ISTANBUL TECHNICAL UNIVERSITY ★ GRADUATE SCHOOL OF SCIENCE ENGINEERING AND TECHNOLOGY

TECHNICAL AND ECONOMIC FEASIBILITY OF THE CONCURRENT DESALINATION AND BORON REMOVAL PROCESS

M.Sc. THESIS

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Seda KAYACI, a M.Sc. student of İTU Graduate School of Science Engineering and Technology student ID 506161069, successfully defended the thesis entitled "TECHNICAL AND ECONOMIC FEASIBILITY OF THE CONCURRENT DESALINATION AND BORON REMOVAL PROCESS", which she prepared after fulfilling the requirements specified in the associated legislations, before the jury whose signatures are below.

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FOREWORD

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February 2019

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ABBREVIATIONS

AMF : Adsorption Membrane Filtration
BP : Booster pump
BWRO : Brackish Water Reverse Osmosis
CAPEX : Capital Cost
CDBR : Countercurrent Desalination and Boron Removal
CMCR : Countercurrent Membrane Cascade with Recovery
EERO : Energy Efficient Reverse Osmosis
ED : Electrodialysis
ERD : Energy Recovery Device
ERI : Isobaric Device
HID : Hybrid Membrane Inter-stage Design
HPP : High-pressure pump
ISD : Internally Staged Design
LPMS : Low-pressure Membrane Stage
MED : Multiple-effect Distillation
MF : Microfiltration
MP : Mixing point
MSF : Multi-stage Flash Distillation
NF : Nanofiltration
OPD : Osmotic Pressure Differential
OPEX : Operating Cost
O&M : Operation and Maintenance
PV : Pressure vessel
RO : Reverse Osmosis
SEC : Specific Energy Consumption
SP : Split Point
SSRO : Single Stage Reverse Osmosis
SWRO : Seawater Reverse Osmosis
TDS : Total dissolved solids
TOTEX : Total Cost

UF **:** Ultrafiltration

VC **:** Vapor Compression

WAVE **:** Water Application Value Engine

WHO **:** World Health Organization

SYMBOLS

A	**:** Annual cost
C	**:** Concentration
$\mathbf{C_{DC}}$	**:** Direct capital cost
Q	**:** Volumetric flow rate
Y	**:** Recovery
Δπ	**:** Osmotic Pressure Differential
σ	**:** Rejection
η	**:** Efficiency
S	**:** Bypass split fraction
$\dot{V}$	**:** Flow rate
ψ_{AF}	**:** RO plant availability parameter
t	**:** Annual lifespan
i	**:** Interest rate

Subscripts

b	**:** Boron
f	**:** Feed
s	**:** Salt
p	**:** Production
c	**:** Capacity

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TECHNICAL AND ECONOMIC FEASIBILITY OF THE CONCURRENT DESALINATION AND BORON REMOVAL PROCESS

SUMMARY

Water is a highly significant source of life. Although water covers a large portion of the earth, the available freshwater reserves are too limited. Due to factors such as ecological deterioration, increasing industrialization and pollution of clean water resources, the existing freshwater reserves have commenced shrinking. For that reason, many people confront difficulty in accessing clean water, and this situation threatens future life as well. Hence, this problem has led scientists to seek alternative options.

Because of the decrease in water supplies in the world, studies have been carried out on the treatment of salt water. The desalination process provides separating the undesired components from water and producing drinking water. Desalination methods are generally composed of two main groups: thermal evaporation and membrane separation processes. In the thermal evaporation method, salt water is boiled, and the produced clean water vapor is condensed to obtain pure water. These systems consist of two main technologies which are Multi-stage flash (MSF) and Multiple-effect distillation (MED). Gulf countries are leaders in the use of these technologies because of low energy costs where they locate. However, over time, the concern that fossil fuels will run out has directed these countries to other areas. Therefore, in recent years, Middle East countries have commenced employing hybrid systems which use both thermal and membrane processes. Reverse osmosis (RO), nanofiltration (NF) and electrodialysis (ED) are examples of membrane processes that use mechanical and chemical methods unitedly. One of the most widely used membrane technologies is the reverse osmosis process. When being applied above the osmotic pressure to the high-concentrated water solution, water molecules pass through a semi-permeable membrane. Thus, this process provides clean water. Since the amount of energy it uses is less than that of the thermal processes, it is one of the most preferred desalination methods in recent years.

The advancement of membrane technology has reduced the cost of desalination processes. Today, desalination cost has been reduced to 0.5 $\$/m^3$. However, many challenges are confronted especially in the treatment of seawater. Some of the components in water endanger the life of the living creatures. Boron, one of these components, is present in many forms in water. Researchers witnessed some harmful influences such as yellowing of leaves and rotting of fruits in some plants grown using irrigation water with high boron concentration. Boron excess has a significant impact on human health. For this reason, in 2004, the World Health Organization (WHO) has suggested that the concentration of salt and boron in the drinkable water should be 350 ppm and 0.5 ppm, respectively.

There are various processes including the ion exchange, two-pass reverse osmosis, pH adjustment and hybrid methods to separate boron from the water. In the ion exchange method, the need for specific resin use, usage of regeneration chemicals, and limited

operating conditions compared to other technologies are the most critical disadvantages. Although new generation membranes enable reaching the desired salt concentration in potable water, they are generally insufficient for boron separation. Therefore, a two-pass reverse osmosis system is needed. In this process, the product water obtained in the first stage (permeate) is treated again in the second stage, and the amount of boron it contains reduces. In this process; however, the recovery is too low, and the pH adjustment is required. Furthermore, the increase in the amount of chemical added to the system increases the cost of the desalination process. Although hybrid processes have been developed which have a positive effect on cost by reducing energy consumption, it has not been possible to get rid of dependence on chemicals.

Energy efficient reverse osmosis (EERO) process has been created to increase the recovery obtained with single stage systems. With this technology, the net specific energy consumption of the system reduces, but the level of boron in the product water cannot lower. Therefore, with the use of countercurrent desalination and boron removal (CDBR) process, the amount of boron and salt in the typical seawater water was reduced to 0.5 ppm and a below level of 100 ppm, respectively. The most important advantage of the CDBR process compared to other boron separation methods is that it achieves high recovery values without the need for pH adjustment.

In this thesis, it was aimed to reduce the net specific energy consumption of the CDBR system by modeling a new configuration of it. In the previous configuration of the CDBR process, the retentate of the low-pressure membrane stage (LPMS) was a waste product. Here, this wastewater was recycled to be blended with the saltwater feed. As a result of all calculations, it was concluded that the SEC_{net} value of the CDBR process was decreased from 3.10 kWh/m^3 to 3.00 kWh/m^3 at 65% recovery to produce 0.5 ppm boron-containing water.

Next, using commercial membranes, the two-pass RO process with concentrate recycling and the CDBR process were designed. For this purpose, Toray DS2 and DuPont WAVE process design software that employ commercial membranes of two companies were utilized. Designing the CDBR process using commercial membranes enabled producing water having a boron composition of 0.495 ppm. Furthermore, the maximum recovery that the CDBR system attained was 62.13%. Although the net SEC was reduced by the CDBR process, it was concluded that the cost was more than the two-pass system.

TERS OZMOS UYGULAMASINI KULLANAN EŞ ZAMANLI BOR VE TUZ AYIRMA PROSESİNİN TEKNİK VE EKONOMİK FİZİBİLİTESİ

ÖZET

Su canlı hayatı için oldukça önemli bir kaynaktır. Yeryüzünün büyük bir kısmının su ile kaplı olmasına rağmen kullanılabilir tatlı su kaynakları oldukça sınırlıdır. Ekolojik dengenin bozulması, artan sanayileşme ve temiz su kaynaklarının kirletilmesi gibi birçok etken nedeniyle mevcut durumdaki su kaynakları da azalmaktadır. Bu nedenle birçok canlı temiz suya ulaşımda zorluklarla karşı karşıya kalmaktadır. Dünya genelinde yaygın olan bu durum dikkate alındığı zaman, su kıtlığı sorunun şu anki hayatı önemli ölçüde tehdit ettiği söylenebilir. Bu problemin ileriki yıllarda daha da ciddi boyutlara ulaşacağının tahmin edilmesi, günümüz bilim insanlarını alternatif çözümler aramaya yöneltmiştir.

Dünya'daki su kaynaklarının azalması nedeni ile tuzlu suyun arıtılması üzerine çeşitli çalışmalar yapılmaktadır. Desalinasyon teknolojisi olarak adlandırılan bu süreç ile su içerinde çözünmüş olarak bulunan istenmeyen bileşenler sudan ayrılarak temiz su eldesi sağlanır. Ortaya çıktığı ilk dönemlerde küçük çaplı sistemler için kullanılan bu teknoloji, günümüzde ise ciddi bir pazar payına sahiptir. Ancak bu yöntemin büyük ölçekli tesisler için kullanılması oldukça yüksek bir enerji ihtiyacı gerektirdiğinden maliyeti de oldukça fazladır.

Desalinasyon yöntemleri genellikle termal buharlaşma ve membran ayırımı süreçleri olmak üzere iki ana grupta toplanmaktadır. Desalinasyon metodu için ilk yıllarda termal buharlaştırma yöntemi kullanılmıştır. Bu teknik ile öncelikle yüksek tuz konsantrasyonuna sahip su kaynatılır ve ardından buharlaştırılır. Üretilen bu temiz su buharının yoğunlaştırılmasıyla içilebilir su elde edilir. Termal buharlaştırma teknolojisi genel olarak çok kademeli ani damıtma (MSF) ve çok etkili damıtma (MED) olmak üzere iki önemli sistemi bünyesinde barındırır. Bu sistemleri kullanan ülkelerin başında ise körfez ülkeleri gelmektedir. Bulundukları bölgedeki enerji maliyetlerinin az olması oldukça yüksek miktarda enerji tüketen termal buharlaştırma süreçlerinde ilerlemelerine olanak sağlamıştır. Ancak zamanla bu enerji kaynaklarının da tükeneceği endişesi bu ülkeleri de başka alanlara yönlendirmiştir. Membran ayırma süreçleri, gelişen teknoloji ile birlikte daha verimli hale gelerek, kullanımda termal buharlaştırma süreçlerinin önüne geçmişlerdir. Hatta Orta Doğu bölgesinde termal buharlaştırma ve membran ayırma süreçlerinin birlikte kullanıldığı hibrit tesisler kurularak desalinasyonda fosil yakıtlara olan bağımlılık azaltılmıştır. Mekanik ve kimyasal yöntemleri birlikte kullanan membran süreçlerine ters ozmos (RO), nanofiltrasyon (NF) ve elektrodiyaliz (ED) örnek olarak gösterilebilir. En yaygın olarak kullanılan membran ayırma süreçlerinden biri olan ters ozmos prosesinde yarı geçirgen bir zar kullanılarak yüksek tuz konsantrasyonuna sahip çözeltiye osmotik basıncın üzerinde bir basınç uygulanması ile su moleküllerinin membrandan karşı tarafa geçmesi sağlanır. Böylece tuzlu sudaki istenmeyen bileşenler uzaklaştırılarak temiz su elde edilir. Kullandığı enerji miktarı termal buharlaştırma süreçlerine kıyasla

daha az olduğu için özellikle de son yıllarda en çok tercih edilen Desalinasyon sistemleri içerisinde yer almaktadır.

Desalinasyon teknolojisindeki artan rekabet maliyetin önemli ölçüde azalmasına neden olmuştur. Ancak maliyet üzerine en önemli etki membran teknolojisinin gelişmesi ile sağlanmıştır. Yüksek tuz ayırımı sağlayan membranlar etkin bir ayırma sağlayarak su kalitesinin artmasına yardımcı olur. Gelişen teknoloji ile ürün suyu miktarında bu artış her geçen gün desalinasyonda membran ayırma teknolojisinin kullanımı için bir umut kaynağı olmuştur. Günümüzde ise deniz suyundan temiz su üretim maliyeti 0,5 $\$/m^3$ seviyesine kadar indirilmiştir. İlerleyen yıllarda ise daha da azalması hedeflenmektedir.

Ancak günümüzde, özellikle deniz suyundan temiz su eldesinde birçok güçlüklerle karşılaşılmaktadır. Çünkü doğası gereği su içerisinde bulunan bazı bileşenler canlı hayatını tehdit etmektedir. Bu bileşenlerden biri olan bor, su içerisinde çeşitli formlarda bulunmaktadır. Son yıllarda yapılan bazı araştırmalara göre değişen bor konsantrasyonuna sahip sulama suları kullanılarak yetiştirilen bazı bitkilerde çeşitli zararlı etkiler gözlemlenmiştir. Bu etkiler bitkilerde yaprak sararmasına ve meyvelerin olgunlaşmadan çürümesine neden olabilmektedir. Her bitkinin bor toleransı farklı seviyelerde olduğu için sulama sularındaki bor konsantrasyonu bitki türüne göre ayarlanmalıdır. Bitkilerin yanı sıra içme suyundaki yüksek bor konsantrasyonu insan ve hayvan sağlığı üzerinde de birçok olumsuz etkiye sahiptir. Bu nedenle Dünya Sağlık Örgütü (WHO) 2004 yılında bir bildiri yayınlayarak, içme suyunda bulunması gereken tuz ve bor konsantrasyonunu sırasıyla 350 ppm ve 0,5 ppm olarak belirlemiştir.

İçme suyundan bor ayırmak için birçok yöntem geliştirilmiştir. Bu yöntemlere örnek olarak iyon değişimi, iki geçişli ters ozmos prosesi, pH ayarlaması ve hibrit yöntemler verilebilir. İyon değişimi metodunun, özel reçine kullanımına olan ihtiyacı ve kısıtlı kullanım şartları diğer teknolojilere kıyasla en önemli dezavantajlarıdır. İleri teknoloji membranların üretilmesi içilebilir suda arzu edilen tuz konsantrasyonuna ulaşmaya imkân verse de, sudan bor ayırımı için aynı başarı elde edilememiştir. Bu amaçla tek geçişli ters ozmos sistemine bir geçiş daha eklenerek, iki geçişli ters ozmos prosesi oluşturulmuştur. Bu süreç bor ayırma için oldukça yaygın kullanılır. Bu proses ile ilk kademede elde edilen ürün suyu (permeate) ikinci aşamada tekrar muamele edilerek içerdiği bor miktarı azaltılır. Ancak ilk kademede arıtılmış suyu tekrar işlemek sürecin geri kazanım oranını azaltır. Ayrıca ürün suyundaki bor konsantrasyonunu düşürmek için ise ikinci kademeye beslenen su için pH ayarlaması yapılması gerekir. Çünkü bor bileşiklerinin düşük pH değerlerindeki yapısı su molekülleri ile yaklaşık olarak aynı çapta olması sebebi ile membran içerisinden geçebilir. Bu nedenle besleme suyuna bir miktar bazik çözelti eklenerek pH artırılır. Böylece bor bileşikleri çapı su moleküllerinkinden daha büyük olan formlara dönüşerek zar içerisinden geçemez hale gelirler. Ayrıca sisteme eklenen kimyasal miktarındaki artış desalinasyon sürecinin maliyetini artırır. Her ne kadar enerji tüketiminin azalmasını sağlayarak maliyet üzerinde olumlu etkiye neden olan hibrit prosesler geliştirilse de yine de kimyasallara olan bağımlılıktan kurtulmak mümkün olmamıştır.

Tek kademeli sistemlerdeki geri kazanım oranını artırmak için enerji etkin ters ozmos (EERO) süreci geliştirilmiştir. Bu proses ile sistemin net özgün enerji tüketimi azaltılarak, geri kazanım oranı %75'e kadar artırılmıştır. EERO sürecinin istenilen bor seviyesine sahip su üretememesi sebebi ile eş zamanlı desalinasyon ve bor giderimi (CDBR) süreci geliştirilmiştir. CDBR prosesi iki geçişli ters ozmos prosesine kıyasla

daha düşük net özgün enerji tüketimi ile daha fazla suyun geri kazanımını sağlar. Ayrıca 35.000 ppm TDS ve 10 ppm bor içeren bir deniz suyunun besleme suyu olarak kullanılması durumunda, ürün suyundaki bor konsantrasyonunu 0,5 ppm seviyesine ve tuz konsantrasyonunu ise 100 ppm seviyesinin altına herhangi bir pH ayarlaması yapmadan indirmeyi başararak deniz suyundan bor ayırma yöntemlerine bir alternatif olmuştur.

Deniz suyundan bor ayırma yöntemlerinde maliyeti azalmak için çeşitli alternatifler kullanılabilir. Bu seçeneklerden biri ise enerji geri kazanım cihazı kullanımıdır. Böylece atık sudaki basıncın yaklaşık olarak yarısı iyileştirilerek sisteme geri gönderilir. Enerji geri kazanım cihazları genel olarak Pelton çarkı ve izobarik sistemler olmak üzere ikiye ayrılırlar. Pelton çarkında mekanik enerjiye dönüşüm var iken, izobarik sistemlerde ise konsantrenin basıncı doğrudan sisteme verilir. Böylece bu cihazların verimliliği 0,95'in üzerine çıkabilmektedir. Bor ayırma sistemlerinde enerji kullanımını azaltmanın bir diğer yöntemi ise hibrit membran konfigürasyonudur. Farklı tuz reddine ve ürün kapasitesine sahip membranlar bir basınç kabı içerisinde belirli bir sırada yerleştirilir. Genellikle en yüksek tuz ve bor reddi sağlayan membranlar ilk sıraya yerleştirilirken, kalan lokasyonlara ise yüksek su üretme kapasitesine sahip membranlar yerleştirilir. Böylelikle hem basınç kabı içerisinde akış eşit olarak dağılır hem de sistemin tükettiği enerjide azalma meydana gelir. Bu sistemin bir diğer avantajı olarak ise membran kirlenmesinin önüne geçmesidir. Enerji tüketimini azaltmada kullanılan bir diğer yöntem ise iki geçişli sistemlerde, ikinci geçişten elde edilen atık suyun prosesin besleme kısmına gönderilmesidir. Bu yöntem ile her ne kadar besleme suyunun tuzluluk oranı artmış olsa da konsantrenin içerdiği basıncın sisteme geri döndürülmesi enerji kullanımını azaltır.

Bu çalışmada ise yeni bir CDBR süreci geliştirilerek sistemin net özgün enerji tüketiminin azaltılması amaçlanmıştır. Standart CDBR prosesinde düşük basınçlı membran kademesinden (LPMS) ayrılan derişik su atık olarak sistemden atılır. Ancak geliştirilmiş CDBR prosesi ile bu su sistemin besleme kısmına gönderilerek tekrar arıtma işlemine dahil edilir. Belirli bir basıncı sahip olarak LPMS'den ayrılan bu suyun sisteme geri döndürülmesi ile bir önceki CDBR prosesine kıyasla daha az net özgün enerji tüketimi ile ürün suyundaki istenilen bor seviyesi elde edilebilmesi amaçlanmıştır. Bu çalışmada, konsantre geri dönüşümlü iki geçişli ters ozmos sistemi ile %48 geri kazanım oranına ulaşılmıştır. Geliştirilmiş CDBR prosesi ise net özgün enerji tüketimini 3,10 kWh/m^3 seviyesinden 3,00 kWh/m^3 seviyesine kadar indirmeyi başarmıştır. Geliştirilmiş CDBR prosesi ile enerji tüketimindeki bu azalışın herhangi bir pH ayarlaması yapılmadan sağlanması sistemin en önemli avantajlarından biridir. Her iki proses için de %85 verimliliğe sahip pompa ve %90 verimliliğe sahip enerji geri kazanım cihazı (ERD) kullanılmıştır.

Çalışmanın ikinci aşamasında ise hem konsantre geri dönüşümlü iki geçişli ters ozmos prosesi hem de geliştirilmiş CDBR prosesi ticari membranlar kullanılarak dizayn edilmiştir. Bunun için Toray DS2 ve DuPont WAVE olmak üzere iki farklı üreticinin membranlarını bünyesinde barındıran proses tasarım programları kullanılmıştır. Yeni nesil %99,8 tuz reddi sağlayan TM820K-440 isimli bir membran ilk geçişte kullanılarak konsantre geri dönüşümlü iki geçişli RO prosesi tasarlanmıştır. Bu tasarımın ikinci kademesinde ise TM710D,BWRO membran tipi seçilmiştir. Bu proses enerji geri kazanımı cihazı ile çalışmaya olanak sağlayan Toray DS2 kullanılarak tasarlanmıştır. Kullanılan enerji geri kazanım cihazı 0,9650, pompalar ise 0,85 verimliliğe sahiptir. Bu sistem için ise geri kazanım yüzdesi olarak %44,68'e ulaşılmıştır. CDBR prosesinin tek aşamalı ters ozmos (SSRO) kademesi ile LPMS

kombinasyonu da Toray DS2 kullanılarak tasarlanmıştır. Her iki aşama için de TM820K-440,SWRO membranı seçilerek ürün suyundaki bor miktarı 0,3 ppm seviyesine kadar düşürülmüştür. Bu sistem için geri kazanım %45,49 olarak kaydedilmiştir. Bu prosesten elde edilen atık, CDBR prosesinin CMCR sistemine besleme suyu olarak gönderilmiştir. Bu kısım ise WAVE tasarım programı kullanılarak tasarlanmıştır. SW30XHR-440i ve BW30HR-440 membranları ile tasarlanan sistemin geri kazanımı %29,2'ye ulaşmıştır. Atık suyun arıtılması ile üretilen su 1,06 ppm bor içermektedir. CDBR prosesinin iki geçişli ters ozmos sistemi ile CMCR kısmının ürettiği ürün suyu karıştırıldığında, 0,495 ppm bor içeren ürün suyu elde edilmiştir. CDBR prosesinin ticari tasarımındaki geri kazanımı ise %62,13 olarak bulunmuştur.

Bu çalışmada her ne kadar net SEC, CDBR işlemi ile azaltılmış olsa da, maliyetinin iki geçişli sistemden daha fazla olduğu sonucuna varıldı. Bu artış kavramsal tasarımda %3 olarak hesaplanırken, bu değer ticari membranların kullanıldığı sistemler için %34 olarak kaydedilmiştir.

1. INTRODUCTION

1.1 Desalination

Water is very crucial for living creatures. Although 70% of the world consists of water, fresh water sources only correspond to 2.5% of it [1]. The total amount of ocean and saline water covers the remaining percentage. On the other hand, 1% of the fresh water is accessible, and the glaciers and snow-covered areas possess the rest of it [2].

Of the world population, 41% inhabit in places where the water scarcity exists. Conventional clean water resources are reducing, or the amount of salt that they hold is increasing due to the extreme usage of them. As a solution to remedy the lack of fresh water on the earth, saline water desalination and reuse of water are the major options to ensure the continuity of world generations [3].

Desalination, the general name for the process of salt removal in water to produce clean water, is a significant source for the production of potable water with the thermal technology developed in the latest 60 years and the membrane technology developed in the last 40 years [3]. In 2005, the thermal technologies (MED and MSF) became a leader in desalination technologies with a share of 60%, while the membrane technologies had a share of 40%. Later, the market share of the membrane technologies approached 60%, and the thermal and hybrid processes held the rest of it. In the present desalination industry, the reverse osmosis process has a share of 63%, while the MSF and MED processes possess a share of 23% and 8%, respectively [4].

While the desalination capacity of plants was 5 million m^3/day in 1980, it raised to 52.8 million m^3/day in 2008. This capacity attained to 85.9 million m^3/day in 2013. 59% of the installed capacity is operated for seawater desalination, 22% for brackish water desalination, 5% for water reuse, 9% for the treatment of river water and 5% for producing pure water [5]. Nowadays globally almost 90 million m^3 water per day has been desalinated, and around 18,500 plants employ to provide drinking water [6]. The countries possessing the maximum desalination capacity are the United States, and Gulf countries -Saudi Arabia, the United Arab Emirates and Kuwait- which produce

40-50% of the desalinated water in the world [6, 7]. The Ras Al Khair plant in Saudi Arabia has the largest desalination capacity in the world and uses both membrane technology and thermal technology. The reverse osmosis capacity of this plant is 308,000 m^3/day, whereas the MSF process works with a capacity of 728,000 m^3/day. Additionally, the Soreq plant in Israel is the largest plant in the world using membrane technology [7]. Table 1.1 indicates a list of some desalination plants in the world.

Table 1.1 : Examples of desalination plants in the world.

Plant	Capacity (m^3/day)	Location	Technology
Ras Al Khair [7]	1,036,000	Saudi Arabia	Membrane & Thermal Desalination
Soreq [7]	510,000	Israel	Membrane Desalination
Magtaa [7]	500,000	Algeria	Membrane Desalination
Victoria [7]	444,000	Australia	Membrane Desalination
Torrevieja [8]	240,000	Spain	Membrane Desalination
Jebel Ali [9]	182,000	United Arab Emirates	Membrane Desalination
Beckton [10]	150,000	Great Britain	Membrane Desalination
Tampa Bay [11]	108,831	United States	Membrane Desalination

The salinity of feed water (brackish water or seawater) for desalination processes varies between about 1,000 ppm to 60,000 ppm [3]. Total dissolved solids (TDS) in seawater varies depending on location [12]. Table 1.2 shows the TDS and ion content of the seawater in the various regions. They are converted into the potable water with the use of appropriate technologies [13]. Besides, the mean ocean pH was around 8.2 before the industrial revolution, but today it is accepted to be approximately 8.1 [14].

The fresh water has the TDS level below 1,000 ppm. Several countries have admitted national drinkable water standards for both certain pollutants and TDS; however, the standard limits differ in countries [3]. To illustrate, the standards for boron content of drinking water change between 0.5 and 5 ppm in worldwide [15]. In 2004, the World Health Organization (WHO) suggested the salt and boron concentration limits in the drinking water as 350 ppm and 0.5 ppm, respectively [16].

Table 1.2 : Major ion components of seawater [13].

Ions	Arabian gulf at Kuwait	Red sea at Jeddah	Eastern Mediterranean	Typical seawater
Chloride (Cl^-)	23,000	22,219	21,200	18,980
Sodium (Na^+)	15,850	14,255	11,800	10,556
Sulfate (SO_4^{2-})	3,200	3,078	2,950	2,649
Magnesium (Mg^{2+})	1,765	742	1,403	1,262
Calcium (Ca^{2+})	500	225	423	400
Potassium (K^+)	460	210	463	380
Bicarbonate (HCO_3^-)	142	146	-	140
Strontium (Sr^{2+})	-	-	-	13
Bromide (Br^-)	80	72	155	65
Borate (BO_3^{3-})	-	-	72	26
Fluoride (F-)	-	-	-	1
Silicate (SiO_3^{2-})	1.5	-	-	1
Iodide (I^-)	-	-	2	<1
TDS	45,000	41,000	38,600	34,483

1.2 Boron

Boron mostly appears in the boric acid form in an aqueous environment. Boric acid is a water-soluble solid, and its solubility is 55 g/L at 25 °C. Boron, an element commonly found in soil, water, and rocks, exists merely at quite low concentrations in nature. The concentration of boron in seawater is between 0.5 to 9.6 ppm, and seawater contains 4.6 ppm boron on average. The boron concentration in clean water changes from 0.01 to 1.5 ppm and enhances considerably in regions with soils that are rich in boron [17].

Recently a considerable rise in the boron concentrations of surface waters has been noticed. Various natural and human-induced factors lead to this increase in different regions [17]. Numerous industrial sectors like glass and cosmetics have widely utilized boron. Due to the high volatility of boron when released to nature, acid rain occurs, and boron accumulates in the soil. As a result, this causes the pollution of potable water resources, and health and environmental problems [18].

Boron, available in vegetables and fruits, is a mineral, which is naturally supplied from nutrition sources. The excess and lack of boron, which is a necessary nutrient for the growth of vegetables and fruits, is detrimental to numerous plants. The shortage of boron disturbs generation of cells, while postponing enzymatic reactions. If the amount of boron is more than needed, lethal effects like yellow leaves, fall of the leaves, and decay of immature fruits can be observed [17].

The endurance levels of diverse plants toward boron differ [17]. Some plants show more sensitivity to boron than other plants. Sensitive plants might withstand irrigation waters containing boron concentration of 0.3 ppm, while highly tolerant plants can endure boron concentration of 4 ppm [19].

Besides, boron deficiency has effects on the function of the animal body. In the absence of boron, which has a positive impact on the metabolism of nutrients, a reduction in the absorption of phosphorus, calcium, and magnesium occurs. The WHO determined that the boron dose of 1 to 13 mg/day is sufficient for healthy individuals. Despite the positive impact of boron on many organ functions, to consume nutrients and waters containing a high amount of boron in the long term causes some problems in the cardiovascular and nervous systems [17].

1.3 Purpose of Thesis

The goal of this study is to reduce the net specific energy consumption (SEC_{net}) of the concurrent desalination and boron removal (CDBR) process by modeling a new configuration of it. In this version, so-called a revised configuration of the CDBR process, apart from the earlier system the concentrate stream of the low-pressure membrane stage (LMPS) is conveyed to the feed side of the process. Besides, to make a comparison, two-pass reverse osmosis (RO) process where the concentrate stream of the 2^{nd} pass is blended with the overall feed flow of the system was also modeled.

Next, the two-pass RO process with the second pass concentrate recycle was designed using the rigorous design software tools in which brand new commercial membranes are employed. Using conventional RO rack design and hybrid membrane inter-stage design (HID), it was aimed to find out the real operating pressure, recovery, and SEC_{net} values of the processes. Besides, as in the two-pass system, the CDBR system was remodeled using the software tools in order to estimate its actual performance.

Finally, the cost estimation analysis was performed for each process so as to explore the change in the cost of desalination processes when CDBR systems are utilized. In all design systems, the common goal is to decrease boron concentration in the final product water to 0.5 ppm or below levels at the reduced SEC_{net} and the advanced recovery. Doing so, it was concluded whether the proposed processes could attain the potable water standards, reported by the WHO.

2. LITERATURE REVIEW

2.1 Desalination Technologies

Commercial desalination methods are mainly grouped as the thermal technologies and the membrane technologies. Besides, there are hybrid desalination systems in which the thermal and membrane technologies are combined [20].

2.1.1 Thermal technologies

Each thermal desalination method uses distillation for the production of water vapor. Subsequently, by condensing this water vapor, water with lower salt content is produced. Thermal evaporation can, therefore, be utilized for desalination of highly saline waters. There are five common flows for thermal desalination systems which are the water supply for desalination, the vapor required for evaporation of the water supply, the cooling water for condensation of clean water vapor, the distilled water with low salinity, and the brine containing salt and impurities [21].

Thermal desalination technologies prevail in the desalination market, especially the Middle East in which energy expenses are low. Two important thermal technologies for desalination are the Multi-stage flash (MSF) and Multiple-effect distillation (MED) systems [20].

In the MSF systems, brine water is boiled and condensed, respectively, in order to produce distilled water. Through the channels that are linked in series the salty feed water is preheated, and then it enters the brine heater. By using thermal energy, the heated saline water within the brine heater is delivered to a vessel that has less ambient pressure than that of the brine heater. The sudden boiling of the salty water due to low pressure causes the formation of vapor. The water vapor condenses on the channels transporting the salty feed water, and then the pure water is produced. Since a small portion of the heated saline water generates vapor, the rest of the water is transported to the second stage having less ambient pressure. This process continues up to cooling down of the brine, and the number of stages in MSF systems is mostly between 18 and 25 [20].

The process of MED utilizes multiple effects (vessels) designed in series, and the next vessel is at lowered ambient pressure as regards the previous one. 8 to 16 vessels are employed in the MED process so as to reduce energy consumption. The saline feed water is sprayed on the external side of the evaporator channels as a thin film in order to aid fast boiling and vaporization. With the use of the heat of the steam generated by evaporation, the remaining brine solution at a lowered pressure and temperature in each subsequent vessel is heated. In this way, the water boils many times without the addition of extra energy later the first vessel. That leads to a decrease in the energy, which is required for distillation, as well as the entire electric power consumption [22].

2.1.2 Membrane technologies

Membrane processes are used for allowing or preventing the pass of specific salt ions [20]. Membrane desalination processes are reverse osmosis (RO), nanofiltration (NF), and electrodialysis (ED). NF extracts divalent ions from slightly brackish water, and RO proses can remove monovalent ions. In ED, ions are used to pass through the membranes owing to electric current, and also they are merely used in brackish water applications [3].

Reverse osmosis, a significant commercial membrane technology, is utilized for brackish water and seawater. RO, which is a pressure-operated process, employs membranes to separate salt and water. In RO membrane processes, the salt water is pressurized towards the membrane. Then, the clean water passes through the film, and the residual saline water is removed. An RO plant consists of a pre-treatment process, high-pressure pump, membrane installation, and a post-treatment process. The most advanced reverse osmosis plants possess an energy recovery system. Because the pressure of the brine solution is high, the pressure energy of this solution is introduced to the feed water with the use of some energy recovery systems. The usage of energy recovery technology and high-level membranes diminishes the expense of reverse osmosis desalination systems [20].

2.1.3 Hybrid technologies

Desalination process can be merged with a power plant in a hybrid system to reduce water production costs. The use of a hybrid system is convenient for desalination if water and power demand is fluctuating significantly. Generally, the MED or MSF process is united with RO and Vapor Compression (VC) or NF units. Combining

processes with power generation systems enables the use of both the produced power and the fuel energy more efficiently [23]. The hybrid desalination systems have many benefits such as providing flexibility in the process, reducing SEC, lowering construction expenses of intake and outfall systems [20].

2.2 Boron Removal Methods

2.2.1 Ion exchange

Ion exchange is one of the processes used for boron removal from seawater. Ion exchange process generally takes place in basic operational needs. In the dilute water and seawater flows, this method might be employed, yet it is often effective at low salt concentrations. The usage of a robust ion-exchanger is required to dissociate boron salts within an aqueous environment, and each anion that will be held within the solution leads to the lower separation capacity. For this reason, boron special ion exchange resins are needed to separate them effectively. A wide range of studies has been carried out on the usage of boron selective chelate ion exchange resins in order to remove boron from the water. Hydroxyl groups in the cis location as ligands referred to as "vis-diols" become involved in synthesizing chelating resin. As these groups have high selectivity against boron, they are unreactive to other components [24].

Darwish et al. [25] studied at different boron concentrations with Amberlite IRA743 resins having different particle sizes. Increased amount of resin caused to improve boron rejection because of increase in surface area. Also, efficiency, in order to remove boron, increased with elevated pH and temperature and decreased with the boron concentration of feed and particle size of resin. They observed that the other ions have no impact on the boron removal efficiency.

2.2.2 pH adjustment

Adjusting the pH level of the process to reduce the concentration of boron in the final permeate stream is one of the major choices. Generally, the first pass of RO ought to run at standard seawater pH in order to prevent scale formation and salt precipitation. However, an optimum pH should be found to reduce scaling and lessen the boron concentration in the product water [12]. In general, boron rejection is approximately between 50% to 75% when the pH value is 7-8, it has higher level than 95% when the pH value is 10.5 [26].

Saif and Almansoori [12], in their study, they added base (caustic soda) to augment the pH value of the second pass. They investigated an RO design system in order to obtain product water at various boron concentrations by using seawater with 40,000 ppm TDS and 13 ppm boron composition. While the pH values of the first and second pass were 7.2 and 9.2, respectively, the boron concentration of the product water was reduced to levels below 1 ppm without permeate splitting.

2.2.3 Two-pass reverse osmosis (RO) process

An SSRO system lowers the salt concentration in the permeate to the required level, but the desalination membranes cannot reject enough boron. A two-pass process which has two RO stages connected in series achieves the needed boron concentration in permeate (0.5 ppm or below). In the two-pass process, the second RO pass uses the permeate produced in the first RO pass as a feed stream [16].

Farhat et al. [27] studied the two-pass RO process without pH adjustment using the membranes of many membrane manufacturers. In the study, they used seawater with high salt concentration. In a single-pass system, they obtained 99% salt rejection and 91% boron rejection performance and produced permeate with a boron concentration of 1.4 ppm. By using commercial membranes, without any pH adjustment, with a two-pass process, they reached boron and salt rejection levels higher than 95% and 99%, respectively, and in the second pass, the boron concentration in permeate decreased to 0.27 ppm. They also found that boron rejection increased from 84% to 91% when the pressure in the first pass increased from 220 psi (15.2 bar) to 800 psi (55.2 bar). They obtained recovery of 40 to 45% when using the pressure of 10 to 15 bar recommended by the membrane manufacturers in the second pass.

2.2.4 Hybrid processes

Treatment of water having low salinity by an electrodialytic process is accepted as a method for boron removal [28]. Turek et al. [28] studied the electrodialytic boron removal method using two different seawater reverse osmosis (SWRO) permeates. Initially, the first permeate contained 604 mg/L TDS and 2.25 mg/L boron, while the other permeate contained 400 mg/L TDS and 1.3 mg/L boron. The process of ED was applied to mitigate the boron concentration to a level of 0.4 mg/L for both systems; however, there was a need to increase the pH value. The pH values varied between 8 and 12, yet there was an increase in boron transport at a pH value of 11.2. Cost

calculations indicated that the cost of boron removal from permeate, which initially included 1.3 mg/L boron, was 0.196 $/m^3$, while the cost of boron removal from permeate, which initially included 2.25 mg/L boron was 0.23 $/m^3$.

Adsorption membrane filtration (AMF) method, which is a hybrid process, integrates the sorption method with the system of membrane separation. Sorbents adsorb solutes, and then the membrane separation process for these sorbents, which are saturated, occurs [26]. This hybrid process employs microspherical sorbents having identical and small sizes, and microfiltration (MF) membranes [18]. Kabay et al. [29] compared the AMF process and RO methods in order to remove boron from seawater in their study. They reduced the boron concentration of permeate from 10 mg/L to less than 2 mg/L.

Güler et al. [30] examined the removal of boron from seawater using a hybrid system developed by combining sorption-membrane filtration method with the RO process. According to them, boron adheres to a particular sorbent. Then a microporous film separates the boron-containing complex from the water. They utilized microfiltration and ultrafiltration (UF) modules so as to remove boron from permeate of SWRO. They used RO permeates with boron concentrations between 1.30 and 1.95 mg/L. When using a resin of 0.5 g/L, the lowest boron concentration was 0.31 mg/L for the MF-based systems and 0.28 mg/L for the UF-based systems. However, over time, the boron concentration for the UF-based system reached 0.62 to mg/L.

2.3 Parameters Affecting RO Boron Rejection

Koseoglu et al. [31] conducted a study in order to examine the effect of feed pH, feed pressure, boron concentration of initial feed, and feed flow rate on boron rejection performance and permeate flow rate. They used commercial SWRO membranes from two different manufacturers. They gathered seawater samples from Mediterranean Sea, Turkey that contains 32,000 TDS and 5 mg/L boron at a pH level of 8.2. After all experiments, they observed that the pH level of feed seawater has a substantial effect on boron removal performance, and increase in pH level leads to enhanced boron rejection since at higher pH values boric acid predominately found in water. These charged boron groups are rejected by reverse osmosis membrane thanks to electrostatic repulsion. Moreover, another reason for boron rejection at higher pH levels is that these species have a larger diameter compared to the neutral form of boron. Moreover, at the same operating conditions, an increase in pH values leads to a decrease in

permeate flow rate due to the membrane fouling and scaling problems, stemmed from magnesium and calcium. Addition to the pH influence, if the feed pressure increase, the permeate flow rate enhances. Yet, the effect of boosted pressure on boron rejection performance depends on membrane characteristics. Besides, there is no correlation between salt and boron rejections at constant design conditions. They also noticed that the boron concentration of feed water and feed flow rate have no considerable effect on final product boron level as well. After all experiments, boron concentration in the final product decreased to the below level of 0.1 ppm.

Magara et al. [32] worked on multi-stage reverse osmosis systems, a 2-stage system containing a stage for seawater desalination and one stage for boron removal, 3-stage system constituting one stage for seawater desalination and two stages for boron removal application. They examined the feed pressure, pH, boron concentration, and recovery effect on permeate water using these systems. They observed that boron rejection values enhance from 90% to 99% with an increase of 9 to 11 in pH level. Increase in feed pressure leads to an increase in boron rejection performance, whereas initial boron concentration has no effect on it. Furthermore, they concluded that in case recovery improves, boron quantity in permeate water increases.

In the study of Cengeloglu et al. [33], they used three different RO membrane type including SWHR, and BW-30 from FILMTEC and AG from GE Osmonics in order to investigate the pH, initial feed concentration and operating pressure influence on the boron removal. They found the same results indicated in other studies explained above. Moreover, they obtained the highest rejection and permeate flow values with the usage of SWHR membrane.

2.4 Cost of Desalination

In the years when the desalination began, the price was not a significant issue since producing potable water from seawater in order to use in ships was the major purpose. In the 1960s and 1970s, desalination methods were widely used in commercial productions, but they were too costly. In the 1970s, production costs commenced declining since the membrane processes started competing. In 1945, costs for seawater desalination were approximately 2.10 $\$/m^3$. In recent years, the cost of water production has decreased to 0.5 $\$/m^3$ for large-scale SWRO plants and 1.0 $\$/m^3$ for MSF technology [23].

The cost of water production is divided into two groups which are capital costs and operation and maintenance costs. Capital costs comprise the equipment, material, and labor costs in order to create a desalination plant, and expenses for the development, approval, design, construction, and financing of the desalination project. Operation and maintenance (O&M) costs constitute labor, energy, chemical, and materials costs such as RO membranes required for plant maintenance and operation, and also the costs related to waste and environmental management [22].

The investment costs of commercial desalination systems of the membrane and thermal processes are considerably different from each other (Table 2.1). For the same plant capacity, the materials and equipment used in the thermal process are more expensive than those used in SWRO. The thermal technology, which utilizes both the thermal and electrical power, deplete a greater quantity of specific energy than that of the RO system, which uses just electrical power, consumes. Furthermore, thermal processes need more chemicals. Additionally, as the plant capacity increases, the total water production cost decreases. Also, there may be environmental and political constraints for such large-scale projects. A desalination plant ought to be installed in a suitable location to forestall extra costs like water transfer [23].

Figure 2.1 demonstrates operating costs including the electricity and thermal energy consumption, replacement of membranes and parts, labor, and chemical costs for the significant desalination technologies. The total operating costs per m^3 for significant desalination technologies were indicated as $ 0.46 for RO, $ 0.60 for MSF and $ 0.50 for MED [34].

Table 2.1 : Energy consumption and total water cost for large-scale desalination processes [23].

Technology	Thermal energy (kWh/m^3)	Electrical energy (kWh/m^3)	Total energy (kWh/m^3)	Investment cost ($/$m^3$/d)	Total water cost ($/$m^3$)
MSF	7.5-12	2.5-4	10-16	1,200-2,500	0.8-1.5
MED	4-7	1.5-2	5.5-9	900-2,000	0.7-1.2
SWRO	-	3-4	3-4	900-2,500	0.5-1.2
BWRO	-	0.5-2.5	0.5-2.5	300-1,200	0.2-0.4

The main expense of RO desalination is about the energy that might correspond to 30-50% of the operating costs. In RO desalination, the energy requirement for seawater feed is 12 kWh/m^3 without energy recovery [35]. With the use of efficient energy

recovery systems, the decline in energy consumption results in decrease the cost of unit water production in RO desalination systems. Energy is recovered by brine using turbo systems which include Pelton turbines, and more. The other new energy recovery systems aim to reduce energy consumption by less than 2 kWh/m^3 for SWRO [23].

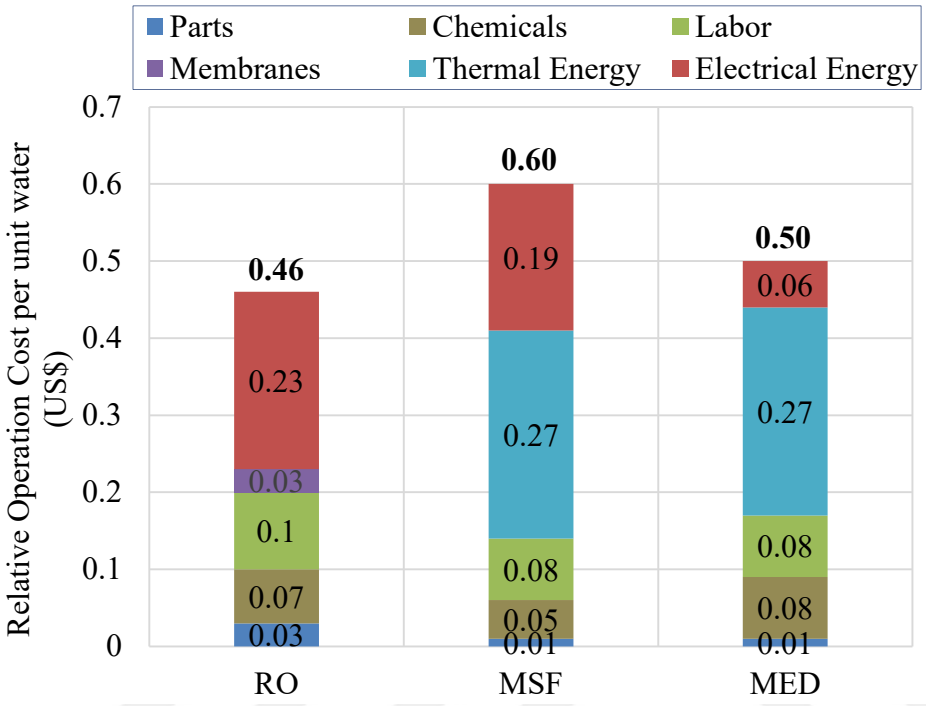

Figure 2.1 : Relative operation costs of the significant desalination technologies (US$/$m^3$), reproduced from [34].

The performance of membrane materials and modules has developed. The enhanced salt rejection, flux, and capacity in order to operate at high pressure causes a reduction in membrane costs. The recovery improved significantly because of developed salt rejection. While the recovery for standard seawater having 35,000 mg/L salinity was nearly 25% in the 1980s, it raised to 35% in the 1990s. Nowadays, the recovery is almost 45%, whereas it is 60% when using a second stage. The enhanced recovery paved the way for reducing both investment and operational costs [23].

2.5 Alternatives for Minimizing Energy Consumption in RO Systems

2.5.1 Energy recovery devices (ERD)

The concentrate of the RO system possesses 40-50% of the energy supplied to the system for the seawater desalination. When employing an energy recovery device (ERD), this energy can return to the system (Figure 2.2). The ERDs can be classified as the centrifugal (Pelton wheel (turbine), etc.) and isobaric (Pressure exchangers, etc.)

energy recovery devices [21]. The pressure exchanger group transfers the pressure of the concentrate directly on the feed of the system [36]. In the other group, the pressure of the concentrate is transferred to an impeller which supplies mechanical energy [21]. In the Pelton system, high-pressure pumps pump the total feed flow of the system. Whereas, in the pressure exchanger system, the high-pressure pumps merely pump the amount of feed equal to the product stream. Due to the decrease in the amount of water to be pumped by the high-pressure pump, the energy consumption is less in these systems because the motor and pump losses reduce. Besides, the efficiencies of pressure exchanger and Pelton wheel are nearly 96%, and 86%, respectively [36].

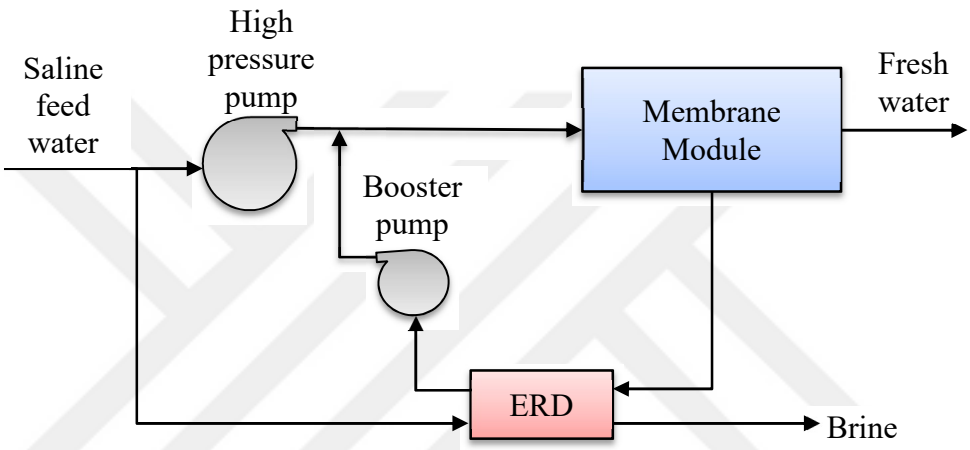

Figure 2.2 : Work exchanger, reproduced from [37].

2.5.2 Hybrid membrane design (HID)

In SWRO plants, where typical spiral wound RO membranes operate, the seawater is served to the first element of the vessel, and the permeate and concentrate flows are obtained from the tail end of the vessel. Since the head element is subjected to all the feed stream and pressure, it works at a considerably higher flux than the next elements.

In a standard design with 7-elements in individual vessels and equal flow distribution between the elements, each membrane would yield 14.3% of the total product water. However, traditional SWRO plants have flow inequality, and generally the lead membrane element yields more than 25% of the total water, while the tail membrane element produces only 6-8% of it. One of the reasons for this decrease in production of permeate is the rise in salinity in the feed water. Another reason is osmotic pressure. Because the permeate is obtained from each element, yet the concentrated water passes through all elements of the vessel, and it is discharged from the end element [38].

The maximum flux at the head element of the pressure vessel, resulting from the minimum osmotic pressure, rises the fouling risk. Therefore, the imbalance between the first and last flows are able to decrease with the use of hybrid membrane inter-stage design (HID) where PV consists of elements having different fluxes. In order to reduce the net pressure throughout the PV, a low flux membrane element is placed in the first position, while high flux membrane elements are placed in the remaining positions of the PV. The DOW (now DuPont) Water Solutions Company introduced this design idea called internally staged design (ISD) [39]. When commercial SWRO membranes of three different companies are employed, this configuration is called the hybrid membrane inter-stage design (HID) [40].

With a low flux/ high salt rejection membrane at first position, merely 14-18% of the total product water is produced. Thus, energy is used more efficiently throughout the vessel. The next membrane element produces 14-16% of the total product when a membrane owning medium permeability/salt rejection is selected [38]. The use of membrane type having high flow/low salt rejection at the last five elements provides the 1-1-5-membrane configuration as demonstrated in Figure 2.3. This hybrid membrane configuration enables both a uniform flux distribution and 5-15% energy-saving by decreasing the fouling risk for each element [38].

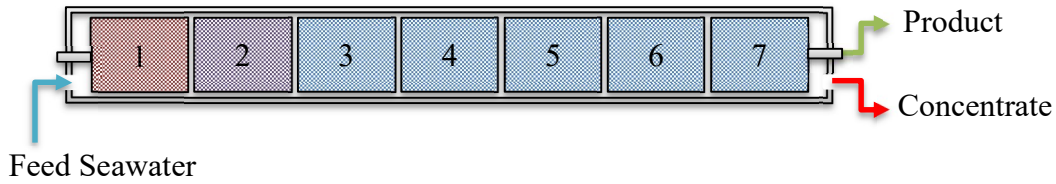

Figure 2.3 : Hybrid membrane configuration, reproduced from [38].

Peñate and García-Rodríguez [39] performed a HID analysis with commercial membranes of three different membrane manufacturers which are Filmtec, Toray, and Hydranautics. They utilized the 7-element standard design, housing the same membranes in a PV, and different HIDs. They observed that HIDs consume less energy than standard PV designs does. For the analysis of HID systems, they placed the elements with high rejection to the first position, energy-saving elements to the middle place, and elements proving high flow to the last positions. According to this study, the general design criterion for the analysis of all membrane brands is the use of membranes that provide high rejection in the first two locations and the membranes providing high flow to the remaining five positions.

Han et al. [37] examined the HID performance by using a single-pass SWRO system in order to reach the desired water quality, 1 ppm boron concentration in the product water. They employed conventional and HID designs (3-4 membrane configuration) by using feed at different concentrations and temperatures. According to the results, temperature is a parameter that effects boron rejection most and if the feed temperature increases the product concentration increases. The HID configurations have reduced SEC than that of conventional designs. If the membranes providing high flow are located in the first three positions, this leads to a decrease in SEC.

2.5.3 Split partial two-pass RO system

Two pass RO processes have two major design options which depend on the desired product quality. In full second pass system, the total permeate obtained from the first pass is treated in a second pass to achieve the desired product quality. For this system, the total recovery is 47% if there is brine recycling from the second pass to feed. In split partial two-pass system, a portion of product produced by the first pass is treated in the second pass, while the rest is mixed with the final product. The total recovery for the split partial two-pass SWRO system is 48.5% if there is brine recycling from the second pass to the raw feed channel. This system enhances the total water recovery and reduces operating costs by providing less energy use. Besides, capital cost decreases because of a reduction in the second pass size [41].

When the desalination design is constructed to remove boron from sea water, the pH of the concentrate of the first pass varies between 9.5 and 11, which can give rise to the calcium carbonate precipitation on the membrane. That difficulty is forestalled by adding of antiscalant to the feed water of the BWRO pass (2^{nd} pass). Furthermore, recycling the concentrate in the second pass to the feed of the first pass is highly influential in averting the scaling problem [38].

Recycling of a small percentage of the concentrate of SWRO enhances lightly the salinity of feed water introduced into the first (SWRO) pass. The increment in energy usage due to the increase in salinity of the water is less than the percentage of energy-saving provided by the whole process. In split-partial two-pass systems, the volume of the second pass can be reduced by 25-50% since the amount of water pumped to the second pass is decreased. With the split partial two pass reverse osmosis configuration, the energy consumption reduces by 12% and 20% for the whole process, and 14% and

16% for practical experimentation with large-scale SWRO systems as displayed in Figure 2.4 [38].

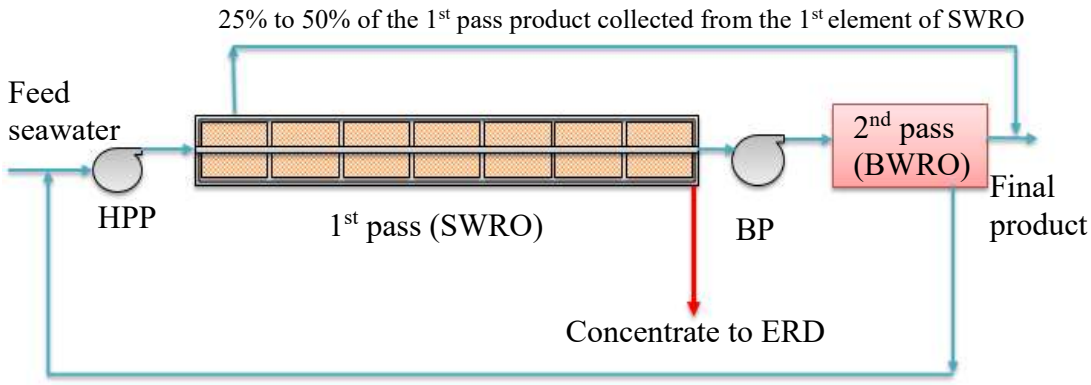

Figure 2.4 : Split partial two-pass SWRO system, reproduced from [38].

Larnaca seawater RO desalination plant located in Cyprus is one of the seawater desalination plants using split partial two-pass RO design and pH adjustment at the second pass. Besides, this plant produces water containing boron below a level of 1 ppm by employing 8-element in a pressure vessel. Here, the permeate produced by the front side of the pressure vessel is split up, whereas permeate obtained from the tail side of the vessel is conveyed to the 2nd pass BWRO element to be treated further. Both of permeates are combined to produce water with lower boron composition [42].

2.5.4 Energy efficient reverse osmosis (EERO) desalination process

Conventional SSRO provides an acceptable recovery at a high OPD value for the standard seawater when operating ideally. Recycling of the retentate improves the recovery despite an increase in SEC. If the RO stages in series operate, the SEC can lower, and the recovery increases, yet inter-stage booster pumps are needed. A countercurrent membrane cascade with recycling (CMCR) using membranes with moderate rejection in some of the stages may improve the recovery, while decreasing the OPD. Nevertheless, the increase in pump costs because of pressurization of the recycle of the permeate augment the SEC [43].

The energy-efficient reverse osmosis (EERO) desalination process, created by the Singapore Membrane Technology Center, combines the SSRO with the CMCR (Figure 2.5). The retentate flow of the SSRO is conveyed to the CMCR. In the CMCR, stage 1 employs nanofiltration (NF), while stage 2 utilizes reverse osmosis (RO) membrane. Doing so, the recovery of water increases, that contributes to reducing the

SEC. Since the SSRO and the CMCR operate at the same OPD in the EERO process, no inter-stage pumps employ on the high-pressure side of the CMCR. Inasmuch as the retentate flow of SSRO is more concentrated than feed seawater, it is necessary to reduce the OPD in both the CMCR and SSRO stages in order to operate the CMCR without using an inter-stage pump. This OPD is less than the optimum value at which the SSRO operates alone. The CMCR achieves this in this way: by combining countercurrent stream of the permeate and the retentate, by recycling of permeate from the stage 1 to stage 2, by using the NF membrane which more salt passes through that providing self-recycling of retentate. These three characteristics of the CMCR cause the concentration difference across the membranes in stages 1 and 2 to decrease. Thus, the system attains high water recovery without increasing the OPD of the EERO process more than that of the SSRO [43].

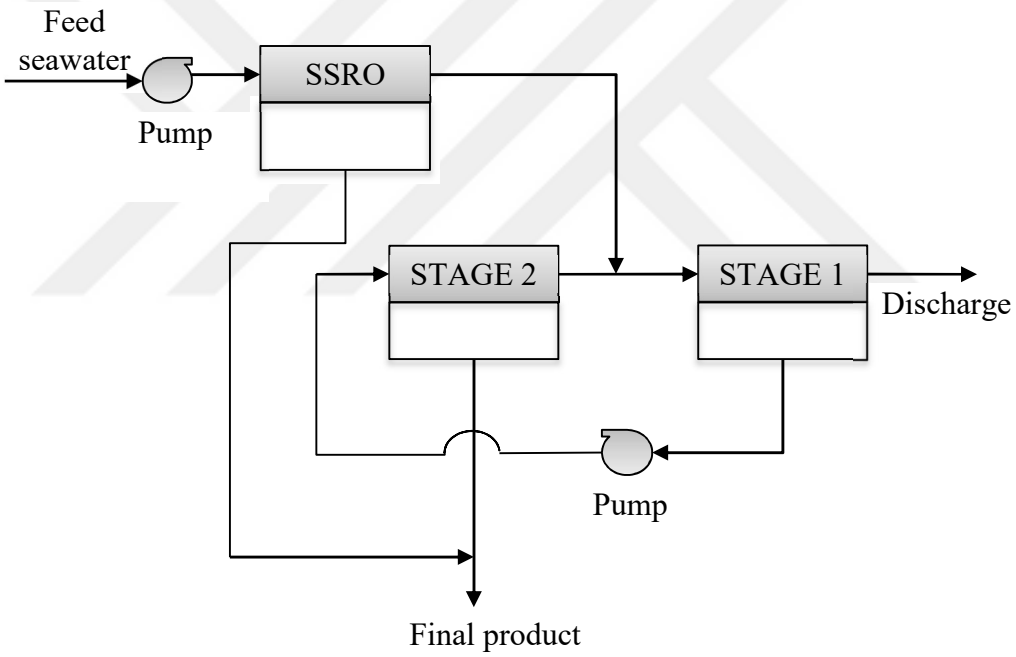

Figure 2.5 : Schematic representation of the EERO desalination process, reproduced from [43].

Chong et al. [43] studied 3-stage EERO (consists of 2-stage CMCR) and 4-stage EERO (consists of 3-stage CMCR). In the study, they produced water with a concentration of 350 ppm using a seawater feed with a concentration of 35,000 ppm. 75% overall recovery was obtained at 111 bar with SSRO, at 74 bar with 3-stage EERO, and at 55.5 bar with 4-stage EERO. And also, the SEC_{net} for SSRO, 3-stage EERO, and 4-stage EERO is 3.086 kWh/m^3, 2.746 kWh/m^3 and 3.086 kWh/m^3, respectively. In the 3-stage EERO, there is 33.2% decrease in OPD and 11% decrease in SEC_{net} compared

to SSRO, while 4-stage EERO provides a 50% decrease in OPD compared to SSRO at the same SEC_{net}.

Chong and Krantz [44] performed a study on 2-2 EERO process which includes 2-stage SSRO and 2-stage CMCR and examined the contribution of the number of the stages in the SSRO and the CMCR to the OPD and SEC. They found out that the OPD of the SSRO system is higher than those of 1-2 EERO, 2-2 EERO, and 1-3 EERO processes, and 1-3 EERO process has the lowest OPD at the same recovery. Besides, they observed that the increased number of SSROs does not change the OPD of the system; however, it leads to a decline in SEC. Beyond the point where total fractional water recovery is 0.6, the 2-2 EERO system has the lower SEC than that of SSRO, but the SEC of SSRO system is higher than the SEC of 1-2 EERO nearly at the point of 0.65.

2.5.5 A novel concurrent desalination and boron removal (CDBR) process

The EERO system reduces OPD according to conventional RO processes but cannot acquire a permeate containing 0.5 ppm boron. The concurrent desalination and boron removal (CDBR) process is a system to provide both boron and salt removal simultaneously with high water recovery running at a lower pressure than other RO processes without changing the pH, and using merely membrane processes [16].

The CDBR combines a CMCR with an SSRO and LPMS (Figure 2.6). The CMCR uses the retentate from the terminal stage of SSRO as feed. The CMCR might involve extra stages so as to increase water recovery. However, the product water from neither the CMCR nor the SSRO has 0.5 ppm boron level that the WHO suggested. Therefore, a portion of the permeate from the SSROs is pumped into the low-pressure membrane stage (LPMS), while the remaining permeate is mixed with the product of the CMCR and the LPMS. Thus, the amount of permeate transferred to the LPMS reduces to reach the required boron concentration [16].

Kürklü et al. [16] designed the CDBR process to produce potable water containing 0.5 ppm boron and less than 100 ppm salt, from seawater containing 10 ppm boron and 35,000 ppm salt. In the study, they used a pump and ERD with an efficiency of 0.85 and 0.90, respectively, with membranes providing salt rejection of 0.997 and boron rejection of 0.900. For individual processes, they calculated the OPD and SEC_{net} for 1-3 CDBR, 2-3 CDBR, and 3-3 CDBR having different split fractions, and they

compared the CDBR performance to that of the two-pass configuration. 1-3 CDBR, 2-3 CDBR, 3-3 CDBR processes, respectively, possess an SSRO, two and three SSRO stages connected in series, in addition to them the CDBR processes contain an LPMS and a 2-stage CMCR. For the 1-3 CDBR, 2-3 CDBR and 3-3 CDBR processes working at a recovery value of 65%, the net specific energy consumption values were 3.10 kWh/m^3, 2.86 kWh/m^3 and 2.70 kWh/m^3, respectively.

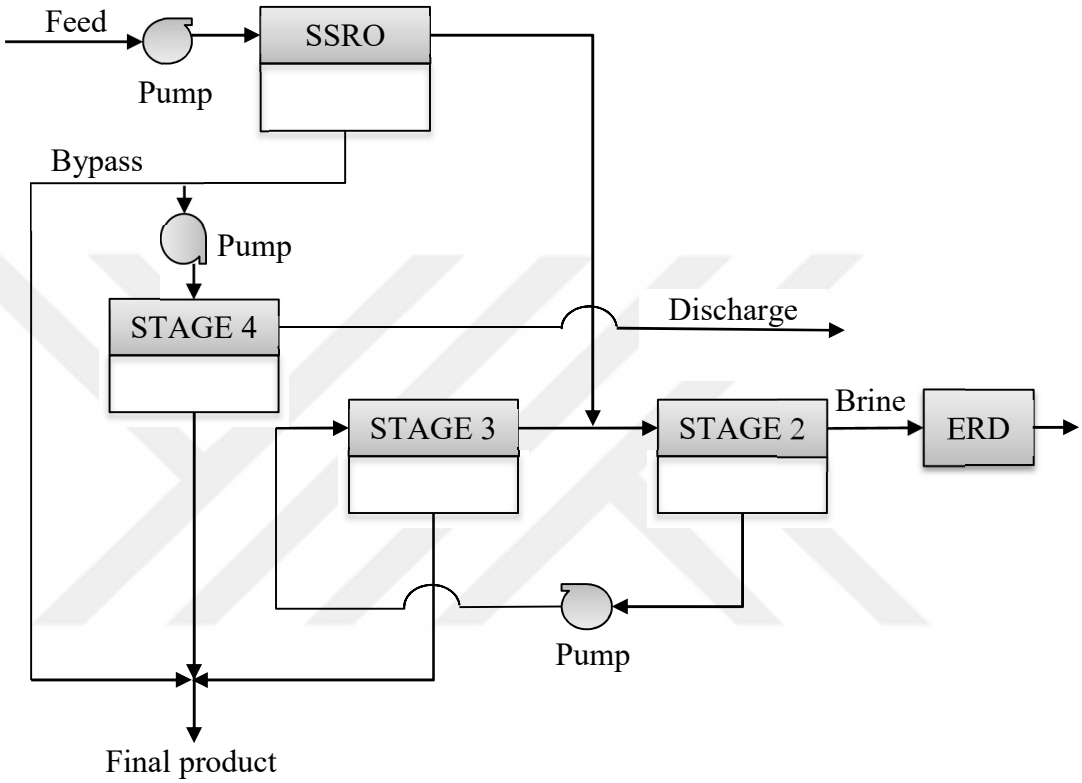

Figure 2.6 : Schematic representation of the CDBR desalination process, reproduced from [16].

3. CONCEPTIONAL DESIGN OF TWO-PASS AND CDBR PROCESSES

In this section, a two-pass RO system with concentrate recycling, in which the brine solution of the second pass recycles to the feed side of the first pass, was designed so as to produce water containing 0.5 ppm boron and maximum 100 ppm salt.

Afterward, a revised configuration of the concurrent desalination and boron removal (CDBR) process was modeled to lessen both boron and salt concentrations in the product water to 0.5 ppm and less than 100 ppm, respectively.

Here, the impact of the concentrate recycling on SEC_{net} of the system was investigated using mathematical analysis. It was projected that these conceptional designs would benefit in determining the operating performances and limits of the actual two-pass and CDBR processes.

3.1 Design Characteristics

According to Suer et al. [16], the below assumptions are needed to be considered in order to model both of the two-pass and CDBR reverse osmosis systems.

- The raw feed water contains 35,000 ppm salt and 10 ppm boron.
- The specified rejections for salt and boron are the values for the standard RO membranes.
- Pressure drop throughout the flow tubes and membranes are omitted.
- The osmotic pressure varies linearly with concentration.
- The influences of the concentration polarization and membrane fouling is not regarded.
- The efficiencies of pump and ERD are specified as 0.85 and 0.90, respectively.
- The retentate recycling is provided with the usage of RO and nanofiltration (NF) membranes having low rejection performance.

- All components of the CDBR system including the SSRO stages, CMCR, and LPMS function at the thermodynamic limits.
- The decrease in the fixed and maintenance costs due to the function at lower operating pressure are not evaluated.
- The decline in the costs of pre-treatment and brine disposal due to the function at the higher water recovery is not regarded.

The average salt rejection value for commercial SWRO membranes from different manufacturers are around 99.7% [45]. Additionally, boron rejection can reach up to 90% at natural pH for commercial membranes providing 99.7% salt rejection [46].

For the CDBR process, because the boron concentration in water is lower than that of salt, it is assumed that only the salt concentration determines the OPD. The linear relation between the OPD and concentration is given by $\pi_i = 0.801 C_{is}$, where π_i is the osmotic pressure (bar), C_{is} is the salt concentration (g/L) [16].

The high-pressure pumps and ERDs operating with 85% and 90% efficiencies, respectively, are appropriate for the CDBR process [16].

The use of RO or NF membranes which provide poor rejection performance enables the pass of some salt through the film, resulting in a reduction of the OPD. Otherwise, in case stage 2 of the CDBR process (shown in Figure 2.5) employs a membrane which provides a high salt rejection, then the recovery decreases, and to attain the desired recovery, the pressure should augment [16].

Concentration polarization and membrane fouling, which are quite common case in RO processes, will be a significant issue in the real performance of the CDBR process. Maintaining one or more safety factors in the SSRO and CMCR stages minimizes these problems. The safety factor is found dividing the retentate volumetric flow by the permeate volumetric flow [16].

According to Zhu et al. [47], the RO system operates close to the thermodynamic limits [47]. The thermodynamic limit means that transmembrane pressure difference across the membrane is equal to the osmotic pressure differential [16]. It can be accepted that new generation membranes reach the thermodynamic limits [47].

The calculation methodology is the same for all processes. In these systems, Q and C designated as the volumetric flow rate in L/h, concentration in g/L, respectively.

Furthermore, f, s, and b signify the abbreviation of feed, salt, and boron, respectively. Using Excel solver, the following steps are followed to obtain the solution:

- The total and solute material balances are established for the individual stages separately.
- The degree of freedom of all system is calculated. Doing so, the parameters that need to be known to solve the equations are specified.
- Then, all unknown parameters in the process including the volumetric flow rates, the salt and boron concentrations, and rejection values, the recovery of each stage are calculated for various total fractional water recovery values.
- After founding all the numerical values of each parameter, the gross specific energy consumption (SEC_{gross}), and net specific energy consumption (SEC_{net}) are evaluated.

3.2 Two-pass RO Process with Concentrate Recycling

Employing of conventional two-pass RO systems for desalination processes enables to attain required boron concentration in the product water. However, this process provides low recovery at high pressure. Here, to augment the system recovery, the concentrate of the 2nd pass is conveyed to the feed side of the 1st pass (Figure 3.1).

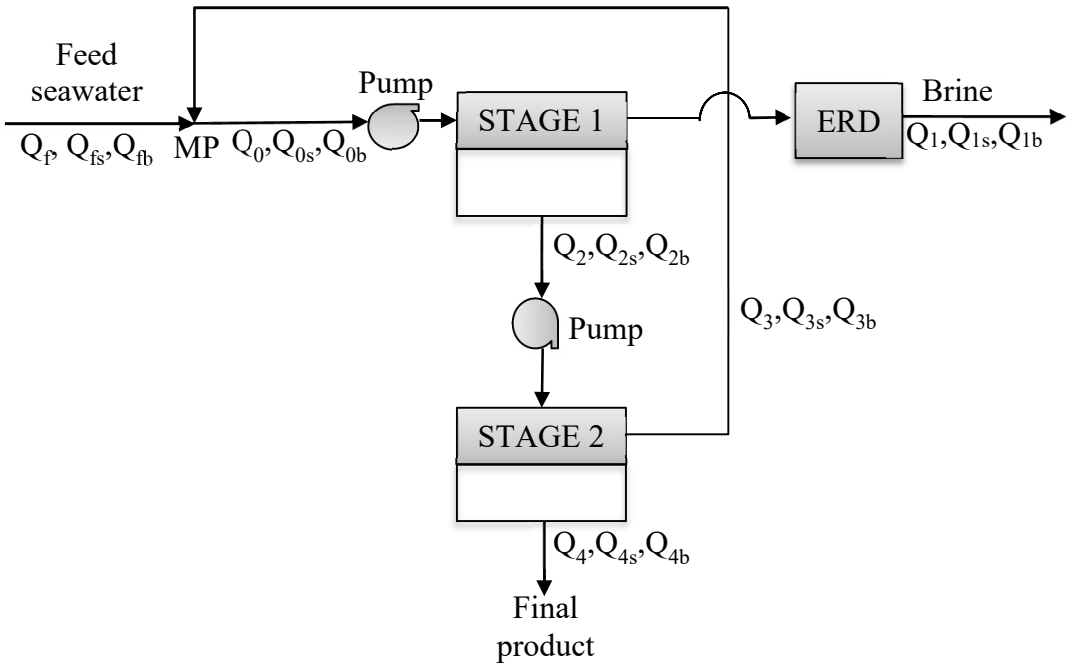

Figure 3.1 : Two-pass RO system with concentrate recycling.

Total and solute material balances for stage 1, stage 2 and the mixing point of the two-pass process are made. If the material balances are established for stage 1, three equations which involve nine unknown variables (Q_0, Q_1, Q_2, C_{os}, C_{1s}, C_{2s}, C_{0b}, C_{1b}, C_{2b}) are obtained. The material balances for stage 2 consist of three equations and six unknowns (Q_3, Q_4, C_{3s}, C_{4s}, C_{3b}, C_{4b}). Finally, in case material balances are established for the mixing point, three equations with three unknowns (Q_f, C_{fs}, C_{fb}) are attained. So the system, entirely, contains nine equations and eighteen unknowns. Namely, this two-pass process has nine-degrees of freedom, these factors are needed to solve the equations.

The input variables in order to satisfy the nine degrees of freedom are listed below.

1) The volumetric flow rate of raw feed water (Q_f)
2) The salt concentration of raw feed water (C_{fs})
3) Boron concentration of raw feed water (C_{fb})
4) Recovery fraction for Pass 2 (Y_2)
5) Total water recovery fraction (Y_{total})
6) Salt rejection in Pass 1 (σ_{1s})
7) Boron rejection in Pass 1 (σ_{1b})
8) Salt rejection in Pass 2, scaled to σ_{2b} (σ_{2s})
9) Boron concentration in desalinated product water (C_{4b})

Total and solute material balances over stage 1 are written as follows:

$$Q_0 = Q_1 + Q_2 \tag{3.1}$$

$$Q_0 C_{0s} = Q_1 C_{1s} + Q_2 C_{2s} \tag{3.2}$$

$$Q_0 C_{0b} = Q_1 C_{1b} + Q_2 C_{2b} \tag{3.3}$$

Total and solute material balances over stage 2 are written as follows:

$$Q_2 = Q_3 + Q_4 \tag{3.4}$$

$$Q_2 C_{2s} = Q_3 C_{3s} + Q_4 C_{4s} \tag{3.5}$$

$$Q_2 C_{2b} = Q_3 C_{3b} + Q_4 C_{4b} \tag{3.6}$$

Total and solute material balances over the mixing point are written as follows:

$$Q_0 = Q_3 + Q_f \tag{3.7}$$

$$Q_0 C_{0s} = Q_3 C_{3s} + Q_f C_{fs} \tag{3.8}$$

$$Q_0 C_{0b} = Q_3 C_{3b} + Q_f C_{fb} \tag{3.9}$$

The recoveries for both stages and the total recovery are written as follows:

$$Y_1 = \frac{Q_2}{Q_0} \tag{3.10}$$

$$Y_2 = \frac{Q_4}{Q_2} \tag{3.11}$$

$$Y_{total} = \frac{Q_4}{Q_f} \tag{3.12}$$

The equations for OPD for individual stages are given as follows:

$$\Delta\pi_1 = K(C_{1s} - C_{2s}) \tag{3.13}$$

$$\Delta\pi_2 = K(C_{3s} - C_{4s}) \tag{3.14}$$

where K=0.801 L.bar/g or 0.0223 kWh.L/m^3 since it is assumed that osmotic pressure varies linearly with concentration [16].

The salt and boron rejections for each stage are calculated as follows:

$$\sigma_{1s} = 1 - \frac{C_{2s}}{C_{0s}} \tag{3.15}$$

$$\sigma_{1b} = 1 - \frac{C_{2b}}{C_{0b}} \tag{3.16}$$

$$\sigma_{2s} = 1 - \frac{C_{4s}}{C_{2s}} \tag{3.17}$$

$$\sigma_{2b} = 1 - \frac{C_{4b}}{C_{2b}} \tag{3.18}$$

Up to this point, all equations that are known for this system were given. The next step is to determine the volumetric flow rates, boron and salt concentrations in individual RO stages. To do this, firstly, all flow rates in the process are evaluated using total material balances.

From Eq. (2.12),

$$Q_4 = Y_{total} Q_f \tag{3.19}$$

Rearranging Eq. (2.11),

$$Q_2 = \frac{Q_4}{Y_2} \tag{3.20}$$

Rearranging Eq. (2.4),

$$Q_3 = Q_2 - Q_4 \tag{3.21}$$

Combining Eqs. (3.1), (3.4) and (3.7),

$$Q_1 = Q_f - Q_4 \tag{3.22}$$

Using Eq. (3.7), Q_0 is determined as well. After all volumetric flow rate calculations, boron concentrations in individual stages are determined using boron material-balance equations.

Then,

Combining Eqs. (3.3), (3.6) and (3.9),

$$C_{1b} = \frac{Q_f C_{fb} - Q_4 C_{4b}}{Q_1} \tag{3.23}$$

Combining Eqs. (3.3), and (3.16),

$$C_{0b} = \frac{Q_1 C_{1b}}{Q_0 - (1 - \sigma_{1b}) Q_2} \tag{3.24}$$

Rearranging Eq. (3.16),

$$C_{2b} = C_{0b}(1 - \sigma_{1b}) \tag{3.25}$$

Rewriting Eq. (3.9),

$$C_{3b} = \frac{Q_2 C_{2b} - Q_4 C_{4b}}{Q_3} \tag{3.26}$$

So far, all boron concentrations were determined. Boron rejection for the second pass (σ_{2b}) is found substituting the known values (C_{2b} and C_{4b}) into Eq. (3.18). The salt rejection in the second RO pass (σ_{2s}) is calculated by being scaled to the boron rejection in the same pass (σ_{2b}) as follows:

$$\sigma_{2s} = \frac{\sigma_{1s}\sigma_{2b}}{\sigma_{1b}} \tag{3.27}$$

Combining Eqs. (3.2), and (3.15),

$$C_{1s} = \frac{[Q_0 - Q_2(1 - \sigma_{1s})]C_{0s}}{Q_1} \tag{3.28}$$

Combining Eqs. (3.2), (3.5), (3.8), (3.15), (3.17) and (3.28),

$$C_{0s} = \frac{Q_f C_{fs}}{Q_0 + (1 - \sigma_{1s})[Q_4(1 - \sigma_{2s}) - Q_2]} \tag{3.29}$$

Rearranging Eq. (3.15),

$$C_{2s} = C_{0s}(1 - \sigma_{1s}) \tag{3.30}$$

Rearranging Eq. (3.17),

$$C_{4s} = C_{2s}(1 - \sigma_{2s}) \tag{3.31}$$

Rewriting Eq. (3.5),

$$C_{3s} = \frac{Q_2 C_{2s} - Q_4 C_{4s}}{Q_3} \tag{3.32}$$

Hence, all the unknown variables of the system are determined.

Then,

The gross specific energy consumption (SEC_{gross}) is written as follows:

$$SEC_{gross} = \frac{Q_f \Delta\pi_1 + Q_3(\Delta\pi_1 - \Delta\pi_2)}{\eta_p Q_4} \tag{3.33}$$

The net specific energy consumption (SEC_{net}) is written as follows:

$$SEC_{net} = SEC_{gross} - \frac{\eta_{ERD} Q_1 \Delta\pi_1}{Q_4} \tag{3.34}$$

3.3 A Revised Configuration of the CDBR processes

The CDBR process, which combines a CMCR system with one or more SSRO stages and LPMS, was created to reach the desired boron concentration in the product water [16]. However, in this study, unlike the initial configuration of the CDBR process, the concentrate of LPMS is conveyed as a feed stream to the SSRO system. In addition to concentrate recycling, this revised configuration of the CDBR process was designed as a split-partial system in order to observe the effect of bypass on the overall recovery of the system and final product boron concentration.

Figure 3.2 demonstrates the revised 1-3 CDBR process with split partial design. In this thesis, firstly, the equations for the 1-3 CDBR process was solved. Then according to this process, the unknown variables are determined for the 2-3 CDBR and 3-3 CDBR processes, respectively. Since these processes apart from 1-3 CDBR process have two or three SSRO stages, the other equations determined for the CMCR and LPMS part are the same in each CDBR system. Hence, the effect adding or removing SSRO stage to the system on total water recovery fraction or specific energy consumption is figured out.

Total and solute material balances are established for each stage and mixing point in the CDBR processes. Material balances for the 1-3 CDBR process consist of nineteen equations containing thirty-five unknowns. That indicates sixteen-degrees of freedom, and all equations can be solved specifying the sixteen input variables. The material balances established for 2-3 CDBR process consist of twenty-five equations with forty-four unknowns, and nineteen input variables are needed to be determined. When the mass balances are made for the 3-3 CDBR process, twenty-eight equations with fifty unknowns are produced, and the number of input variables needed to be defined is twenty-two.

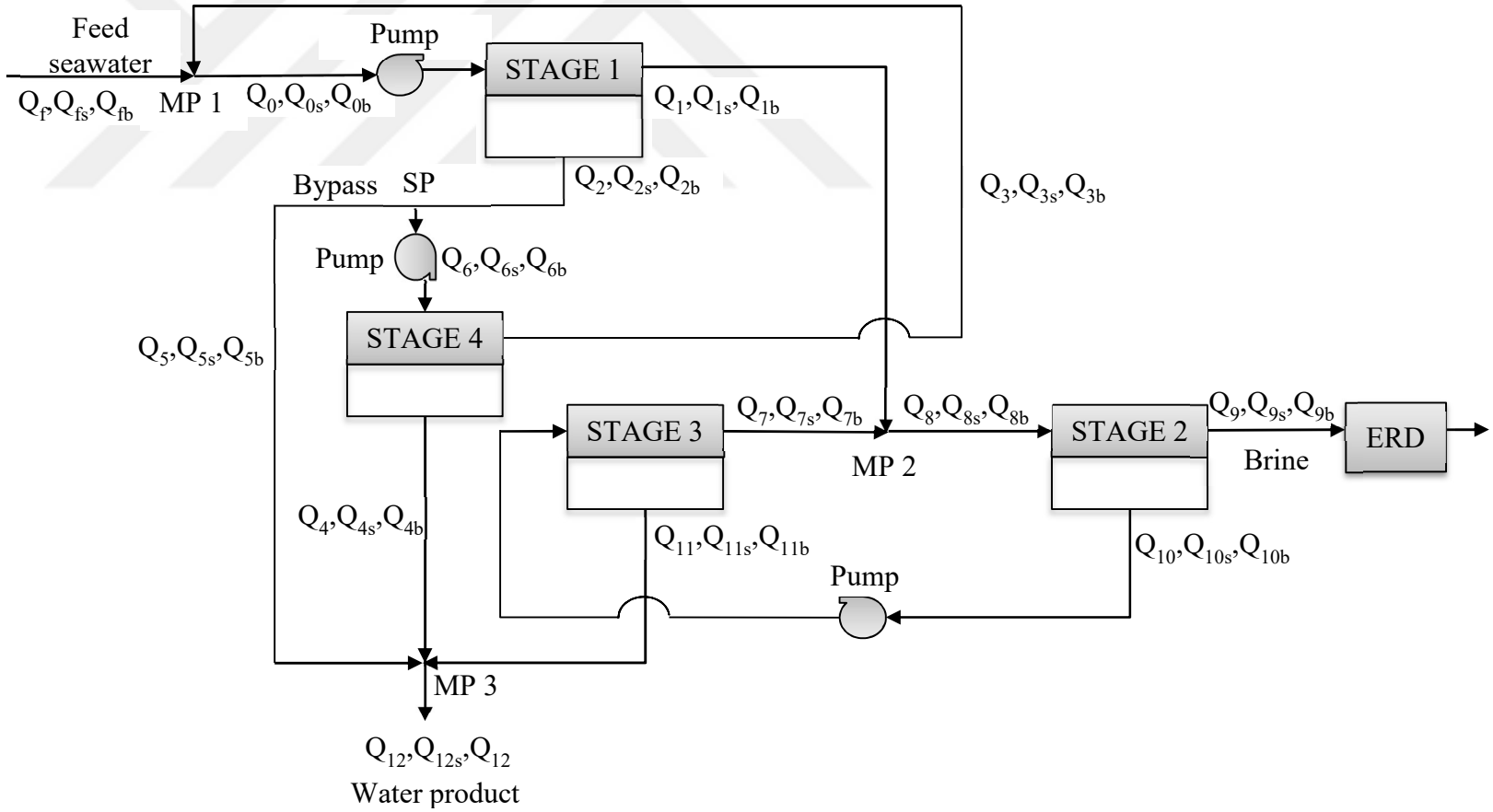

Figure 3.2 : Schematic representation of the revised configuration of CDBR process.

The input variables in order to satisfy the degrees of freedom for each process are listed below.

For the 1-3 CDBR process,

1) The volumetric flow rate of raw feed water (Q_f)
2) The salt concentration of raw feed water (C_{fs})
3) Boron concentration of raw feed water (C_{fb})
4) Recovery fraction for stage 2 (Y_2)
5) Recovery fraction for stage 3 (Y_3), $Y_2+Y_3=1$ in order to ensure $C_{1s}=C_{7s}$
6) Recovery fraction for stage 4 (Y_4)
7) Total water recovery fraction (Y_{total})
8) Salt rejection in SSRO (σ_{1s})
9) Boron rejection in SSRO (σ_{1b})
10) Boron rejection of stage 2 (σ_{2b}), scaled to σ_{2s}
11) Boron rejection of stage 3 (σ_{3b}), scaled to σ_{3s}
12) Salt rejection of stage 4 (σ_{4s}), scaled to σ_{4b}
13) Boron concentration in product water (C_{12b})
14) Equal salt concentrations in permeates of the final SSRO stage and Stage 3 ($C_{2s}=C_{11s}$)
15) OPD of stage 2 equal to that of stage 3 ($\Delta\pi_2 = \Delta\pi_3$)
16) Bypass split fraction (S)

For the 2-3 CDBR process,

17) Recovery for stage 1A (Y_{1A})
18) Salt rejection for the second SSRO stage (σ_{1s})
19) Boron rejection for the second SSRO stage (σ_{1b})

For 2-3 CDBR process,

20) Recovery for stage 1C (Y_{1C})
21) Salt rejection for the final stage of SSRO (σ_{1s})

22) Boron rejection in the final stage of SSRO (σ_{1b})

Total and solute material balances over stage 1 are written as follows:

$$Q_0 = Q_1 + Q_2 \tag{3.35}$$

$$Q_0 C_{0s} = Q_1 C_{1s} + Q_2 C_{2s} \tag{3.36}$$

$$Q_0 C_{0b} = Q_1 C_{1b} + Q_2 C_{2b} \tag{3.37}$$

Total and solute material balances over stage 2 are written as follows:

$$Q_8 = Q_9 + Q_{10} \tag{3.38}$$

$$Q_8 C_{8s} = Q_9 C_{9s} + Q_{10} C_{10s} \tag{3.39}$$

$$Q_8 C_{8b} = Q_9 C_{9b} + Q_{10} C_{10b} \tag{3.40}$$

Total and solute material balances over stage 3 are written as follows:

$$Q_{10} = Q_7 + Q_{11} \tag{3.41}$$

$$Q_{10} C_{10s} = Q_7 C_{7s} + Q_{11} C_{11s} \tag{3.42}$$

$$Q_{10} C_{10b} = Q_7 C_{7b} + Q_{11} C_{11b} \tag{3.43}$$

Total and solute material balances over stage 4 are written as follows:

$$Q_6 = Q_3 + Q_4 \tag{3.44}$$

$$Q_6 C_{6s} = Q_3 C_{3s} + Q_4 C_{4s} \tag{3.45}$$

$$Q_6 C_{6b} = Q_3 C_{3b} + Q_4 C_{4b} \tag{3.46}$$

Total and solute material balances over mixing point 1 are written as follows:

$$Q_0 = Q_f + Q_3 \tag{3.47}$$

$$Q_0 C_{0s} = Q_f C_{fs} + Q_3 C_{3s} \tag{3.48}$$

$$Q_0 C_{0b} = Q_f C_{fb} + Q_3 C_{3b} \tag{3.49}$$

Total and solute material balances over mixing point 2 are written as follows:

$$Q_8 = Q_1 + Q_7 \tag{3.50}$$

$$Q_8 C_{8s} = Q_1 C_{1s} + Q_7 C_{7s} \tag{3.51}$$

$$Q_8 C_{8b} = Q_1 C_{1b} + Q_7 C_{7b} \tag{3.52}$$

Total and solute material balances over mixing point 3 are written as follows:

$$Q_{12} = Q_4 + Q_5 + Q_{11} \tag{3.53}$$

$$Q_{12} C_{12s} = Q_4 C_{4s} + Q_5 C_{5s} + Q_{11} C_{11s} \tag{3.54}$$

$$Q_{12} C_{12b} = Q_4 C_{4b} + Q_5 C_{5b} + Q_{11} C_{11b} \tag{3.55}$$

Total material balance over the split point is written as follows:

$$Q_2 = Q_5 + Q_6 \tag{3.56}$$

The recoveries for individual stages, the total recovery and the split ratio are written as follows:

$$Y_1 = \frac{Q_2}{Q_0} \tag{3.57}$$

$$Y_2 = \frac{Q_{10}}{Q_8} \tag{3.58}$$

$$Y_3 = \frac{Q_{11}}{Q_{10}} \tag{3.59}$$

$$Y_4 = \frac{Q_4}{Q_6} \tag{3.60}$$

$$Y_{total} = \frac{Q_{12}}{Q_f} \tag{3.61}$$

$$S = \frac{Q_5}{Q_2} \tag{3.62}$$

The salt and boron rejections for each stage are evaluated as follows:

$$\sigma_{1s} = 1 - \frac{C_{2s}}{C_{0s}} \tag{3.63}$$

$$\sigma_{1b} = 1 - \frac{C_{2b}}{C_{0b}} \tag{3.64}$$

$$\sigma_{2s} = 1 - \frac{C_{10s}}{C_{8s}} \tag{3.65}$$

$$\sigma_{2b} = 1 - \frac{C_{10b}}{C_{8b}} \tag{3.66}$$

$$\sigma_{3s} = 1 - \frac{C_{11s}}{C_{10s}} \tag{3.67}$$

$$\sigma_{3b} = 1 - \frac{C_{11b}}{C_{10b}} \tag{3.68}$$

$$\sigma_{4s} = 1 - \frac{C_{4s}}{C_{6s}} \tag{3.69}$$

$$\sigma_{4b} = 1 - \frac{C_{4b}}{C_{6b}} \tag{3.70}$$

The equations for OPD for each stage are expressed as follows:

$$\Delta\pi_1 = K(C_{1s} - C_{2s}) \tag{3.71}$$

$$\Delta\pi_2 = K(C_{9s} - C_{10s}) \tag{3.72}$$

$$\Delta\pi_3 = K(C_{7s} - C_{11s}) \tag{3.73}$$

$$\Delta\pi_4 = K(C_{3s} - C_{4s}) \tag{3.74}$$

Firstly, all volumetric flow rates in the system are determined using entire material balances as shown in two-pass RO design.

From Eq. (3.61),

$$Q_{12} = Y_{total} Q_f \tag{3.75}$$

Eqs. (3.35), (3.38), (3.41), (3.44), (3.47), (3.50), (3.53) and (3.56) are combined to obtain the overall material-balance equation. Then, using this equation,

$$Q_9 = Q_f - Q_{12} \tag{3.76}$$

Substituting Eq. (3.38) into (3.58),

$$Q_8 = \frac{Q_9}{(1 - Y_2)} \tag{3.77}$$

Rewriting Eq. (3.58),

$$Q_{10} = Y_2 Q_8 \tag{3.78}$$

Combining Eqs. (3.78) and (3.59),

$$Q_{11} = Y_2 Y_3 Q_8 \tag{3.79}$$

Combine Eqs. (3.79), (3.41) and (3.58),

$$Q_7 = Y_2(1 - Y_3)Q_8 \tag{3.80}$$

Substituting Eq. (3.80) into (3.50),

$$Q_1 = [1 - Y_2(1 - Y_3)]Q_8 \tag{3.81}$$

From Eq. (3.60),

$$Q_4 = Y_4 Q_6 \tag{3.82}$$

Substituting Eq. (3.82) into (3.44),

$$Q_3 = (1 - Y_4)Q_6 \tag{3.83}$$

Substituting Eq. (3.83) into (3.47)

$$Q_0 = Q_f + (1 - Y_4)Q_6 \tag{3.84}$$

Combining Eqs. (3.56) and (3.62),

$$Q_2 = \frac{Q_6}{1 - S} \tag{3.85}$$

Substituting Eqs. (3.81), (3.84) and (3.85) into (3.35),

$$Q_6 = \frac{[1 - Y_2(1 - Y_3)]Q_8 - Q_f}{1 - Y_4 - \frac{1}{1 - S}} \tag{3.86}$$

Combining Eqs. (3.62) and (3.85),

$$Q_5 = \left(\frac{S}{1 - S}\right) Q_6 \tag{3.87}$$

All equations required to calculate the flow rates of the entire process were found out. When solving the salt and boron equations together, the concentration values needed for each stage are determined.

From Eq. (3.51),

$$C_{12s} = \frac{Q_1 C_{1s} + Q_7 C_{7s}}{Q_{12}} \tag{3.88}$$

From Eq. (3.39),

$$C_{10s} = \frac{Q_8 C_{8s} - Q_9 C_{9s}}{Q_{10}} \tag{3.89}$$

Substituting Eqs. (3.88) and (3.89) into (3.39),

$$C_{9s} = \frac{Q_1 C_{1s} - Q_{11} C_{11s}}{Q_9} \tag{3.90}$$

Equalizing Eq. (3.72) to (3.73),

$$C_{9s} - C_{10s} = C_{7s} - C_{11s} \tag{3.91}$$

Combining Eqs. (3.88), (3.90) and (3.91),

$$C_{7s} = \frac{C_{1s}Q_1(Q_{10} - Q_9) + Q_9^2 C_{9s} + C_{11s}Q_{10}(Q_9 - Q_{11})}{Q_9(Q_{10} + Q_7)} \tag{3.92}$$

Combining Eqs. (3.77), (3.78), (3.79), (3.80), (3.81), (3.91) and (3.92),

$$C_{7s} = \frac{C_{1s}(1 - Y_2 + Y_2Y_3) + C_{11s}(1 - Y_2 - Y_3)}{(1 - Y_2)(2 - Y_3)} \tag{3.93}$$

If C_{1s} is equal to C_{7s}, then,

$$\frac{1 - Y_2 + Y_2Y_3}{(1 - Y_2)(2 - Y_3)} = 1 \tag{3.94}$$

$$Y_2 + Y_3 = 1 \tag{3.95}$$

That ensuring both stages of the CMCR function at the equal OPDs avoids using the inter-stage pumps ($\Delta\pi_2 = \Delta\pi_3$). Similarly, in order to avert the use of additional pumps between the CMCR and SSRO stages, the OPD of them should be equal to each other. That is, the permeate salt concentrations produced from both components of the CDBR must be equalized ($C_{2s} = C_{11}$). In case C_{2s} is equal to C_{11s}, by equalizing the OPDs of stage 1 to 2 as indicated in Eqs. (3.71) and (3.73), it is concluded that C_{1s} is equal to C_{7s}. Namely, at the point where the SSRO retentate meets the CMCR, the entropy of mixing is zero.

As seen in the equation above, in order to obtain the equality of ($C_{1s} = C_{7s}$), the sum of Y_2 and Y_3 must be equal to 1. Given that the safety factor must be at least 1, then, Y_2 and Y_3 must be equal to 0.5.

Combining Eq. (3.51) with $C_{1s} = C_{7s}$ equality,

$$C_{8s} = C_{1s} \tag{3.96}$$

Substituting Eq. (3.65) into (3.36),

$$C_{1s} = \frac{C_{0s}[Q_0 - Q_2(1 - \sigma_{1s})]}{Q_1} \tag{3.97}$$

Substituting Eqs. (3.65) and (3.97) into (3.42),

$$C_{10s} = \frac{Q_{11}C_{0s}(1 - \sigma_{1s}) + \frac{Q_7}{Q_1}C_{0s}[Q_0 - Q_2(1 - \sigma_{1s})]}{Q_{10}} \tag{3.98}$$

Substituting Eqs. (3.97) and (3.98) into (3.65), and also it's known that C_{1s}, C_{7s} and C_{8s} are equal to each other. Then,

$$\sigma_{2s} = 1 - \frac{Q_1\left[Q_{11}(1 - \sigma_{1s}) + \frac{Q_7}{Q_1}[Q_0 - Q_2(1 - \sigma_{1s})]\right]}{Q_{10}[Q_0 - Q_2(1 - \sigma_{1s})]} \tag{3.99}$$

Substituting Eqs. (3.65) and (3.98) into (3.67),

$$\sigma_{3s} = 1 - \frac{(1 - \sigma_{1s})Q_{10}}{Q_{11}(1 - \sigma_{1s}) + \frac{Q_7}{Q_1}[Q_0 - Q_2(1 - \sigma_{1s})]} \tag{3.100}$$

σ_{2b} and σ_{3b} are obtained using scaling to salt rejections which are found via analysis.

$$\sigma_{2b} = \frac{\sigma_{1b}\sigma_{2s}}{\sigma_{1s}} \tag{3.101}$$

$$\sigma_{3b} = \frac{\sigma_{1b}\sigma_{3s}}{\sigma_{1s}} \tag{3.102}$$

Rearranging Eq. (3.66),

$$C_{10b} = C_{8b}(1 - \sigma_{2b}) \tag{3.103}$$

Rearranging Eq. (3.68),

$$C_{11b} = C_{10b}(1 - \sigma_{3b}) \tag{3.104}$$

Using total boron material balance,

$$C_{9b} = \frac{Q_f C_{fb} - Q_{12} C_{12b}}{Q_9} \tag{3.105}$$

Substituting Eqs. (3.103) and (3.104) into (3.67),

$$C_{8b} = \frac{Q_9 C_{9b}}{Q_8 - Q_{10}(1 - \sigma_{2b})} \tag{3.106}$$

Substituting Eqs. (3.105) and (3.106) into (3.43),

$$C_{7b} = \frac{Q_{10} C_{10b} - Q_{11} C_{11b}}{Q_7} \tag{3.107}$$

From Eq. (3.52),

$$C_{1b} = \frac{Q_8 C_{8b} - Q_7 C_{7b}}{Q_1} \tag{3.108}$$

Rearranging Eq. (3.64),

$$C_{2b} = C_{0b}(1 - \sigma_{1b}) \tag{3.109}$$

Combining Eqs. (3.37) and (3.109),

$$C_{0b} = \frac{Q_1 C_{1b}}{Q_0 - Q_2(1 - \sigma_{1b})} \tag{3.110}$$

If S is not zero, all boron concentrations before and after the split point are equal to each other ($C_{2b} = C_{5b} = C_{6b}$). In case S is equal to zero then C_{5b} must be equal to zero.

From Eq. (3.55),

$$C_{4b} = \frac{Q_{12} C_{12b} - Q_{11} C_{11b} - Q_5 C_{5b}}{Q_4} \tag{3.111}$$

From Eq. (3.49),

$$C_{3b} = \frac{Q_0 C_{0b} - Q_f C_{fb}}{Q_3} \tag{3.112}$$

From Eq. (3.70), σ_{4b} can be calculated,

$$\sigma_{4b} = \frac{C_{2b} - C_{4b}}{C_{2b}} \tag{3.113}$$

Scaling,

$$\sigma_{4s} = \frac{\sigma_{1s}\sigma_{4b}}{\sigma_{1b}} \tag{3.114}$$

Substituting Eq. (3.69) into (3.45),

$$C_{3s} = \frac{C_{6s}[Q_6 - Q_4(1 - \sigma_{4s})]}{Q_3} \tag{3.115}$$

Substituting Eqs. (3.65) into (3.36) and using $C_{1s} = C_{8s}$ equality,

$$C_{0s} = \frac{Q_1 C_{8s}}{Q_0 - Q_2(1 - \sigma_{1s})} \tag{3.116}$$

Substituting Eqs. (2.81) and (2.82) into (2.14),

$$C_{6s} = \frac{\left[\frac{Q_1 Q_0}{Q_0 - Q_2(1 - \sigma_{1s})}\right] C_{8s} - Q_f C_{fs}}{Q_6 - Q_4(1 - \sigma_{4s})} \tag{3.117}$$

Substituting Eqs. (3.116) and (3.117) into (3.43), using $C_{2s} = C_{6s}$, equality then,

$$C_{8s} = \frac{Q_f C_{fs}[Q_0 - Q_2(1 - \sigma_{1s})]}{Q_1[Q_0 - (Q_6 - Q_4(1 - \sigma_{4s}))(1 - \sigma_{1s})]} \tag{3.118}$$

Rearranging Eq. (3.69),

$$C_{4s} = C_{6s}(1 - \sigma_{4s}) \tag{3.119}$$

From Eq. (3.39),

$$C_{9s} = \frac{Q_8 C_{8s} - Q_{10} C_{10s}}{Q_9} \tag{3.120}$$

From Eq. (3.54),

$$C_{12s} = \frac{Q_4 C_{4s} + Q_{11} C_{11s} + Q_5 C_{5s}}{Q_{12}} \tag{3.121}$$

The gross specific energy consumption (SEC_{gross}) is written as follows:

$$SEC_{gross} = \frac{Q_f \Delta\pi_1 + Q_3(\Delta\pi_1 - \Delta\pi_4) + Q_{10}\Delta\pi_3 + Q_6\Delta\pi_4}{\eta_p Q_{12}} \tag{3.122}$$

The net specific energy consumption (SEC_{net}) is written as follows:

$$SEC_{net} = SEC_{gross} - \frac{\eta_{ERD} Q_9 \Delta\pi_2}{Q_{12}} \tag{3.123}$$

So far, 1-3 CDBR process was analyzed. However, the 2-3 CBDR process includes two SSRO stages connected in series unlike the previous CDBR system. Therefore, for the 2-3 CDBR process, it is sufficient to establish mass balances between SSRO stages only (Figure 3.3).

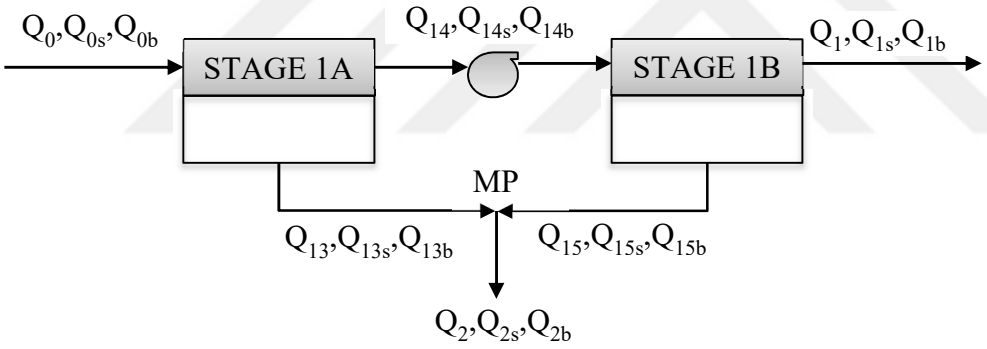

Figure 3.3 : Schematic representation of the two-stage RO system.

Total and solute material balances over stage 1A are written as follows:

$$Q_0 = Q_{13} + Q_{14} \tag{3.124}$$

$$Q_0 C_{0s} = Q_{13} C_{13s} + Q_{14} C_{14s} \tag{3.125}$$

$$Q_0 C_{0b} = Q_{13} C_{13b} + Q_{14} C_{14b} \tag{3.126}$$

Total and solute material balances over stage 1B are written as follows:

$$Q_{14} = Q_1 + Q_{15} \tag{3.127}$$

$$Q_{14}C_{14s} = Q_1C_{1s} + Q_{15}C_{15s} \tag{3.128}$$

$$Q_{14}C_{14b} = Q_1C_{1b} + Q_{15}C_{15b} \tag{3.129}$$

Total and solute material balances on the permeate mixing point are written as follows:

$$Q_2 = Q_{13} + Q_{15} \tag{3.130}$$

$$Q_2C_{2s} = Q_{13}C_{13s} + Q_{15}C_{15s} \tag{3.131}$$

$$Q_2C_{2b} = Q_{13}C_{13b} + Q_{15}C_{15b} \tag{3.132}$$

The recoveries for each SSRO stage are rewritten as follows:

$$Y_{1A} = \frac{Q_{13}}{Q_0} \tag{3.133}$$

$$Y_{1B} = \frac{Q_{15}}{Q_{14}} \tag{3.134}$$

The equations for OPD for each stage are expressed as follows:

$$\Delta\pi_{1A} = K(C_{14s} - C_{13s}) = [K(C_{0s} - C_{13s})\Delta\pi_{1B}]^{0.5} \tag{3.135}$$

$$\Delta\pi_{1B} = K(C_{1s} - C_{15s}) \tag{3.136}$$

The salt and boron rejections for each SSRO stage are calculated as follows:

$$\sigma_{1s} = 1 - \frac{C_{13s}}{C_{0s}} \tag{3.137}$$

$$\sigma_{1b} = 1 - \frac{C_{13b}}{C_{0b}} \tag{3.138}$$

$$\sigma_{1s} = 1 - \frac{C_{15s}}{C_{14s}} \tag{3.139}$$

$$\sigma_{1b} = 1 - \frac{C_{15b}}{C_{14b}} \tag{3.140}$$

Firstly, all volumetric flow rates in the two-stage RO system are evaluated using the total material balances of stage 1A and 1B.

From Eq. (3.133),

$$Q_{13} = Y_{1A} Q_0 \tag{3.141}$$

Rewriting Eq. (3.130),

$$Q_{15} = Q_2 - Q_{13} \tag{3.142}$$

Rewriting Eq. (3.124),

$$Q_{14} = Q_0 - Q_{13} \tag{3.143}$$

Rearranging Eq. (3.137),

$$C_{13s} = C_{0s}(1 - \sigma_{1s}) \tag{3.144}$$

Rewriting Eq. (3.125),

$$C_{14s} = \frac{Q_0 C_{0s} - Q_{13} C_{13s}}{Q_{14}} \tag{3.145}$$

Rewriting Eq. (3.128),

$$C_{15s} = \frac{Q_{14} C_{14s} - Q_1 C_{1s}}{Q_{15}} \tag{3.146}$$

Rearranging Eq. (3.138),

$$C_{13b} = C_{0b}(1 - \sigma_{1b}) \tag{3.147}$$

Rewriting Eq. (3.126)

$$C_{14b} = \frac{Q_0 C_{0b} - Q_{13} C_{13b}}{Q_{14}} \tag{3.148}$$

Rewriting Eq. (3.129),

$$C_{15b} = \frac{Q_{14}C_{14b} - Q_1C_{1b}}{Q_{15}} \tag{3.149}$$

The gross specific energy consumption (SEC_{gross}) is written as follows:

$$SEC_{gross} = \frac{\begin{matrix} Q_f\Delta\pi_{1A} + Q_3(\Delta\pi_{1A} - \Delta\pi_4) + Q_{14}(\Delta\pi_{1B} - \Delta\pi_{1A}) + \\ Q_{10}\Delta\pi_3 + Q_6\Delta\pi_4 \end{matrix}}{\eta_p Q_{12}} \tag{3.150}$$

The net specific energy consumption (SEC_{net}) is written as follows:

$$SEC_{net} = SEC_{gross} - \frac{\eta_{ERD}Q_9\Delta\pi_2}{Q_{12}} \tag{3.151}$$

Initially, a random value is given to the recovery of stage 2B to solve the equations. However, there is a ratio between the OPD values of the SSRO stages as indicated in Eq. (135). Therefore, this equality cannot be achieved when substituting randomly assigned recovery value into the equations of OPD. So, Eq. (3.135) is rearranged in order to obtain Eq. (3.152), and the new equation, called θ, is equalized to 1. By making a solution with Excel solver, it is estimated the SEC in the appropriate Y_{2B} by specifying that the θ value is equal to 1 considering the safety factor for individual stages in the constraints menu.

$$\theta = \frac{[K(C_{0s} - C_{13s})\Delta\pi_{1B}]^{0.5}}{K(C_{14s} - C_{13s})} = 1 \tag{3.152}$$

The 3-3 CBDR process includes three SSRO stages linked in series apart from the 2-3 CDBR system. Therefore, as in the 2-3 CDBR system, it is enough to establish mass balances between three SSRO stages only for the 3-3 CDBR process (Figure 3.4).

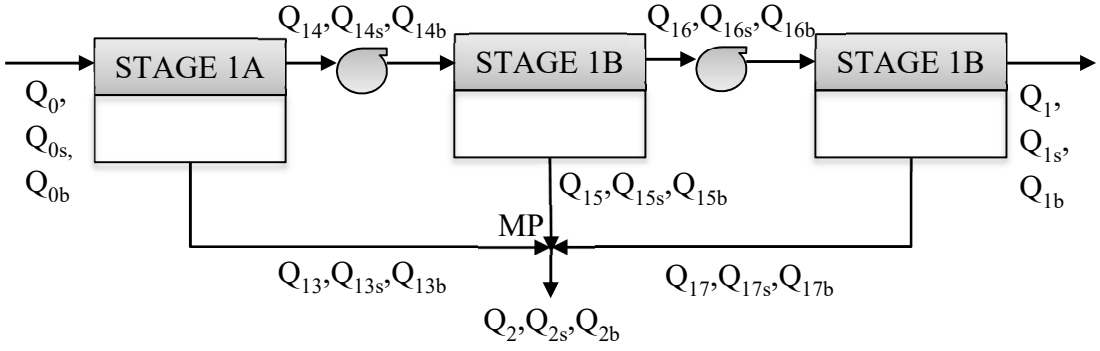

Figure 3.4 : Schematic representation of the three-stage RO system.

Total and solute material balances over stage 1A are written as follows:

$$Q_0 = Q_{13} + Q_{14} \tag{3.153}$$

$$Q_0 C_{0s} = Q_{13} C_{13s} + Q_{14} C_{14s} \tag{3.154}$$

$$Q_0 C_{0b} = Q_{13} C_{13b} + Q_{14} C_{14b} \tag{3.155}$$

Total and solute material balances over stage 1B are written as follows:

$$Q_{14} = Q_{16} + Q_{15} \tag{3.156}$$

$$Q_{14} C_{14s} = Q_{16} C_{16s} + Q_{15} C_{15s} \tag{3.157}$$

$$Q_{14} C_{14b} = Q_{16} C_{16b} + Q_{15} C_{15b} \tag{3.158}$$

Total and solute material balances over stage 1C are written as follows:

$$Q_{16} = Q_1 + Q_{17} \tag{3.159}$$

$$Q_{16} C_{16s} = Q_1 C_{1s} + Q_{17} C_{17s} \tag{3.160}$$

$$Q_{16} C_{16b} = Q_1 C_{1b} + Q_{17} C_{17b} \tag{3.161}$$

Total and solute material balances on the mixing point are written as follows:

$$Q_2 = Q_{13} + Q_{15} + Q_{17} \tag{3.162}$$

$$Q_2 C_{2s} = Q_{13} C_{13s} + Q_{15} C_{15s} + Q_{17} C_{17s} \tag{3.163}$$

$$Q_2 C_{2b} = Q_{13} C_{13b} + Q_{15} C_{15b} + Q_{17} C_{17b} \tag{3.164}$$

The recoveries for each SSRO stage are rewritten as follows:

$$Y_{1A} = \frac{Q_{13}}{Q_0} \tag{3.165}$$

$$Y_{1B} = \frac{Q_{15}}{Q_{14}} \tag{3.166}$$

$$Y_{1C} = \frac{Q_{17}}{Q_{16}} \tag{3.167}$$

The equations for OPD for each stage are expressed as follows:

$$\Delta\pi_{1A} = K(C_{14s} - C_{13s}) = [K(C_{0s} - C_{13s})\Delta\pi_{1C}]^{0.5} \tag{3.168}$$

$$\Delta\pi_{1B} = K(C_{16s} - C_{15s}) \tag{3.169}$$

$$\Delta\pi_{1C} = K(C_{1s} - C_{17s}) \tag{3.170}$$

The salt and boron rejections for each SSRO stage are calculated as follows:

$$\sigma_{1s} = 1 - \frac{C_{13s}}{C_{0s}} \tag{3.171}$$

$$\sigma_{1b} = 1 - \frac{C_{13b}}{C_{0b}} \tag{3.172}$$

$$\sigma_{1s} = 1 - \frac{C_{15s}}{C_{14s}} \tag{3.173}$$

$$\sigma_{1b} = 1 - \frac{C_{15b}}{C_{14b}} \tag{3.174}$$

$$\sigma_{1s} = 1 - \frac{C_{17s}}{C_{16s}} \tag{3.175}$$

$$\sigma_{1b} = 1 - \frac{C_{17b}}{C_{16b}} \tag{3.176}$$

Firstly, all volumetric flow rates in the three-stage RO system are determined using the total material balances of stage 1A, 1B and 1C.

From Eq. (3.165),

$$Q_{13} = Y_{1A}Q_0 \tag{3.177}$$

From Eq. (3.153),

$$Q_{14} = Q_0 - Q_{13} \tag{3.178}$$

From Eq. (2.156),

$$Q_{15} = Q_{14} - Q_{16} \tag{3.179}$$

Rewriting Eq. (3.167),

$$Q_{17} = Y_{1C} Q_{16} \tag{3.180}$$

Combining Eqs. (3.159) and (3.180),

$$Q_{16} = \frac{Q_1}{1 - Y_{1C}} \tag{3.181}$$

After obtaining the volumetric flow rates, solving the salt and boron equations together, the concentration values are determined.

Rewriting Eq. (3.171),

$$C_{13s} = C_{0s}(1 - \sigma_{1s}) \tag{3.182}$$

Rearranging Eq. (3.154),

$$C_{14s} = \frac{Q_0 C_{0s} - Q_{13} C_{13s}}{Q_{14}} \tag{3.183}$$

Rearranging Eq. (3.157),

$$C_{15s} = \frac{Q_{14} C_{14s} - Q_{16} C_{16s}}{Q_{15}} \tag{3.184}$$

Rewriting Eq. (3.176),

$$C_{17s} = C_{16s}(1 - \sigma_{1s}) \tag{3.185}$$

Combining Eqs. (3.160) and (3.185),

$$C_{16s} = \frac{Q_1 C_{1s}}{Q_{16} - Q_{17}(1 - \sigma_{1s})} \tag{3.186}$$

From Eq. (3.172),

$$C_{13b} = C_{0b}(1 - \sigma_{1b}) \tag{3.187}$$

Rearranging Eq. (3.155),

$$C_{14b} = \frac{Q_0 C_{0b} - Q_{13} C_{13b}}{Q_{14}} \tag{3.188}$$

Rewriting Eq. (3.158),

$$C_{15b} = \frac{Q_{14} C_{14b} - Q_{16} C_{16b}}{Q_{15}} \tag{3.189}$$

Rearranging Eq. (3.176),

$$C_{17b} = C_{16b}(1 - \sigma_{1b}) \tag{3.190}$$

Combining Eqs. (3.190) and (3.161),

$$C_{16b} = \frac{Q_1 C_{1b}}{Q_{16} - Q_{17}(1 - \sigma_{1b})} \tag{3.191}$$

Rearranging Eq. (3.168),

$$\theta = \frac{[K(C_{0s} - C_{13s})\Delta\pi_{1C}]^{0.5}}{K(C_{14s} - C_{13s})} = 1 \tag{3.192}$$

As in the 2-3 CDBR process, with setting Eq. (3.192) to 1 using the excel solver, and Y_{1A} and Y_{1C} values are provided.

Then,

The gross specific energy consumption (SEC_{gross}) is written as follows:

$$SEC_{gross} = \frac{\begin{matrix} Q_f\Delta\pi_{1A} + Q_3(\Delta\pi_{1A} - \Delta\pi_4) + Q_{14}(\Delta\pi_{1B} - \Delta\pi_{1A}) + \\ Q_{16}(\Delta\pi_{1C} - \Delta\pi_{1B}) + Q_{10}\Delta\pi_3 + Q_6\Delta\pi_4 \end{matrix}}{\eta_p Q_{12}} \tag{3.193}$$

The gross specific energy consumption (SEC_{net}) is written as follows:

$$\mathrm{SEC_{net}} = \mathrm{SEC_{gross}} - \frac{\eta_{ERD} Q_9 \Delta\pi_1}{Q_{12}} \tag{3.194}$$

4. TECHNICAL FEASIBILITY OF TWO-PASS AND CDBR PROCESSES

In this section, a two-pass process containing concentrate recycling, and a revised configuration of CDBR process were modeled employing commercial reverse osmosis membranes of different manufacturers so that the actual performance of these systems could be estimated. The aim for both design scenarios is to mitigate the boron concentration of the final product water to 0.5 ppm or below levels. In order to obtain this water quality, Toray Design system 2.0 (Toray DS2), and The Water Application Value Engine (WAVE) were utilized as commercial RO design software tools.

4.1 Design Characteristics

The following conditions are considered in common when designing commercially for all reverse osmosis desalination systems constructed in this work.

- The raw feed water to be desalinated contains the ion composition of the typical seawater outlined in Table 1.2. Besides, 10 ppm boron is added to the content of the feed stream, and the TDS value of this seawater is set to 35,000 ppm.
- The feed seawater pH level is specified as 8.10, and the design temperature of the process is maintained at a constant value of 25 ℃. The membranes used in this work provides salt and boron rejection higher than 99.75% and 91%, respectively.
- As for ERD, Isobaric Device (ERI) with a yield of 0.965 is selected, and the efficiency of pumps is set to 0.85 as well.
- Membrane types vary according to design requirements, but the element age is generally set to 1 for all systems.
- The pH value of the raw feed water is adjusted to 7 by adding sulfuric acid for Pre-treatment application.
- Post-treatment application is the beyond scope of this study.

4.2 Two-pass RO Systems with Concentrate Recycling

All two-pass RO designs for seawater desalination applications are modeled in the above-mentioned design conditions using Toray DS2 design software which allows working with ERD. These processes were designed using the conventional RO rack model incorporating the same SWRO membrane in each segment of the pressure vessel, and the HID concept in which pressure vessel houses membrane elements having different flow and rejection characteristics. All processes mainly involve concentrate recycling. In order to fulfill this design, the brine flow of the second stage was combined with the initial feed stream of the process, and this blended stream was fed to the first pass (Figure 4.1). To do this, SWRO membrane type was used for the first pass, while in the other pass, a BWRO membrane was employed.

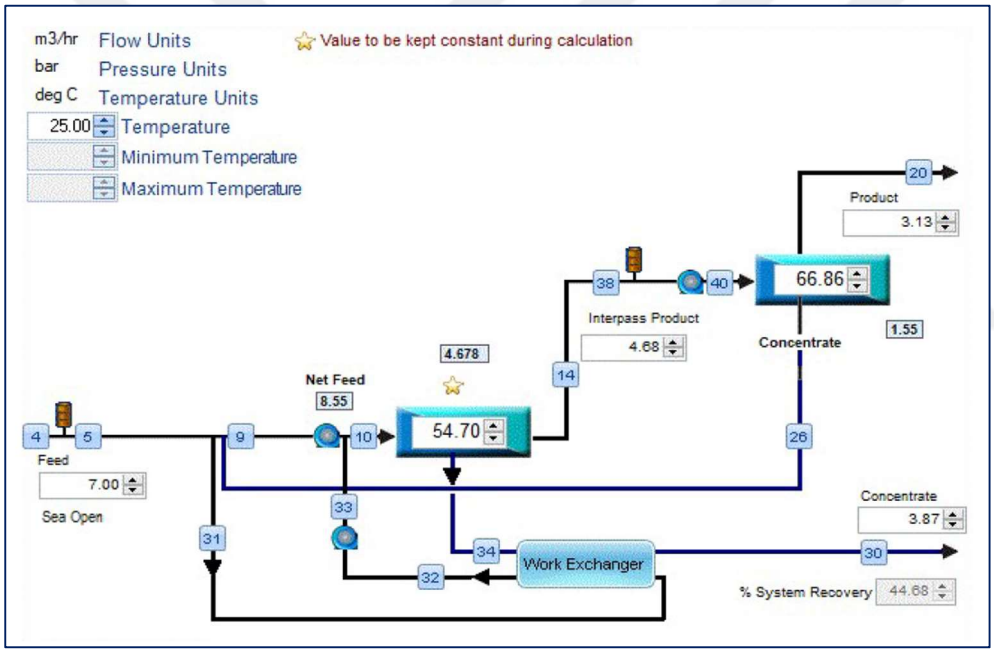

Figure 4.1 : Flow diagram of the two-pass RO system with ERD and pH adjustment.

In this work, the parameters affecting boron rejection such as operating pressure, pH level, feed flow rate, overall recovery, were also studied. For this purpose, firstly, the flow rate of the raw feed seawater to be desalinated was examined by taking into consideration the errors of the system and the amount of boron content in the product water. The flow rate of the raw feed water was tested as 6, 7, and 8 m^3/h, respectively, for different scenarios under the constant process conditions such as pH level, temperature, and recovery. Thus, the tendency of the process to provide maximum overall recovery for each scenario was examined. For this purpose, the recovery of the first pass was changed between 40% and 55%, while this range for the second pass

was arranged to be between 55% and 90%. As described in Section 2, the pH of the stream fed to the second pass must be changed according to the desired product water quality in commercial two-pass processes. Based upon the boron rejection capacities of the membranes, the appropriate working conditions were determined by trying different levels in the pH range of 9 and 11.

This study encompasses the use of both conventional SWRO model and HID concept to perform two-pass RO design commercially. For both systems, individual pressure vessels used for Pass 1 and 2 holds 7-elements to boost process recovery as much as possible. Initially, 7, 2-5, and 1-1-5 membrane configurations were examined for the first pass, 7-element membrane configuration indicates that there are seven same elements in a PV. The 3-4 membrane configuration represents the same membrane type in the first three positions, and the identical element type in the last four positions which is different from the first four elements. The 1-1-5 membrane configuration is achieved by combining a membrane that provides high rejection in the first position, an energy-saving membrane in the second position, and membranes providing high product flow in the last five positions. For Pass 2, the same BWRO element was used for all processes. In this study in order to model the first pass, the TM820M-400, and TM820K-440 membrane types for the 7-element configuration, the TM820M-400 and TM820R-400 membrane types for the 2-5 element membrane configuration, and the TM820M-400, TM820E-400 and TM820R-400 membranes for the 1-1-5 membrane configuration were used. TM710D, BWRO membrane type was employed for Pass 2. Table 4.1 shows the major features of the membranes used in this study.

Table 4.1 : Specifications of BWRO and SWRO membranes used in two-pass processes (Test conditions: 32,000 mg/L NaCl, 55.2 bar, 25 °C).

Model	Diameter (Inch)	Membrane area (m^2)	Product flow rate (m^3/d)	Salt rejection (%)	Boron rejection (%)
TM820M-400 [48]	8”	37	26.5	99.8	95
TM820E-400 [49]	8”	37	28.3	99.75	91
TM820R-400 [50]	8”	37	32.2	99.8	95
TM820K-440 [51]	8”	41	24.2	99.86	96
TM710-D [52]	4”	8	9.8	99.8	-

4.3 Revised Concurrent Desalination and Boron Removal Process

In order to model a commercial 1-3CDBR (S = 0) process for the purpose of seawater desalination, the system consisting of the SSRO and LPMS components was created with Toray DS2 design software, while the CMCR part was constructed using Dupont WAVE. However, in the CDBR process, unlike the previous two-pass designs, SWRO membranes were employed in the second pass (LPMS) to further reduce the boron content in the product water approximately to the level of the permeate stream of the conceptional CDBR design. Thus, the need for pH adjustment was eliminated. For both passes, the TM820K-440 membrane providing the highest rejection performance was employed. One of the most important matters to be considered in this procedure is to increase the overall recovery, while reducing the boron content in the product water and maintain the operating pressure of the second pass between 10 and 15 bar. Another part of the CDBR process, the CMCR that desalinated the remaining brine flow from the SSRO, was created in the DuPont Wave design software. As shown in Figure 4.2, a CMCR was commercially modeled for wastewater treatment by merging the concentrate of the 2nd pass with the 1st pass feed. In this step, since NF membranes could not provide enough removal performance practically, BWRO membranes instead of them and SWRO membranes were used for Pass 1, and Pass 2, respectively. Table 4.2 lists the specifications of BWRO and SWRO membranes benefited in this work. Additionally, there was no need for any pH adjustment and chemical addition when constructing the CMCR.

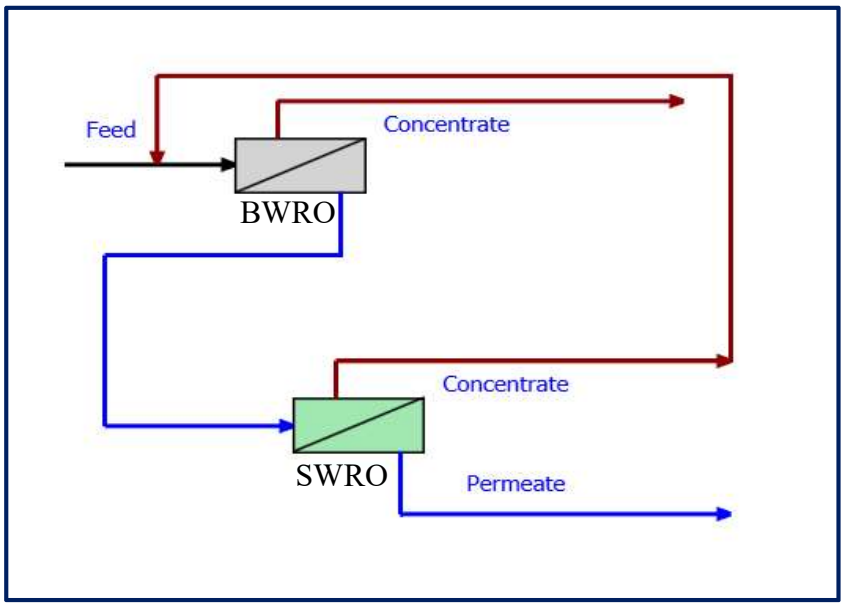

Figure 4.2 : Flow diagram of the CMCR system without pH adjustment.

Since the previous process configuration was designed to operate at 25℃ in the Toray DS2, this temperature value was also preserved in the design of the CMCR system. The pH value of 7.22 of the brine stream of SSRO was also used for the feed stream of the CMCR. However, the brine flow rate obtained from SSRO is too insufficient to function this CMCR system. Therefore, various levels of this value were tested in the DuPont wave design program, and an appropriate flow rate was assigned. The feed flow rate for this work was 22.93 m^3/h. Moreover, SSRO and CMCR must be run at the same operating pressure in order to eliminate the need for pumps at the part of the joint. Therefore, the pressure in both stages of the CMCR system was set to 57.27 bar, that was the concentrate pressure of the SSRO. Furthermore, since the system was equipped with an Isobaric device (ERI) as for ERD, flow equality around the process should be ensured. That is, the amount of raw water corresponding to the percent recovery is conveyed to the high-pressure pump in order to be pressurized. The rest is transferred to the ERD to take the energy of the effluent solution. This change must be considered especially in calculating SEC_{net} of the overall process.

Table 4.2 : Specifications of BWRO and SWRO membranes used in the CMCR. (Test conditions: 32,000 mg/L NaCl and 55.2 bar for SWRO membrane, 2,000 mg/L and 15.5 bar for BWRO membrane at 25 ℃).

Model	Diameter (Inch)	Membrane area (m^2)	Product flow rate (m^3/d)	Salt rejection (%)	Boron rejection (%)
SW30XHR-440i [53]	8”	41	25	99.82	93
BW30HR-440 [54]	8”	41	48	99.7	-

5. COST ANALYSIS

In this section, cost analysis of the two-pass and CDBR processes were performed based on the studies of Mistry and Lienhard [56], Bilton et.al [57], and Chong and Krantz [44]. A plant with a potable water production of 10,000 m^3/d was considered as the basis of design.

In this study, the sum of capital (CAPEX) and operating cost (OPEX) constitutes the total cost (TOTEX) of the system. The annual total cost for water desalination (A_{total}), which is obtained by summing up the annual capital and operating costs, is divided by annual production quantity of water to calculate the total cost per unit water produced as follows:

$$C(\text{per } m^3) = \frac{A_{CAPEX} + A_{OPEX}}{\dot{V}_p} = \frac{A_{TOTEX}}{\dot{V}_p} \tag{5.1}$$

The following equation determines the annual water production amount denominated in volume (Vp).

$$\dot{V}_p = 365 \times \psi_{AF} \times \dot{V}_c \tag{5.2}$$

where ψ_{AF} is the RO plant availability parameter, and $\dot{V}_c$ is the water production capacity of the plant in m^3/d. Annual capital cost is given by the following formula:

$$A_{CC} = \left[\frac{i(1+i)^t}{(1+i)^t - 1}\right] C_{DC} \tag{5.3}$$

where t is the annual lifespan of the system, C_{DC} is the direct capital cost of the RO system, and i is the rate of interest. In this study, the interest rate and annual lifespan of the system are specified as 7.5% and 20 years, respectively.

The CAPEX is calculated from the sum of the infrastructure and RO system costs. The RO system cost is divided into sub-groups as pre-treatment cost, post-treatment cost, RO element cost and piping cost, and the cost of the RO elements is expressed with

the sum of cost of membranes, pressure vessels, pumps, motors, and ERDs. The representative prices of some of the RO elements are shown in Table 5.1.

Table 5.1 : The representative prices and capacities of RO elements [56].

RO elements	Price [$]	Capacity [m^3/d]
Membrane	550	25
Pressure vessel	1,945	-
Pumps	50,000	720
Motors	12,000	-
ERD	24,000	1,000

The flows required for the two-pass are listed in Table 5.2. However, in commercial two-pass design apart from the two-pass designed via mathematical analysis, pressure exchanger was used as ERD. In the system to pressurize the raw seawater feed, the portion of the feed that is as same as the volumetric flow amount of the total brine is conveyed to pressure exchanger and high-pressure pump pumps the rest of the water. Since the pressure exchanger system includes an additional pump as shown in Figure 2.2, this requirement is added to the cost of the system. Therefore, the flows for pumps and motors are rewritten. In Table 5.2, volumetric flow rates for each pass are obtained from Figure 3.1.

Table 5.2 : Flows used to determine the RO element costs for the two-pass processes.

RO elements	Conceptional	Commercial
Membrane and pressure vessel	$Q_2 + Q_4$	$Q_2 + Q_4$
HPPs and motor	Q_f	$Y_{total} \times Q_f$
BPs and motor	Q_2	$Q_2 + (1 - Y_{total}) \times Q_f$
ERD	Q_1	Q_1

As in the commercial two-pass process, the flows for the CDBR system are rewritten as shown in Table 5.3. Volumetric flow rates for each stage are provided in Figure 3.2.

Table 5.3 : Flows used to determine the RO element costs for the CDBR processes.

RO elements	Conceptional	Commercial
Membrane and pressure vessel	$Q_2 + Q_{10} + Q_{11} + Q_4$	$Q_2 + Q_{10} + Q_{11} + Q_4$
HPPs and motors	$Q_f + Q_{10}$	$Y_{total} \times Q_f + Q_{10}$
BPs and motors	Q_2	$Q_2 + (1 - Y_{total}) \times Q_f$
ERD	Q_9	Q_9

In the work of Mistry and Lienhard [56], the cost of the RO systems was associated with the cost of RO elements with the use of scaling parameters. However, these parameters change according to the operating conditions of the system. In this study, the cost of pre-treatment was calculated according to the overall recovery since the CDBR process required less pre-treatment to achieve the same amount of product compared to the two-pass process. Besides, electricity cost was changed considering a decrease in the feed flow rate. The other scaling constants were kept at the same values.

6. RESULTS AND DISCUSSIONS

In the previous sections, to obtain the needed water quality, a revised configuration of the two-pass and CDBR processes was modeled separately via mathematical analysis and using commercial RO design software tools. Then the cost of each system was calculated. In this section, the results are presented and discussed in detail.

6.1 Conceptional Designs of the Two-pass and CDBR Processes

In this study, a two-pass RO system, which recycles the concentrate of Pass 2 to the feed side of Pass 1, employs. In addition to the two-pass RO system, the new CDBR process, including the concentrate flow transfer of the LPMS to the SSRO feed to lessen the SEC_{net} further was developed.

For both RO systems, seawater containing 35,000 ppm salt, and 10.0 ppm boron was desalinated. Using standard RO membranes with 0.997 salt and 0.90 boron rejection, it was possible to produce potable water containing less than 100 ppm salt and 0.5 ppm boron.

In order to increase the amount of water provided by the LPMS, the recovery of this stage was adjusted to 0.95. The total fractional water recovery of both CMCR stages was set to 0.5 to avoid membrane fouling problems. Additionally, standard pump and ERD providing a yield of 0.85 and 0.90, respectively were employed for both systems.

Figure 6.1 illustrates the OPD in the total fractional water recovery that varies between 0.55 and 0.75 for both processes. In each system, in case the water recovery boosts, the OPD of the system increases as well. If the two-pass process and the revised CDBR processes are compared, the revised CDBR process seems to operate at about 33.4% less OPD than the two-pass process. Besides, the OPD of all CDBR processes is equal for the same recovery. That is, adding or removing one or two SSRO stages to the process does not change the OPD at which the system functions. Since the last one of the SSRO stages determines the OPD of the system. This terminal stage always has the highest OPD since it treats highly concentrated water compared to the others. So, when solving the system, the OPD of the final is equalized to that of the CMCR.

Additionally, it was observed that the 3-3CBR S=0.17 system could not be operated at the recovery values of 74% and 75% due to negative salt concentration values in the permeate of the LPMS. Further information for the volumetric flow rate and concentration values for each process was provided in Appendix A.

Figure 6.2 demonstrates the SEC_{net} values at the different total fractional water recovery for each process. Up to the point where the water recovery is 0.585 the two-pass RO system has a lower SEC_{net} value than all CDBR processes. Above this value, however, the 3-3 CDBR process is more favorable at the same recovery to obtain lower SEC_{net} in comparison with the two-pass system. Likewise, above the point where the water recovery is 0.59, and 0.625, respectively, the 2-3 CDBR process and the 1-3 CDBR process have a reduced SEC_{net} than that of the two pass system.

As shown in Figure 6.2, the revised 3-3 CDBR process achieves the same total water recovery at a lower SEC_{net} compared to the other CDBR systems. Due to the increase in the number of SSRO stages enables more saline water to be treated. Hence, while the amount of dilute water fed to the LPMS increases, the amount of concentrated water used as feed conveyed to the CMCR stage decreases. Reducing the amount of concentrate feed water to be handled contributes to the decline of the SEC_{net}. The increase in the amount of feed water (permeate of Pass 1) delivered to the LPMS has no significant effect on the SEC_{net} since it is dilute water.

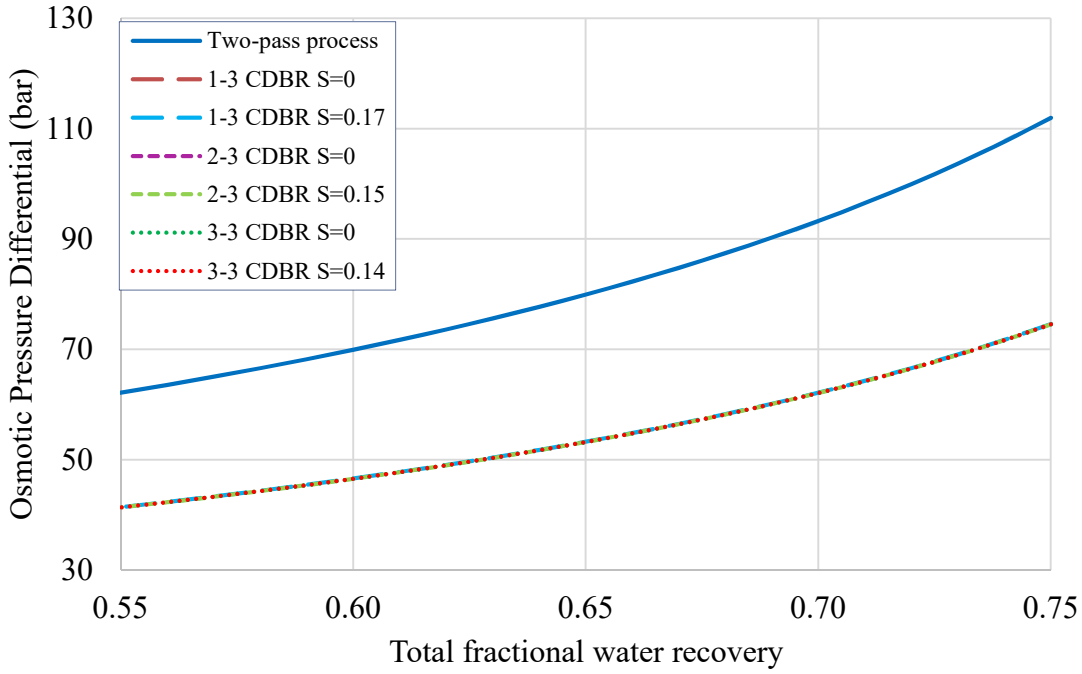

Figure 6.1 : Osmotic pressure differential as a function of total water recovery.

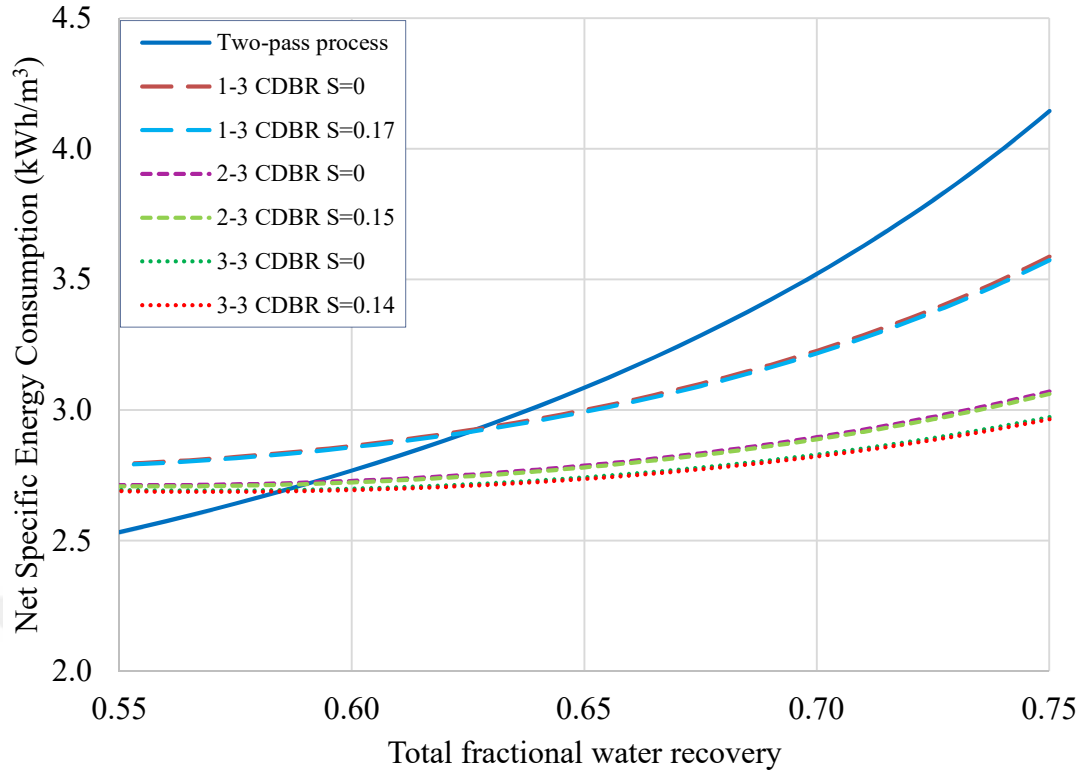

Figure 6.2 : Change of the net SEC with total water recovery.

Table 6.1 demonstrates the OPD and SEC_{net} outcomes for each process for the total recovery values of 55%, 65%, and 75%. Table 6.2 shows all the volumetric flow rates for each stage of the two-pass process. If the fouling and scaling problems are considered, safety factors must be at least one. In this case, the maximum recovery, that the system reaches, is 0.48. For a plant with a potable water production capacity of 10,000 m^3/day, the seawater to be fed into the two-pass RO system with concentrate recycling is 20,833 m^3 per day. Besides, this system operates at 53.80 bar and has SEC_{net} of 2.30 kWh/m^3.

Table 6.1 : Summary of results for the two-pass and CDBR systems.

Process	55% recovery		65% recovery		75% recovery	
	OPD (bar)	SEC_{net} (kWh/m^3)	OPD (bar)	SEC_{net} (kWh/m^3)	OPD (bar)	SEC_{net} (kWh/m^3)
Two-pass system	62.2	2.53	80.0	3.09	112.0	4.15
1-3 CDBR S=0.0	41.4	2.79	53.3	3.00	74.6	3.59
2-3 CDBR S=0.0	41.4	2.71	53.3	2.79	74.6	3.07
3-3 CDBR S=0.0	41.4	2.70	53.3	2.74	74.6	2.97

Table 6.2 : Operating conditions of the two-pass RO process.

Parameter	Stage 1	Stage 2	Overall
Recovery (%)	49	95	48
Feed flow (m^3/day)	21,360	10,526	20,833
Pump flow (m^3/day)	21,360	10,526	31,886
Flow to ERD (m^3/day)	10,833	-	10,833
Feed pressure (bar)	53.80	0.91	53.80
Net SEC (kWh/m^3)	-	-	2.30
Product boron level (ppm)	1	0.5	0.5

Table 6.3 displays all the volumetric flow rates at the total system recovery of 65% for each stage of the 1-3 CDBR process. For this CDBR process yielding potable water of 10,000 m^3/day, the raw seawater to be fed into the system is 15,769 m^3/day. The SEC_{net} value is 3.00 kWh/m^3 for this system which performs at 53.30 bar. The boron concentration in the permeate of LPMS and CMCR stage was calculated around 0.3 ppm and 10.5 ppm, respectively. That is because of the highly concentrated feed of the CMCR. Namely, raise in salinity of feed flow causes low boron rejection performance.

To sum up, all outcomes explained above are compared, it can be concluded that the total recovery of the system by adding CMCR to the process raises from 48% to 75%.

Table 6.3 : Operating conditions of the 1-3 CDBR, S=0 process at 65% recovery.

Parameter	Stage 1	Stage 2	Stage 3	Stage 4	Overall
Recovery (%)	49	50	50	95	65
Feed flow (m^3/day)	15,769	10,769	5,385	7,308	15,385
Pump flow (m^3/day)	15,769	-	5,385	7,308	28,462
Flow to ERD (m^3/day)	-	5,385	-	-	5,385
Feed pressure (bar)	53.30	53.30	53.30	1.28	53.30
Net SEC (kWh/m^3)	-	-	-	-	3.00
Product boron level (ppm)	1	10.5	1	0.3	0.5

6.2 Technical Feasibility of Two-pass and CDBR Processes

6.2.1 Two-pass RO systems with concentrate recycling

Here, the performance of commercial two-pass process with concentrate recycling that was designed using Toray DS2 was evaluated. Seawater possessing 10,000 ppm TDS and 10 ppm boron was desalinated. In this step, an isobaric device with a yield of 0.9650 as ERD is employed and pump efficiency is adjusted to 0.85. Considering the warnings and errors reported by the software, the suitable membrane types are determined to lower boron quantity below a level of 0.5 ppm. 3-pressure vessels were used to decrease operating pressure in the 2nd pass while the 1st pass contains only one.

In order to figure out the effect of the feed stream rate on boron rejection, various values of it were tested. For this reason, three different scenarios were created in which the flow rates of the raw seawater feed are specified as 6, 7 and 8 m^3/h, respectively. Additionally, at different pH conditions, the change in boron level was observed using a 7-element membrane configuration that possesses the same membrane type. For this step of the study, SWRO membrane labeled TM820K-440 for the first pass, and TM710D-BWRO membrane for the second pass is employed. The recoveries of the first and second passes for each case are set to 54 and 55, respectively, to examine the effect of only flow quantity and pH.

Figure 6.3 indicates that for each case if the pH level of the second pass feed elevates, the boron quantity of the desalinated water lowers. It's known that an increase in pH level induces the formation of a boron complex whose diameter is larger than that of water molecules. Hence, the membrane does not allow the passage of these complexes. Namely, when pH increases boron rejection improves, and it is a parameter that has a significant effect on boron rejecting performance. Moreover, thanks to membranes yielding high boron rejection, the system could provide fresh water containing boron of 0.5 ppm or less which is recommended by WHO at lower pH conditions.

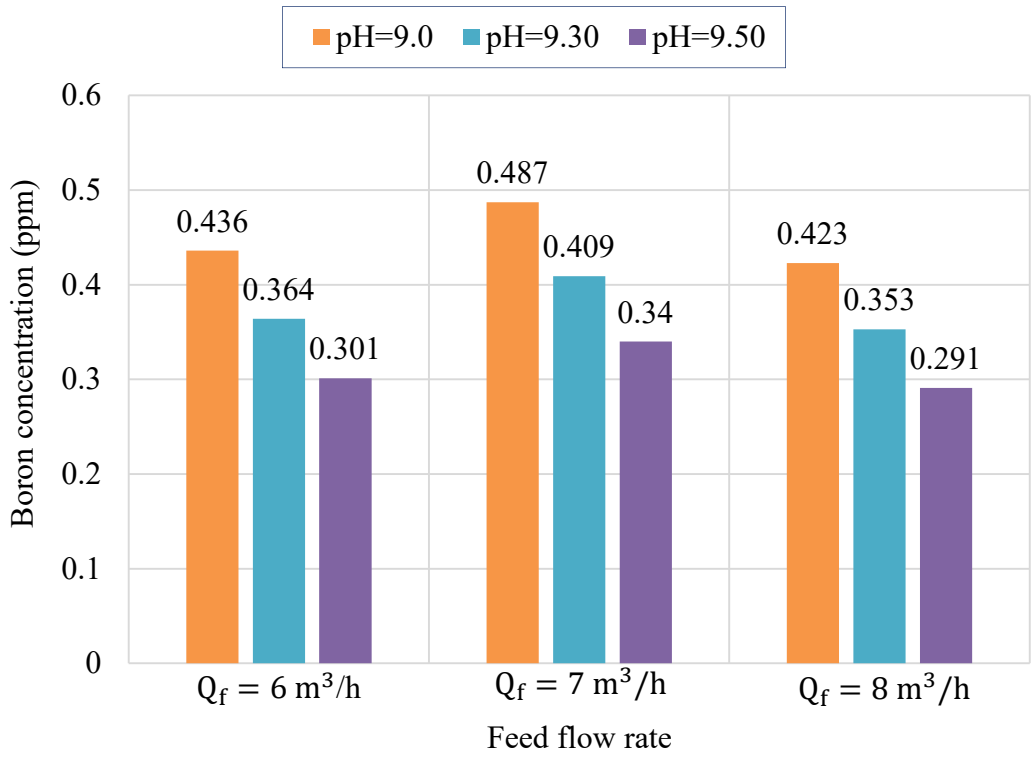

Figure 6.3 : Effect of the pH and feed flow rate on boron rejection performance (Operating conditions: 1st pass recovery: 54%, 2nd pass recovery: 55% at 25°C).

Additionally, Figure 6.3 demonstrates that at the same pH value there is no effect of the feed flow rate on boron rejection performance. Since in the first case, the product water boron concentration is 0.436 ppm at pH=9. This value ascended to 0.487 ppm at the same pH level in the second case but decreased to 0.423 ppm for the third case. That means that there is no correlation between the feed flow rate and boron rejection.

The relationship between recovery and feed flow rate can be explained based upon the results shown in Figure 6.4 and Table 6.4. Figure 6.4 shows the maximum recovery and Table 6.4 indicates the flow rates obtained for each case. In Case 3, the amount of the feed fed to Pass 1 is higher than the others. However, the first pass recovery cannot increase to the level of that is reached in Case 2. Since there is a permeate flow limit for each membrane, increase in recovery is restricted for Case 3. In case the recovery of the first pass decreases, total system recovery decreases. That's why in Case 2, higher recovery could be obtained compared to that of Case 3. In Case 1 in which system has the lowest first pass recovery, when increasing it the limit for the brine flow is exceeded since the feed flow is low compared to the others.

Table 6.4 : Streams of the system containing various feed flow rates at pH=9, 25°C.

Flow information	Case 1	Case 2	Case 3
Raw feed water (m^3/h)	6.00	7.00	8.00
Feed to Pass 1 (m^3/h)	7.68	8.55	9.52
Brine to ERD (m^3/h)	3.61	3.87	4.66
Feed to Pass 2 (m^3/h)	4.07	4.68	4.85
Pass 2 concentrate (m^3/h)	1.68	1.55	1.51
Total product (m^3/h)	2.39	3.13	3.35

From all reasons, for 7-element membrane configuration, the appropriate pH and feed flow rate is determined as 9, and 7 m^3/h, respectively considering the boron rejection, and recovery performance.

Figure 6.5 illustrates the relationship between overall recovery and feed pressure for Pass 1 and 2 for the same design conditions, so it can be concluded that recovery changes with pressure linearly. However, the relationship between total recovery and boron rejection given with an exponential function as shown in Figure 6.6. This figure shows that boron concentration in product water decreases with increasing recovery. Namely, an increase in recovery leads to better boron rejection performance. One of the most important factors that enhance the productivity of the system is the increase in recovery of the first pass. Generally, if the recovery of a stage improves, a decrease

in the boron rejection performance is observed. Conversely, the SWRO membrane named TM820K-440, employed in the first stage, shows a high boron rejection performance and allows the passage of less boron component with water as much as possible. That leads to a reduction in the amount of boron in the stream fed to the second stage. That is, the boron removal performance depends on the membrane characteristics, and then the boron quantity of the final product decreases with increasing recovery for a two-pass process with concentrate recycling.

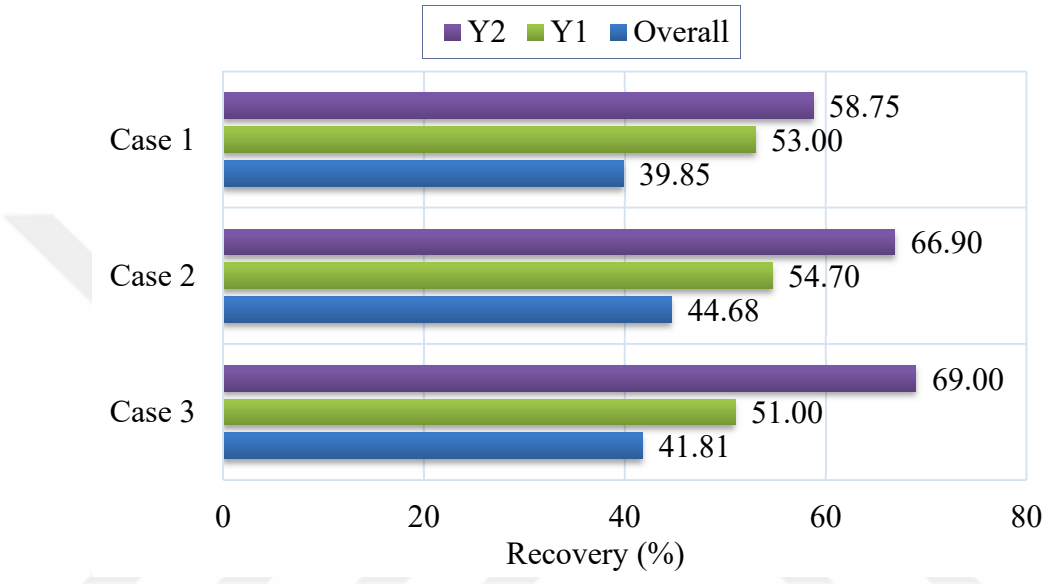

Figure 6.4 : The relationship between the recovery of the first (Y1) and second pass (Y2) and maximum recovery of the overall system (Design conditions: conventional RO rack design with TM820K-440, feed flow rate of 7 m^3/h at pH=9, 25°C).

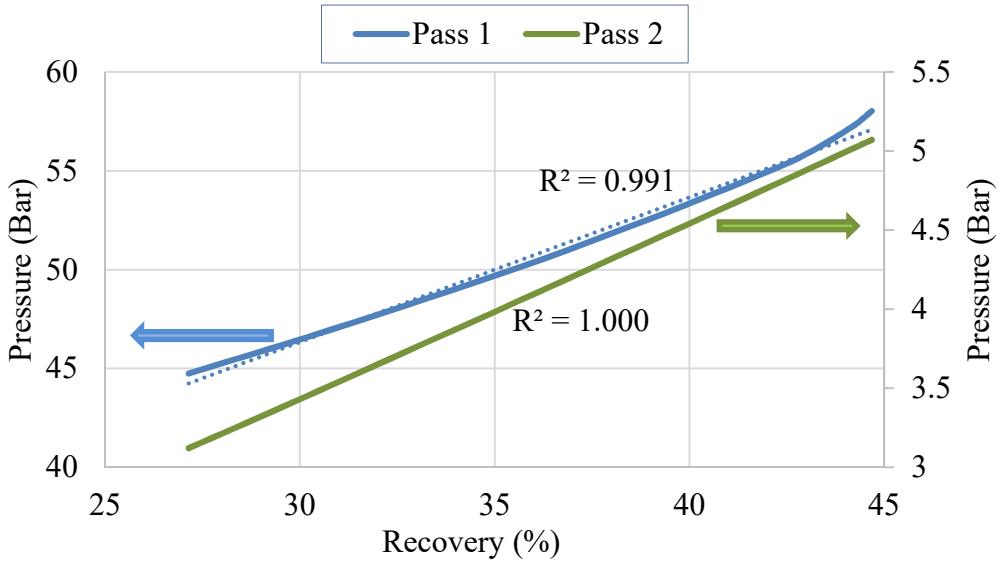

Figure 6.5 : The relationship between feed pressure of each pass and total recovery of the system (Operating conditions: conventional RO rack design with TM820K-440, feed flow rate of 7 m^3/h at pH=9, 25°C).

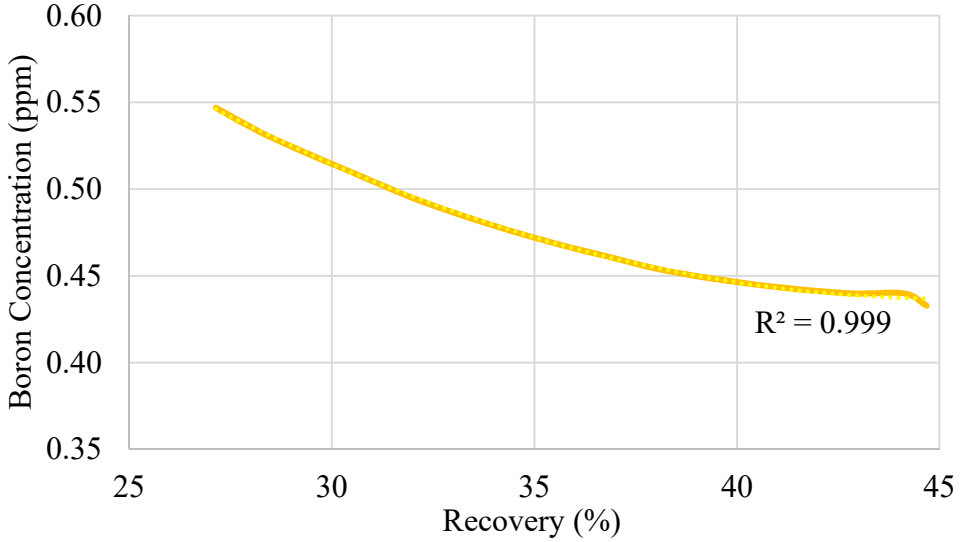

Figure 6.6 : The relationship between boron rejection performance and total recovery of the system (Operating conditions: conventional RO rack design with TM820K-440, feed flow rate of 7 m^3/h at pH=9, 25°C).

In Figure 6.7, the net specific energy consumption (SEC_{net}) was plotted as a function of total recovery. From this figure, it can be drawn a conclusion that up to the point where recovery is 35%, the SEC_{net} lowers exponentially with raise in recovery. However, after this recovery value, increase in the SEC_{net} was observed.

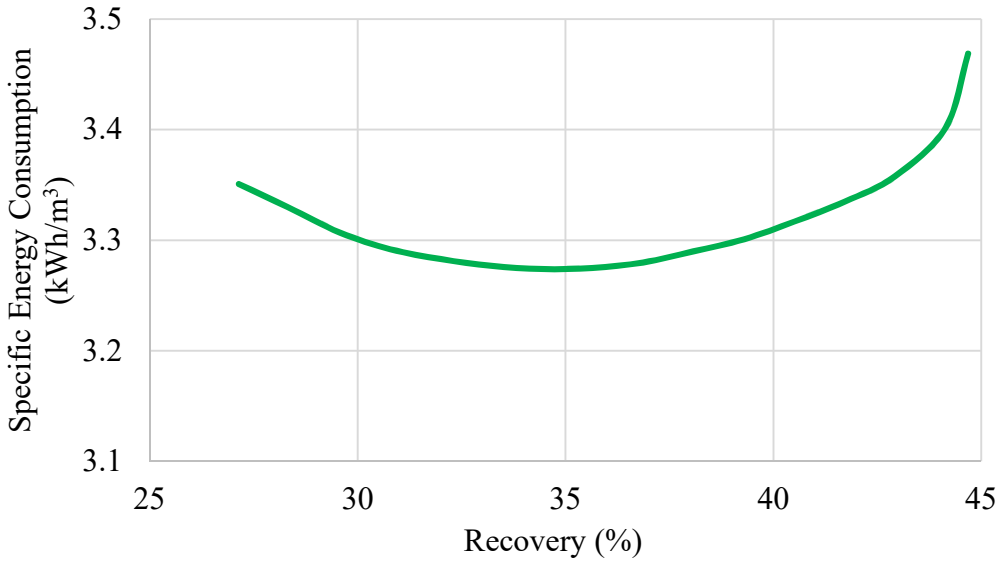

Figure 6.7 : The relationship between net specific energy consumption and total recovery of the system (Operating conditions: conventional RO rack design with TM820K-440, feed flow rate of 7 m^3/h at pH=7, 25°C).

Hybrid membrane inter-stage design (HID) was studied, in addition to the standard pressure vessel designs. The aim is to explain the advantages and disadvantages of

HID in the point of pressure, recovery and SEC compared to conventional commercial design (Figure 6.8-9). Three different configurations were created for HID systems. 7-element membrane configuration (Design 1) possesses the same SWRO membrane named TM820M-400. In the 2-5 membrane configuration (Design 2), SWRO membranes designated TM820M-400 and TM820R-400 locate in the two front and last five positions, respectively. The 1-1-5 membrane configuration apart from the previous design has an energy saving SWRO membrane named TM820E-400 in the second position. As known from Table 4.1 in Section 4, permeate flow rates of selected membranes are very close to each other. Therefore, these three designs provided the same recovery percentage. Design 1 has the highest operating pressure. Because in this design, there is just a membrane named TM820M-400 which provides more salt and boron rejection and lower productivity compared to the others in each location of the pressure vessel. In case using this membrane, it is needed to apply more pressure to increase the total recovery of the system, and that leads to higher operating pressure as explained above. The lowest SEC_{net} was calculated in design 3. That is due to the usage of the TM820E-400, the energy-saving membrane located at the second front location of the pressure vessel.

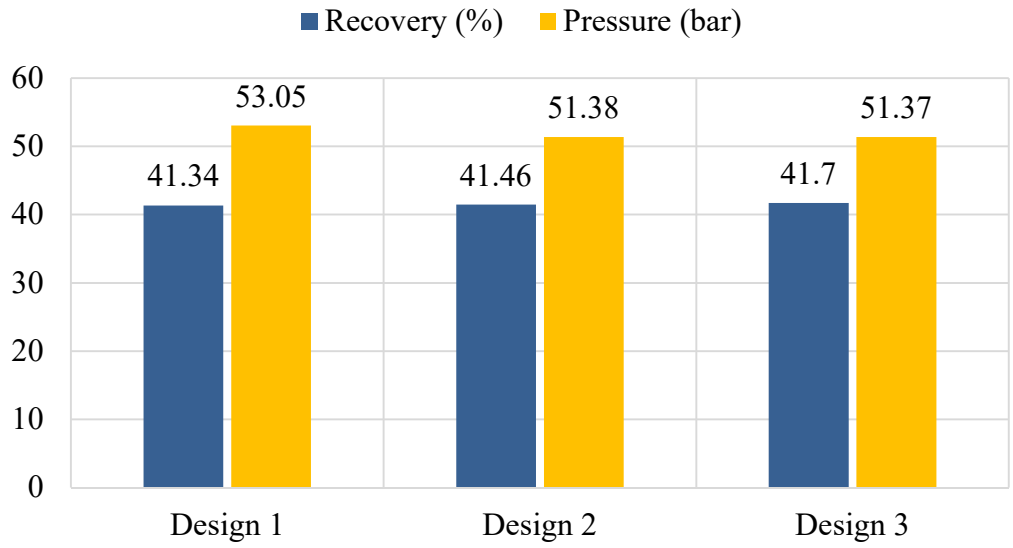

Figure 6.8 : The relationship between recovery and pressure, Design 1: (7)TM820M-400, Design 2: (2)TM820M-400 + (5)TM820R-400, Design 3: (1)TM820M-400 + (1)TM820E-400 + (5)TM820R-400 (Operating conditions: feed flow rate of 7 m^3/h at pH=9, 9.4 and 9.5, respectively, 25°C).

Table 6.5 represents the operating conditions for a plant producing around 10,000 m^3 potable water in a day. This plant is the best design that was built so as to achieve the

highest overall recovery in this study. This design was constructed using an SWRO membrane named TM820K-440 in Pass 1 and BWRO membrane labeled TM710D in Pass 2. Briefly, employing commercial membranes the ultimate recovery that two-pass reverse osmosis system with concentrate recycle is 44.68%. In this case, that system operates at a feed pressure of 58.03 bar and provides a decrease in product boron concentration to 0.433 ppm. This value meets the potable water requirement which was suggested by WHO.

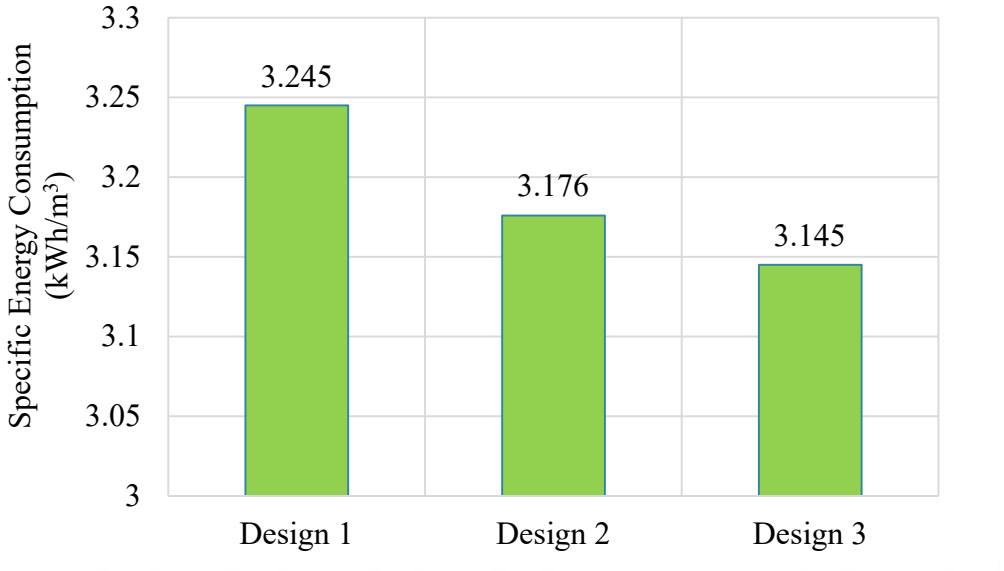

Figure 6.9 : Net specific energy consumption, Design 1: (7)TM820M-400, Design 2: (2)TM820M-400 + (5)TM820R-400, Design 3: (1)TM820M-400 + (1)TM820E-400 + (5)TM820R-400 (Operating conditions: feed flow rate of 7 m^3/h at at pH=9, 9.4 and 9.5, respectively, 25°C).

To summarize, the desired water quality can be attained with the new generation commercial membranes. However, when the HID model is used, the recovery achieved by the system does not meet expectations even if it requires low working pressure.

Table 6.5 : Operating conditions of the two-pass process with a production capacity of 10,000 m^3/day at pH=9 and 25 °C.

Parameter	Stage 1	Stage 2	Overall
Recovery (%)	54.70	66.86	44.68
Feed flow (m^3/day)	27,503	15,044	22,512
Pump flow (m^3/day)	27,503	15,044	42,548
Flow to ERD (m^3/day)	12,459	-	12,459
Feed pressure (bar)	58.03	5.071	58.03
Net SEC (kWh/m^3)	-	-	3.469
Product boron level (ppm)	0.992	0.433	0.433
Number of elements	933	2,799	3,732
Number of pressure vessels	134	400	534

6.2.2 Revised concurrent desalination and boron removal process

As indicated before, the design of a commercial CDBR process was completed in two steps. Initially, the part containing SSRO and LMPS using Toray DS2 was modeled. When configuring the two-pass RO process with concentrate recycling, the same membrane type (TM820K-440) for both passes was utilized. With this module which yields high boron and salt rejection, a system providing recovery of 45.50% was designed, and the boron level in the product water lessened to 0.286 ppm. Then, brine water collected from this process was utilized as feed water for the CMCR system. However, since this amount of water was insufficient for the CMCR process, the salt collected from 6-RO train was combined and fed into the CMCR. In other words, the total wastewater of 6-RO train designed with Toray DS2 was fed to 1 CMCR system. The concentrate flow removed from a single two-pass RO system is 3.822 m^3/h, whereas the feed flow to CMCR is 22.93 m^3/h. Table 6.6 illustrates the operating conditions of a commercial CDBR process with a production capacity of around 10,000 m^3/day. In this process the CMCR system achieved a recovery of 29.2%, reducing the boron concentration in the product water to 1.06 ppm. The end products produced in both Toray DS2 and WAVE are mixed, and the final product water contains 0.495 ppm of boron. This revised CDBR process, designed using commercial membranes, attained an ultimate recovery of 62.13%.

Table 6.6 : Operating conditions of the commercial CDBR process.

Parameter	Stage 1	Stage 2	Stage 3	Stage 4	Overall
Recovery (%)	54.0	37.8	68.5	71.0	62.13
Feed flow (m^3/day)	19,139	9,984	3,779	10,336	16,273
Pump flow (m^3/day)	19,139	-	3,779	10,336	33,254
Flow to ERD (m^3/day)	-	6,217	-	-	6,217
Feed pressure (Bar)	57.99	57.27	57.27	9.97	57.99
Net SEC (kWh/m^3)	-	-	-	-	3.07
Product boron level (ppm)	0.974	9.82	1.09	0.286	0.495
Number of elements	2,325	2,325	1,329	277	6,256
Number of pressure vessels	333	167	56	2,325	2,881

To sum up, for a desalination system whose production capacity is around 10,000 m^3/d, adding a CMCR to the process enables to improve total recovery from 44.68% to 62.13%. Both processes provide clean water with boron level below 0.5 ppm. Unlike the two-pass RO process, the CDBR achieves this value without the need for pH adjustment. Additionally, Table 6.7 and 6.8 exhibits a relative comparison of the

conceptional and commercial designs for the two-pass RO and CDBR systems, respectively. In commercial designs, although the systems operate at high pressure compared to the conceptional design, its productivity is very low.

Table 6.7 : A relative comparison of the conceptional and commercial designs for the two-pass process.

Stages	Conceptional design			Technical feasibility analysis		
	Recovery (%)	Pressure (Bar)	SEC_{net} (kWh/m^3)	Recovery (%)	Pressure (Bar)	SEC_{net} (kWh/m^3)
Stage 1	49	53.80	-	54.70	58.03	-
Stage 2	95	0.91	-	66.86	5.071	-
Overall	48	53.80	2.30	44.68	58.03	3.47

Table 6.8 : A relative comparison of the conceptional and commercial designs for the CDBR process.

Stages	Conceptional design			Technical feasibility analysis		
	Recovery (%)	Pressure (Bar)	SEC_{net} (kWh/m^3)	Recovery (%)	Pressure (Bar)	SEC_{net} (kWh/m^3)
Stage 1	49	53.30	-	54.0	57.99	-
Stage 2	50	53.30	-	37.8	57.27	-
Stage 3	50	53.30	-	68.5	57.27	-
Stage 4	95	1.28	-	71.0	9.97	-
Overall	65	53.30	3.00	62.13	57.99	3.07

6.3 Cost Estimation Analysis

In this part of the study, the cost estimation results for the processes which are designed in previous sections were assessed in detail. Cost analyses are performed according to a plant that operates with a production capacity of 10,000 m^3/day and produces fresh water containing boron of 0.5 ppm or less.

6.3.1 Cost estimation analysis for the conceptional RO designs

The outcomes for the conceptional systems are given in Table 6.9-10. The capital and operating costs in $ are listed for the two-pass RO system operating at 48% recovery in Table 6.9 and for the CDBR system working with 75% recovery in Table 6.10. Table 6.9 clarifies that infrastructure cost has the largest share of capital costs (63.10%) for the two-pass system. Electricity provides the most contribution to operation costs with a share of 68.51% in operating cost. In CDBR process (Table 6.10), as calculated in the two-pass systems, the infrastructure cost possesses the

maximum share of 63.10% among the other capital costs. The operating cost is highly affected by electricity with a share of 70.82%.

Table 6.9 : Total cost for the conceptional two-pass RO system.

Costs	Percent	Million $	$/m^3
Membranes and PVs	3.84	0.72	
Pumps and ERDs	14.65	2.74	
Pre-treatment	5.65	1.06	
Piping & connections	12.21	2.29	
Post-treatment	0.55	0.10	
Infrastructure	63.10	11.81	
Total capital cost	100.0	18.72	
Annual capital cost	-	0.53	0.43
Electricity	68.51	1.67	
Replacement	19.70	0.48	
Labor	7.10	0.17	
Chemical	4.69	0.11	
Operating cost	100.0	2.44	
Annual operating cost	-	0.70	0.57
Ratio of annual cost relative to two-pass RO			1.00

Table 6.10 : Total cost for the conceptional CDBR system.

Costs	Percent	Million $	$/m^3
Membranes and PVs	4.35	0.78	
Pumps and ERDs	12.01	2.14	
Pre-treatment	9.26	1.65	
Piping & connections	10.79	1.93	
Post-treatment	0.49	0.09	
Infrastructure	63.10	11.26	
Total capital cost	100.0	17.84	
Annual capital cost	-	0.51	0.41
Electricity	70.82	1.88	
Replacement	18.32	0.49	
Labor	6.54	0.17	
Chemical	4.32	0.11	
Operating cost	100.0	2.65	
Annual operating cost	-	0.76	0.62
Ratio of annual cost relative to two-pass RO			1.03

6.3.2 Cost estimation analysis for the technical feasibility analysis

The cost estimation results for the commercial systems are demonstrated in Table 6.10-11. The infrastructure costs hold a maximum share of 63.10% for both processes.

However, the electricity cost has the biggest share in operating costs, around 57.52% for the two-pass design, while these values calculated as 55.22% for the CDBR design.

Table 6.11 : Total cost for the commercial two-pass RO system.

Costs	Percent	Million $	$/m^3
Membranes and PVs	9.31	3.09	
Pumps and ERDs	10.84	3.60	
Pre-treatment	2.85	0.94	
Piping & connections	13.30	4.41	
Post-treatment	0.60	0.20	
Infrastructure	63.10	20.94	
Total capital cost	100.0	33.19	
Annual capital cost	-	0.86	0.45
Electricity	57.52	2.07	
Replacement	24.58	0.88	
Labor	10.78	0.39	
Chemical	7.12	0.26	
Operating cost	100	3.60	
Annual operating cost	-	1.04	0.55
Ratio of annual cost relative to two-pass RO			1.00

Table 6.12 : Total cost for the commercial CDBR system.

Costs	Percent	Million $	$/m^3
Membranes and PVs	15.53	9.04	
Pumps and ERDs	4.98	2.90	
Pre-treatment	2.24	1.30	
Piping & connections	13.54	7.88	
Post-treatment	0.62	0.36	
Infrastructure	63.10	36.71	
Total capital cost	100	58.18	
Annual capital cost	-	1.51	0.79
Electricity	55.22	1.99	
Replacement	31.77	1.14	
Labor	7.84	0.28	
Chemical	5.17	0.19	
Operating cost	100	3.60	
Annual operating cost	-	1.04	0.55
Ratio of annual cost relative to two-pass RO			1.34

6. CONCLUSIONS AND RECOMMENDATIONS

This study, as explained in the previous sections, focuses on the remodeling of the two-pass and CDBR processes providing potable water product in an energy efficient way. The results and recommendations obtained from the study are as follows:

- Based on the outcomes that are provided with the design of the conceptional two-pass and CDBR processes, it can be concluded that these systems enable to desalinate typical seawater and reduce its boron, and salt content to 0.5 ppm, and lower than 100 ppm, respectively. In other words, the desalinated water generated by these systems complies with the water quality standards, suggested by WHO.
- According to the results obtained from conceptional design, the CDBR systems always work at less OPD compared to the other system for any overall recovery values. A reduction of 33.4% in the OPD is the same for all CDBR processes. Namely, the design of the process as a split-partial system does not have any contribution to the OPD.
- Additionally, at higher recovery values, to employ the CDBR process is more convenient as against the two-pass process in terms of SEC_{net}. Even though an increase in the number of SSRO stages does not change the OPD, it provides a reasonably decrease in the SEC_{net}.
- In the commercial designs, the conventional two-pass process provided product water with a boron content of 0.433 ppm by adjusting the pH level of the second pass feed to 9 at a design temperature of 25°C. Moreover, the HID configurations enabled to decrease boron composition in water to around 0.485 ppm at a pH range of 9 and 9.5. That indicates that new brand membranes are good enough for boron removal to meet the potable water standards. Besides, there is no need for elevated pH value in order to obtain a such boron level.
- Similarly, in the commercial CDBR design, water with a boron concentration of 0.495 ppm was produced without the need for pH adjustment in any stage

of the CDBR except pre-treatment step. In order to achieve that result, the SWRO membranes, exhibiting rejection performance higher than 99.75% for salt and 91% for boron, were utilized for both design parts.

- The conceptional two-pass process with concentrate recycling succeeded to reach a maximum recovery of 48% when fouling and scaling problems are taken into account. However, in the commercial two-pass design, this value was found as 44.68% despite the use of new generation membranes providing high salt rejection and permeate flow rate. With the use of HID model, it was not reached to this level since Toray RO membranes are not convenient for this design type. Therefore, the conventional 7-element membrane configuration was selected as the appropriate two-pass RO design. Besides, it was discovered that the only advantage of HID model is its significant effect on decreasing feed pressure accordingly power consumption.
- In the conceptional design of the revised 1-3CDBR (S=0) process, maximum recovery of 75% was obtained, whereas this value is 62.13% in the commercial design.
- Design calculations for 7-element conventional two-pass RO systems utilizing TM420K-440,SWRO membrane type demonstrated that the pH level of the feed water is the most effective parameter on boron rejection capability. Also, the designs examples showed that the raw feed flow rate has no significant effect on boron rejection. However, for a two-pass system containing SWRO membrane with high rejection named TM820K, 440 if feed pressure increases boron quantity in the product water decreases because of the high boron removal efficiency of the membrane.
- Additionally, for the same system, it was observed that the total recovery enhances with increasing operating pressure. Thus, if the overall recovery value increases boron rejection performance enhances for a design employing a membrane providing high salt and boron rejection. Furthermore, to the point where recovery is 35%, the SEC_{net} decreases with increasing recovery. However, after this period, the SEC_{net} rises gradually.
- The results of the cost estimation analysis showed that the parameter contributing the most to the capital costs was the infrastructure, while it was

observed that this parameter was electricity for the operating costs. The scaling factors used in the study belong to a system operating at 40% recovery and could not be adapted to systems working at higher recovery values. Therefore, these percents will differ in real designs.

- In the conceptual designs, the CDBR process is costing 3% more than the two-pass systems. This value increased to 34% in commercial designs. In the CDBR systems, this increase in the cost was an unexpected situation despite a reduction in energy consumption. This problem is due to the fact that the unit costs of the RO components cannot be reached, and the factors other than the pre-treatment and electricity costs at the increasing recovery were ignored in the calculations.
- Based on these results, it can be concluded that the new configuration of the CDBR process is suitable for pilot scale testing.

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APPENDICES

APPENDIX A: Flow rate and concentration values.

APPENDIX A

Table A.1 : Flow rate and concentration values for two-pass RO system operating at a recovery of 55%, 65%, and 75%.

Stream number	55% Recovery			65% Recovery			75% Recovery		
	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)
f	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100
0	1.0289	34.0485	0.0100	1.0342	33.8810	0.0100	1.0395	33.7152	0.0100
1	0.4500	77.7222	0.0216	0.3500	99.9160	0.0276	0.2500	139.8650	0.0385
2	0.5789	0.1021	0.0010	0.6842	0.1016	0.0010	0.7895	0.1011	0.0010
3	0.0289	1.1787	0.0105	0.0342	1.1732	0.0105	0.0395	1.1678	0.0105
4	0.5500	0.0455	0.0005	0.6500	0.0452	0.0005	0.7500	0.0450	0.0005

Table A.2 : Flow rate and concentration values for 1-3 CDBR S=0 system operating at a recovery of 55%, 65%, and 75%.

Stream number	55% Recovery			65% Recovery			75% Recovery		
	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)
f	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100
0	1.0171	34.4397	0.0101	1.0250	34.1860	0.0101	1.0329	33.9359	0.0101
1	0.6750	51.8422	0.0147	0.5250	66.6463	0.0188	0.3750	93.2940	0.0261
2	0.3421	0.1033	0.0010	0.5000	0.1026	0.0010	0.6579	0.1018	0.0010
3	0.0171	1.6864	0.0150	0.0250	1.6243	0.0146	0.0329	1.5880	0.0144
4	0.3250	0.0200	0.0003	0.4750	0.0225	0.0003	0.6250	0.0236	0.0003
5	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
6	0.3421	0.1033	0.0010	0.5000	0.1026	0.0010	0.6579	0.1018	0.0010
7	0.2250	51.8422	0.0155	0.1750	66.6463	0.0199	0.1250	93.2940	0.0277
8	0.9000	51.8422	0.0149	0.7000	66.6463	0.0191	0.5000	93.2940	0.0265
9	0.4500	77.7117	0.0216	0.3500	99.9182	0.0276	0.2500	139.8901	0.0385
10	0.4500	25.9728	0.0082	0.3500	33.3744	0.0105	0.2500	46.6979	0.0146
11	0.2250	0.1033	0.0008	0.1750	0.1026	0.0010	0.1250	0.1018	0.0014
12	0.5500	0.0541	0.0005	0.6500	0.0440	0.0005	0.7500	0.0366	0.0005

Table A.3 : Flow rate and concentration values for 1-3 CDBR S=0.17 system operating at a recovery of 55%, 65%, and 75%.

Stream number	55% Recovery			65% Recovery			75% Recovery		
	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)
f	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100
0	1.0141	34.5426	0.0101	1.0206	34.3343	0.0102	1.0271	34.1285	0.0102
1	0.6750	51.8422	0.0147	0.5250	66.6464	0.0188	0.3750	93.2941	0.0261
2	0.3391	0.1036	0.0010	0.4956	0.1030	0.0010	0.6521	0.1024	0.0010
3	0.0141	2.0360	0.0180	0.0206	1.9642	0.0176	0.0271	1.9237	0.0174
4	0.2674	0.0019	0.0001	0.3908	0.0050	0.0001	0.5141	0.0065	0.0002
5	0.0576	0.1036	0.0010	0.0842	0.1030	0.0010	0.1109	0.1024	0.0010
6	0.2814	0.1036	0.0010	0.4113	0.1030	0.0010	0.5412	0.1024	0.0010
7	0.2250	51.8422	0.0155	0.1750	66.6464	0.0199	0.1250	93.2941	0.0277
8	0.9000	51.8422	0.0149	0.7000	66.6464	0.0191	0.5000	93.2941	0.0265
9	0.4500	77.7115	0.0216	0.3500	99.9181	0.0276	0.2500	139.8900	0.0385
10	0.4500	25.9729	0.0082	0.3500	33.3747	0.0105	0.2500	46.6983	0.0146
11	0.2250	0.1036	0.0008	0.1750	0.1030	0.0010	0.1250	0.1024	0.0014
12	0.5500	0.0542	0.0005	0.6500	0.0441	0.0005	0.7500	0.0367	0.0005

Table A.4 : Flow rate and concentration values for 2-3 CDBR S=0 system operating at a recovery of 55%, 65%, and 75%.

Stream number	55% Recovery			65% Recovery			75% Recovery		
	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)
f	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100
0	1.0171	34.4438	0.0101	1.0250	34.1952	0.0102	1.0329	33.9538	0.0103
1	0.6750	51.8430	0.0147	0.5250	66.6483	0.0188	0.3750	93.2990	0.0262
2	0.3421	0.1139	0.0011	0.5000	0.1195	0.0012	0.6579	0.1270	0.0013
3	0.0171	1.9260	0.0170	0.0250	2.0043	0.0178	0.0329	2.1501	0.0194
4	0.3250	0.0185	0.0003	0.4750	0.0203	0.0003	0.6250	0.0206	0.0003
5	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
6	0.3421	0.1139	0.0011	0.5000	0.1195	0.0012	0.6579	0.1270	0.0013
7	0.2250	51.8430	0.0156	0.1750	66.6483	0.0199	0.1250	93.2990	0.0277
8	0.9000	51.8430	0.0149	0.7000	66.6483	0.0191	0.5000	93.2990	0.0265
9	0.4500	77.7010	0.0216	0.3500	99.9007	0.0276	0.2500	139.8641	0.0385
10	0.4500	25.9849	0.0082	0.3500	33.3958	0.0105	0.2500	46.7340	0.0146
11	0.2250	0.1269	0.0008	0.1750	0.1434	0.0011	0.1250	0.1690	0.0015
12	0.5500	0.0628	0.0005	0.6500	0.0535	0.0005	0.7500	0.0453	0.0005
13	0.1895	0.1033	0.0010	0.2922	0.1026	0.0010	0.4111	0.1019	0.0010
14	0.8276	42.3087	0.0164	0.7328	47.7906	0.0138	0.6218	56.3346	0.0164
15	0.1526	0.1269	0.0016	0.2078	0.1434	0.0014	0.2468	0.1690	0.0016

Table A.5 : Flow rate and concentration values for 2-3 CDBR S=0.15 system operating at a recovery of 55%, 65%, and 75%.

Stream number	55% Recovery			65% Recovery			75% Recovery		
	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)
f	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100
0	1.0144	34.5345	0.0101	1.0211	34.3261	0.0102	1.0277	34.1238	0.0103
1	0.6750	51.8430	0.0147	0.5250	66.6483	0.0188	0.3750	93.2991	0.0262
2	0.3394	0.1141	0.0011	0.4961	0.1199	0.0012	0.6527	0.1276	0.0013
3	0.0144	2.2686	0.0200	0.0211	2.3638	0.0210	0.0277	2.5392	0.0228
4	0.2741	0.0007	0.0001	0.4006	0.0018	0.0001	0.5271	0.0006	0.0001
5	0.0509	0.1141	0.0011	0.0744	0.1199	0.0012	0.0979	0.1276	0.0013
6	0.2885	0.1141	0.0011	0.4217	0.1199	0.0012	0.5548	0.1276	0.0013
7	0.2250	51.8430	0.0156	0.1750	66.6483	0.0199	0.1250	93.2991	0.0277
8	0.9000	51.8430	0.0149	0.7000	66.6483	0.0191	0.5000	93.2991	0.0265
9	0.4500	77.7009	0.0216	0.3500	99.9006	0.0276	0.2500	139.8640	0.0385
10	0.4500	25.9850	0.0082	0.3500	33.3960	0.0105	0.2500	46.7343	0.0146
11	0.2250	0.1271	0.0008	0.1750	0.1436	0.0011	0.1250	0.1694	0.0015
12	0.5500	0.0629	0.0005	0.6500	0.0535	0.0005	0.7500	0.0453	0.0005
13	0.1879	0.1036	0.0010	0.2897	0.1030	0.0010	0.4075	0.1024	0.0010
14	0.8265	42.3645	0.0122	0.7314	47.8821	0.0139	0.6202	56.4756	0.0165
15	0.1515	0.1271	0.0012	0.2064	0.1436	0.0014	0.2452	0.1694	0.0016

Table A.6 : Flow rate and concentration values for 3-3 CDBR S=0 system operating at a recovery of 55%, 65%, and 75%.

Stream number	55% Recovery			65% Recovery			75% Recovery		
	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)
f	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100
0	1.0171	34.4449	0.0101	1.0250	34.1980	0.0102	1.0329	33.9596	0.0103
1	0.6750	51.8432	0.0147	0.5250	66.6490	0.0188	0.3750	93.3009	0.0262
2	0.3421	0.1168	0.0011	0.5000	0.1245	0.0012	0.6579	0.1351	0.0013
3	0.0171	1.9956	0.0176	0.0250	2.1193	0.0188	0.0329	2.3316	0.0210
4	0.3250	0.0179	0.0003	0.4750	0.0195	0.0003	0.6250	0.0195	0.0003
5	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
6	0.3421	0.1168	0.0011	0.5000	0.1245	0.0012	0.6579	0.1351	0.0013
7	0.2250	51.8432	0.0156	0.1750	66.6490	0.0199	0.1250	93.3009	0.0277
8	0.9000	51.8432	0.0149	0.7000	66.6490	0.0191	0.5000	93.3009	0.0265
9	0.4500	77.6946	0.0216	0.3500	99.8888	0.0276	0.2500	139.8426	0.0385
10	0.4500	25.9919	0.0082	0.3500	33.4092	0.0105	0.2500	46.7592	0.0146
11	0.2250	0.1406	0.0008	0.1750	0.1694	0.0011	0.1250	0.2175	0.0015
12	0.5500	0.0681	0.0005	0.6500	0.0599	0.0005	0.7500	0.0525	0.0005
13	0.1882	0.1033	0.0010	0.2909	0.1026	0.0010	0.4100	0.1019	0.0010
14	0.8289	42.2406	0.0122	0.7341	47.7120	0.0139	0.6229	56.2409	0.0165
15	0.0818	0.1267	0.0012	0.1139	0.1431	0.0014	0.1401	0.1687	0.0016
16	0.7471	46.8533	0.0134	0.6201	56.4527	0.0162	0.4829	72.5064	0.0208
17	0.0721	0.1406	0.0013	0.0951	0.1694	0.0016	0.1079	0.2175	0.0021

Table A.7 : Flow rate and concentration values for 3-3 CDBR S=0.14 system operating at a recovery of 55%, 65%, and 75%.

Stream number	55% Recovery			65% Recovery		
	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)	Flow rate (L/h)	Salt concentration (g/L)	Boron concentration (g/L)
f	1.0000	35.0000	0.0100	1.0000	35.0000	0.0100
0	1.0146	34.5297	0.0101	1.0213	34.3202	0.0102
1	0.6750	51.8432	0.0147	0.5250	66.6490	0.0188
2	0.3396	0.1170	0.0011	0.4963	0.1249	0.0012
3	0.0146	2.3227	0.0204	0.0213	2.4694	0.0219
4	0.2775	0.0009	0.0001	0.4055	0.0015	0.0001
5	0.0475	0.1170	0.0011	0.0695	0.1249	0.0012
6	0.2921	0.1170	0.0011	0.4269	0.1249	0.0012
7	0.2250	51.8432	0.0156	0.1750	66.6490	0.0199
8	0.9000	51.8432	0.0149	0.7000	66.6490	0.0191
9	0.4500	77.6945	0.0216	0.3500	99.8888	0.0276
10	0.4500	25.9919	0.0082	0.3500	33.4093	0.0105
11	0.2250	0.1406	0.0008	0.1750	0.1695	0.0011
12	0.5500	0.0681	0.0005	0.6500	0.0599	0.0005
13	0.1867	0.1036	0.0010	0.2886	0.1030	0.0010
14	0.8279	42.2927	0.0122	0.7327	47.7973	0.0139
15	0.0813	0.1269	0.0012	0.1132	0.1434	0.0014
16	0.7466	46.8823	0.0134	0.6196	56.5032	0.0162
17	0.0716	0.1406	0.0013	0.0946	0.1695	0.0016

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- **S. Kayaci, S. Kürklü, S. Velioglu, M. G. Ahunbay, S.B. Tantekin-Ersolmaz, W.B. Krantz,** A Novel Process for Concurrent Desalination and Boron Removal (CDBR), Euromembrane 2018, 9th - 13th July 2018, Valencia, Spain.