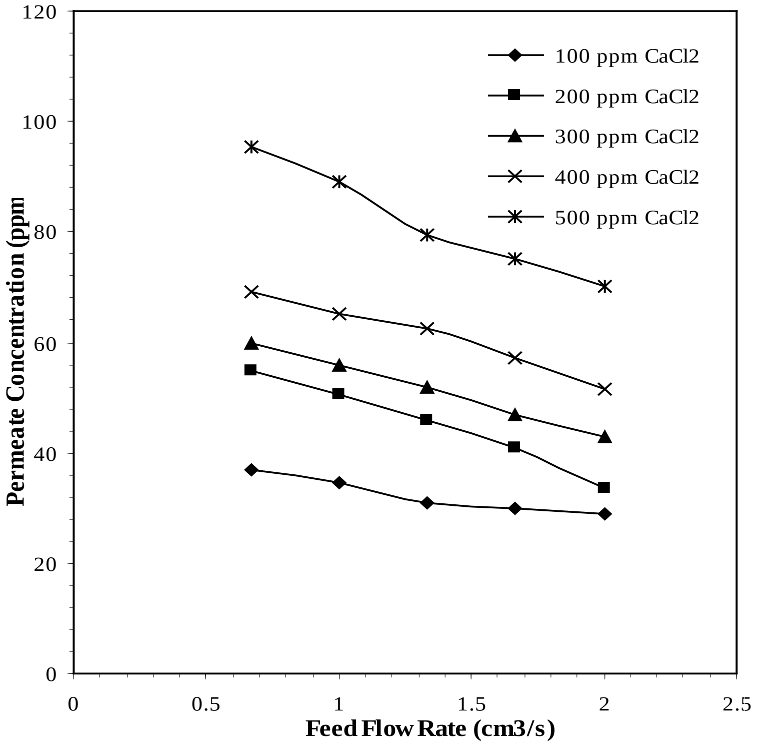


Fig. 4-2 Effect of Feed Flow Rate on Permeate Concentration $CaCl_2$

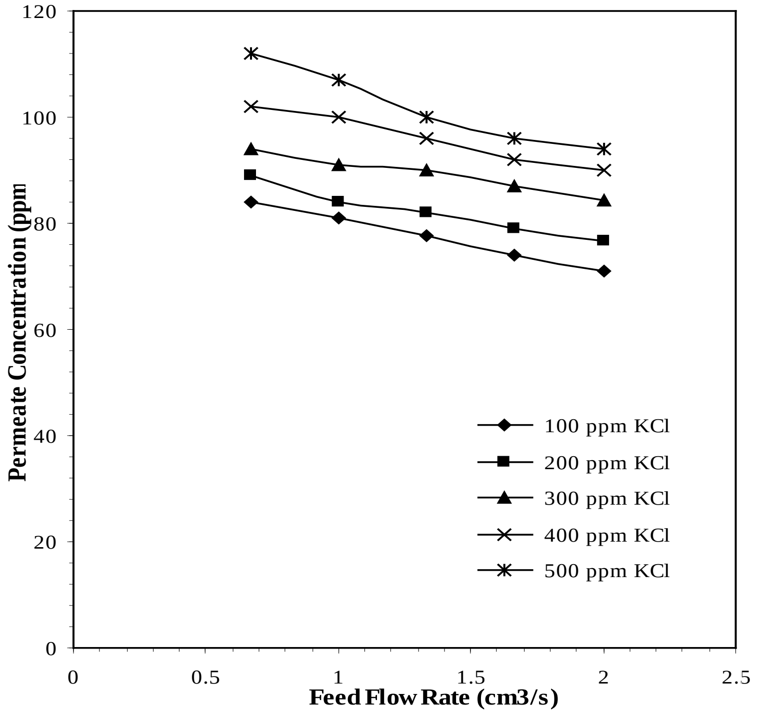


Fig. 4-3 Effect of Feed Flow Rate on Permeate Concentration KCl

4.2.2 Effect of Feed Flow Rate on Mass Transfer Coefficient:

Mass transfer coefficient increases with the increase in feed Flow Rate as shown in figures. (4-4 to 4-6) for Sodium chloride, calcium chloride and potassium chloride respectively.

The figures show that increasing in the feed Flow Rate leads to an increase in the driving pressure for fluid flow and increase in turbulent flow respectively. This would prevent or decrease the concentration buildup in the solution at the surface membrane. In other words, the thickness of the concentration boundary solution will be less while the diffusivity of solute (D_{AB}) stays constant. Therefore, the mass transfer coefficient increases steadily with the increase in the Feed Flow Rate.

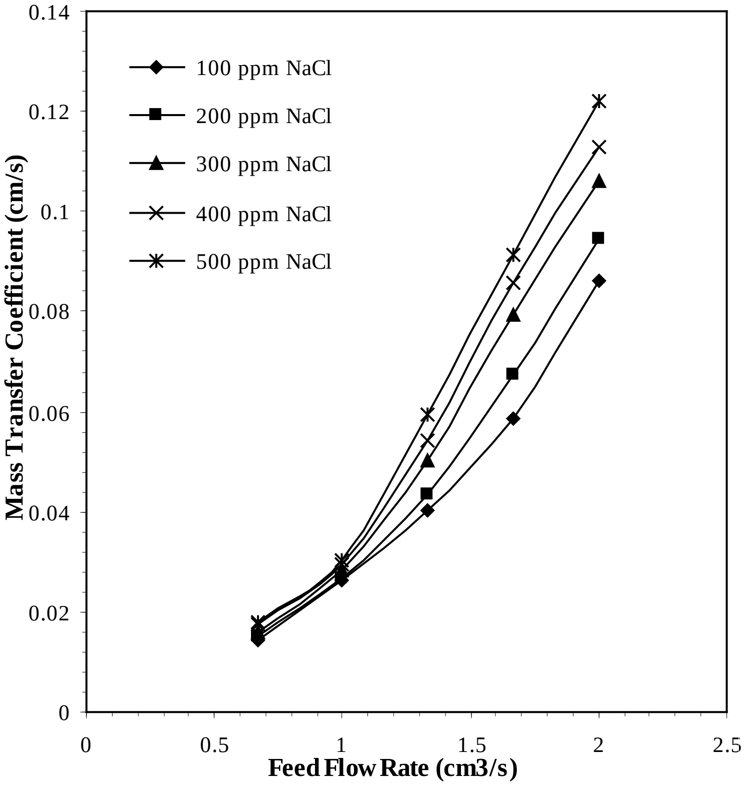


Fig. 4-4 Effect of feed Flow Rate on Mass Transfer Coefficient NaCl

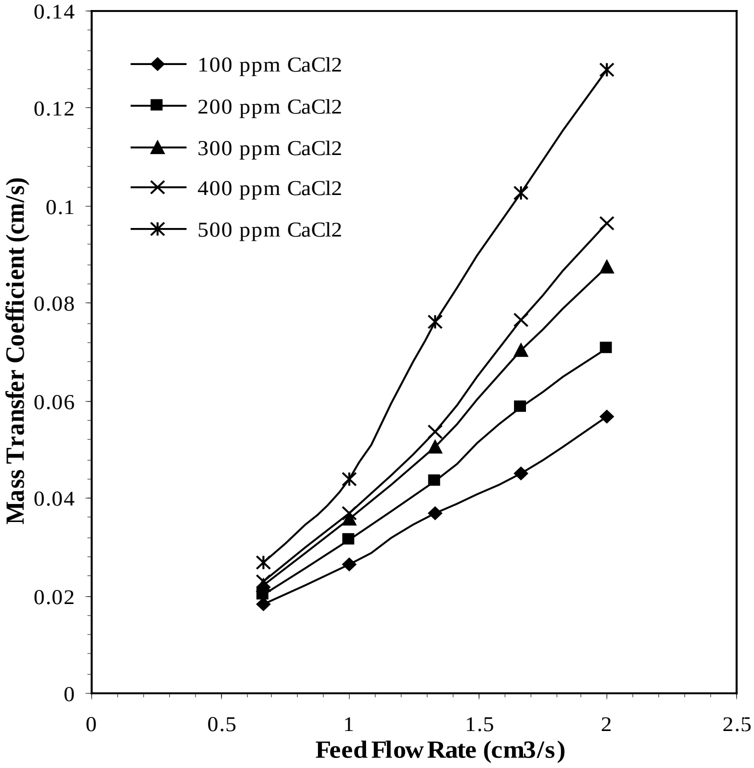


Fig. 4-5 Effect of feed Flow Rate on Mass Transfer Coefficient $CaCl_2$

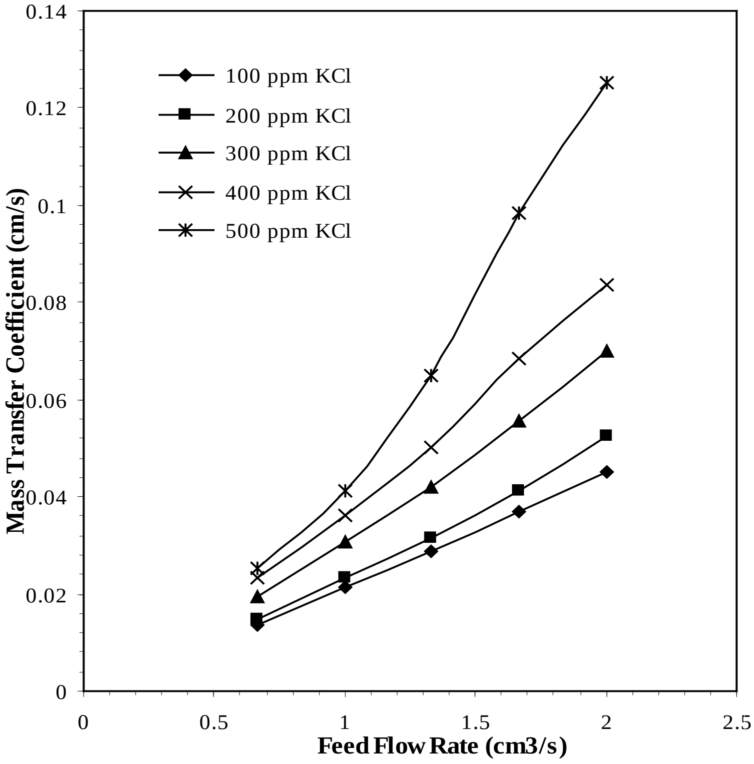


Fig. 4-6 Effect of feed Flow Rate on Mass Transfer Coefficient KCl

4.3 Effect of Feed Temperature:

4.3.1 Effect of Feed Temperature on permeates Concentration:

It noticed that the rate of water permeation through the membrane increase as the feed water temperature increases since the viscosity of the solution is reduced and higher diffusion rate of water through the membrane is obtained. Increasing feed water temperature will yield lower salt rejection or higher salt passage due to higher diffusion rate for salt through the membrane.

The effect of temperature on membrane performance is the most important parameter. The experiment data shows that when temperature of feed water is increased from 15 to 35 ºC, the permeate salinity ppm TDS increases from 50 to 100 ppm TDS for (NaCl).

Permeate and salt passage increase with increasing the feed water temperature as show in figures (4-7to4-9) for sodium chloride, calcium chloride and potassium chloride respectively.

The rate of water permeation through the membrane increase as the feed water temperature increases since the viscosity of the solution is reduced and higher diffusion rate of water through the membrane is obtained (Sourirajan, 1979). Increasing feed water temperature will yield lower salt rejection or higher salt passage due to higher diffusion rate for salt through the membrane.

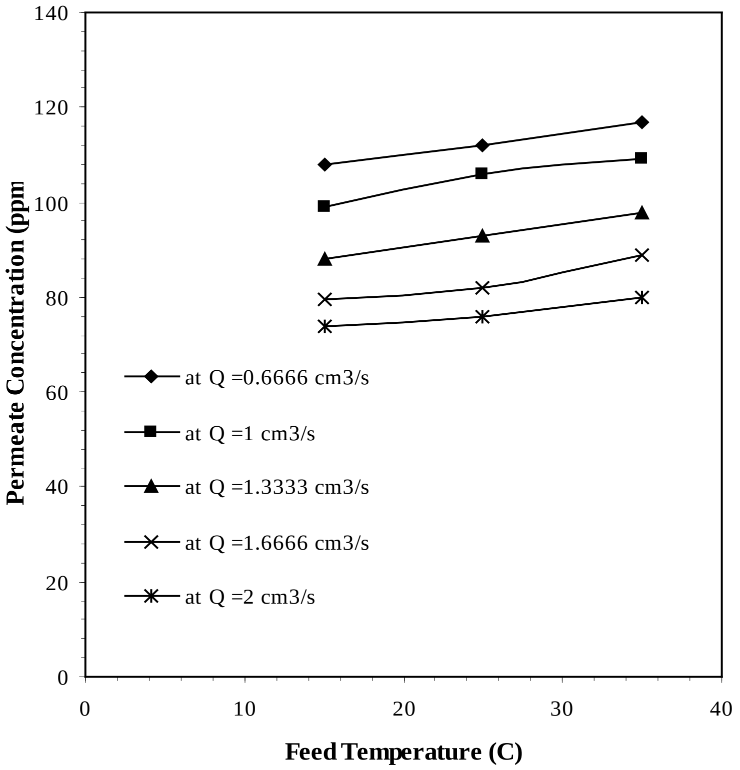


Fig. 4-7 Effect of Feed Temperature on permeates Concentration NaCl

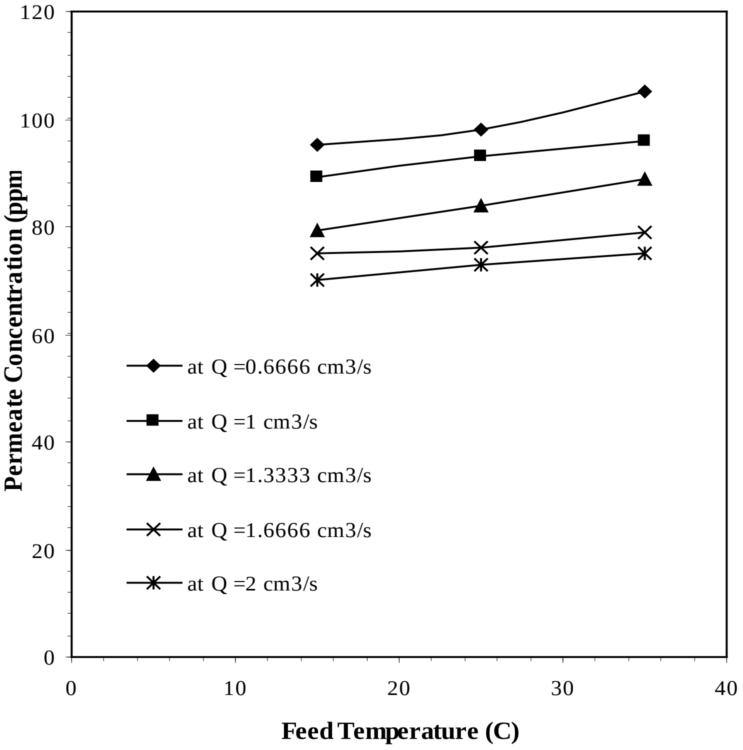


Fig. 4-8 Effect of Feed Temperature on permeates Concentration $CaCl_2$

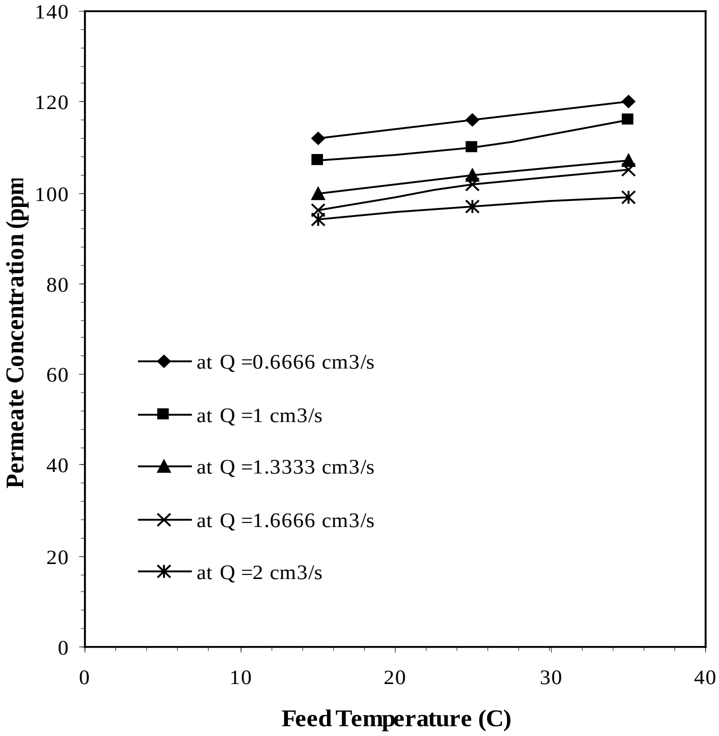


Fig. 4-9 Effect of Feed Temperature on permeates Concentration KCl

4.3.2 Effect of Feed Temperature on Mass Transfer Coefficient:

The rate of water permeation through the membrane increase as the feed water temperature increases since the viscosity of the solution is reduced and higher diffusion rate of water through the membrane is obtained.

The experimental results show that the spiral wound polymer membrane is very sensitive to the feed temperature. The membrane productivity is very sensitive to changes in feed water temperature. As water temperature increases, water flux increases almost linearly, due primarily to the higher diffusion rate of water through the membrane. Therefore, the mass transfer coefficient increases with the increase in feed temperature as shows in figures (4-10 to 4-12) for sodium chloride, calcium chloride and potassium chloride respectively.

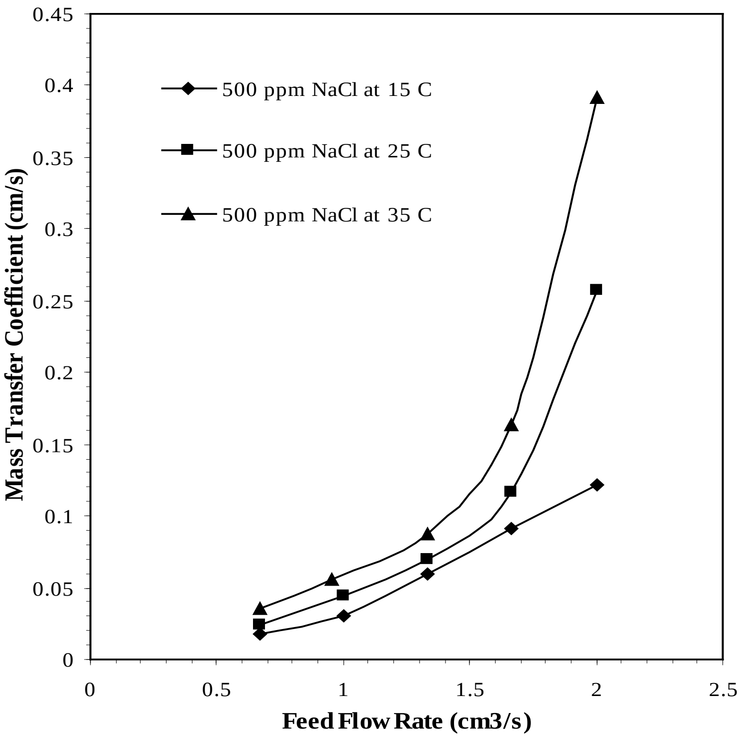


Fig. 4-10 Effect of feed Temperature on Mass Transfer Coefficient NaCl

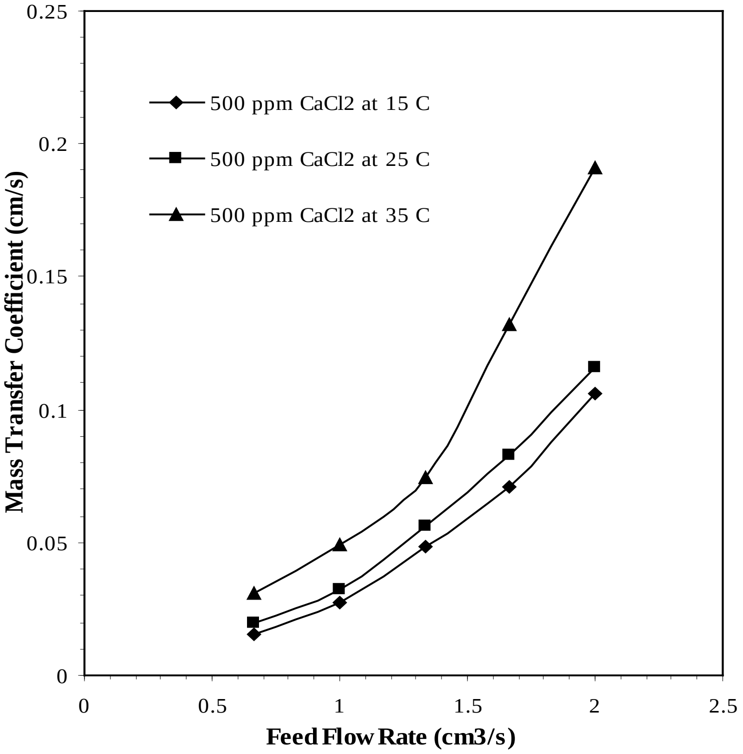


Fig. 4-11 Effect of feed Temperature on Mass Transfer Coefficient $CaCl_2$

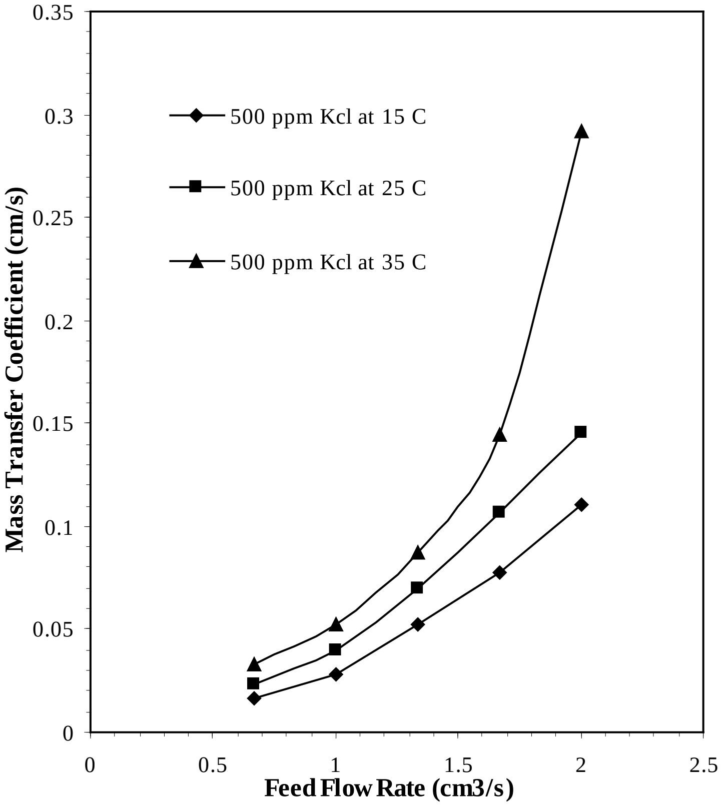


Fig. 4-12 Effect of feed Temperature on Mass Transfer Coefficient KCl

4.4 Effect of Different Salt Feed Concentration on Mass Transfer Coefficient and Permeate Concentration:

Figures (4-13 and 4-14) illustrate the effect of feed solution concentration on mass transfer coefficient and permeates concentration at different types of solution ($NaCl$,$CaCl_2$ and KCl).The selectivity of reverse osmosis is acknowledged not to be based on a sieving mechanism; the next most intuitive parameter to be considered for discussion is the size of the ions. It was found that the permeation of salts with a common anion is dictated by cation intrinsic properties such as size and charge.

$NaCl<KCl<CaCl_2$

From the results it was found that Sodium chloride, NaCl has high mass transfer coefficient and high permeate concentration because of it high osmotic pressure (driving force) than other materials studied.

Increasing the osmotic pressure of the draw solution will increase the driving force ($\Delta\pi$) for water flux. At the same concentration of feed solutions, osmotic pressure depends on the molecular weight of solute and has an additive effect on osmotic pressure.

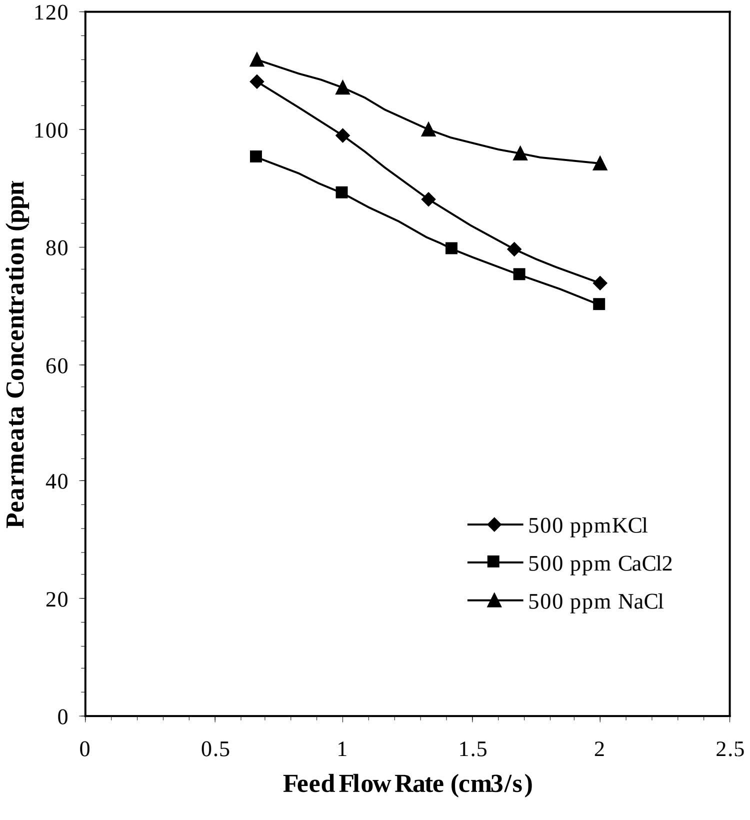


Fig. 4-13 Effect of Feed Concentration on permeates Concentration for Different Ions.

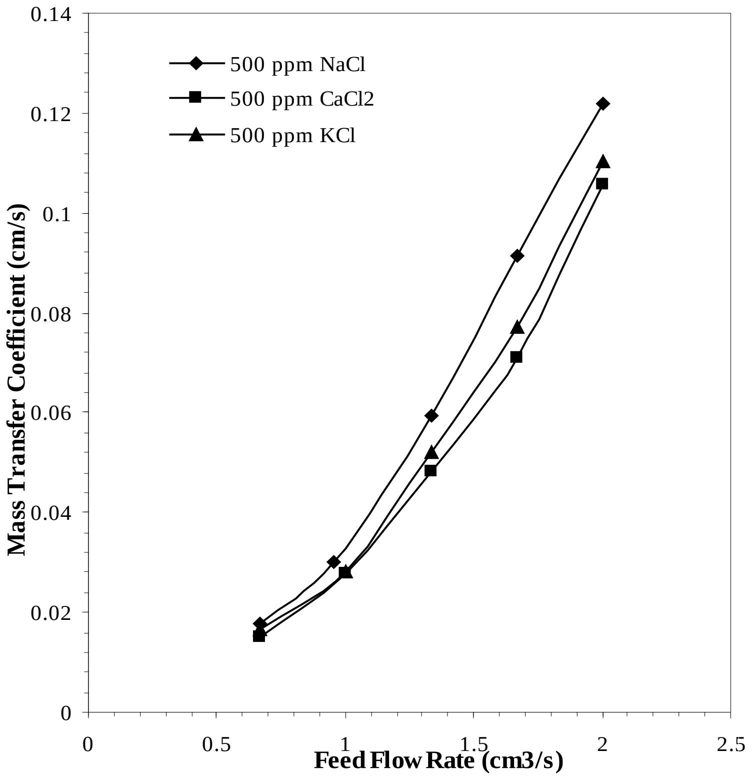


Fig. 4-14 Effect of Feed Concentration on Mass Transfer Coefficient for Different Ions.

4.5 Results of Simulation:

All the above discussed results are obtained also from the designed simulation model and this sample of obtained results: effects of feed flow rate on permeate concentration, by using least square method.

Least square method calculates the statistics for a line to calculate a straight line that best fits the data, and then returns an array that describes the line. It also possible to combine least square method with other functions to calculate the statistics for other types of models that are linear in the unknown parameters, including polynomial, logarithmic, exponential, and power series. Because this function returns an array of values, it must be entered as an array formula for example (e.g. NaCl) table (4-1).

Table 4-1 Simulation Equation:

For 100 ppm NaCl	y = 10.413x2 - 44.026x + 75.036
For 200 ppm NaCl	y = 1.2216x2 - 33.227x + 98.878
For 300 ppm NaCl	y = -2.5064x2 - 17.946x + 95.72
For 400 ppm NaCl	y = 0.643x2 - 30.214x + 114.2
For 500 ppm NaCl	y = 5.8491x2 - 41.846x + 133.82

The above equations were used in building simulation models for all salts used in this work (NaCl, $CaCl_2$ and KCl), it was builded two models for each salt, one for the effect of feed flow rate on permeat concentration and the other for the effect of temperature on permeat concentration fig.(4-16 to 4-22).

The results obtained from simulation models mentioned with comparison with expermental results in table (B-1 to B-18) and fig.(4-15) showes an example for NaCl at feed flow rate effect on mass transfer coefficient.

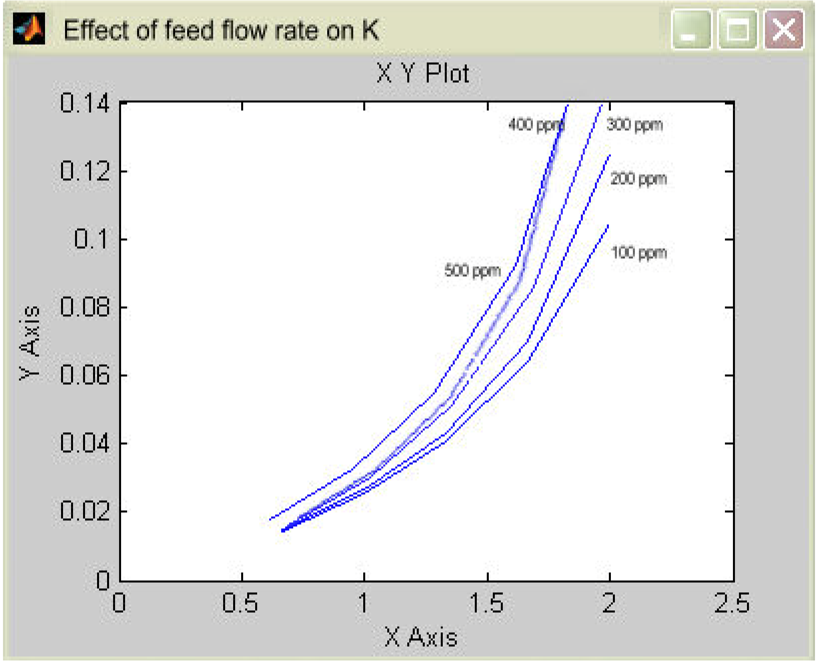


Fig. 4-15 Effect of Feed Flow rate on Mass Transfer Coefficient NaCl

Chapter five

Conclusions and Recommendations for Further Works

5.1 Conclusions:

1. The increase in the feed flow rate reduces concentration polarization value and decreases the permeate concentration.
2. The experiment data shows that when temperature of feed water is increased from 15 to 35 ºC, the permeate salinity ppm TDS increases from 70 to 115 ppm TDS for (NaCl).
3. Mass transfer coefficient increase with the increase in feed Flow Rate.
4. The mass transfer coefficient increase with the increase in feed temperature. As water temperature increases, water flux increases almost linearly, due primarily to the higher diffusion rate of water through the membrane.
5. For different salt feed the selectivity of reverse osmosis is acknowledged not to be based on a sieving mechanism. It was founded that the permeation of salts with a common anion is dictated by cation intrinsic properties such as size and charge.

$$NaCl<KCl<CaCl_2$$

6. Thermal expansion occurs in pore size of a polymeric membrane when the feed temperature increases that lead to an increase in permeate concentration and mass transfer coefficient.

5.2 Recommendations for Further Work

1. Using another types of membranes (e.g. cellulose triacetate CTA) to investigate their behavior in the same module type.
2. Using different types of salts can be used as osmotic agent to investigate their solutions behavior i.e. solute and water flux, in same module (like magnesium).
3. Studying the effect of fouling on the module performance and the rate of foulants growth along the time.
4. Studying the effect of the pretreatment system on reverse osmosis unit efficiency.

References

[1] Abd-Alsattar, K.," Modeling of Reverse Osmosis Systems ", Department of Chemical Engineering, University of Basrah (2003). http://www.desline.com/articoli/5202.pdf

[2] Al-Rubaie Ahmed Mohammed, "Evaluation of Reverse Osmosis unit Performance of Al-daura Refinery Water Treatment Plant", thesis University of Technology (2006).

[3] Al-Malack, M., and Anderson G.K., "Water Research", 31, 3064, No.12, (1997).

[4] Barbara Kneen, Ann Lemley and Linda Wagenet, "Reverse Osmosis treatment of Drinking Water", Cornell Cooperative Extension, College of Human Ecology (November 2005).

[5] Bhattacharyya and Williams, M "Membrane Handbook, Introduction", Van Nostrand Reinhold, New York, NY (1992).

[6] Bhattacharjee, C., Sarkar. P., Datta, S., Gupta, B.B., Bhattacharya, P.K., "Separation and Purification Technology", (2001).

[7] Buros, O.K. "The ABCs of Desalting, Second ed", International Desalination Association, Topsfield, Mass, (2000).

[8] Dale, M.C., and Okos, M.R., "Ind. Eng. Chem. Prod. Res. Dev.", 22, 452, (1983)

[9] Degremont Company, "Water Treatment Handbook", 5th Ed. John Wiley and Sons, New York, (1979)

[10] Dharmesh, S. Bhanushali, "Solvent- Resistant Nanofiltaration Membranes: Separation Studies and Mideling", (2002)

[11] Dow Company, "Dow Chemical Corporation Filmtec Membrane Elements Guide", (2005).

[12] El-Dessouky, H.T., Ettouney, H.M., and Al-Roumi, Y., Chemical Engineering Journal 73 (1999) 173.

[13] Eric,M.V.Hoek,Albert,S.Kim,and Elimelech, Environmental Engineering Science 19(2002) .

[14] George, S.,"Basic water Treatment", Britain, (1997).

[15] Goosen, M. F. A., Sablani, S. S., Al-Hinai, H., Al Obeidani, S., Al-Belushi, R., and Jackson, D., "Separation Science and Technology", 39, 1, (2004).

[16] Hussein, M., and James, M.D., "Ind. Eng. Chem. Res.", 28, 6, (1989).

[17] Hauge, L.J. "Desalination" 102 (1995) 219.

[18] Ibrahim Khaled El-Sayed Gabr, "Experimental and Numerical Optimization of Reverse Osmosis Desalination Plant", Mansoura University (2007)

[19] Ibrahim S. Al-Mutaz1, Mohammad A. Al-Ghunaimi, " Performance of Reverse Osmosis Units at High Temperatures", Chemical Eng. Dept., College of Engineering, King Saud University, October (2001).

[20] Jarusutthirak, C., Amy, G., and Croue, J-P., "Desalination", 145, 247, (2002).

[21] Jeffrey R. McCutcheon and Menachem Elimelech, American Institute of Chemical Engineers, 53, July (2007)

[22] John, A. B., "Feasibility of Cross-Flow Microfiltration for Combined Sewer Overflows", M.Sc. thesis, Pittsburgh University, (2003)

[23] Judson, K.C., "Separation Processes", McGraw-Hill, New York, (1971).

[24] Khan, A.H.," Desalination Processes and Multistage Flash Distillation Practice ", Book (1986).

[25] Khawla, A.M., "Modeling, Simulation, and Optimization of Large-

Scale Commercial Desalination Plants", Ph.D. thesis, State University, (1998).

[26] Kimura, S. and Sourirajan, S. (1967), Lache Journal, Vol.13, No.3, p (497-503).

[27] Kirk and Othmer, "Encyclopedia of Chemical Technology", 15, 92, (1981).

[28] Krishna, H.J, "Introduction to Desalination Technologies", (2003). http://www.twdb.state.tx.us/desalination

[29] Labbez, C., Fievet, P., Szymczyk, A., Vidonne, A., Foissy, A., and Pagetti, J., "Separation and Purification Technology", 30, 47, (2003).

[30] Lacey, R.E., and Leob, S., "Industrial Processing with Membrane", Wiley and Sons, (1972).

[31] Mario, "An Introduction to Desalination", California University (2005) http://www.wescinc.com.

[32] Mark Wilf et al, "Effect of Feed Temperature and Flow Rate on Permeate Flux in Spiral Wound Reverse Osmosis Systems", "Challenges and Opportunities for Engineering Education, Research and Development", Miami, Florida, USA, (2004)

[33] Mohammed Al-hewayzi, "Designing Study and Economic Comparison for both RO and Electrical Analysis Desalination Processes" thesis University of Technology, (1986).

[34] Mousa, S.M., "Brackish Water Desalination an Alternative for Water Supply Enhancement in Jordan", Department of Mechanical and Industrial Engineering, Applied Science University (1999) http://www.desaline.com/articoli/3724.pdf

[35] Merten, U., "Desalination by Reverse Osmosis", M.I.T. Press, Cambridge, Mass., (1966).

[36] Mulder, M.,“Basic Principles of Membrane Technology”, 2nd ed., Kluwer Academic Publishers, Dordrecht, The Netherlands, (1997).

[37] McCutcheon, J.R., R.L. McGinnis, M. Elimelech, and Desalination by a novel ammonia–carbon dioxide forward osmosis process: influence of draw and feed solution concentrations on process performance, J. Member. Sci. 278 (2006) 114–123

[38] Ohya, H. and Taniguchi, Y. (1975),” Desalination “, Book.

[39] Oakland, Pacific Institute, California, “Desalination, With a Grain of Salt”, August (2006).
www.pacinst.org/reports/desalination.

[40] Patrick, M., “Membrane Seperation Processes”, Elsevier, Amstrdam, (1976).

[41] Peter, S., “Desalination”, 56, 17 (1985).

[42] Pique, G.G. Water Conditioning and Purification, July (2000).

[43] Planners, U.S., “Desalting Handbook for Planners” Department of the Interior, Bureau of Reclamation ,3rd Edition, (2003)

[44] Potts, D.E., hlert, R.C. and Wang, S.S., “A Critical Review of Fouling of Reverse Osmosis Membranes”, Desalination, 36, pp. 235-264 (1981).

[45] Sablani, S.S., Goosen, M.F.A. Al-Belushi, R. and Wilf, M, “Concentration polarization in ultrafiltration and reverse osmosis”, Desalination 141 (2001) 269–289

[46] Semiat, R. Water International 25, 54 (2000).

[47] Shaffer, L.H., and Mintz, M.S., “Principles of Desalination”, Academic Press, Inc., New York, (1980)

[48] Srikanth, G., “Membrane Separation Processes– Technology and Business Opportunities”, March (2005).

<http://www.tifac.org.in/news/memb.htm>.

[49] Sourirajan, S., “Pure and Applied Chemistry”, 50, 593, (1979).

[50] Spiegler, K.S. and Y.M. El-Sayed, “A Desalination Primer”, Balaban Desalination Publications, Santa Maria Imbaro, Italy (1994).

[51] Syed Ali, Paul Boblak, Efrem Capili and Stanislav Milidovich, “Membrane Separation and Ultrafiltration” CHE-396 Senior Design (2000).

[52] Suhan Kim, Eric M.V. Hoek, Desalination 186 (2005) 111–128.

[53] Taha, Y.K., “Performance of Reverse Osmosis”, M.Sc. thesis, Baghdad University, (2000).

[54] Taniguchi, Y. (1978), Desalination Journal p. (71-88).

[55] Tweddle, T.A. and Thayer, W.L. (1981), Desalination Journal, P.(71-88).

[56] Tzahi Y. Cath, Amy E. Childress, Menachem Elimelech, Review Forward osmosis: Principles, applications, and recent developments May (2006).

[57] Vassilis, G., “Desalination”, 68, 77, (1988).

[58] Van de Lisdonk, C.A.C., Rietman, B.M., Sterk, G.R., Schippers, J.C., “Proceeding AWWA Memberane Technology Conference”, San Antonio (2001).

[59] Vineet K.Gupta, Sun-Tak Hwang, William B. Krantz,Alan R. Greenberg, Desalination 208(2007).

[60] Wang Xuesong, J. of chem. Ind. and Eng, 6 (1978) 15-27

[61] Wang Xuesong, Desalination, 62 (1987) 211-220.

[62] Williams E. Michael, “A Brief Review of Reverse Osmosis Membrane Technology”, Williams Engineering Services Company (2003) http://www.wescinc.com.

[63] Xiaohua, Z. and Menachem, E.,” Colloidal Fouling of Reverse Osmosis Membranes “, Department of Civil and Environmental Engineering, University of California (1997).

Appendix A

Experimental Data

Table (A-1) Variation of Feed Flow Rate with Permeate Concentration at 100 ppm NaCl, 15 ^{0}C

NaCl 100 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	50	1200
1	42	1141
1.3333	35	1102
1.6666	29.8	1071
2	29	1000

Table (A-2) Variation of Feed Flow Rate with Permeate Concentration at 200 ppm NaCl, 15 ^{0}C

NaCl 200 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	77	1315
1	67	1259
1.3333	58	1210
1.6666	45.1	1178
2	38	1104

Table (A-3) Variation of Feed Flow Rate with Permeate Concentration at 300 ppm NaCl, 15 ^{0}C

NaCl 300 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	83	1342
1	74	1259
1.3333	69	1218
1.6666	57.9	1178
2	50	1120

Table (A-4) Variation of Feed Flow Rate with Permeate Concentration at 400 ppm NaCl, 15 ^{0}C

NaCl 400 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	94	1321
1	85	1298
1.3333	76	1222
1.6666	64	1185
2	57	1142

Table (A-5) Variation of Feed Flow Rate with Permeate Concentration at 500 ppm NaCl, 15 ^{0}C

NaCl 500 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	108	1325
1	99	1304
1.3333	88	1217
1.6666	79.7	1190
2	73.9	1144

Table (A-6) Variation of Feed Flow Rate with Permeate Concentration at 100 ppm KCl, 15 ^{0}C

KCl 100 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	84	1310
1	81	1300
1.3333	77.8	1297
1.6666	74	1293
2	71	1290

Table (A-7) Variation of Feed Flow Rate with Permeate Concentration at 200 ppm KCl, 15 ^{0}C

KCl 200 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	89	1291
1	84	1280
1.3333	82	1275
1.6666	79	1262
2	76.8	1242

Table (A-8) Variation of Feed Flow Rate with Permeate Concentration at 300 ppm KCl, 15 ^{0}C

KCl 300 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	94	1397
1	91	1381
1.3333	90	1372
1.6666	87	1359
2	84.2	1346

Table (A-9) Variation of Feed Flow Rate with Permeate Concentration at 400 ppm KCl, 15 ^{0}C

KCl 400 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	102	1480
1	100	1472
1.3333	96	1465
1.6666	92	1447
2	90	1430

Table (A-10) Variation of Feed Flow Rate with Permeate Concentration at 500 ppm KCl, 15 ^{0}C

KCl 500 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	112	1675
1	107	1619
1.3333	100	1489
1.6666	96	1484
2	94	1440

Table (A-11) Variation of Feed Flow Rate with Permeate Concentration at 100 ppm $CaCl_2$, 15 ^0C

$CaCl_2$100 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.623	37	1010
1	34.5	1030
1.3333	31	1010
1.6666	30	1015
2	29	1000

Table (A-12) Variation of Feed Flow Rate with Permeate Concentration at 200 ppm $CaCl_2$, 15 ^0C

$CaCl_2$200 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	55	1170
1	50.6	1162
1.3333	45.9	1157
1.753	40.8	1143
2	33.6	1130

Table (A-13) Variation of Feed Flow Rate with Permeate Concentration at 300 ppm $CaCl_2$, 15 ^0C

$CaCl_2$ 300 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	60	1220
1	56	1200
1.3333	52	1189
1.6666	47	1172
2	43	1169

Table (A-14) Variation of Feed Flow Rate with Permeate Concentration at 400 ppm $CaCl_2$, 15 ^0C

$CaCl_2$ 400 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	69.1	1327
1	65	1310
1.3333	62.4	1305
1.6666	57.3	1296
2	51.4	1288

Table (A-15) Variation of Feed Flow Rate with Permeate Concentration at 500 ppm $CaCl_2$, 15 ^{0}C

$CaCl_2$ 500 ppm		
Feed Flow Rate (cm3/s)	Permeate Concentration (ppm)	Drain Concentration (ppm)
0.6666	95.3	1405
1	89.1	1391
1.3333	79.4	1312
1.6666	75	1267
2	70	1238

Table (A-16) Variation of Feed Flow Rate with Mass Transfer Coefficient at 15 ^{0}C NaCl

NaCl					
Feed Flow Rate	Mass Trans. Coeff. (cm/s) 100 ppm	Mass Trans. Coeff. (cm/s) 200 ppm	Mass Trans. Coeff. (cm/s) 300 ppm	Mass Trans. Coeff. (cm/s) 400 ppm	Mass Trans. Coeff. (cm/s) 500 ppm
0.6666	0.01442	0.01499	0.01607	0.01774	0.0178
1	0.0262	0.0267	0.02822	0.02941	0.03016
1.3333	0.0404	0.04365	0.05032	0.05407	0.05955
1.6666	0.05875	0.06746	0.07921	0.0859	0.0913
2	0.0863	0.0946	0.1061	0.1127	0.122

Table (A-17) Variation of Feed Flow Rate with Mass Transfer Coefficient at 15 ^{0}C KCl

KCl					
Feed Flow Rate	Mass Trans. Coeff. (cm/s) 100 ppm	Mass Trans. Coeff. (cm/s) 200 ppm	Mass Trans. Coeff. (cm/s) 300 ppm	Mass Trans. Coeff. (cm/s) 400 ppm	Mass Trans. Coeff. (cm/s) 500 ppm
0.6666	0.0138	0.01486	0.01926	0.02336	0.01647
1	0.02129	0.02327	0.03059	0.03626	0.02829
1.3333	0.02885	0.03162	0.04207	0.05034	0.052
1.6666	0.03698	0.04131	0.05551	0.06839	0.0771
2	0.04513	0.05257	0.06992	0.0836	0.11029

Table (A-18) Variation of Feed Flow Rate with Mass Transfer Coefficient at 15 ^{0}C $CaCl_2$

$CaCl_2$					
Feed Flow Rate	Mass Trans. Coeff. (cm/s) 100 ppm	Mass Trans. Coeff. (cm/s) 200 ppm	Mass Trans. Coeff. (cm/s) 300 ppm	Mass Trans. Coeff. (cm/s) 400 ppm	Mass Trans. Coeff. (cm/s) 500 ppm
0.6666	0.01837	0.02009	0.02203	0.02294	0.0151
1	0.02628	0.03157	0.03567	0.03694	0.02763
1.3333	0.03692	0.04349	0.05054	0.0537	0.0482
1.6666	0.04511	0.0589	0.0702	0.0766	0.071
2	0.05668	0.0707	0.08754	0.0966	0.1057

Table (A-19) Variation of Feed Flow Rate with Mass Transfer Coefficient for 500 ppm $CaCl_2$, at (15, 25 and 35 ^0C)

$CaCl_2$ 500 ppm			
Feed Flow Rate	Mass Trans. Coeff. (cm/s) 15 ^0C	Mass Trans. Coeff. (cm/s) 25 ^0C	Mass Trans. Coeff. (cm/s) 35 ^0C
0.6666	0.0151	0.02	0.03084
1	0.02763	0.0325	0.04887
1.3333	0.0482	0.05604	0.07442
1.6666	0.071	0.08287	0.13235
2	0.1057	0.1157	0.19088

Table (A-20) Variation of Feed Flow Rate with Mass Transfer Coefficient for 500 ppm NaCl, at (15, 25 and 35 ^0C)

NaCl 500 ppm			
Feed Flow Rate	Mass Trans. Coeff. (cm/s) 15 ^0C	Mass Trans. Coeff. (cm/s) 25 ^0C	Mass Trans. Coeff. (cm/s) 35 ^0C
0.6666	0.0178	0.024	0.036
1	0.03016	0.044	0.056
1.3333	0.05955	0.07	0.088
1.6666	0.0913	0.117	0.163
2	0.122	0.25795	0.39141

Table (A-21) Variation of Feed Flow Rate with Mass Transfer Coefficient for 500 ppm KCl, at (15, 25 and 35 ^{0}C)

KCl 500 ppm			
Feed Flow Rate	Mass Trans. Coeff. (cm/s) 15 ^{0}C	Mass Trans. Coeff. (cm/s) 25 ^{0}C	Mass Trans. Coeff. (cm/s) 35 ^{0}C
0.6666	0.01647	0.02323	0.03337
1	0.02829	0.03944	0.05246
1.3333	0.052	0.06954	0.08773
1.6666	0.0771	0.10667	0.14444
2	0.11029	0.1457	0.29225

Appendix B:

Simulink Data

Table (B-1) Effect of Flow Rate on Mass Transfer Coefficient and Permeate Concentration for $CaCl_2$ at Different Temperate.

$CaCl_2$						
Feed Flow Rate (cm^3/s)	15 ^0C		25 ^0C		35 ^0C	
	Mass Transfer Coefficient (cm/s) experement	Mass Transfer Coefficient (cm/s) Simulink	Mass Transfer Coefficient (cm/s) experement	Mass Transfer Coefficient (cm/s) Simulink	Mass Transfer Coefficient (cm/s) experement	Mass Transfer Coefficient (cm/s) Simulink
0.0666	0.0151	0.01693	0.02	0.01789	0.03084	0.02989
1	0.02763	0.02961	0.0325	0.03274	0.04887	0.05335
1.333	0.0482	0.04716	0.05604	0.05351	0.07442	0.08569
1.666	0.071	0.07307	0.08287	0.08163	0.13235	0.1311
2	0.1057	0.1153	0.1157	0.1169	0.19088	0.1958

Table (B-2) Effect of Flow Rate on Mass Transfer Coefficient and Permeate Concentration for NaCl at Different Temperate

NaCl						
Feed Flow Rate (cm^3/s)	15 ^{0}C		25 ^{0}C		35 ^{0}C	
	Mass Transfer Coefficient (cm/s) experement	Mass Transfer Coefficient (cm/s) Simulink	Mass Transfer Coefficient (cm/s) experement	Mass Transfer Coefficient (cm/s) Simulink	Mass Transfer Coefficient (cm/s) experement	Mass Transfer Coefficient (cm/s) Simulink
0.0666	0.0178	0.01782	0.024	0.01854	0.036	0.01862
1	0.03016	0.03246	0.044	0.03383	0.056	0.03535
1.333	0.05955	0.05465	0.07	0.05842	0.088	0.06457
1.666	0.0913	0.09198	0.117	0.1066	0.163	0.1303
2	0.122	0.1673	0.25795	0.2535	0.39141	0.4289

Table (B-3) Effect of Flow Rate on Mass Transfer Coefficient and Permeate Concentration for KCl at Different Temperate.

KCl						
Feed Flow Rate (cm^3/s)	15 ^{0}C		25 ^{0}C		35 ^{0}C	
	Mass Transfer Coefficient (cm/s) experement	Mass Transfer Coefficient (cm/s) Simulink	Mass Transfer Coefficient (cm/s) experement	Mass Transfer Coefficient (cm/s) Simulink	Mass Transfer Coefficient (cm/s) experement	Mass Transfer Coefficient (cm/s) Simulink
0.0666	0.01647	0.01615	0.02323	0.0228	0.03337	0.0329
1	0.02829	0.0304	0.03944	0.04145	0.05246	0.05433
1.333	0.052	0.05098	0.06954	0.06747	0.08773	0.08555
1.666	0.0771	0.07855	0.10667	0.1031	0.14444	0.1423
2	0.11029	0.1096	0.1457	0.1497	0.29225	0.2989

Table (B-4) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for $CaCl_2$ at 15 ^{0}C.

Feed Concentration at (0.666 cm^3/s) for $CaCl_2$	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.01837	0.01752
200 ppm	0.02009	0.02012
300 ppm	0.02203	0.0215
400 ppm	0.02294	0.02285
500 ppm	0.0151	0.01693

Table (B-5) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for $CaCl_2$ at 15 ^{0}C

Feed Concentration at (1cm^3/s) for $CaCl_2$	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.02628	0.02614
200 ppm	0.03157	0.0314
300 ppm	0.03567	0.03536
400 ppm	0.03694	0.03634
500 ppm	0.02763	0.02961

Table (B-6) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for $CaCl_2$ at 15 ^{0}C.

Feed Concentration at (1.333cm^3/s) for $CaCl_2$	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.03692	0.03527
200 ppm	0.04349	0.04422
300 ppm	0.05054	0.05127
400 ppm	0.0537	0.05131
500 ppm	0.0482	0.04716

Table (B-7) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for $CaCl_2$ at 15 ^{0}C.

Feed Concentration at (1.666cm^3/s) for $CaCl_2$	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.04511	0.0455
200 ppm	0.0589	0.05951
300 ppm	0.0702	0.06901
400 ppm	0.0766	0.06791
500 ppm	0.071	0.07307

Table (B-8) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for $CaCl_2$ at 15 ^{0}C.

Feed Concentration at (2cm^3/s) for $CaCl_2$	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.05668	0.05764
200 ppm	0.0707	0.07878
300 ppm	0.08754	0.08806
400 ppm	0.0966	0.08627
500 ppm	0.1057	0.1153

Table (B-9) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for NaCl at 15 ^{0}C.

Feed Concentration at (0.666 cm^3/s) for NaCl	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.01442	0.01458
200 ppm	0.01499	0.0151
300 ppm	0.01607	0.01632
400 ppm	0.01774	0.01738
500 ppm	0.0178	0.01782

Table (B-10) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for NaCl at 15 ^{0}C.

Feed Concentration at (1 cm^3/s) for NaCl	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.0262	0.02548
200 ppm	0.0267	0.02647
300 ppm	0.02822	0.0301
400 ppm	0.02941	0.0311
500 ppm	0.03016	0.03246

Table (B-11) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for NaCl at 15 ^{0}C.

Feed Concentration at (1.333 cm^3/s) for NaCl	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.0404	0.04072
200 ppm	0.04365	0.04295
300 ppm	0.05032	0.05114
400 ppm	0.05407	0.0516
500 ppm	0.05955	0.05465

Table (B-12) Effect of Feed Flow Rate on Mass .Transfer Coefficient and Permeate Concentration for NaCl at 15 ^{0}C

Feed Concentration at (1.666 cm^3/s) for NaCl	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.05875	0.06392
200 ppm	0.06746	0.06997
300 ppm	0.07921	0.08589
400 ppm	0.0859	0.08618
500 ppm	0.0913	0.09198

Table (B-13) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for NaCl at 15 ^{0}C.

Feed Concentration at (2 cm^3/s) for NaCl	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.0863	0.1045
200 ppm	0.0946	0.1248
300 ppm	0.1061	0.1505
400 ppm	0.1127	0.1586
500 ppm	0.122	0.1673

Table (B-14) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for KCl at 15 ^{0}C.

Feed Concentration at (0.666 cm^3/s) for KCl	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.0138	0.01381
200 ppm	0.01486	0.01491
300 ppm	0.01926	0.0193
400 ppm	0.2336	0.02333
500 ppm	0.01647	0.01615

Table (B-15) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for KCl at 15 ^{0}C.

Feed Concentration at (1 cm^3/s) for KCl	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.02129	0.02128
200 ppm	0.02327	0.02301
300 ppm	0.03059	0.0303
400 ppm	0.03626	0.03621
500 ppm	0.02829	0.0304

Table (B-16) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for KCl at 15 ^{0}C.

Feed Concentration at (1.333 cm^3/s) for KCl	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.02885	0.02902
200 ppm	0.03162	0.03177
300 ppm	0.04207	0.04233
400 ppm	0.06839	0.05069
500 ppm	0.052	0.05098

Table (B-17) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for KCl at 15 ^{0}C.

Feed Concentration at (1.666 cm^3/s) for KCl	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.03698	0.03698
200 ppm	0.04131	0.0415
300 ppm	0.05551	0.05558
400 ppm	0.06839	0.06775
500 ppm	0.0771	0.07855

Table (B-18) Effect of Feed Flow Rate on Mass Transfer Coefficient and Permeate Concentration for KCl at 15 ^{0}C.

Feed Concentration at (2 cm^3/s) for KCl	Mass Transfer Coefficient experiment (cm/s)	Mass Transfer Coefficient simulink (cm/s)
100 ppm	0.04513	0.0451
200 ppm	0.05257	0.05262
300 ppm	0.06992	0.07032
400 ppm	0.0836	0.08899
500 ppm	0.11029	0.1096