

**PERFORMANCE EVALUATION OF A NOVEL HYBRID
ADSORPTION DESALINATION WITH DIRECT
CONTACT MEMBRANE DISTILLATION**

BY

AYMAN ABDULHAK AHMED IBRAHIM AL-AZAB

A Thesis Presented to the
DEANSHIP OF GRADUATE STUDIES

KING FAHD UNIVERSITY OF PETROLEUM & MINERALS

DHAHRAN, SAUDI ARABIA

In Partial Fulfillment of the
Requirements for the Degree of

MASTER OF SCIENCE

In

CHEMICAL ENGINEERING

MAY 2022

KING FAHD UNIVERSITY OF PETROLEUM & MINERALS

DHAHRAN- 31261, SAUDI ARABIA

DEANSHIP OF GRADUATE STUDIES

This thesis, written by **AYMAN ABDULHAK AHMED IBRAHIM AL-AZAB** under the direction of his thesis advisor and approved by his thesis committee, has been presented and accepted by the Dean of Graduate Studies, in partial fulfillment of the requirements for the degree of **MASTER OF SCIENCE IN CHEMICAL ENGINEERING**.



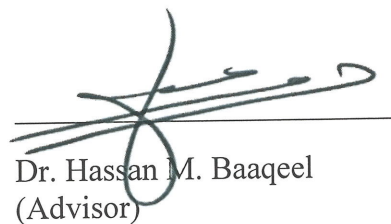
Dr. Hassan M. Baaqeel
Department Chairman



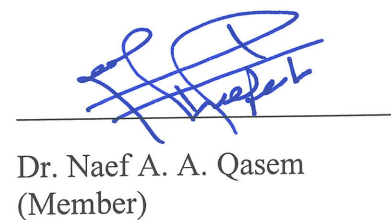
Dr. Suliman Al-Homidan
Dean of Graduate Studies

21/6/2022

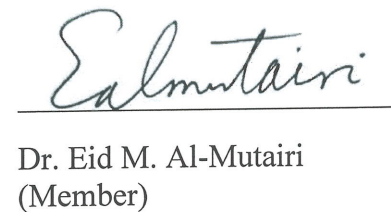
Date



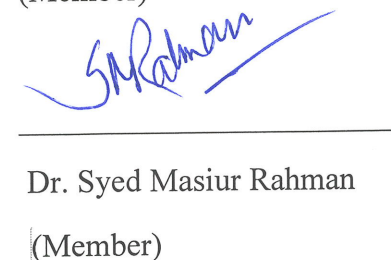
Dr. Hassan M. Baaqeel
(Advisor)



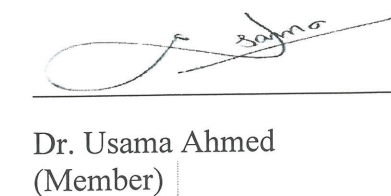
Dr. Naef A. A. Qasem
(Member)



Dr. Eid M. Al-Mutairi
(Member)



Dr. Syed Masiur Rahman
(Member)



Dr. Usama Ahmed
(Member)

© Ayman Abdulhak Ahmed Ibrahim Al-Azab

2022

To my beloved mother: Sameera Abdullah Al-Azab.

ACKNOWLEDGMENTS

First of all, I'd like to say: Alhamdullilah.

My profound gratitude goes to my advisor Dr. Hassan Baaqeel for his support, guidance, effort, encouraging words and valuable comments during our journey working on this thesis.

I would like also to express my gratitude and deep regards to my committee member Dr. Naef Qasem for his constant encouragement, exemplary guidance and monitoring throughout my thesis research work. I can't thank him enough for his time, efforts, and valuable input to achieve the objectives of my thesis project. The encourage, guidance and help provided by him time to time will carry me a long way in my future life on which I am about to embark.

My profound and special thanks go to my thesis committee members Dr. Syed Masiur Rahman, Dr. Eid Al-Mutairi, and Dr. Usamah Ahmed, for their help and comments.

My deep thankfulness, gratitude, and appreciation go to Dr. Tawfik A. Saleh, Eng. Meriem Merzoug and Dr. Jameel Almulaiki for their efforts and valuable inputs throughout my research.

With a special mention to Mr. Ahmed Saleh Al-Azzani (Abo Saleh), Mr. Ahmed Al-Sofi, and Mr. Husien Al-Sarari for their constant encouragement.

Thanks extend to my wonderful friends Karem Al-Garadi, Akram Al-Absi, Monther Al-Koshab, Ismaiel Mudhafar, Mohammed Abdulsalam, Mohammed Al-Zuhairi, Haitham Al-

Khawlani, Ala'a Al-Ghafori, Muneer Al-Jabeen, Khaldoon Saif Nasr, Khaldoon Kaleem, Redhwan A. A. Saleh and Anwar Al-Sharjabi.

Lastly, I would like to thank my family, who taught me the value of hard work by their own example. They rendered me enormous support whenever I needed it the most.

|

|

TABLE OF CONTENTS

ACKNOWLEDGMENTS	V
TABLE OF CONTENTS	VII
LIST OF TABLES	IX
LIST OF FIGURES	X
NOMENCLATURE.....	XIII
ABSTRACT.....	XVII
ملخص الرسالة.....	XIX
CHAPTER 1 INTRODUCTION.....	1
CHAPTER 2 LITERATURE REVIEW.....	12
2.1 Desalination Technologies	12
2.1.1 Membrane Desalination	13
2.1.2 Thermal Desalination.....	14
2.2 Emerging Desalination Technologies	17
2.3 Adsorption Desalination (AD)	18
2.3.1 AD System Description	19
2.3.2 Optimal AD Cycle Time.....	20
2.3.3 Selection of Silica Gel.....	20
2.4 Direct Contact Membrane Distillation (DCMD).....	24
2.4.1 DCMD System Description	24
2.4.2 Optimal Membrane Properties	26
2.5 Hybrid Desalination Technologies.....	27

2.6 Objectives	29
CHAPTER 3 SYSTEM MODELING	30
3.1 AD-DCMD Hybrid System Description	30
3.2 Mathematical Modeling	32
3.2.1 Direct Contact Membrane Distillation (DCMD)	32
3.2.2 Adsorption Desalination (AD)	38
3.2.3 Integrated System Performance	43
3.3 Constructive and Operating Parameters	45
3.4 Economic Model	47
CHAPTER 4 RESULTS AND DISCUSSION.....	50
CHAPTER 5 CONCLUSION.....	80
APPENDIX 1: MODEL VALIDATION	81
APPENDIX 2: AD-DCMD MODEL ALGORITHM	83
REFERENCES.....	84
VITAE	99

LIST OF TABLES

Table 1 Thermo-physical properties of different silica gel types.	21
Table 2 Recommended membrane properties for DCMD system.....	26
Table 3 Some reported integrations of thermal desalination technologies.	27
Table 4 Design and operating parameters of the AD-DCMD hybrid system.....	44
Table 5 Adsorption properties of Silica-gel/water pair [84].....	46
Table 6 Initial values of operating conditions used in calculation.....	46
Table 7 The capital investment cost for AD-DCMD hybrid system.	49

LIST OF FIGURES

Figure 1 Projected water stress by country: 2040 [4].	2
Figure 2 Total water usage in Saudi Arabia (Categorized by beneficiaries) [5].	4
Figure 3 Drinking water sources in Saudi Arabia [5].	5
Figure 4 Operational desalination facilities by feed water type [13,14].	8
Figure 5 Operational desalination facilities by technology [13,14].	9
Figure 6 Categories of desalination technologies.	12
Figure 7 Adsorption desalination system.	19
Figure 8 (a) The percentage of water desorbed at different regeneration temperatures. (b) The percentage of water desorbed at different times [54].	23
Figure 9 Direct contact membrane distillation process [58].	25
Figure 10 AD-DCMD hybrid system (a) configuration 1, (b) configuration 2, and (c) configuration 3.	31
Figure 11 Temperature profiles' history for the AD system components under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{cw,in} = 0.3\text{ kg/s}$, $m_{hw,in} = 0.4\text{ kg/s}$, $m_{chw,in} = 0.5\text{ kg/s}$, $m_{cond,in} = m_{DCMD}$, $f = 0.08\text{ kg/s}$, and $T_{bp} = 20\text{ }^{\circ}\text{C}$ for configuration 1.	51
Figure 12 Effect of half cycle time on GOR values of AD-DCMD hybrid system under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{cw,in} = 0.3\text{ kg/s}$, $m_{hw,in} = 0.4\text{ kg/s}$, $m_{chw,in} = 0.5\text{ kg/s}$, $m_{cond,in} = m_{DCMD}$, $f = 0.08\text{ kg/s}$, and $T_{bp} = 20\text{ }^{\circ}\text{C}$ for configuration 1.	52
Figure 13 Effect of mass flowrate of the inlet seawater to the condenser ($m_{cond,in} = m_{DCMD}$, f) on the freshwater production at different heating water flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{cw,in} = 0.3\text{ kg/s}$, $m_{chw,in} = 0.5\text{ kg/s}$, $T_{bp} = 20\text{ }^{\circ}\text{C}$, and half-cycle time= 700 s for configuration 1.	54
Figure 14 Effect of mass flowrate of the inlet seawater to the condenser ($m_{cond,in} = m_{DCMD}$, f) on the system GOR at different heating water flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{chw,in} = 0.5\text{ kg/s}$, $T_{bp} = 20\text{ }^{\circ}\text{C}$, and half-cycle time = 700 s for configuration 1.	55
Figure 15 Effect of temperature of the inlet seawater to the condenser on the freshwater production at different heating water temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{cw,in} = 0.3\text{ kg/s}$, $m_{hw,in} = 0.4\text{ kg/s}$, $m_{chw,in} = 0.5\text{ kg/s}$, $m_{bp} = 0.06\text{ kg/s}$, $T_{bp} = 20\text{ }^{\circ}\text{C}$, and half-cycle time= 700 s for configuration 1.	58

- Figure 16 Effect of temperature of the inlet seawater to the condenser on the system GOR at different heating water temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{cw,in} = 0.3\text{ kg/s}$, $m_{hw,in} = 0.4\text{ kg/s}$, $m_{chw,in} = 0.5\text{ kg/s}$, $m_{bp} = 0.06\text{ kg/s}$, $T_{bp} = 20\text{ }^{\circ}\text{C}$, and half-cycle time= 700 s for configuration 1. 59
- Figure 17 Effect of mass flowrate of the inlet seawater as a cooling water and a feed for DCMD ($m_{cw,in} = m_{DCMD}, f$) on the freshwater production at different heating water flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{hw,in} = 0.4\text{ kg/s}$, $m_{chw,in} = 0.5\text{ kg/s}$, $m_{cond,in} = 0.3\text{ kg/s}$, $T_{bp} = 20\text{ }^{\circ}\text{C}$, and half-cycle time= 700 s for configuration 2. 61
- Figure 18 Effect of mass flowrate of the inlet seawater as a cooling water and a feed for DCMD ($m_{cw,in} = m_{DCMD}, f$) on the system GOR at different heating water flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{chw,in} = 0.5\text{ kg/s}$, $m_{hw,in} = 0.4\text{ kg/s}$, $m_{cond,in} = 0.3\text{ kg/s}$, $T_{bp} = 20\text{ }^{\circ}\text{C}$, and half-cycle time= 700 s for configuration 2..... 62
- Figure 19 Effect of temperature of the inlet seawater as a cooling water and a feed for DCMD ($T_{cw,in}$) on the freshwater production at different heating water temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{hw,in} = 0.4\text{ kg/s}$, $m_{cw,in} = m_{DCMD}, f = 0.08\text{ kg/s}$, $m_{chw,in} = 0.5\text{ kg/s}$, $m_{cond,in} = 0.3\text{ kg/s}$, $T_{bp} = 20\text{ }^{\circ}\text{C}$, and half-cycle time= 700 s for configuration 2. 65
- Figure 20 Effect of temperature of the inlet seawater as a cooling water and a feed for DCMD ($T_{cw,in}$) on the system GOR at different heating water temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{chw,in} = 0.5\text{ kg/s}$, $m_{hw,in} = 0.4\text{ kg/s}$, $m_{cond,in} = 0.3\text{ kg/s}$, $m_{cw,in} = m_{DCMD}, f = 0.08\text{ kg/s}$, $T_{bp} = 20\text{ }^{\circ}\text{C}$, and half-cycle time= 700 s for configuration 2. 66
- Figure 21 Effect of membrane length and effective area on the outlet temperature of the DCMD hot channel..... 68
- Figure 22 Effect of inlet feed temperature of DCMD system on the freshwater production at different DCMD permeate temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $m_{hw,in} = m_f, DCMD = 0.08\text{ kg/s}$, $m_{cw,in} = 0.3\text{ kg/s}$, $m_{chw,in} = 0.5\text{ kg/s}$, $m_{cond,in} = 0.3\text{ kg/s}$, $m_{bp} = 0.06\text{ kg/s}$, and half-cycle time= 700 s for configuration 3..... 70
- Figure 23 Effect of inlet feed temperature of DCMD system on the system GOR at different DCMD permeate temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$,

mchw, in = 0.5 kg/s, mhw, in = mf, DCMD = 0.08 kg/s, mcw, in = 0.3 kg/s, mcond, in = 0.3 kg/s, mbp = 0.06 kg/s, and half-cycle time= 700 s for configuration 3.....	71
Figure 24 Effect of inlet feed flow rate of DCMD system (mf, DCMD = mhw, in) on the freshwater production at different DCMD permeate mass flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under Tcw, in = 30 °C, Tcond, in = 30 °C, Tchw, in = 30 °C, mcw, in = 0.3 kg/s, mchw, in = 0.5 kg/s, mcond, in = 0.3 kg/s, Tbp = 20 °C, Tbf = 80 °C, and half-cycle time= 700 s for configuration 3.	74
Figure 25 Effect of inlet feed flow rate of DCMD system on the system GOR at different DCMD permeate mass flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under Tcw, in = 30 °C, Tcond, in = 30 °C, Tchw, in = 30 °C, mchw, in = 0.5 kg/s, mcw, in = 0.3 kg/s, mcond, in = 0.3 kg/s, Tbp = 20 °C, Tbf = 80 °C, and half-cycle time= 700 s for configuration 3.	75
Figure 26 SDWP of the proposed configurations at different heat source temperatures..	77
Figure 27 COP of the proposed configurations at different heat source temperatures.	78
Figure 28 Freshwater costs at different heat source temperatures.	79
Figure 29 A comparison of experimental results values and those of the present simulation.	82

Nomenclature

Abbreviations

AD: adsorption desalination

t: time (s)

COP: coefficient of performance

U: overall heat transfer coefficient ($\text{kW}/\text{m}^2 \cdot \text{K}$)

COE: cost of electricity

x: salinity concentration (ppt)

GOR: gained output ratio

X: Mole fraction

DCMD: direct contact membrane distillation

Y: life expectancy (yr)

SDWP: specific daily water production ($\text{m}^3/\text{ton} \cdot \text{day}$)

ΔH : heat of adsorption; enthalpy or latent Heat of vaporization of water (kJ/kg)

Symbols

A: heat transfer surface area (m^2)

A_o : a constant

a : Salt activity coefficient

Cc: capital cost (\$)

CP: annual power cost (\$/yr)

CR: Capital recovery ratio

C_F : annual capital cost (\$/yr)

C_L : annual labor cost (\$/yr)

Greek letters

α : Contribution of Knudsen diffusion to mass transfer

ρ : Density kg/m^3

β : Concentration Polarization coefficient

γ : a constant is either 0 or 1

δ : Membrane thickness, m

ε : Porosity, %

C_{Mg} : annual management cost (\$/yr)

θ : a constant is either 0 or 1

C_p : specific heat capacity at constant pressure (kJ/kg·K)

μ : Viscosity Pa-s

C_{pw} : freshwater cost (\$/Liter)

τ : Membrane tortuosity

C_v : specific heat capacity at constant volume (kJ/kg·K)

D_c : diffusion coefficient (m^2/s)

Subscripts

D_e : effective diffusion coefficient, m^2/s

ads: adsorbed

D_h : Hydraulic diameter, m

b : Brine

d_{pore} : Pore diameter, m

bed : adsorbent bed

E: characteristic energy (kJ/kg)

bf: bulk Feed

E_a : activation energy (kJ/kg)

bp: bulk Permeate

f: plant availability

c: Conduction

H : Overall heat transfer coefficient, $W/m^2 K$

chw: chilled water

h_g : vaporization/condensation latent heat (kJ/kg)

cond: condenser

h : Convective heat transfer coefficient $W/(m^2 K)$

cw: cooling water

J : Mass flux, $kg/(m^2 s)$

e: exit

k : Thermal conductivity, $\frac{W}{m K}$

evap : evaporator

K_n : Knudsen number

f: Feed side

k_s : Solute mass transfer coefficient, m/s

hw: heating water

K_L : LDF constant (1/s)

hx : heat exchanger

l : specific labor cost (\$/m ³)	i : inlet
m : mass (kg)	k : Knudsen
M : Molecular weight, g/mol	m : membrane
$\dot{m}$: mass flowrate (kg/s)	mem : membrane material
n : a constant	mf : Membrane feed side
Nu : Nusselt number	mp : Membrane permeate side
P : operating pressure (kPa) in AD model, or Pa in MD model	M : Molecular
P : Pressure, Pa	p : Permeate side
q : actual adsorption amount (kg/kg)	pw : produced freshwater
q_{eq} : equilibrium adsorption amount (kg/kg)	s : silica gel (adsorbent material)
Q : Heat transfer, W/m ²	$sorp$: adsorption or desorption
R : gas constant, 8.314 kJ/kmol.K in AD model, or 8314 J/kmol.K in MD model	sw : seawater
r : Radius of membrane pore, m	v : Vapors/Vaporization
Re : Reynolds number	w : water
r_p : adsorbent particle radius (m)	$w.a$: Water in Air
Sc : Schmidt number	wv : water vapor
Sh : Sherwood number	wvp : water vapor inside the pores

T: temperature (°C) in AD model, or (K) in MD model

]

ABSTRACT

Full Name : [Ayman Abdulhak Ahmed Ibrahim Al-Azab]
Thesis Title : [Performance Evaluation of a Novel Integration of Adsorption Desalination System with Direct Contact Membrane Distillation System]
Major Field : [Chemical Engineering]
Date of Degree : [May 2022]

Water desalination is an important and effective approach to meeting the growing demand for freshwater and supplementing freshwater resources. The existing desalination technologies' main disadvantage is their excessive energy consumption. Adsorption desalination (AD) and direct contact membrane distillation (DCMD) are emerging desalination technologies that gained interest due to their ability to operate at low temperatures (below 85 °C). This thesis proposes a novel integration of adsorption desalination (AD) and direct contact membrane distillation (DCMD) processes, designated as the AD-DCMD system, in which the waste heat of the AD system is utilized to drive the DCMD system and vice versa. Three AD-DCMD configurations are proposed and studied. In configuration 1, the feed stream (seawater) of the DCMD system is heated by the waste heat inside the AD condenser. Configuration 2 proposes that the DCMD feed is heated by the waste heat in the adsorption bed. Configuration 3 has an opposite arrangement in which the AD system is driven by the DCMD hot brine. The main performance parameters such as gain output ratio (GOR), freshwater production rate, specific daily water production (SDWP), and freshwater cost are investigated. The key parameters investigated to influence the hybrid system's performance include AD cooling water mass flow and temperature, AD heating water mass flow and temperature, and cycling time. Results show

that GOR values reach 1.48, 1.6, and 1.25 for configurations 1, 2, and 3, respectively. At a source temperature of 90 °C, the freshwater production of the DCMD-AD hybrid system attains SDWP values of 26, 53 and 93 m³/ton.day for configurations 1, 2, and 3, respectively. Under the same conditions of the optimal performance, the cooling effect was obtained as a by-product for the three configurations. This was demonstrated by the coefficient of performance (COP) values which are 0.76, 0.70, and 0.52 for configurations 1, 2, and 3, respectively, at a heat source temperature of 60 °C. Moreover, all configurations have proved a capability of achieving a remarkable reduction in freshwater cost against standalone AD systems (0.06 \$/L). The freshwater cost has reached 0.049, 0.026 and 0.028 (\$/L) for configurations 1, 2 and 3, respectively. |

ملخص الرسالة

الاسم الكامل: أيمن عبدالحق أحمد إبراهيم العزب

عنوان الرسالة: دراسة وتقييم أداء تكامل نظام تحلية المياه باستخدام الامتزاز (Adsorption) مع نظام التقطير باستخدام غشاء التلامس المباشر (DCMD)

التخصص: هندسة كيميائية

تاريخ الدرجة العلمية: مايو 2022

تعد تحلية المياه طريقة مهمة وفعالة لتلبية الطلب المتزايد على المياه الصالحة للشرب، كما أنها تعتبر مكملة لموارد المياه العذبة الطبيعية. يتمثل العيب الرئيسي لتقنيات التحلية الحالية في استهلاكها المفرط للطاقة. تعد تحلية المياه باستخدام الامتزاز (Adsorption) و التقطير باستخدام الأغشية (MD) من تقنيات تحلية المياه الناشئة التي اكتسبت اهتماماً كبيراً من قبل الباحثين والمهتمين في تحلية المياه نظراً لقدرتها على تحلية الماء المالح عند درجة حرارة منخفضة (أقل من 85 درجة مئوية).

تحتوي هذه الأطروحة على دراسة متكاملة لمحاكاة تكامل جديد من نوعه بين التحلية بالامتزاز (AD) و التقطير باستخدام الغشاء من نوع (DCMD)، حيث تم الإشارة إليها باسم النظام الهجين (DCMD) والذي فيه يتم استخدام الحرارة المهدرة داخل أجزاء نظام الامتزاز (AD) لتشغيل نظام التقطير (DCMD) والعكس صحيح.

تم تكوين ودراسة نظام تحلية المياه الهجين (AD-DCMD) بثلاث طرق مختلفة. في الطريقة الأولى يتم تسخين مياه البحر المغذية لنظام (DCMD) بواسطة الحرارة المهدرة داخل مكثف نظام الامتزاز (AD). وفي الطريقة الثانية يتم استخدام الحرارة المهدرة في مفاعل الامتزاز (Desorption bed reactor) لتسخين تيار ماء البحر المالح المتوجه للتحلية في نظام التقطير (DCMD). في الطريقة الثالثة تم اقتراح ترتيب معاكس للطريقتين الأولى والثانية حيث يتم تشغيل نظام الامتزاز (AD) بواسطة المحتوى الحراري للمحلول الملحي الساخن الخارج من نظام التقطير (DCMD). تم التحقق من أداء وكفاءة الطرق المقترحة من خلال دراسة مؤشرات الأداء الرئيسية مثل (GOR) والذي يشير إلى كفاءة نظام التحلية الحراري في استخدام الطاقة، ومعدل إنتاج المياه الصالحة للشرب، والإنتاج اليومي للمياه لكل طن من مادة السليكا جل (Silica gel) والتي تعتبر المادة المستخدمة لأداء مهمة الامتزاز.

تظهر النتائج أن قيم GOR للطرق 1 و 2 تصل إلى 1.48 و 1.6 على التوالي. وتمثل هذه النسب حوالي الضعف لنسبة نظام الامتزاز (AD) وحيداً بنفس كمية الطاقة . كما أن عند دخول المياه المالحة بدرجة حرارة 90 درجة مئوية يبلغ الإنتاج اليومي من المياه الصالحة للشرب للنظام الهجين 26 و 53 و 93 متر مكعب لكل طن واحد من السيليكا جل (Silica Gel) .

أثبتت الطرق المقترحة قدرتها على تحقيق انخفاض ملحوظ في تكلفة المياه الصالحة للشرب مقارنة بسعر المياه المنتجة من نظام الامتزاز (AD) والتي تبلغ \$ 0.06 للتر الواحد، حيث بلغت تكلفة المياه الصالحة للشرب 0.049 و 0.026 و 0.028 \$ للتر الواحد باستخدام الطرق الأولى والثانية والثالثة على التوالي. |

CHAPTER 1

INTRODUCTION

Water is essential for all life forms to exist. Although water covers almost 70% of the earth's surface, roughly 97% of it is salt water. Although it is considered that the small percentage of fresh water obtainable is appropriate for supporting life forms on Earth, consumable water is not fairly distributed around the planet, and in some regions of the world, it is scarce. Thermal energy is the earliest source of producing fresh water naturally. Thermal energy is absorbed by the earth's surface, which supplies enough energy to evaporate sea water. Water vapor rises, generating clouds at various elevations. Hence, fresh water is obtained as rain. This fundamental principle, which is as natural as rain, is commonly applied in seawater desalination [1,2].

As of today, over 40% of the world's population has water scarcity. This percentage is predicted to rise to the 60s, owing primarily to the world's exponential population growth. The use of impure water is also a common cause of the majority of diseases and a significant number of deaths worldwide [1]. Nearly 1 % of the world's population relies on desalinated water, and by 2025, 14% of the world's population is expected to encounter water scarcity, according to the United Nations. The water desalination sector, therefore, has great potential unless water conservation is drastically improved. It is clear that seawater desalination is the only renewable source for producing fresh water [3]. Figure 1 illustrates the extensive water stress predicted by the ensemble projection of

climate models and socioeconomic scenarios using SSP2 (shared socio-economic pathway) and RCP8.5 (radiative forcing level stabilizes at 8.5 W/m² before 2100 by employment of a range of technologies and strategies for reducing greenhouse gas emissions). By 2040, a total of 33 countries, including Arabian Gulf countries, are expected to encounter extremely high water stress [4].

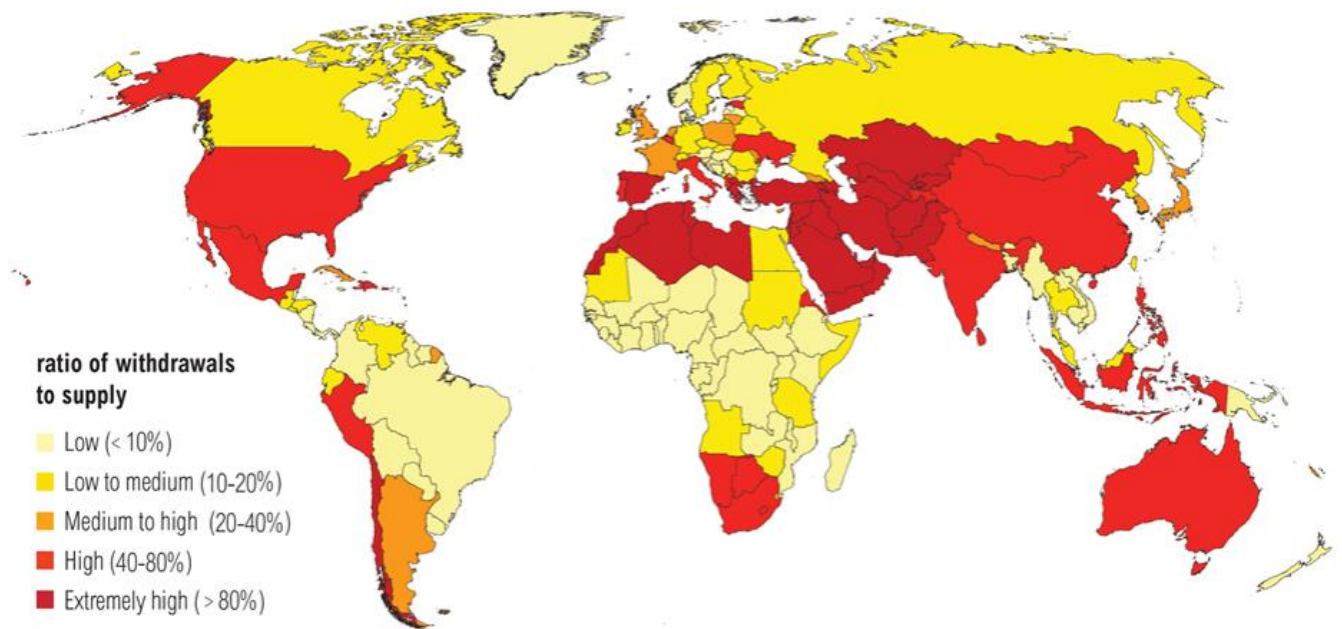


Figure 1 Projected water stress by country: 2040 [4].

The combined consequences of the world's continuously growing population, changing lifestyles, and limited natural supplies for pure consumable water have been significantly increasing the demand for desalination and water recycling. Desalination is a widely used method for supplying fresh water to dry zones all across the world, attributed to the fact that over 70% of the world's population lives within 70 kilometers of seas or oceans.

Desalination is commonly considered the most feasible and cost-effective solution to the world's fresh water shortage [1].

Saudi Arabia is classified as one of the most water-scarce nations on the planet. In Saudi Arabia, total urban water usage was calculated to be 3.39 cubic kilometers per year in 2018. Eighty-three percent of water use is consumed by agriculture and just 4 percent by industry (Figure 2) [5].

Next to the United States and Canada, Saudi Arabia is the world's largest water user, with a daily water intake of 250 liters per capita. Saudi Arabia has been the world's largest source of desalinated water to meet demand, accounting for 25 % of the global supply, and is home to the largest desalination plants in the world [6].

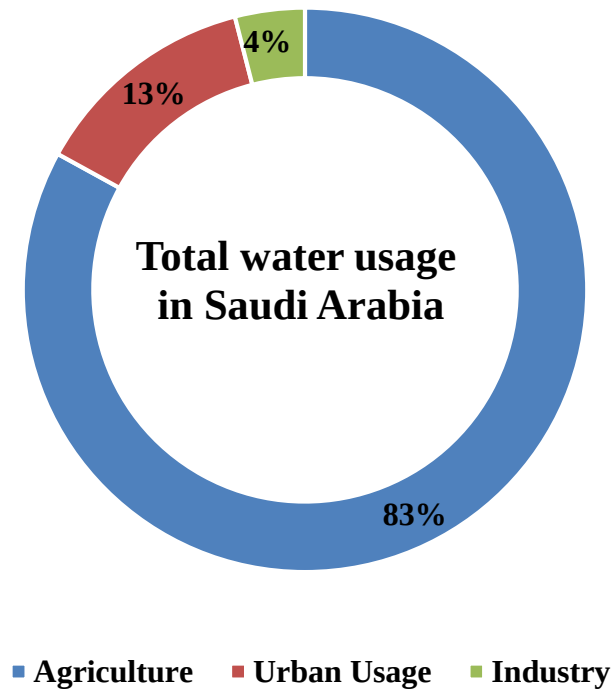


Figure 2 Total water usage in Saudi Arabia (Categorized by beneficiaries) [5].

Approximately 50% of drinking water today comes from desalination, 40% from non-renewable groundwater extraction, and just 10% from surface water in the mountains in the southern region of the country (Figure 3). Through the desalination process, the natural water supply is improved in various regions and cities of the Kingdom of Saudi Arabia, where there are extreme freshwater shortages. One of the key strategic priorities for overcoming the water shortage crisis is to develop a range of desalination plants, along

with support facilities in regions suffering from shortages of freshwater resources, based on the outcomes of technical feasibility studies [5].

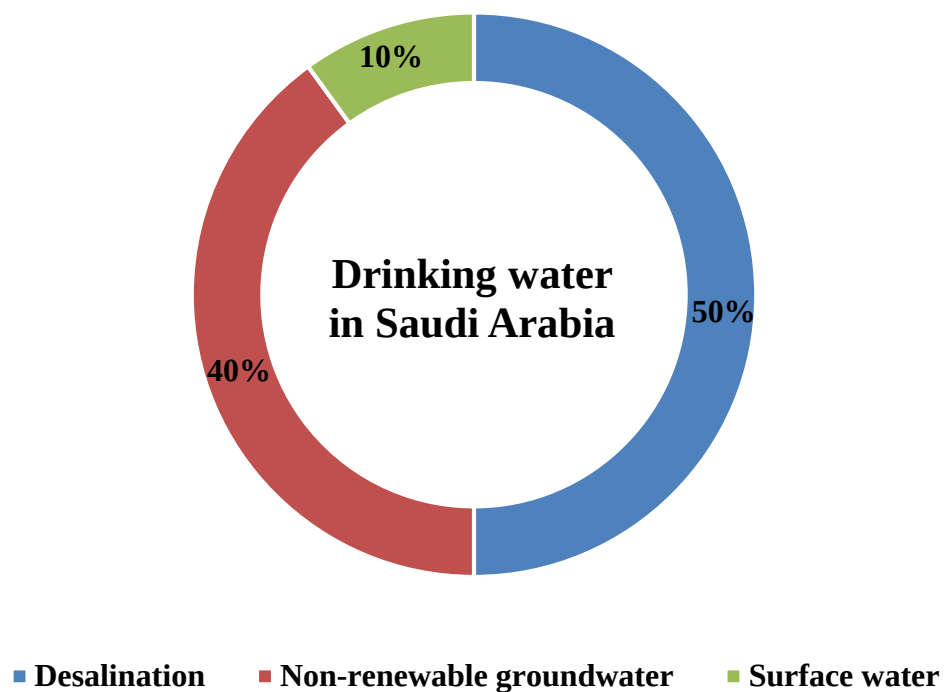


Figure 3 Drinking water sources in Saudi Arabia [5].

Desalination plants capacity increased from 47.6 million m^3/d at the end of 2008 to more than 78.4 million m^3/d in 2015. Groundwater, which has a tendency to become brackish with time, is the primary source of fresh water for inland communities, whereas coastal communities are more likely to use seawater. In the growing desalination market, roughly

40% of the desalination capacity treats brackish less saline water, whereas 60% of the desalination capacity treats seawater [2].

Historically, desalination plants have been established in locations where there is no alternative to public water supply, such as in the Gulf region. Desalination plants have moved to new locations around the world due to the improved desalination facilities (e.g. low cost of membrane desalination) and also the growing scarcity of fresh water around the world. It is stated that the output of the Magtaa desalination plant in Algeria is above 500,000 m³/d. At 1,025,000 m³/d, the Ras Al Khair project in Saudi Arabia, which uses both thermal and membrane technologies in tandem, is one of the world's largest thermal desalination plants. 150 countries throughout the world are currently using desalination, including Europe, the Middle East, and North Africa [9].

Nearly half of the world's total desalination capacity (48%) is generated in the Middle East and North Africa (MENA). Saudi Arabia, United Arab Emirates, and Kuwait account for the majority of the world's production with ratios of 15.5%, 10.1%, and 3.7%, respectively. On a global scale, East Asia and Pacific-North America regions account for 18.4% of desalination capacity, respectively, with China (7.5%) and the United States (11.2%) leading the production of the world's desalinated water. Gulf Cooperation Council (GCC) countries consume 65% more water than the annual world average (about 500 m³ per capita to 820 m³ per capita)

Groundwater supplies in several regions of the Gulf Cooperation Council (GCC) have decreased significantly in recent years, and the water table of aquifers has fallen below 200 m depth in some areas leading to water scarcity [10,11].

Seawater desalination has emerged as a primary solution to meet the water needs in the Kingdom of Saudi Arabia (KSA), which has limited freshwater resources. Groundwater is the primary source of freshwater in Saudi Arabia. In 1962, the country's shallow aquifers contained 557 billion m³ of water, making it one of the largest reserves on the world. Groundwater faces draining as time passing; only 84 billion m³ of water were left in 2010 after three decades of irresponsible agricultural development [10]. Only 3.9 billion m³ of groundwater is renewable. Therefore, Saudi Arabia reduced wheat production in 2008 and totally imported it starting in 2016 in order to conserve the country's limited water source. One of Saudi Arabia's water management policies is to reduce the amount of groundwater used for agriculture. In 1970, the Saudi government began implementing seawater desalination to meet the country's domestic water demand, and it quickly replaced groundwater as the primary domestic water source. Due to groundwater restrictions, the demand for desalination has increased. Because of the country's increasing population and domestic water demand, the Saudi government has already placed seawater desalination at the top of the national water management plan's priority list [10,12].

Membrane-based systems are gaining traction in countries such as Algeria, Australia, and Spain, while thermal processes continue to be widely used throughout the Middle East, particularly in Saudi Arabia, the United Arab Emirates, Bahrain, and Kuwait, where Saudi Arabia accounts for 34.8 % of all thermal desalination processes worldwide[2].

As indicated in Figure 4, in 2018, seawater desalination accounted for 61% of total desalination, and brackish water contributed 21% of the feed water source. Brackish water desalination is used to treat water with total dissolved salts (TDS) levels of 3,000 to 10,000

mg/L. Saline water has a TDS of more than 10,000 mg/L, whereas seawater has an average TDS of 35,000 mg/L. [13,14].

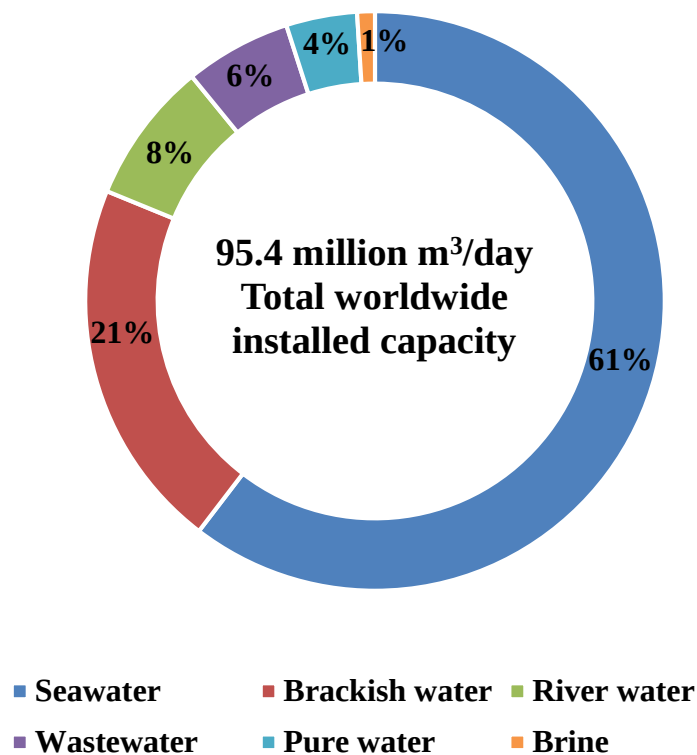


Figure 4 Operational desalination facilities by feed water type [13,14].

Desalination plants over the world use many technologies. Reverse osmosis (RO), multi-stage flash (MSF), and multi-effect distillation (MED) are the technologies that have been mostly spread in the marketplace. As a first step, desalination plants around the world began using MSF and MED thermal technologies. It is more likely that MED will take over MSF in thermal processes because it is less expensive and more efficient. However,

RO desalination capacity around the world has increased dramatically in recent decades due to the improvements in membrane performance and energy recovery systems. Compared to 2015, there was a 4% increase in the percentage of operational desalination facilities using RO, which became 69%, while the percentage of facilities employing emerging technologies such as forward osmosis (FO), Membrane distillation (MD) and adsorption desalination (AD) is 1% (Figure 5) [13,14].

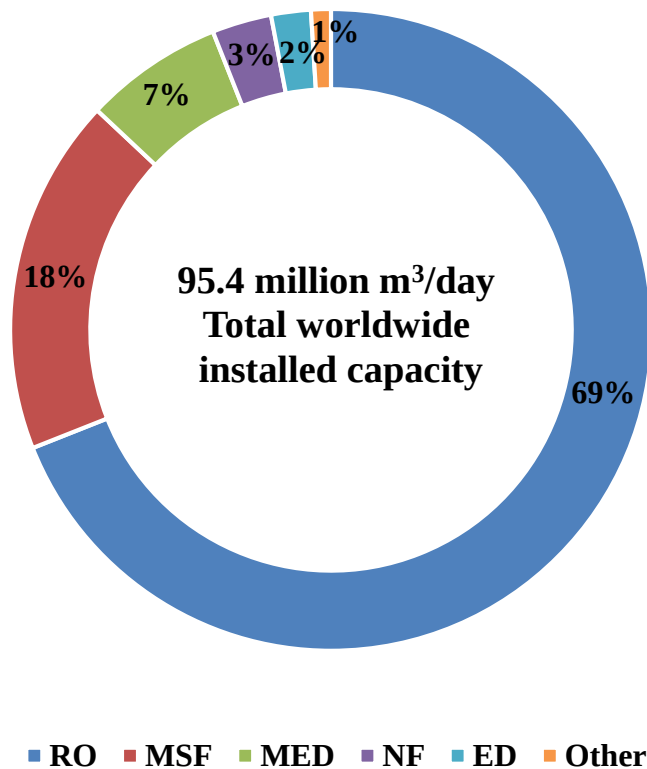


Figure 5 Operational desalination facilities by technology [13,14].

For all currently available practical desalination technologies, it takes 3.5 to 7 kW_{elec}/m³ (electrical energy consumption per water production). As a result, the Saudi Arabian government estimated that desalination produces 150 million tons of CO₂ annually. Desalination's high energy demand is intrinsically tied to global warming and climate change [13,16,17].

IDA Secretary-General Patricia A. Burke also stated that the growth in desalination is nonlinear, with other factors such as the availability of financing and the global oil prices influencing slow growth, but growth accelerators such as pollution of traditional water resources, industrial development, population growth, and climate change are still playing a significant role. Since its beginnings, the desalination industry has consistently supplied low-cost desalination options by developing energy-efficient technology, employing efficiency-improving techniques, and implementing environmentally friendly practices [2].

Based on the aforementioned research, The AD and DCMD systems can be driven at a low-energy source and produce a good amount of freshwater for a small (portable) design. To the best of authors' knowledge, the hybridization of the DCMD and AD system is not conducted in the literature so far. Furthermore, the performance of the AD system and traditional DCMD system (each as a standalone system) is not high and needs improvements. Therefore, the primary objective of this research is to assess the hybridization of DCMD and AD systems to improve the energy consumption per unit of freshwater production.

In this thesis, a hybrid AD-DCMD desalination system is presented in three different configurations. This hybrid system includes DCMD and AD units, which were integrated synergistically in a way that maximizes their performance and minimizes freshwater costs. In the presented AD-DCMD hybrid system, waste heat released from the AD system is used to drive the DCMD system, such as in configurations 1 and 2, or vice versa, such as in configuration 3. In configuration 1, the heat released inside the AD condenser resulted from the condensation of the water vapor coming from the desorption bed has been employed in heating the DCMD seawater feed stream. Configuration 2 has proposed that the DCMD feed stream passes through the adsorption bed to work as cooling water and gain the heat released inside the adsorption bed. The relatively high outlet brine temperature of the DCMD system (which can be enough for the desorption process) has proven a capability of being used as a heating source (hot water) for the AD system, and this arrangement resulted in configuration 3. The developed integrated AD-DCMD system configurations have demonstrated improved performance in terms of gain output ratio (GOR), specific daily water production (SDWP), and cost reduction per unit of freshwater. In addition, a cooling effect is obtained as a by-product in the three configurations. |

CHAPTER 2

LITERATURE REVIEW

2.1 Desalination Technologies

Two desalination technologies are extensively used worldwide: membrane desalination and thermal desalination (Figure 6).

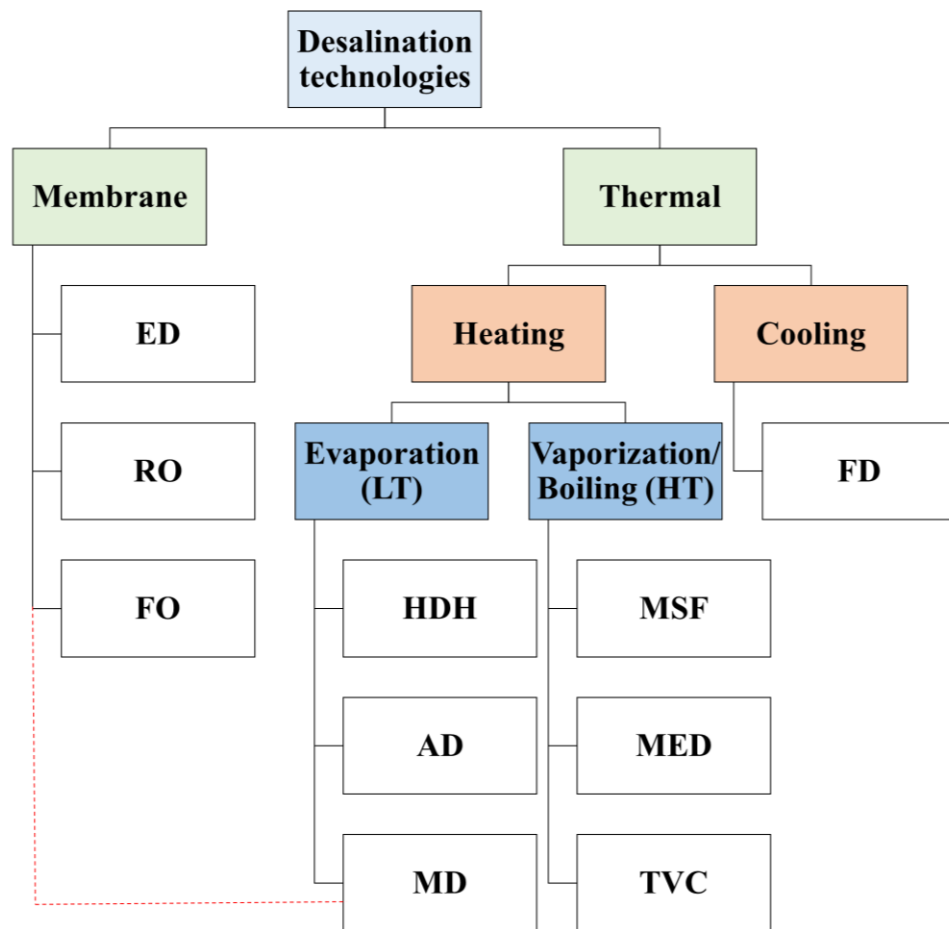


Figure 6 Categories of desalination technologies.

2.1.1 Membrane Desalination

Membrane desalination overtook thermal desalination due to the increase in energy costs and the improvements in membrane research, as well as the growing awareness of the environmental effects of fossil fuels. Membrane desalination and RO are preferred due to their low energy usage as well as their ability to handle a wide range of feed salinity [18]. When a semipermeable membrane is employed in membrane desalination, it can be viewed as pressure-driven filtration. Water to a certain extent, the membrane allows water to pass through, while solutes are prevented. Based on the degree of solute rejection or passage, the membrane processes are different. RO and Nanofiltration (NF) are the primary commercial membrane techniques currently in use.

In the late 1960s, RO desalination was firstly used for the desalination of brackish water. Its application was broadened to seawater desalination with additional advances and became a major competitor in the desalination market by the 1980s [19]. RO membranes have improved significantly over the past few decades, allowing for substantial water production at a reasonable cost [20]. Additionally, these advances allowed for reductions in energy consumption, providing customizable solutions to a wide range of water supply needs from a variety of water sources [21].

Polymeric materials, such as cellulose triacetate (CTA) or polyamide thin-film composites (TFC), have been used to make RO membranes. With a typical salt rejection rate of 99.4–99.8%, the key properties of the membrane are its water permeability and its impermeability of dissolved salts and ions. Applying hydraulic pressure to overcome the

osmotic pressure of feed water serves as the RO's primary driving force [22,23]. As a result, the pressure ranges from 15 to 25 bar for brackish water desalination (BWD) and 60 to 80 bar for saline water desalination (SWD) [24,25]. RO desalination accounts for almost 69% of the world's desalination capacity and for 95% of the total desalination capacity of membrane desalination [13]. In many ways, NF is similar to RO, but it uses lower pressures, and it's only suitable for desalinating brackish groundwater with certain low-salinity [26].

2.1.2 Thermal Desalination

Thermal desalination processes, including multistage flash distillation (MSF) and multiple-effect distillation (MED), were widely viable for a long time. Prior to the recent improvements in membrane technologies, the majority of the world's desalination capacity used to be provided by thermal processes [27]. Thermal desalination processes are typically used for high salinity feed-waters such as seawater, which can reach salinities as high as 45 g/l, as the case in the Arab Gulf Cooperation Council (GCC) countries and the Middle East and North Africa (MENA) region, which is considered extremely high compared to the standard seawater salinity is just 32 g/l. Thermal desalination's major drawbacks include low recoveries, high feed water temperatures, and, most critically, high energy consumption, which results in a high energy cost [22]. Desalination by thermal means is a form of phase change desalination, and as such, it is a relatively simple process that closely simulates the natural water cycle. Here, the water is heated until it evaporates, producing water vapor, which is then condensed and returns to a liquid state [15].

Multi-stage flash distillation (MSF):

The fundamental principle of the multi-stage flash process is flash evaporation, where the heated seawater boils once it is subjected to reduced pressure. Typically, saline water is heated by low-pressure steam up to 90 to 120 °C and delivered into the flashing chamber. Flashing action causes the fed seawater to boil and evaporate [28]. Brine that is not flashed moves to the next flash chamber or stage, where it gets flashed at a lower pressure. Therefore, the water can be flashed in stages without having to apply any additional heat to it along the process. The pressure in each stage should be maintained below the pressure which corresponds to the stage feed saturation temperature. Water vapor produced from each stage is condensed and cooled by preheating the seawater that enters the process, resulting in a higher MSF's economic efficiency [15].

Multiple-effect distillation (MED):

According to historical records, multi-effect distillation was the first desalination process to be reported in the literature in the 1840s. Thermal desalination was largely dominated by MSF technology during the 1950s and 1960s. A new wave of technological and economic advancements in the MED process has re-established the competition with the MSF process again for the MED process [29]. With low-temperature multi-effect distillation (LT MED) being the most efficient thermal distillation process, the MED process is more thermodynamically efficient than MSF. Accordingly, it consumes less power and provides a better performance ratio than MSF [30].

In MED, steam is used to bring seawater up to a boiling point in the first evaporator, which is also called the "effect." The seawater is preheated and then sprayed onto the evaporator's surface to accelerate the process of evaporation. Brine that hasn't been evaporated in the first effect is fed into the second effect, where further evaporation occurs [31]. In the second effect, un-boiled brine is heated by vapors from the first effect, and the process is repeated, with decreased temperature and pressure, in successive effects [28].

Vapor compression desalination (VC):

In vapor compression, the heat source is obtained by condensing vapor. Condensation occurs as a result of the compression process, which raises the temperature and pressure of vapor. The developed latent heat can be utilized to heat water for further vapor production [32]. There are two methods to compress the vapor and supply the necessary heat. Mechanical vapor compression, or MVC, is the first type of vapor compression, which employs an electrically driven compressor [33]. The second is a thermo-compressor (TVC), in which a venture orifice at the steam jet creates the water vapor and captures it from an evaporator, hence lowering the ambient pressure of the system [28]. In most cases, the VC process is used on a small or medium-sized scale. Though in practice, big VC units are often used in an integrated form with other thermal desalination technologies, such as MSF and MED [34,35]. In thermal desalination, pretreatment of the feed water is mandatory, with less strict pretreatment requirements for MSF and MED processes [19]. Considering the greater temperature required for thermal desalination, some chemicals are usually added to avoid scaling. Antiscalants, antifoam, and biocides are some examples.

Previously mentioned chemicals should be able to withstand temperatures as high as 120–160 °C as the case in MSF plants [36].

2.2 Emerging Desalination Technologies

Previous sections examined the two primary desalination technologies, membrane, and thermal desalination technologies. Research is incredibly active in the desalination field. This is attributed to the critical need to secure a reliable global water supply, especially with increasing water stress worldwide and the high energy consumption of current desalination systems. About 75.2 TWh per year is consumed by the existing worldwide desalination capacity of 38 billion cubic meters per year (BCM/y), and this consumption is expected to increase by 40% by 2030 and then another 55% by 2050 [37]. The present research works in the field of desalination take care of creating innovative desalination technologies, continuing advancement of underdevelopment desalination technologies, and new system designs combining two or more desalination processes. The primary goals of the previously mentioned efforts include increasing system recovery, reducing energy consumption, reducing overall costs, utilizing shared facilities, and, perhaps most importantly, reducing the impact on the environment [38].

Herein, several promising emerging desalination technologies are first introduced, followed by a brief discussion of hybrid desalination technologies. Currently, water desalination, as well as renewable energy, have a direct impact on human welfare and sustainability, as well as their connection with the environment and food in what are known as water-energy-food and water-energy-environment nexuses [39,40]. It is therefore critical to develop innovative desalination methods with the ultimate goal of reducing

energy consumption and increasing water productivity. Several promising new desalination technologies are now being developed at bench, lab, and pilot scales. Some of these processes are novel, while others are being reintroduced as a result of recent scientific advancements [41].

Several top new desalination techniques have been developed, including AD, MD (e.g., DCMD), solar desalination, forward osmosis (FO), capacitive deionization (CDI), electrodialysis and electrodialysis reversal (ED/EDR), and humidification dehumidification (HDH) [38,41]. Other desalination technologies are being researched and developed, although they are still in the early stages of development. Newly developed bioelectrochemical-based desalination systems, including microbial desalination, are more energy efficient because they are bio-based; they are also more environmentally friendly [42,43].

2.3 Adsorption Desalination (AD)

Due to their low impact on the environment, adsorption technologies have been first used as an alternative to conventional air conditioning systems [44,45]. Since the late 1980s, water and silica gel systems have been sold commercially in Japan [46]. Adsorption as a water desalination technology gained interest due to its ability to operate at low temperatures (below 85 °C), supplied either by exhaust gases from other processes or solar energy, as designed by K. C. Ng et al. (2011) [47].

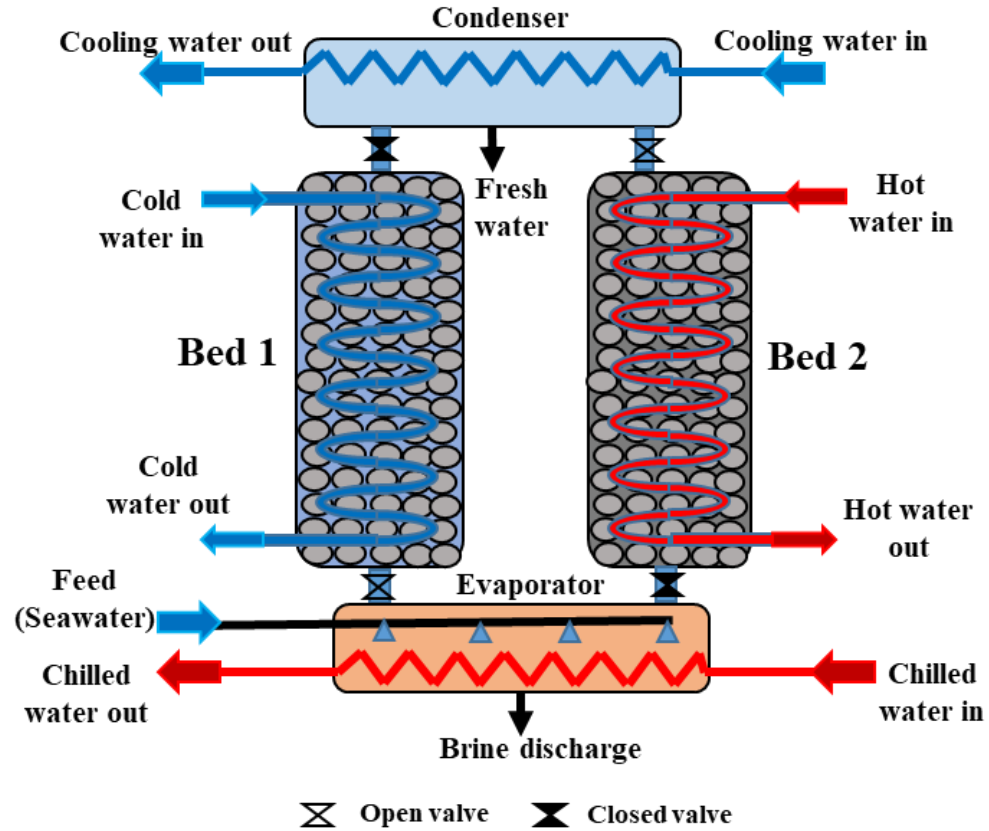


Figure 7 Adsorption desalination system.

2.3.1 AD System Description

As illustrated in Figure 7, The main components of the AD process are an evaporator, an adsorption bed, and a condenser. The evaporator evaporates sea or brackish water, which is then transported to silica gel for adsorption (the evaporation-adsorption phase). Adsorption heat is rejected using a cooling water circuit, and the evaporation temperature is maintained using an external heat load. After the preset time for the evaporation-adsorption phase has passed, hot water powered by an external heat source is used for the condensation-desorption phase, and the potable water is finally collected from the condenser [48]. To maximize the efficiency of the adsorption desalination cycle, low-cost

hydrophilic porous materials are used as adsorbents. High surface area and low regeneration temperatures make Silica gel an excellent option as an adsorbent [49].

The AD desalination technology has many advantages, such as the production of potable water free of biological contamination and lowered scaling and fouling. In addition, to be energy efficient, the AD cycle has almost no moving parts, which ensures that it is extremely low-maintenance by design [50].

2.3.2 Optimal AD Cycle Time

In the reported literature, the cycle time was not simulated or modeled; instead, it was determined experimentally. An experimental investigation of the optimal cycle time was performed for a two-bed/four-bed system for the temperature range of 65-85 °C, and the operative cycle time was varied between 1240 s and 1960 s for the two-bed/four-bed AD system [21,51]. This means the optimum half-cycle time is in the range between 620 s and 880 s.

2.3.3 Selection of Silica Gel

The AD system uses silica gel as an adsorbent; however, there are numerous types of silica gel in the marketplace, and not all of them are appropriate for use in the AD system. For instance, silica gel with pore size ranges from 2.0 to 3.0 mm is used as desiccants packed in vapor-permeable plastic. It works well to control local humidity and prevent items from spoiling or degrading. This type of silica gel with pore-size ranges between 2.0 to 3.0 mm is inappropriate for AD purposes [52]. Silica gel is essential for the AD system due to its

high adsorption capacity, rapid regeneration time, low regeneration temperature, and good thermal conductivity. As an AD adsorbent, the typical pore diameter needs to be between 2.0 and 4.0 nm in size [53]. Table 1 shows the thermodynamic properties of different types of silica gel, including Fuji Davison Type A, Type 3A, and Type RD.

Table 1 Thermo-physical properties of different silica gel types.

Property	Type A	Type 3A	Type RD
Specific surface area (m^2/g)	650	606	650
Porous volume (ml/g)	0.36	0.45	0.35
Average pore diameter (nm)	2.2	3.0	2.1
Apparent density (kg/ m ³)	730	770	800
pH value	5.0	3.9	4.0
Water content (wt. %)	< 2.0	0.87	—
Specific heat capacity (kJ/kg K)	0.921	0.921	0.921
Thermal conductivity (W/m K)	0.174	0.174	0.198
Mesh size	10-40	60-200	10-20

Source: [54].

The rate of desorption, the rate at which water vapour is removed from the adsorbent, is affected by temperature and time. Fuji Davison Type A, Type 3A, and Type RD silica gels were regenerated at various temperatures and times, and the percentage desorption

accomplished is shown in Figure 8, a. To achieve 95% regeneration, a 90 °C temperature heat source was sufficient for all three varieties of silica gel (Figure 8, b). Type 3A silica gel takes 2.5 minutes to desorb 95% of its original mass of water at 140 °C, while Type RD and Type A silica gel need 4 minutes and 5 minutes, respectively, to desorb the same amount of water. Due to its higher porous volume, Type 3A silica gel has the fastest response time [54].

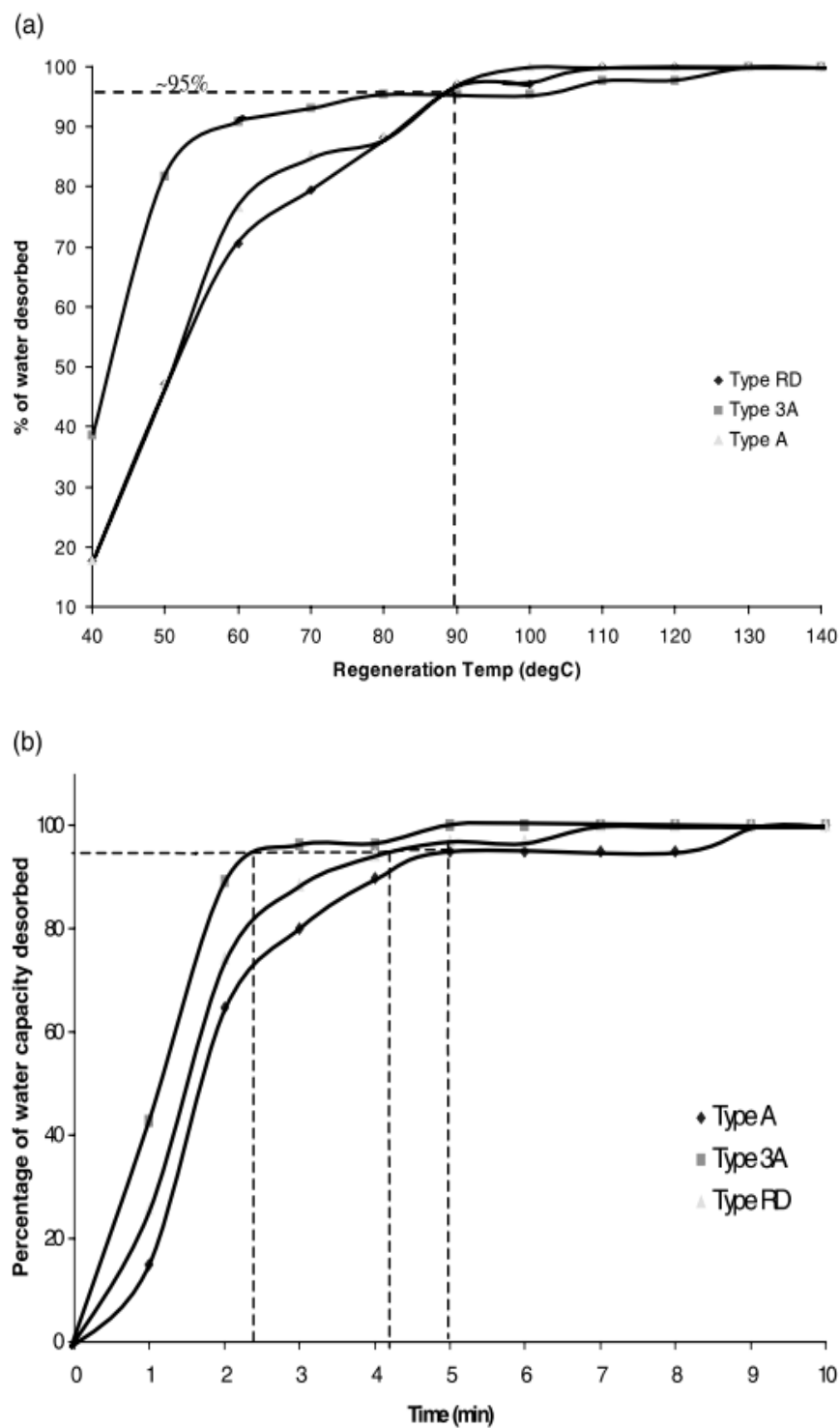


Figure 8 (a) The percentage of water desorbed at different regeneration temperatures. (b) The percentage of water desorbed at different times [54].

2.4 Direct Contact Membrane Distillation (DCMD)

Membrane distillation (MD) is an emerging thermally driven method of separation that only transports water vapor through a hydrophobic, microporous membrane [50]. The partial vapor pressure difference maintained at the membrane's two interfaces (hot feed and cold permeate sides) serves as the driving force for MD's operation [55].

Based on the condition applied on the cold permeate side of the membrane, different MD configurations have been proposed; direct contact membrane distillation (DCMD), vacuum membrane distillation (VMD), air gap membrane distillation (AGMD), and sweeping gas membrane distillation (SGMD), whose cold permeate sides include cooling water, vacuum, cold plate for stagnant air, and inert gas, respectively [56]. Other MD configurations, such as liquid-gap MD and material-gap MD, have been recently created for the purpose of improving the permeate flux [55].

2.4.1 DCMD System Description

DCMD is the most commonly used among MD configurations because it doesn't require additional equipment such as vacuum pumps or condensers. It's called direct contact MD because both the hot and cold channels directly interact with the membrane on opposite sides (Figure 9). The permeate is obtained from the evaporation of hot inlet volatile compounds, which, later on, condensate in the cold side of the membrane module [57].

The pressure difference across the membrane transports the vapor to the permeate side, where it condenses inside the membrane module. Due to the hydrophobic nature of the

membrane, water in its liquid phase cannot pass through the membrane; however, only the gas phase can penetrate through the membrane pores.

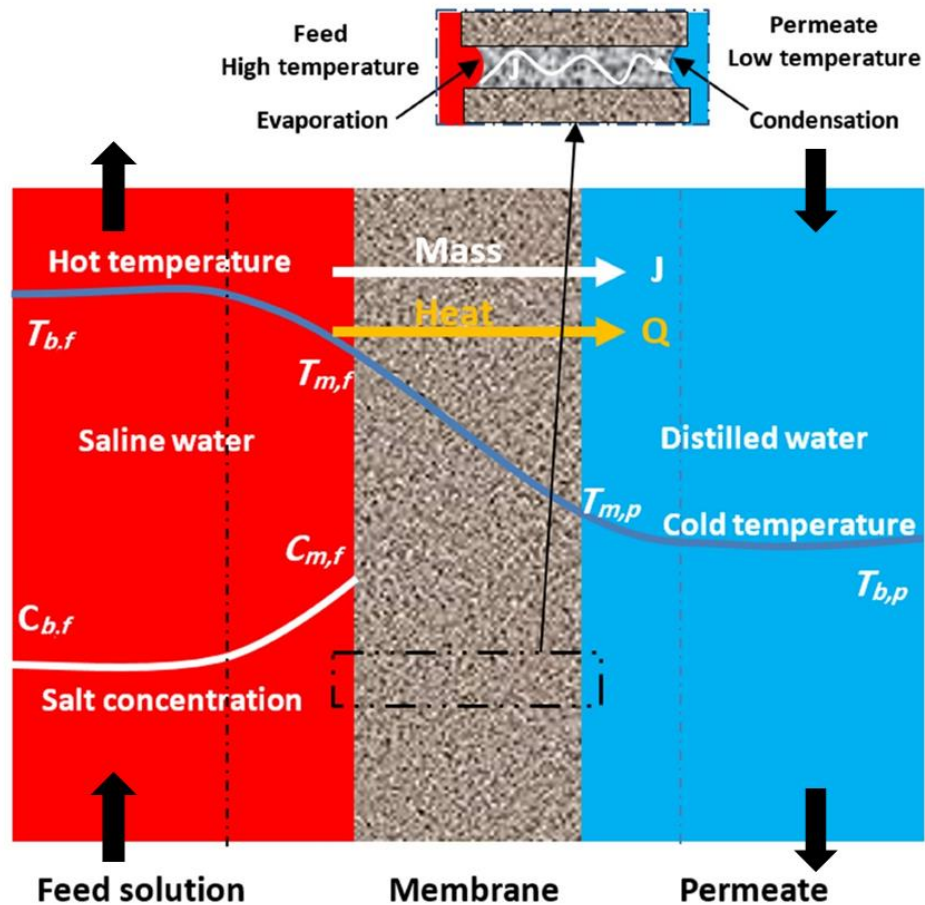


Figure 9 Direct contact membrane distillation process [58].

DCMD relies on a dynamic interaction between the simultaneous transfer of mass and heat since a temperature gradient leads to mass transport from the feed side to the permeate side [59,60]. In this configuration, the main challenge is heat loss through the membrane [55].

2.4.2 Optimal Membrane Properties

To date, DCMD has been developed with a variety of membrane types, including organic (polymeric), inorganic (ceramic), and composite/hybrid (organic and inorganic). Polymeric materials are the most commonly used in DCMD due to their inherent hydrophobicity, commercial availability, and low production costs. Polymeric membranes, on the other hand, have some drawbacks, including low mechanical, chemical, and temperature stability. Table 2 summarizes the best membrane properties for DCMD [57].

Table 2 Recommended membrane properties for DCMD system.

Property	Recommendation
Contact angle	>105°, as high as possible
Pore diameter	0.1–1 μm
Liquid entry pressure	>2.5 bar
Porosity	80–90%
Thickness	2–700 μm
Tortuosity	The lowest possible
Thermal conductivity	The lowest possible
Tensile strength	The highest possible

Sources: [57,61,62].

2.5 Hybrid Desalination Technologies

Previously, a single type of desalination technology, e.g., MSF, MED, or RO, used to be utilized in one desalination plant. Hybrid systems, on the other hand, are becoming more popular as a way to combine several desalination processes at a single location. This strategy is employed in order to reduce energy consumption, minimize the cost, and maximize the overall recovery, hence reducing the brine discharge [63]. Table 3 shows how waste heat of some reported thermal desalination technologies was utilized to enhance water recovery and energy performance. Based on the type of the integrated processes, there are two hybrid types, homogeneous and heterogeneous hybridizations. Homogeneous hybridization combines two processes of the same type; whereas heterogeneous hybridization combines two processes of

different types (e.g., membrane with thermal processes) [15].

Table 3 Some reported integrations of thermal desalination technologies.

Hybrid system	Configurations / features	Major findings	Ref.
MED-DCMD	• Multi-generation of cooling-power-desalination. • The DCMD unit was integrated to produce additional freshwater by using the heat released from the	• Total freshwater capacity of 165 m³/day • Optimal membrane evaporation efficiency, mass flux through the membrane DCMD and GOR were 46.2%, 9.8×10^{-6} kg/m²s and 5.6, respectively	[64]

	absorber while the rejected brine from MED was discharged.	
MED-DCMD	The MED brine is treated by the DCMD allowing at the same time a preheating of the MED feed water.	[65] MED-DCMD configurations outperformed standalone MED, with PR increased by >25%
HDH-AD	• Air heated HDH cycle. • The brine rejected from the HDH system was fed into the AD evaporator. • The heat released from the AD condenser is used to drive the HDH system.	[66] • Improved GOR from <3.5 for standalone HDH to 8 for HDH-AD • Reduced freshwater cost from 2.01 ¢/L for standalone HDH to 0.65 ¢/L for HDH-AD.
MED-AD	• The last stage of parallel feed MED was connected to the AD adsorbent bed to take up the vapor. • RSM-based optimization.	[67] • Last stage temperature reduced to 10 °C • Compared to standalone MED, water production increased by 14.7 times, PR increased by 12.9, and specific energy consumption reduced by 11.3%.

MED-AD	The last stage of parallel feed MED was connected to the AD adsorbent bed to take up the vapor.	• Temperature range extended from 50–38 °C to 50–3 °C • Freshwater production 3-fold of standalone MED
--------	---	---

2.6 Objectives

The main objectives of this thesis include:

1. Synergistically integrate direct contact membrane distillation (DCMD) and adsorption desalination (AD) units in a way that maximizes the system performance and minimizes freshwater costs.
2. Efficient utilization of the waste heat released from the AD system to drive the DCMD system.
3. To develop a simulation-based model which evaluates the performance of the hybrid AD-DCMD system in terms of:
 - Gain output ratio (GOR).
 - Specific daily water production (SDWP).
 - Cost reduction per unit of freshwater.

CHAPTER 3

SYSTEM MODELING

3.1 AD-DCMD Hybrid System Description

Due to the waste heat inside the AD condenser and the adsorption bed, the AD-DCMD system is proposed as a hybrid desalination system composed of AD and DCMD units. The DCMD and AD are integrated synergistically in three configurations. In configuration 1 (Figure 1, a), the waste heat in the AD condenser (from the hot water stream desorbed from the desorption bed) is used to drive the DCMD system. Therefore, pre-heat the seawater stream before entering the DCMD process, hence eliminating the DCMD heater. Configuration 2 (Figure 1, b) is similar to configuration 1, with the only difference that the DCMD seawater feed passes through the sorption bed to get used of the waste heat, which is rejected during the desorption process.

In the third configuration (Figure 10, c), the hot brine stream discharged from the DCMD process is used to pre-heat the desorption bed. This results in maximizing the water production of the hybrid system. Moreover, by eliminating the AD heater and minimizing the heat usage, the produced freshwater cost is simultaneously reduced.

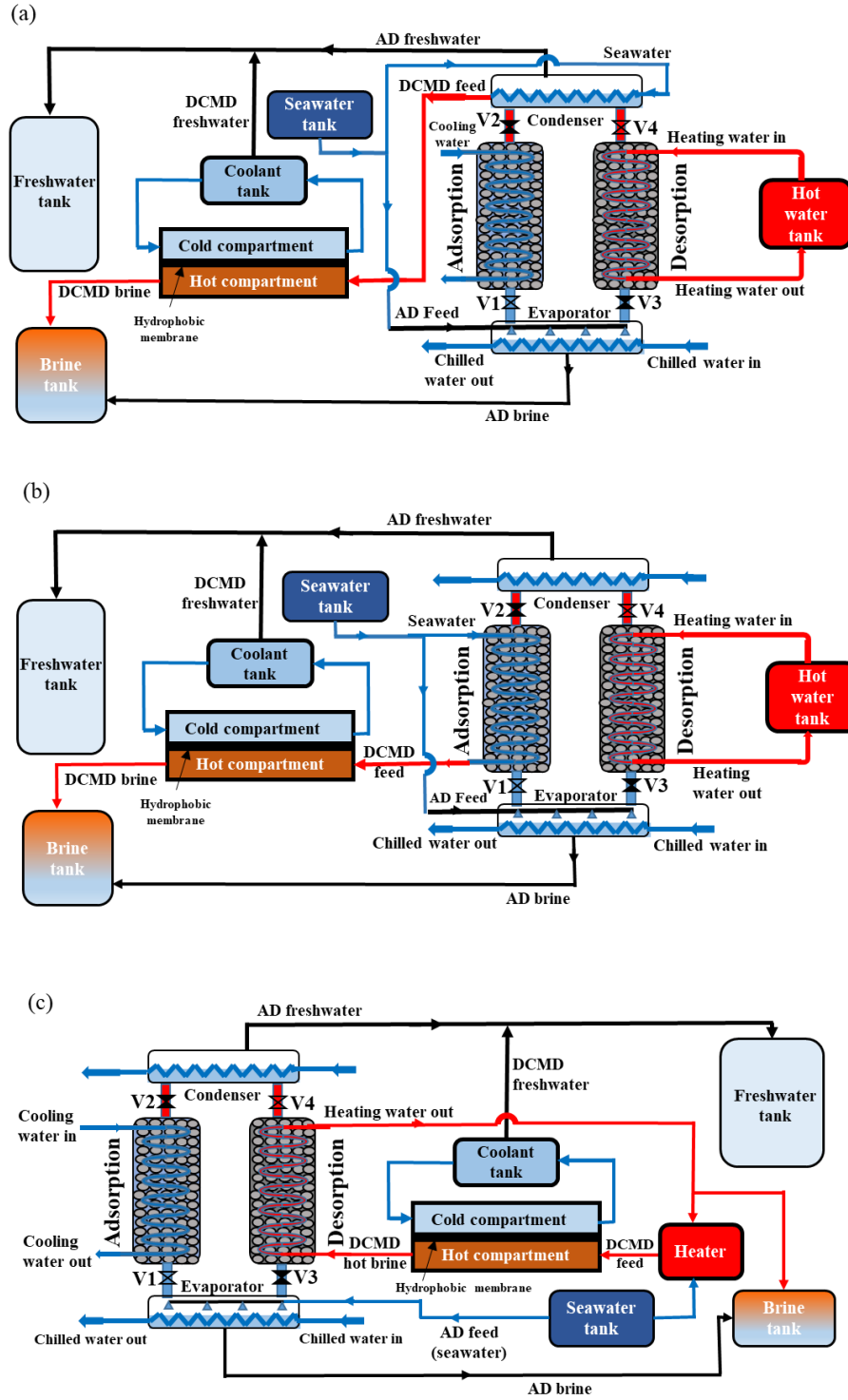


Figure 10 AD-DCMD hybrid system (a) configuration 1, (b) configuration 2, and (c) configuration 3

3.2 Mathematical Modeling

3.2.1 Direct Contact Membrane Distillation (DCMD)

- Mass transfer

The permeate mass flux through the membrane is primarily determined by the difference in vapor pressure between the feed and permeate sides of the membrane. The permeate mass flux can be calculated as follows [62,69,70];

$$J = D_e \Delta p_m = D_e (p_{mf} - p_{mp}^*) \quad (1)$$

where J is the permeate mass flux and D_e is the equivalent diffusion coefficient. p_m denotes the vapor pressure difference at transmembrane surfaces. p_{mf} and p_{mp}^* are the partial pressures at the feed and the permeate sides of the, respectively. The partial pressures are calculated based on the assumption that the membrane surfaces are in thermodynamic equilibrium [71]. The salinity of sodium chloride affects the partial pressure at the feed side, which must be taken into account when calculating the pressure [72].

$$p_{mf} = p_{mf}^* X_{wf} a_{wf} \quad (2)$$

where a_{wf} and X_{wf} denote the activity coefficient of the water mole fraction in the feed, respectively. The mole fraction is the ratio of the number of moles of any specie to the total number of moles present in the solution, and the activity coefficient shows the variation of substances from their ideal behavior due to impurities.

The activity coefficient for an aqueous solution of NaCl is given as [69,72–74]:

$$a_{wf} = 1 - 0.5 X_{NaCl} - 10 X_{NaCl}^2 \quad (3)$$

p_{mf}^* and p_{mp}^* are the vapor pressures of water and can be calculated at the membrane surface temperatures T_{mf} and T_{mp} , using the Antoine equation as follows:

$$p_{mf}^* = \exp\left(23.1964 - \frac{3816.44}{(T_{mf} - 46.13)}\right) \quad (4)$$

$$p_{mp}^* = \exp\left(23.1964 - \frac{3816.44}{(T_{mp} - 46.13)}\right) \quad (5)$$

The transport of gases and vapor through porous media is explained by three different mechanisms. They are the Knudsen model, molecular diffusion model, and Poiseuille flow model. The Knudsen flow and molecular diffusion models are applicable for DCMD. There is no trans-membrane hydrostatic pressure exerted because both the feed and permeate solutions are kept under constant pressure (about 1 atm) inside the membrane module. Therefore, the Poiseuille flow is insignificant in this case [75]. The ratio of the Knudsen diffusion to the molecular diffusion was used to calculate the combined influence of the molecular and Knudsen diffusions. The controlling mechanism in the mass transfer is determined by a ratio α whose value varies from 0 to 1 and can be given as [69]:

$$\alpha = 0.9469 \left(\frac{d_{pore}}{d_i}\right) + 0.0796 \quad (6)$$

Where d_{pore} and d_i are the mean fiber diameter and the size of inter-fiber space of the membrane, respectively. The value of d_i was taken 51.1×10^{-7} [69].

$$D_e = \left(\left(\frac{\alpha}{D_k}\right) + \left(\frac{1-\alpha}{D_m}\right)\right)^{-1} \quad (7)$$

D_e , D_k , and D_m represent the effective, Knudsen, and molecular diffusion coefficients, respectively. D_k , and D_m are determined using the following expressions [76]:

$$D_k = \left(\left(\frac{3 \varphi}{2 \varepsilon d_{pore}} \right) \left(\frac{\pi R T_m}{8 M_w} \right)^{0.5} \right)^{-1} \quad (8)$$

$$D_M = \left(\frac{R T_m \varphi P_{air,pore}}{M_w \varepsilon P D_{w,a}} \right)^{-1} \quad (9)$$

Where ε is the membrane porosity, R is the universal gas constant, T_m is the mean or average temperature across the membrane surfaces, d_{pore} is the pore diameter, M_w is the molecular weight of water, P is the total pressure inside the pore (assumed to be constant and equal to the sum of the air and water partial pressures), and $D_{w,a}$ is pressure independent molecular diffusion coefficient of water in air. φ is the hypothesis path through the membrane, which is given as [58]:

$$\varphi = \delta \tau \quad (10)$$

where δ is the membrane thickness and τ is the membrane tortuosity which can be calculated as [77]:

$$\tau = \frac{1}{\varepsilon} \quad (11)$$

$PD_{w,a}$ (Pa m²/s) is calculated using the following formula, which can be used in the temperature range of 273–373 K [73]:

$$PD_{w,a} = 1.895 \times 10^{-5} T_m^{2.072} \quad (12)$$

$$T_m = \frac{T_{mf} + T_{mp}}{2} \quad (13)$$

When air and water vapors fill the pores, the partial pressure of air inside the membrane pores are written as:

$$P_{air,pore} = P_{pore} - P_{wvp} \quad (14)$$

Where P_{wvp} is the partial pressure of water vapors inside the pores, which can be determined using the Antoine equation based on the mean temperature across the membrane surfaces (T_m), and P_{pore} is the total pressure inside the pores, which is assumed to be the average of the feed and permeate side pressures:

$$P_{pore} = \frac{P_f + P_p}{2} \quad (15)$$

- Heat transfer

Different types of heat transfer are identified in the DCMD system. On the feed side, where water is flowing over the membrane surface at high temperature and salinity, the heat transfer is due to convection. The fluid then flows on the permeate side, resulting in convective heat transfer. Heat is transferred across the membrane via conduction and mass flux. Hence, heat transfer in a DCMD module occurs in three zones:

(a) Convective heat transfer from the feed side to the membrane surface (Q_f)

It is governed by Newton's law of cooling [58]:

$$Q_f = h_f (T_{bf} - T_{mf}) \quad (16)$$

Where T_{bf} is the mean feed bulk temperature of the inlet and outlet of the hot feed stream and h_f is the convective heat transfer coefficient in the feed side which depends on the type of the flow (laminar or turbulent).

(b) Heat transfer through the membrane layer (Q_m)

The total heat flux across the membrane is calculated by adding the heat transfer through the membrane material by conduction (Q_c) to the evaporative mass flux through the pores (Q_v) [78].

$$Q_m = Q_c + Q_v \quad (17)$$

$$Q_c = \left(\frac{k_m}{\delta}\right) (T_{mf} - T_{mp}) \quad (18)$$

$$Q_v = J \Delta H_v \quad (19)$$

where k_m is the effective thermal conductivity of the membrane, which is predicted on both gas and membrane polymer conductivities [55,79]. Depending on the isostress series model, the effective thermal conductivity of the membrane matrix consists of the thermal conductivity of gas (air and water vapor) k_g and membrane solid k_{mem} and can be determined as follows [55,80]:

$$k_m = \left(\left(\frac{\varepsilon}{k_g} \right) + \left(\frac{1-\varepsilon}{k_{mem}} \right) \right)^{-1} \quad (20)$$

ΔH_v is the Enthalpy of vaporization of water; and it is given as:

$$\Delta H_v = 1.7535 T_{mf} + 2024.3 \quad (21)$$

(c) Convective heat transfer from membrane surface to permeate stream Q_p

$$Q_p = h_p (T_{mp} - T_{bp}) \quad (22)$$

Where T_{bp} is the average permeate bulk temperature of the inlet and outlet of the cold permeate stream and h_p is the convective heat transfer coefficient in the permeate side,

which depends on the type of the flow (laminar or turbulent) and can be calculated by using different correlations. The following energy conservation is valid under steady-state conditions:

$$Q_f = Q_m = Q_p \quad (23)$$

The temperatures of the feed and permeate sides of the membrane are given as [69,74]:

$$T_{mf} = \frac{k_m \left(T_{bp} + \frac{h_f}{h_p} T_{bf} \right) + \delta (h_f T_{bf} - J \Delta H_v)}{k_m + h_f \left(\delta + \frac{k_m}{h_p} \right)} \quad (24)$$

$$T_{mp} = \frac{k_m \left(T_{bf} + \frac{h_p}{h_f} T_{bp} \right) + \delta (h_p T_{bp} + J \Delta H_v)}{k_m + h_p \left(\delta + \frac{k_m}{h_f} \right)} \quad (25)$$

These temperatures can be used back to calculate the vapor pressures using Eqs. (4) and (5), and then the permeate flux can be determined using Eq. (1).

The convective heat transfer coefficient is defined as follows:

$$h = \frac{Nu \, k}{D_h} \quad (26)$$

where k , D_h , and Nu represent the average thermal conductivity of the fluid on the feed or permeate sides, the hydraulic diameter of the flow channels, and the dimensionless Nusselt number, respectively. In the situation of laminar channel flow, Graetz-L  v  que provided the following correlation for Nusselt Number [81,82], which is valid for both feed and permeate sides:

$$Nu = 1.86 \left(\frac{Re \times Pr \times D_h}{L} \right)^{0.33} \quad (27)$$

Pr is the Prandtl number which is defined as the ratio of viscous diffusion rate to thermal diffusion rate:

$$Pr = \frac{\nu}{\alpha} = \frac{\mu C_p}{k} \quad (28)$$

where μ and C_p are the dynamic viscosity and the specific heat of the fluid, respectively.

For turbulent channel flow, Nusselt number of the feed (hot side) and permeate (cold side) streams are calculated as follows [69,71]:

$$Nu_f = 0.027 (Re_f)^{0.8} (Pr_f)^{0.4} \left(\frac{\mu_{bf}}{\mu_{mf}} \right)^{0.14} \quad (29)$$

$$Nu_p = 0.027 (Re_p)^{0.8} (Pr_p)^{0.3} \left(\frac{\mu_{mp}}{\mu_{bp}} \right)^{0.14} \quad (30)$$

The total heat transfer in the module membrane can be expressed in terms of the overall heat transfer coefficient, H , as follows:

$$Q_m = H (T_{bf} - T_{bp}) \quad (31)$$

The overall heat transfer coefficient can be calculated based on the temperature difference between the feed and permeate streams through the following expression [83]:

$$H = \left[\frac{1}{h_f} + \frac{1}{\left(\frac{k_m}{\delta} \right) + \frac{J \Delta H_v}{(T_{mf} - T_{mp})}} + \frac{1}{h_p} \right]^{-1} \quad (32)$$

3.2.2 Adsorption Desalination (AD)

Following assumptions are considered when modeling the AD system,

- adsorbent material is homogenous.
- The system components are well-insulated.
- Thermal resistance between adsorbent tubes and the adjacent adsorbent is neglected.
- Pumping energy of the heating and cooling water is neglected compared to the amount of energy provided for heating.

- Equilibrium adsorption capacity and adsorption kinetics

The adsorption uptake and kinetics of water vapor are employed to evaluate the adsorption uptake. According to Alsaman. Et.al, the Dubnin-Astackov (D-A) model is more appropriate for water vapor sorption [84].

$$q_{eq} = q_{max} \exp \left(- \left(\frac{RT}{E} \ln \left(\frac{P_{bed,sat}}{P_{sorp}} \right) \right)^n \right) \quad (33)$$

where, q_{eq} and q_{max} denote the equilibrium adsorption capacity as determined by isotherms of adsorption and maximum adsorption amount, respectively.

The linear driving force (LDF) model is used to estimate the actual adsorption uptake (q) as follows [85–88]:

$$\frac{dq}{dt} = k_L (q_{eq} - q) \quad (34)$$

Here, k_L (1/s) is the LDF constant which can be determined as [84]:

$$k_L = \frac{A_0 D_c}{r_p^2} \exp \left(\frac{-E_a}{RT} \right) \quad (35)$$

- **Sorption beds:**

The energy conservation of sorption (adsorption or desorption) bed takes into account the heat stored in the bed (adsorbent, metal (tubes and fins), and adsorbed water vapor), as well as heat transfer between the bed and heating or cooling water, and adsorption or desorption heat. It is expressed as:

$$\left[(mC_P)_{tubes} + (mC_P)_{fins} + (mC_P)_s + m_s C_v q \right]_{bed} \frac{dT_{bed}}{dt} = m_s \Delta H \frac{dq}{dt} + \dot{m}_w C_{P,w} (T_{w,i} - T_{w,e}) \quad (36)$$

The subscript 'w' can refer to either heating water (hw) used in desorption and pre-heating processes or cooling water (cw) used in adsorption and pre-cooling operations. The first term on the left-hand side (Eq. 36) is zero for preheating and precooling switching processes. In Eq. 36, ΔH denotes the adsorption heat, which is determined using adsorption isotherm curves as [89]:

$$\Delta H = h_{fg} + E \ln \left(\frac{q_{max}}{q_{eq}} \right)^{\frac{1}{n}} + \frac{E \times T \times CR}{n} \ln \left(\frac{q_{max}}{q_{eq}} \right)^{\frac{(1-n)}{n}} \quad (37)$$

where h_{fg} (in kJ/kg) stands for vaporization/condensation latent heat and CR Capital recovery ratio or amortization charge.

- **Condenser:**

The AD condenser is used to condense the desorbed water vapor produced during the desorption process; the heat it rejects can be used to heat the DCMD feed stream (seawater) like in (Figure1 a). The condenser inputs and outputs mass balance is expressed as follows:

$$\frac{dm_{cond}}{dt} = m_s \left(\frac{dq}{dt} \right)_{des} \quad (38)$$

where m_{cond} is considered the freshwater product of the AD system.

$$\dot{m}_{pw,AD} = \frac{dm_{cond}}{dt} \quad (39)$$

The condenser's energy conservation includes the heat stored in the condenser (metals (tubes and fins (if any), and the accumulated water vapor), the cooling energy of the desorbed vapor, the latent heat of condensation, and the heat exchange between the condensed water and another stream (e.g. DCMD feed as in configuration 1). It is expressed as follows:

$$\begin{aligned} & \left[(mC_P)_{tubes} + (mC_P)_{fins} + (mC_P)_w \right]_{cond} \frac{dT_{cond}}{dt} = m_s C_{Pwv} (T_{des} - \\ & T_{cond}) \left(\frac{dq}{dt} \right)_{des} + m_s h_{fg}(T_{cond}) \left(\frac{dq}{dt} \right)_{des} + \dot{m}_{cond,in} C_{P,w} (T_{cond,in} - T_{cond,out}) \quad (40) \end{aligned}$$

- Evaporator

The AD evaporator is a major element of the hybrid AD-DCMD system due to its role in desalination and cooling. Thus, the mass balance equation of the evaporator is critical because some water evaporates from it to be adsorbed by an adsorption bed, and some brine exits at high salinity. The evaporator's inlet saline water can be supplied intermittently in order to maintain a steady amount of liquid (saline water) inside the evaporator. While the exit, extremely saline brine is periodically removed from the evaporator when it reaches a concentration between 180-220 ppt. It is recommended that it be further treated to produce salt as a second byproduct in addition to the cooling effect. The mass and salt concentration conservation equations for the evaporator are as follows:

$$\frac{dm_{sw,evap}}{dt} = \gamma \dot{m}_{sw,AD} - \theta \dot{m}_{b,AD} - m_s \left(\frac{dq}{dt} \right)_{ads} \quad (41)$$

$$\dot{m}_{sw,evap} \frac{dx_{sw,evap}}{dt} = \gamma x_{sw,AD} \dot{m}_{sw,AD} - \theta x_{b,AD} \dot{m}_{b,AD} - x_{ads} \dot{m}_s \left(\frac{dq}{dt} \right)_{ads} \quad (42)$$

The evaporator energy conservation equation is a balance of heat stored in the evaporator (metals (tubes and fins (if any) and water), the energy content of entering saline water (AD feed), energy removed by AD brine, latent heat of evaporation, and heat exchange between evaporated water and chilled stream. It is stated as,

$$\begin{aligned} &[(mC_P)_{tubes} + (mC_P)_{fins} + (mC_P)_{sw}]_{evap} \frac{dT_{evap}}{dt} = \\ &\gamma \dot{m}_{sw,AD} h_{sw,AD}(T_{sw,AD}, x_{sw,AD}) - \theta \dot{m}_{b,AD} h_{b,AD}(T_{evap}, x_{evap}) - \\ &\dot{m}_s h_{fg}(T_{evap}, x_{evap}) \left(\frac{dq}{dt} \right)_{ads} + \dot{m}_{chw} C_{P,chw}(T_{chw,in} - T_{chw,out}) \end{aligned} \quad (43)$$

Symbols γ and θ equal 1 when the feeding saline water and AD brine enter and leave the evaporator, respectively. Otherwise, they are zero.

- Heat exchangers

The LMTD method is used to define the heating and cooling processes for desorption and adsorption beds, as well as the heat exchange with the secondary fluid in the condenser and the evaporator as following:

$$T_{y,out} = T_{hx} + (T_{y,in} - T_{hx}) \exp\left(\frac{-U A_{hx}}{(m C_P)_y}\right) \quad (44)$$

where "y" represents heating water, cooling water, or chilled water, and "hx" denotes the heat exchanger, which includes Bed#1, Bed#2, condenser, and evaporator.

3.2.3 Integrated System Performance

The water produced by the hybrid AD-DCMD system can be calculated as follows:

$$\dot{m}_{pw} = \dot{m}_{pw,AD} + \dot{m}_{pw,DCMD} \quad (45)$$

The parameter "gained output ratio" (GOR) is widely used to assess the system's efficiency. It is critical to note that the GOR values for both AD and DCMD systems are quite low; therefore, being both in one hybrid system significantly increases the GOR value (to almost double). GOR is calculated for individual systems (AD [66] and DCMD [76]) and for the hybrid system as a whole in order to demonstrate the contribution of each system. It is determined as follows:

$$GOR_{AD} = \int_0^{t_{cycle}} \frac{\dot{m}_{pw,AD} h_{fg}}{\dot{m}_{hw}(h_{hw,in} - h_{hw,out})} dt \quad (46)$$

$$GOR_{DCMD} = \frac{\dot{m}_{pw,DCMD} \Delta H_v}{Q_t} \quad (47)$$

Where Q_t is the total heat energy removed from the hot water in the DCMD hot channel.

$$GOR_{Total} = \frac{\dot{m}_{pw,AD} h_{fg} + \dot{m}_{pw,DCMD} \Delta H_v}{\dot{m}_{hw}(h_{hw,in} - h_{hw,out})} \quad (48)$$

Calculating the cooling effect of the AD system in the two configurations is important. This can be expressed mathematically as the coefficient of performance (COP), which is the ratio of cooling effect to heat input as [66],

$$COP = \int_0^{t_{cycle}} \frac{\dot{m}_{chw}(h_{chw,i} - h_{chw,o})}{\dot{m}_{hw}(h_{hw,in} - h_{hw,out})} dt \quad (49)$$

Table 4 Design and operating parameters of the AD-DCMD hybrid system.

Property	Value
DCMD system:	
Membrane effective area (A_{mem})	0.175 m ²
Membrane feed rate ($\dot{m}_{f,DCMD}$)	300 kg/h
Channel width (D)	24 mm
Channel height (Ht)	5 mm
Membrane porosity (ϵ)	0.85
Membrane pore size (d_{pore})	1.2 μ m
Membrane thickness (δ)	151.8 μ m
Cold stream inlet temperature (T_{bp})	20 °C
Hot stream inlet temperature (T_{bf})	60 °C - 90 °C
AD system [66,84]:	
Overall heat transfer coefficient of the condenser (UA_{cond})	0.5 kW/K
Overall heat transfer coefficient of the evaporator (UA_{evap})	0.35 kW/K
Overall heat transfer coefficient of the bed (UA_{bed})	0.6 kW/K
Mass of bed tubes (m_{tubes})	2.97 kg
Mass of bed fins (m_{fin})	0.72 kg
Mass of bed heat exchanger cover (m_{iron})	15 kg
Mass of condenser metal (m_{cond})	1.533 kg
Mass of evaporator metal (m_{evap})	1.3 kg

Mass of silica-gel in each bed (m_s)	6.75 kg
Bed tubes' specific heat capacity ($C_{p,tube}$)	0.386 kJ/kg·K
Bed fins' specific heat capacity ($C_{p,fin}$)	0.905 kJ/kg·K
Silica-gel's specific heat capacity ($C_{p,s}$)	0.924 kJ/kg·K
Mass of saline water in the evaporator ($m_{sw,evap}$)	3 kg
Universal gas constant (R)	8.314 kJ/kmol·K
Heating water inlet temperature ($T_{HW,in}$)	80 °C
Cooling water inlet temperature ($T_{cw,in}$)	30 °C
Chilled water inlet temperature ($T_{chw,in}$)	$T_{cw,in}$

3.3 Constructive and Operating Parameters

In this work, the beds are regarded as similar to those reported in the experimental study [84] due to an optimized bed filled with silica-gel, which is readily available in the market as a low-cost, and stable material. Table 5 presents the operating parameters of the DCMD system, and it also lists the parameters of the two identical beds, the condenser and the evaporator. The equilibrium and kinetic adsorption parameters of the silica-gel and H₂O pair which are necessary for calculating the equilibrium and actual uptakes (q_{eq} and q) were listed in Table 5. Ambient pressure (101.3 kPa) is assumed as the operating pressure for DCMD system. It is crucial to notice that thermal property fluctuation is taken into account. The thermodynamic properties of water are calculated using the REFPROP software [90]. Initial values for the operating conditions are listed in Table 6. The simulation is carried out using the MATLAB software with computational tolerance error of $< 10^{-6}$.

Table 5 Adsorption properties of Silica-gel/water pair [84].

Property	Value
q_{max}	0.36 kg/kg
E	167.7 kJ/kg
n	1.68
A_0	32
E_a	28.67 kJ/kg
D_c	$22.5 \times 10^{-9} \text{ m}^2/\text{s}$
r_p	0.00175 m

Table 6 Initial values of operating conditions used in calculation

Operating condition	Initial value
Effective diffusion coefficient (D_e)	$1.5 \times 10^{-6} \text{ m}^2/\text{s}$
Temperature of the adsorption bed (T_{ads})	79 °C
Temperature of the desorption bed (T_{des})	34 °C
Uptake of the adsorption bed (q_{ads})	0.15 kg/kg
Uptake of the desorption bed (q_{des})	0.26 kg/kg
Temperature of the condenser (T_{cond})	45 °C
Temperature of the evaporator (T_{evap})	30 °C
Evaporator salinity (x_{evap})	35 ppt
Brine salinity (x_b)	35 ppt

3.4 Economic Model

The cost of freshwater produced by the two AD-DCMD configurations is calculated in accordance with El-Dessouky and Ettouney [91], and Qasem and Zubair [66].

Table 7 summarizes the capital costs of the analyzed components of DCMD and AD systems based on the real purchasing prices and some assumptions. The assumptions made are as following:

- Electrical heaters are used to heat the water, and the unit cost of electricity (COE) is 0.07\$/kWh [92] .
- Specific cost of labor (l) = \$0.1/m³ [91].
- Annual management cost is 20% of the labor cost [93].
- Annual maintenance cost account for 1.5% of the capital investment cost [93].
- Interest rate (i) is 5% [94].
- Plant life expectancy (Y) is 30 years [91] .
- Availability of the plant (f) is 0.9 [91].
- The land cost is neglected due to the outdoor location of the plant and operating in a rural area [95].

The designs of the three configurations (Figure 10) are used to reasonably assume the capital cost of the AD-DCMD hybrid system as stated in

Table 7. The following equations can be used to estimate the total costs, including the operating costs:

Capital recovery ratio (CR), or amortization charge, [96], is calculated as:

$$CR = \frac{i(i+1)^Y}{(i+1)^Y - 1} \quad (50)$$

The annual capital cost (in \$/yr):

$$C_A = C_c \times CR \quad (51)$$

The annual power cost (in \$/yr)

$$CP = COE \times \frac{e}{3600} \times f \times \dot{m}_{pw} \times 3600 \times 24 \times 365 \quad (52)$$

Where, “e” (in kJ/kg) denotes the energy per unit mass of produced freshwater. The annual labor cost (in \$/yr) is:

$$C_L = l \times f \times \frac{\dot{m}_{pw}}{1000} \times 3600 \times 24 \times 365 \quad (53)$$

The annual maintenance cost (in \$/yr) can be calculated as:

$$C_M = 0.015 \times C_A \quad (54)$$

The annual management cost (in \$/yr) is:

$$C_{mg} = 0.2 \times C_L \quad (55)$$

The total annual cost of the freshwater (in \$/yr) is then determined as:

$$C_T = C_A + CP + C_L + C_M + C_{mg} \quad (56)$$

And finally, the cost of the produced freshwater (in \$/L) is calculated as follows:

$$C_{pw} = \frac{C_T}{f \times \dot{m}_{pw} \times 3600 \times 24 \times 365} \quad (57)$$

Table 7 The capital investment cost for AD-DCMD hybrid system.

Item	Price (US \$)	Ref.
Membrane	60 (\$/m ²)	[93]
Pumps	300	
Fittings/pipes	70	[97]
Accessories	66	[97]
Control devices	80	[97]
Condenser	120	[66]
Evaporator	150	[66]
Water tanks	260	[66]
Silica gel (high quality)	5 × 2 × 6.75	[66]
Adsorbent bed with fins	2 × 120	[66]

CHAPTER 4

RESULTS AND DISCUSSION

Hybrid AD-DCMD systems in configurations 1, and 2 are used to demonstrate the general thermal behavior of these desalination processes. Since the temperatures of the AD cycle components fluctuate during the adsorption/desorption process, temperature history of the AD system was investigated for 14 cycles. Figure 11 depicts the temperature-time history of the AD cycle components including adsorption bed, desorption bed, evaporator, and condenser at steady-state conditions. Heating and cooling water temperatures are maintained at 80 °C and 30°C while the half-cycle duration is kept 700 s with a switching time of 20 s.

During the adsorption and desorption processes, the temperature of the bed varies between 36 to 76 °C. The condenser temperature is also fluctuating between 37 and 47 °C which is relatively low. This can be attributed to the cold seawater (DCMD feed in configuration 1) passing through the condenser to get use of the waste heat of the AD system in operating the DCMD system. Temperature of the evaporator is reduced to around 22 °C due to the exchange of heat between the chilled water and the fed seawater which then gets sucked by means of pressure difference between the adsorption bed and the evaporator.

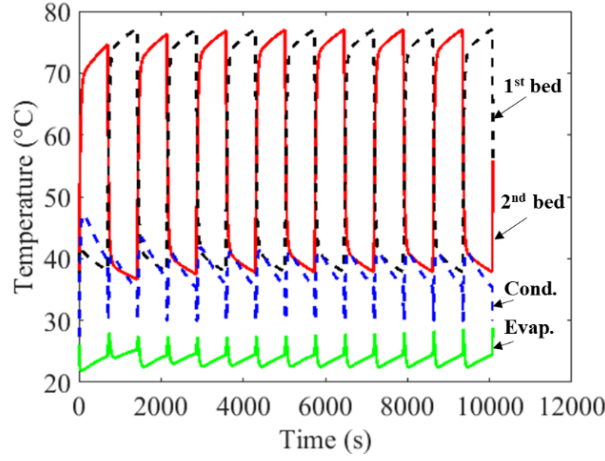


Figure 11 Temperature profiles' history for the AD system components under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $\dot{m}_{cw,in} = 0.3\text{ kg/s}$, $\dot{m}_{hw,in} = 0.4\text{ kg/s}$, $\dot{m}_{chw,in} = 0.5\text{ kg/s}$, $\dot{m}_{cond,in} = \dot{m}_{DCMD,f} = 0.08\text{ kg/s}$, and $T_{bp} = 20\text{ }^{\circ}\text{C}$ for configuration 1.

In order to demonstrate the relevance of cycling time in enhancing the hybrid system's performance, GOR values behavior of AD, DCMD and the hybrid system are illustrated in Figure 12. Total GOR values slightly increase with increasing cycle time due to a decrease in the required heating energy while the bed is hot. However, it is clear that total GOR is almost the same for half cycle time $\geq 700\text{ s}$. Increasing the heating time reduces the temperature difference between the hot bed and the heater. According to the presented results, the half cycle time for water production and GOR can be set to 700 s and, hence used to test the other operating parameters for the two configurations.

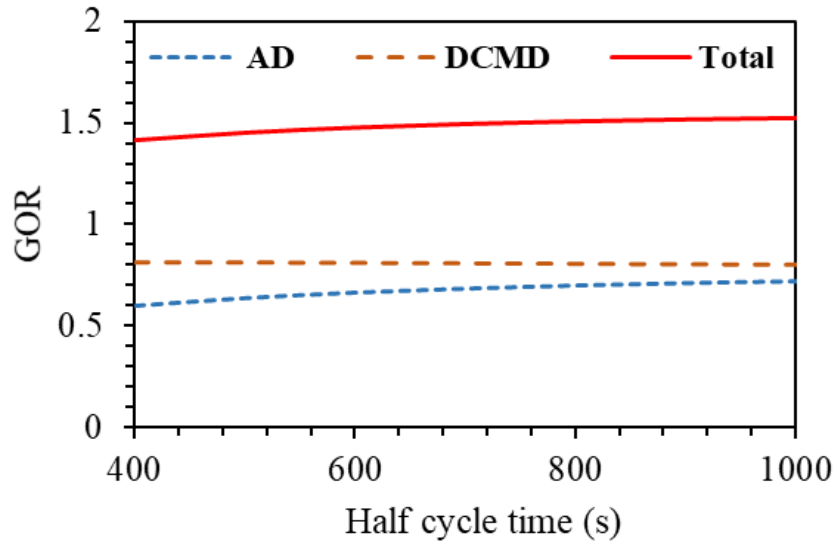


Figure 12 Effect of half cycle time on GOR values of AD-DCMD hybrid system under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{hw,in} = 80\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $\dot{m}_{cw,in} = 0.3\text{ kg/s}$, $\dot{m}_{hw,in} = 0.4\text{ kg/s}$, $\dot{m}_{chw,in} = 0.5\text{ kg/s}$, $\dot{m}_{cond,in} = \dot{m}_{DCMD,f} = 0.08\text{ kg/s}$, and $T_{bp} = 20\text{ }^{\circ}\text{C}$ for configuration 1.

As shown in Figure 13, the hybrid system's water production is a function of both flowrates of seawater entering the condenser to feed DCMD system ($\dot{m}_{sw-DCMD}$) and heating water of AD system ($\dot{m}_{hw}$). A larger amount of seawater inlet to the condenser generally enhances the water production of AD system, which in turn enhances the hybrid system's freshwater production. The adding of seawater could improve the condensation of water content (in the condenser), thereby maximizing AD freshwater output. On the other hand, the freshwater produced from the DCMD system decreases with increasing seawater inlet to the condenser (Figure 13, c). This can be attributed to the fact that temperature of the condenser outlet stream decreases with increasing the flow rate of the inlet stream (cold seawater), hence slightly decreasing the DCMD permeate flux. This little decrease in the DCMD flux has a negligible effect on the hybrid system's overall freshwater production

as illustrated by Figure 13, a. According to the results, optimum flow rate of seawater entering the condenser to feed DCMD system is 0.08 kg/s. Hybrid AD-DCMD system performance is not affected significantly by the amount of heating water inlet. As long as the flow rate of heating water inlet is ≥ 0.4 kg/s, the amount of water vapor that can be desorbed is about equal to the amount that was adsorbed onto the adsorbent material. Therefore, $\dot{m}_{hw}$ of 0.4 kg/s is taken for further calculations. Figure 13, b illustrates that increasing the AD system's heating water amounts from 0.2 to 0.8 kg/s has a negligible effect on the total freshwater production, which increases only by < 0.2 kg/h.

As illustrated in Figure 14, GOR follows a similar pattern as freshwater production when changing the flow rates of seawater passing through the condenser and heating water entering the desorption bed. The hybrid system's maximum GOR reaches 2.02 at $\dot{m}_{sw-DCMD}$ of 0.08 kg/s. It is clear from Figure 14(a) that GOR maximum value is the same at all heating water flow rates. This means changing the heating water flow rate doesn't significantly affect the total GOR of configuration 1. Results are presented in independent sub-Figures to avoid data interference and to shed additional light on the effect of stream flow rates on freshwater production and GOR of the hybrid system.

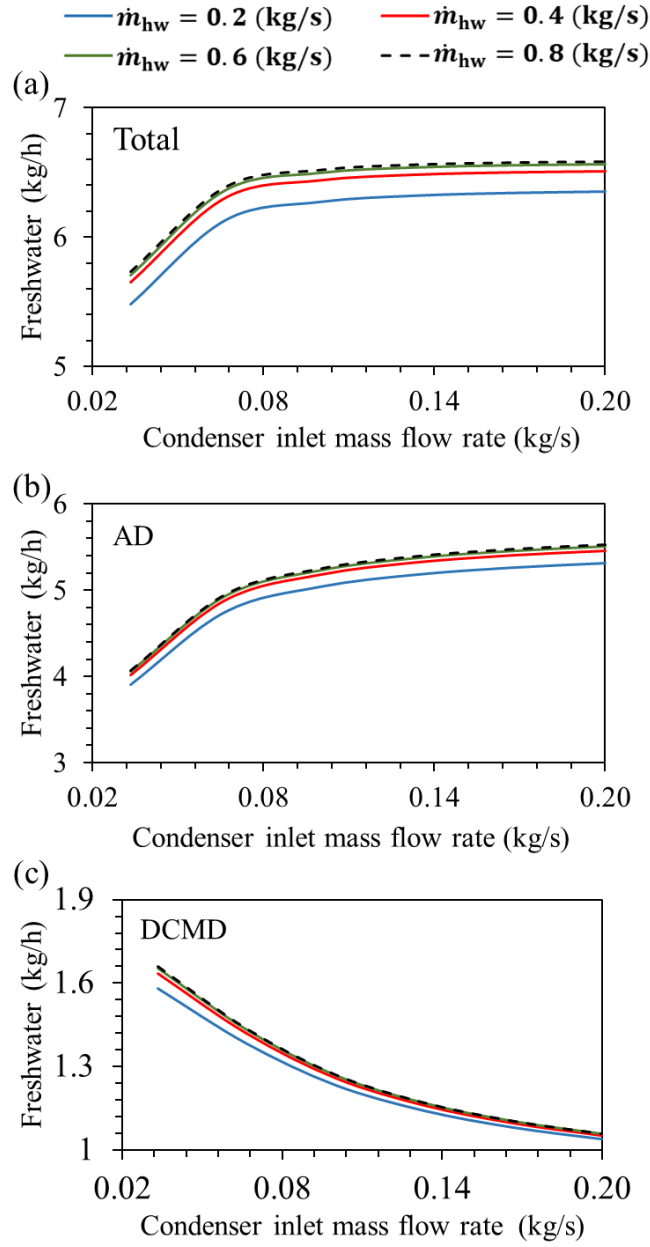


Figure 13 Effect of mass flowrate of the inlet seawater to the condenser ($\dot{m}_{cond,in} = \dot{m}_{DCMD,f}$) on the freshwater production at different heating water flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30^\circ\text{C}$, $T_{hw,in} = 80^\circ\text{C}$, $T_{cond,in} = 30^\circ\text{C}$, $T_{chw,in} = 30^\circ\text{C}$, $\dot{m}_{cw,in} = 0.3 \text{ kg/s}$, $\dot{m}_{chw,in} = 0.5 \text{ kg/s}$, $T_{bp} = 20^\circ\text{C}$, and half-cycle time= 700 s for configuration 1.

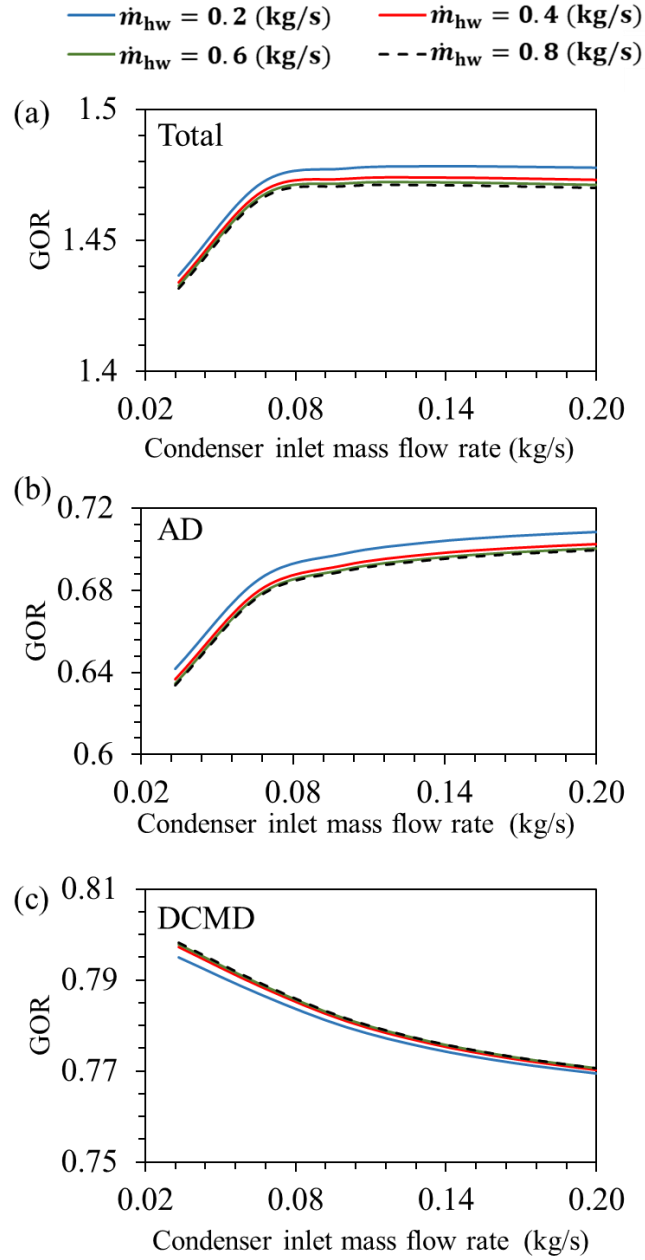


Figure 14 Effect of mass flowrate of the inlet seawater to the condenser ($\dot{m}_{cond,in} = \dot{m}_{DCMD,f}$) on the system GOR at different heating water flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30^\circ\text{C}$, $T_{hw,in} = 80^\circ\text{C}$, $T_{cond,in} = 30^\circ\text{C}$, $T_{chw,in} = 30^\circ\text{C}$, $\dot{m}_{chw,in} = 0.5 \text{ kg/s}$, $T_{bp} = 20^\circ\text{C}$, and half-cycle time = 700 s for configuration 1.

According to Figure 15, a, temperature of seawater inlet to the condenser ($T_{cond,in}$) has a significant effect on water production of AD-DCMD hybrid system, as can be seen when it is increased from 20 °C to 40 °C at selected heating water inlet temperature ($T_{hw,in}$). As $T_{cond,in}$ increases, the temperature difference between the hot and cold sides of the condenser get minimized. This results in a decrease in the condensed amount of freshwater inside the condenser, leading to a decrease in the freshwater production of the AD system. On the other hand, at a fixed heating water inlet temperature, an increase in freshwater production of the DCMD system is achieved by increasing the temperature of seawater inlet to the condenser which is in turn increasing the feed temperature of DCMD system. The increased DCMD freshwater production is a result of high permeate flux which is due in part to the high temperature difference across the membrane in addition to the other physical properties of the membrane which were used in the model such as high porosity, high contact angle, and large pore size. This is explained by Antoine Eqs.4 and 5 in which the effect of temperature on vapor pressure is minimal at low feed temperatures, but becomes much more significant at higher feed temperatures. The results also show that as the temperature of the AD heating water raises, freshwater outputs of DCMD, AD, and hybrid system increase. This can be attributed to the influence of higher heating water temperature in maximizing the desorption amounts. Then, the amount of desorbed water vapor (from the desorption bed) and the cold seawater entering the condenser (to feed DCMD) exchange heat in the condenser leading to a significant increase in the condensed amount (AD freshwater) and a raise in the DCMD inlet feed temperature, hence increasing DCMD distillate (freshwater).

GOR values of the hybrid and individual systems in configuration 1 are also influenced by condenser seawater inlet temperature and heating water inlet temperature (Figure 16). Two scenarios are noticed for the AD GOR values (Figure 16, b). The first one is when $T_{cond,in} > 30\text{ }^{\circ}\text{C}$, AD GOR values increase with increasing $T_{hw,in}$. This might be attributed to the enhancement in the desorbed water vapor amounts at higher heating temperatures, which leads to higher GOR values according to Equation 46. While, in the second scenario, at $T_{cond,in} < 30\text{ }^{\circ}\text{C}$, low heating water temperatures result in better AD GOR values. For instance, at $T_{cond,in} = 20\text{ }^{\circ}\text{C}$ (fixed), AD GOR values are 0.7 and 0.77 at $T_{hw,in}$ of 80 and $60\text{ }^{\circ}\text{C}$, respectively. Higher DCMD GOR values are noticed with increasing either $T_{hw,in}$ or $T_{cond,in}$ (Figure 16, c). This might be due to the higher heat amount which is exchanged between the desorbed water vapor (hot) and the seawater (DCMD feed) inside the condenser resulting in increased DCMD feed temperatures and hence higher GOR values.

DCMD-AD hybrid system has made a significant jump forward, with a total GOR > 2 at $T_{cond,in} \geq 30\text{ }^{\circ}\text{C}$ and $T_{hw,in} \geq 80\text{ }^{\circ}\text{C}$ (Figure 16, a). However, relatively lower GOR values are obtained at $T_{hw,in} = 60\text{ }^{\circ}\text{C}$ as a direct result of the low AD GOR at the same conditions. Insignificant increase in total GOR values occurs when increasing $T_{hw,in}$ from 80 to 90; therefore, setting $T_{hw,in} = 80\text{ }^{\circ}\text{C}$ is efficient. Based on the previously stated analysis for the effect of $T_{cond,in}$ and $T_{hw,in}$ on freshwater production and GOR values, the optimum $T_{cond,in}$ and $T_{hw,in}$ are 30 and $80\text{ }^{\circ}\text{C}$, respectively.

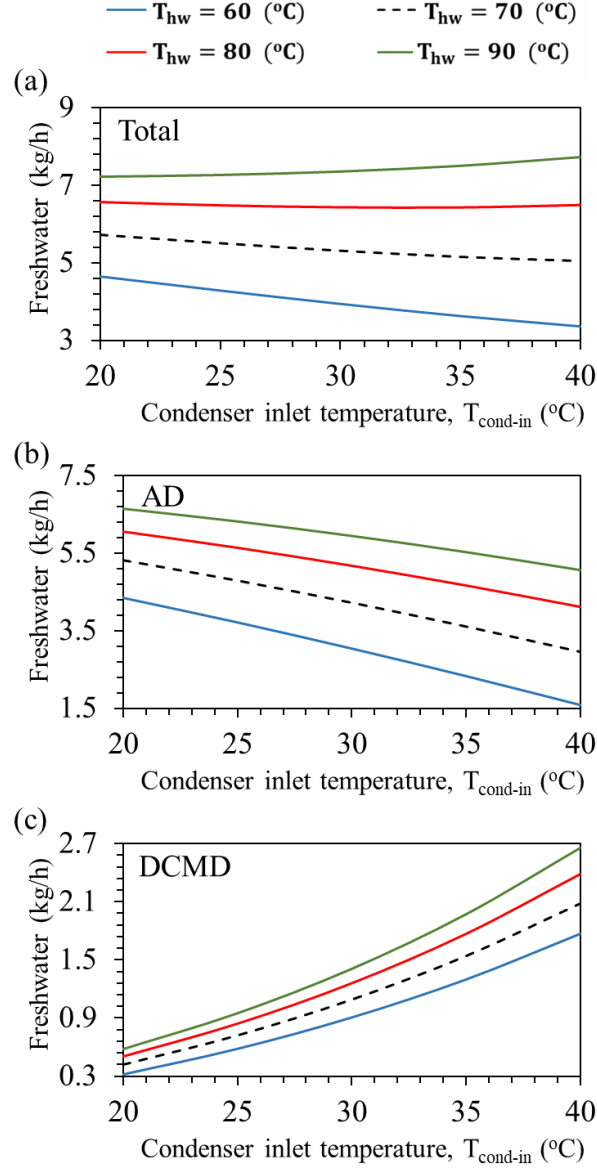


Figure 15 Effect of temperature of the inlet seawater to the condenser on the freshwater production at different heating water temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{\text{cw,in}} = 30$ °C, $T_{\text{cond,in}} = 30$ °C, $T_{\text{chw,in}} = 30$ °C, $\dot{m}_{\text{cw,in}} = 0.3$ kg/s, $\dot{m}_{\text{hw,in}} = 0.4$ kg/s, $\dot{m}_{\text{chw,in}} = 0.5$ kg/s, $\dot{m}_{\text{bp}} = 0.06$ kg/s, $T_{\text{bp}} = 20$ °C, and half-cycle time= 700 s for configuration 1.

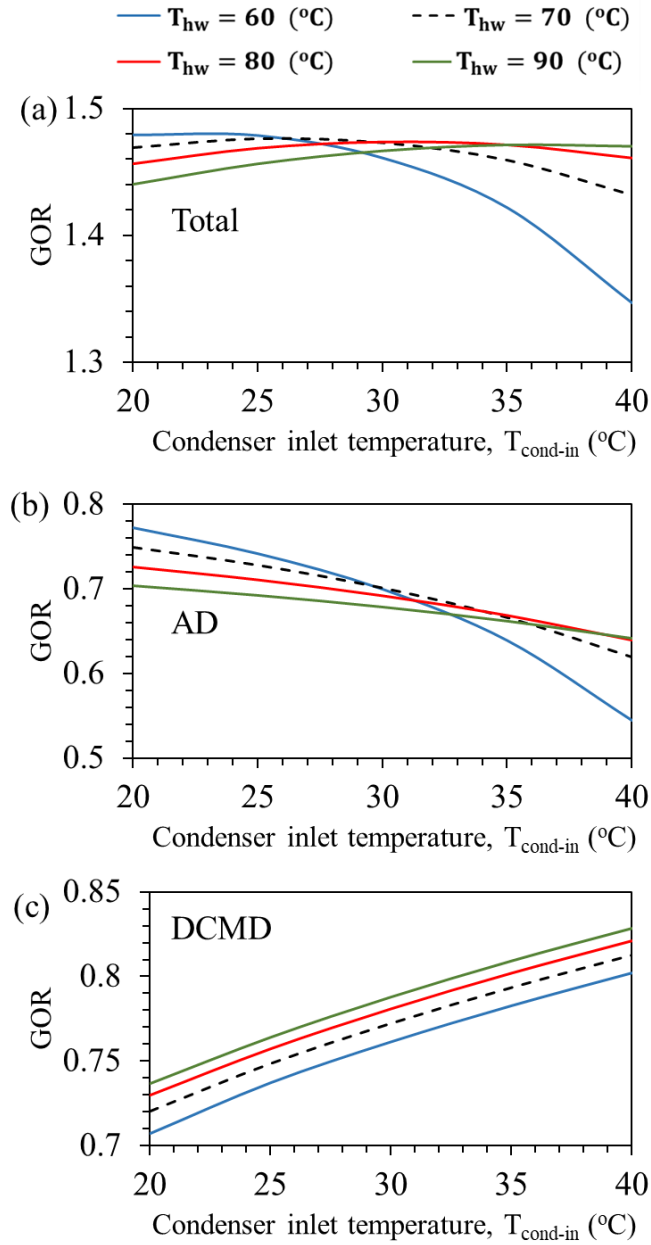


Figure 16 Effect of temperature of the inlet seawater to the condenser on the system GOR at different heating water temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{\text{cw,in}} = 30$ °C, $T_{\text{hw,in}} = 80$ °C, $T_{\text{chw,in}} = 30$ °C, $\dot{m}_{\text{cw,in}} = 0.3$ kg/s, $\dot{m}_{\text{hw,in}} = 0.4$ kg/s, $\dot{m}_{\text{chw,in}} = 0.5$ kg/s, $\dot{m}_{\text{bp}} = 0.06$ kg/s, $T_{\text{bp}} = 20$ °C, and half-cycle time= 700 s for configuration 1.

In configuration 2 the seawater (DCMD feed) is heated inside the adsorption bed (Figure 10, b) instead getting heated inside the AD condenser as in configuration 1 (Figure 10, a).

Seawater (DCMD feed) is used as a pre-cooler for the adsorption bed before entering the DCMD system. As shown in Figure 17, the freshwater production of the AD-DCMD hybrid system is shown for various cooling water mass flow rates ($\dot{m}_{cw}$) at selected heating water mass flow rate ($\dot{m}_{hw}$). Increasing the cooling water mass flow rate of the AD system while maintaining a constant heating water mass flow rate significantly boosts freshwater production of DCMD system until reaching its maximum value at $\dot{m}_{cw}$ of 0.08 kg/s before it decreases (Figure 17, c). For example, by increasing $\dot{m}_{cw}$ from 0.03 to 0.08 kg/s, produced freshwater of DCMD increases by almost 1 kg/h which contributes in enhancing the overall system productivity. Freshwater production of the AD system shows almost a constant trend while changing $\dot{m}_{cw}$ (Figure 17, b). Compared with configuration 1, configuration 2 produce around double freshwater amount (Figure 17, a). In addition increasing $\dot{m}_{hw}$, to some extent, increases the total freshwater production. However, almost the same freshwater amounts are produced when $\dot{m}_{hw} \geq 0.4$ kg/s.

DCMD GOR increases with increasing $\dot{m}_{cw}$ before reaching its maximum value at $\dot{m}_{cw}$ of 0.08 kg/s; whereas AD GOR decreases until reaching its minimum value at the same $\dot{m}_{cw}$ (Figure 18, b and c). GOR values of the hybrid system in Figure 18, a are slightly more than those in Figure 14, a by around 0.35 indicating the better performance of configuration 2. Figure 17 and Figure 18 indicate that $\dot{m}_{cw}$ of 0.08 kg/s and $\dot{m}_{hw}$ of 0.4 kg/s are the reasonable operational mass flowrates for an optimal performance.

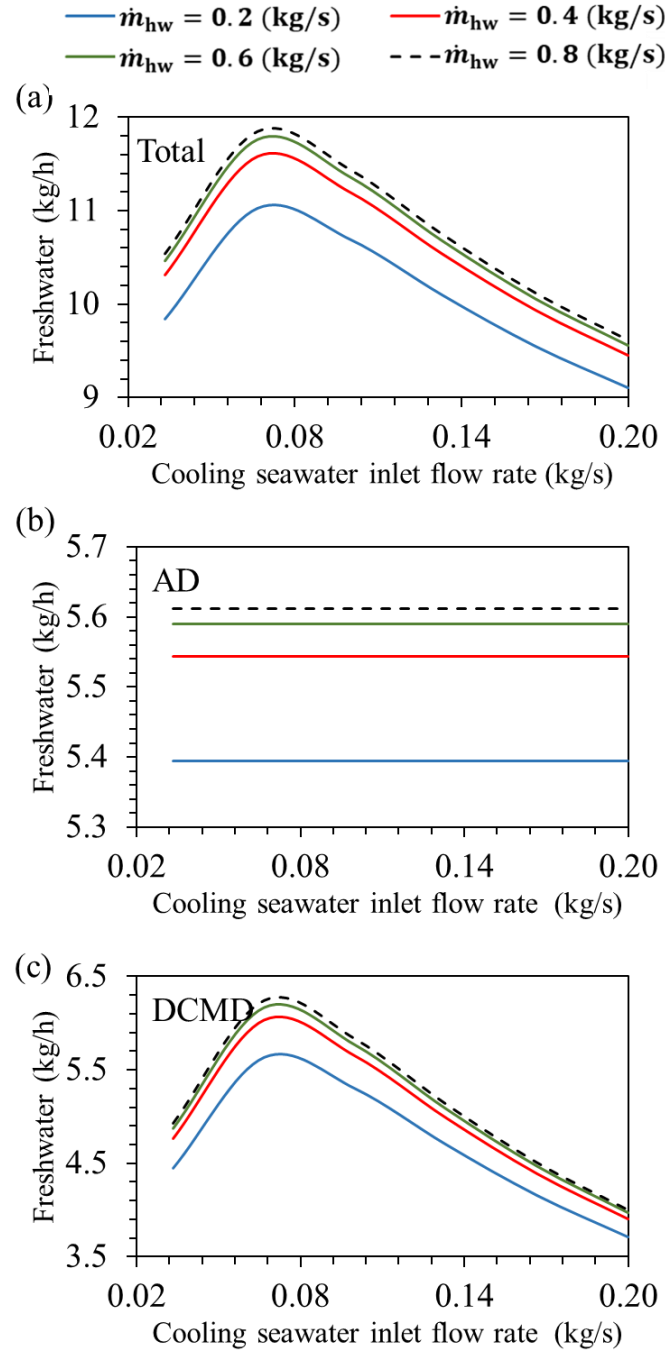
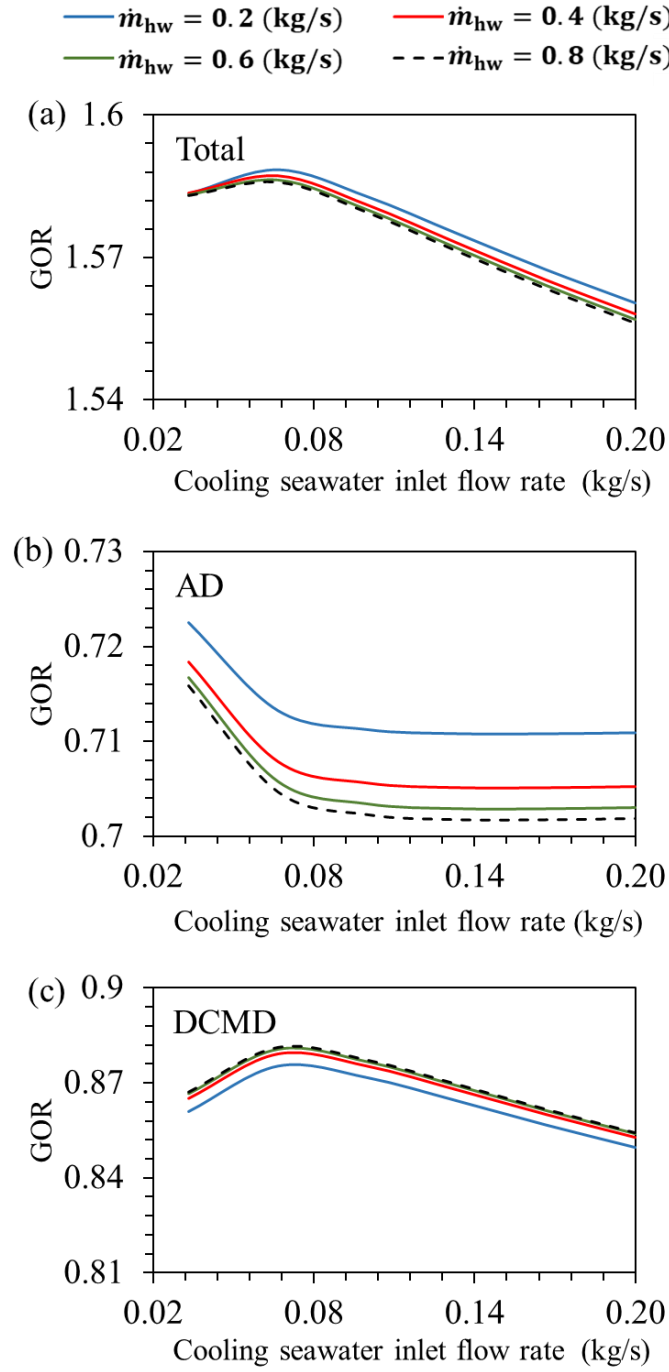


Figure 17 Effect of mass flowrate of the inlet seawater as a cooling water and a feed for DCMD ($\dot{m}_{cw,in} = \dot{m}_{DCMD,f}$) on the freshwater production at different heating water flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30^\circ\text{C}$, $T_{hw,in} = 80^\circ\text{C}$, $T_{cond,in} = 30^\circ\text{C}$, $T_{chw,in} = 30^\circ\text{C}$, $\dot{m}_{hw,in} = 0.4 \text{ kg/s}$, $\dot{m}_{chw,in} = 0.5 \text{ kg/s}$, $\dot{m}_{cond,in} = 0.3 \text{ kg/s}$, $T_{bp} = 20^\circ\text{C}$, and half-cycle time= 700 s for configuration 2.



[Figure 18 Effect of mass flowrate of the inlet seawater as a cooling water and a feed for DCMD ($\dot{m}_{cw,in} = \dot{m}_{DCMD,f}$) on the system GOR at different heating water flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30^\circ\text{C}$, $T_{hw,in} = 80^\circ\text{C}$, $T_{cond,in} = 30^\circ\text{C}$, $T_{chw,in} = 30^\circ\text{C}$, $\dot{m}_{chw,in} = 0.5 \text{ kg/s}$, $\dot{m}_{hw,in} = 0.4 \text{ kg/s}$, $\dot{m}_{cond,in} = 0.3 \text{ kg/s}$, $T_{bp} = 20^\circ\text{C}$, and half-cycle time= 700 s for configuration 2.]

The temperatures of AD cooling water ($T_{cw,in}$) (which will be fed to DCMD) and heating water ($T_{hw,in}$) also influence the freshwater production and GOR values of configuration 2. At a fixed $T_{hw,in}$, the freshwater production of the AD system is almost constant when increasing $T_{cw,in}$ from 20 to 40 °C (Figure 19, b), while a different trend appears in configuration 1 where the freshwater production of the AD system decreases with increasing the temperature of seawater inlet to the condenser (to feed DCMD system). Increasing the temperature of seawater input to the adsorption bed (cooling water), which thus contributes in raising the feed temperature of the DCMD system, increases freshwater production of the DCMD system at a constant heating water inlet temperature. Similar to the behavior of configuration 1, freshwater production of AD and DCMD systems in configuration 2 is enhanced with increasing heating water inlet temperature. This is attributed to that the higher the heating water temperature, the higher the desorbed amount of water vapor in AD system and also the more heat transferred to the DCMD feed stream passing through the adsorption bed.

Figure 19, a shows that the total freshwater production of AD-DCMD hybrid system significantly increases with increasing the heating water temperature at a fixed cooling water inlet temperature. For instance, when heating water temperature is raised from 60 to 90 °C, produced freshwater from the AD-DCMD hybrid system increases about 9 kg/h, which is much higher than the improvement occurs in configuration 1 under almost the same conditions.

GOR values of configuration 2 are also investigated. Figure 20, a shows that AD GOR increases with raising the cooling water inlet temperature. For instance, at constant $T_{hw,in}$

of 80 °C, AD GOR increases from 0.69 to 0.72 when $T_{cw,in}$ is raised from 20 °C to 40 °C, respectively. Moreover, significantly high DCMD GOR values are obtained with increasing $T_{hw,in}$ and $T_{cw,in}$. Figure 20, c shows that DCMD GOR increases by almost 0.3 when $T_{hw,in}$ is increased from 60 °C to 90 °C at a fixed cooling water inlet temperature. Similarly, at $T_{hw,in}$ of 80 °C (fixed), DCMD GOR increases by almost 0.06 with increasing $T_{cw,in}$ from 20 °C to 40 °C. Because of the larger heat exchanged between the adsorbent (silica-gel) and seawater (DCMD feed) in the adsorption bed, higher DCMD feed temperatures and, therefore, higher DCMD GOR values are obtained. The total GOR values of configuration 2 are greater than those reported in Figure 16, a (for configuration 1) by around 0.5. Based on the analysis of freshwater production and GOR values for configuration 2 at different heating water temperatures and cooling water temperature, the optimum performance occurs at $T_{hw,in}$ and $T_{cw,in}$ of 90 °C and 40 °C, respectively.

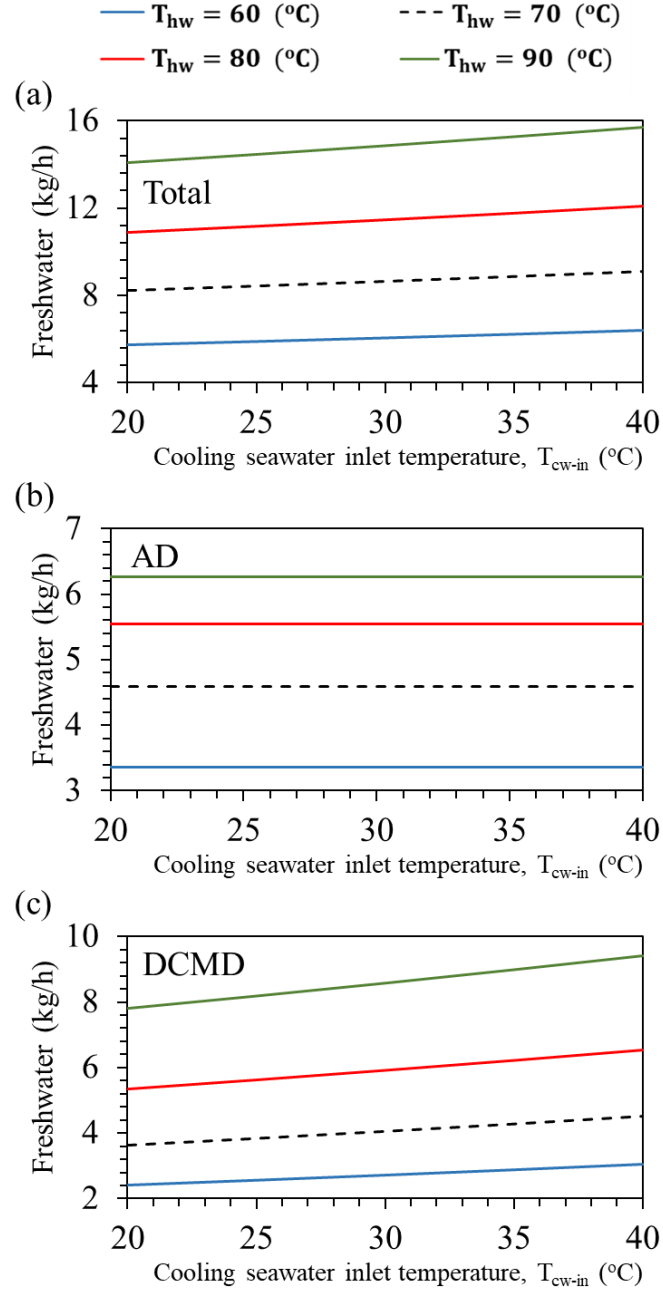


Figure 19 Effect of temperature of the inlet seawater as a cooling water and a feed for DCMD ($T_{cw,in}$) on the freshwater production at different heating water temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{hw,in} = 80$ °C, $T_{cond,in} = 30$ °C, $T_{chw,in} = 30$ °C, $\dot{m}_{hw,in} = 0.4$ kg/s, $\dot{m}_{cw,in} = \dot{m}_{DCMD,f} = 0.08$ kg/s, $\dot{m}_{chw,in} = 0.5$ kg/s, $\dot{m}_{cond,in} = 0.3$ kg/s, $T_{bp} = 20$ °C, and half-cycle time = 700 s for configuration 2.

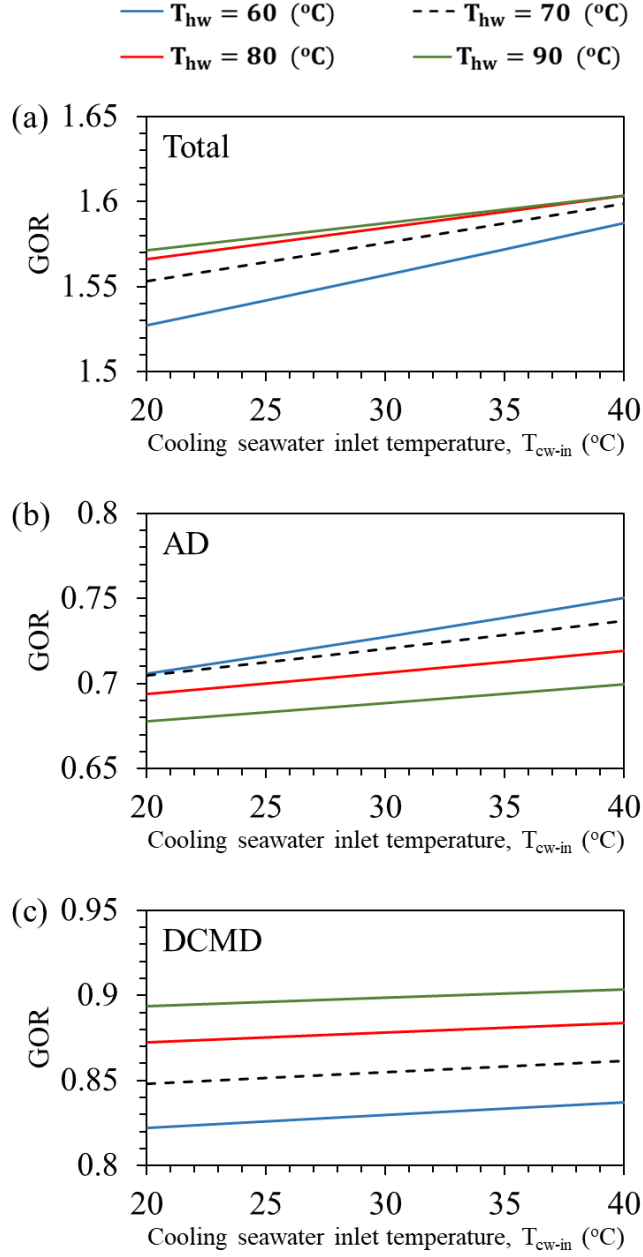


Figure 20 Effect of temperature of the inlet seawater as a cooling water and a feed for DCMD ($T_{cw,in}$) on the system GOR at different heating water temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{hw,in} = 80$ °C, $T_{cond,in} = 30$ °C, $T_{chw,in} = 30$ °C, $\dot{m}_{chw,in} = 0.5$ kg/s, $\dot{m}_{hw,in} = 0.4$ kg/s, $\dot{m}_{cond,in} = 0.3$ kg/s, $\dot{m}_{cw,in} = \dot{m}_{DCMD,f} = 0.08$ kg/s, $T_{bp} = 20$ °C, and half-cycle time= 700 s for configuration 2.

In configuration 3's arrangement (Figure 10, c), the brine stream exiting from DCMD system is used as a heating source for the AD system; therefore, maintaining its temperature between 55 and 85 °C is critical to achieve excellent performance of the hybrid system. DCMD hot channel outflow temperature ($T_{f,out}$) was calculated based on mass and energy balance of the hot side of the membrane as follows,

$$T_{f,out} = \frac{\dot{m}_{f,in} c_{p,f} - Q_t}{\dot{m}_{f,out} c_{p,f}} \quad (58)$$

Where $\dot{m}_{f,in}$ and $\dot{m}_{f,out}$ represent the inlet and outlet mass flow rates of the hot water, respectively; and Q_t stands for the total heat energy removed from the hot water in the DCMD hot channel.

Maximizing $T_{f,out}$ occurs when decreasing the length of channel because smaller the membrane length results in less resident time for hot water in the feed channel. However, this may lead to decreasing the surface area, hence decreasing the permeate flux. Thus, choosing optimum length and surface area is needed to maintain a high DCMD permeate flux and a high outlet temperature (enough to pre-heat the desorption bed) at the same time. Figure 21 depicts how membrane length and effective area (A_{mem}) influence the outlet temperature of the DCMD hot channel. Results show that the higher the membrane length and effective area, the lower the outlet temperature of the DCMD hot channel. For instance, at DCMD inlet feed temperature ($T_{f,in}$) of 80 °C, $A_{mem} \leq 0.175 \text{ m}^2$, and $L_{mem} \leq 0.5 \text{ m}$, $T_{f,out}$ is $\geq 65 \text{ °C}$ (which is enough to pre-heat the desorption bed) ; while, $T_{f,out}$ value is $< 50 \text{ °C}$ at $A_{mem} > 0.2 \text{ m}^2$ and $L_{mem} > 0.5 \text{ m}$.

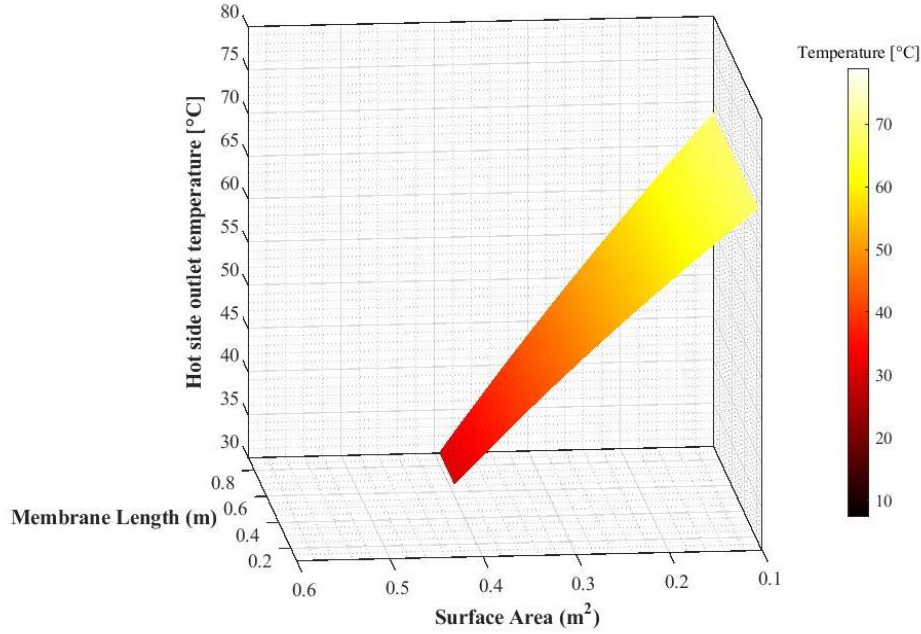


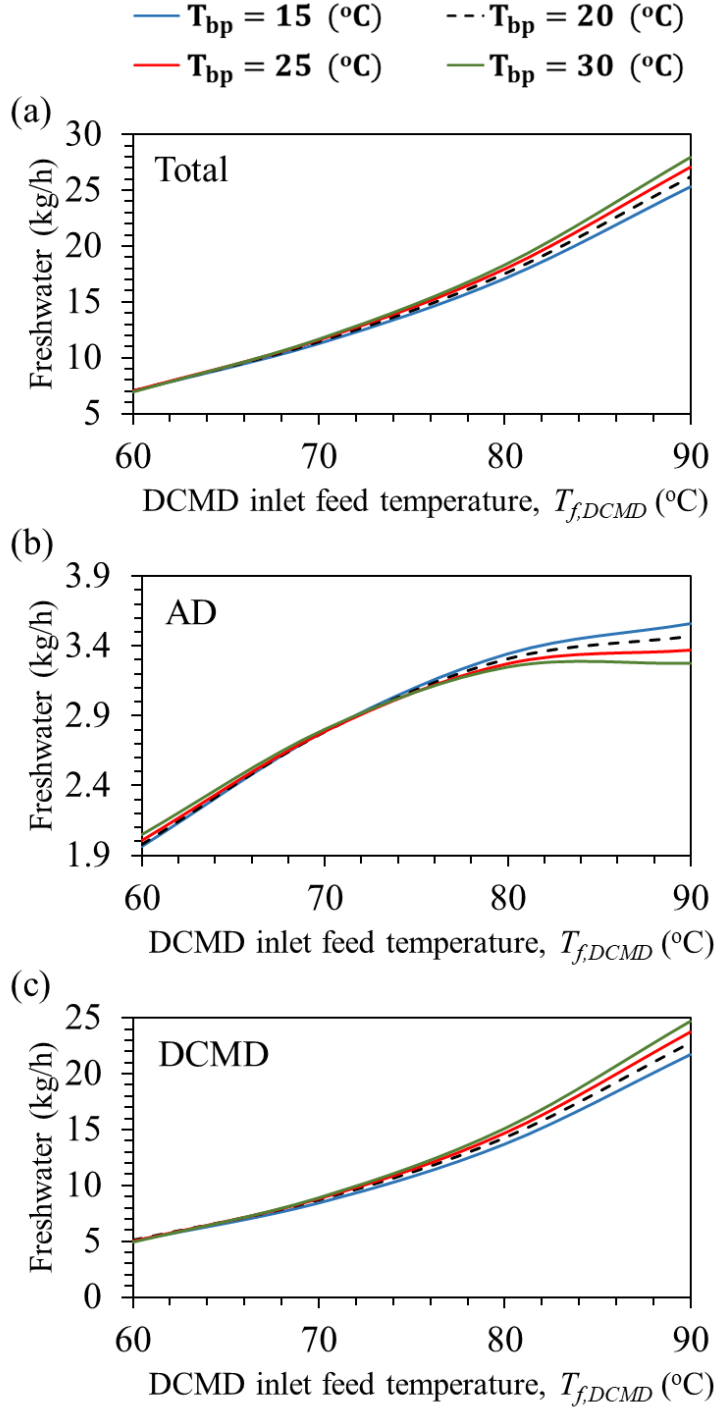
Figure 21 Effect of membrane length and effective area on the outlet temperature of the DCMD hot channel.

The arrangement of configuration 3 differs from those of configurations 1 and 2. The hot brine stream exiting from DCMD is utilized to pre-heat the desorption bed as illustrated in Figure 10, c. Therefore, the influence of the feed inlet temperature ($T_{f,DCMD}$) and the permeate temperature (T_{bp}) on the performance of the AD-DCMD hybrid system is investigated. Figure 22, c shows that at fixed cold permeate temperature, freshwater production of DCMD system increases when the feed (seawater) temperature rises from 60 °C to 90 °C. This can be attributed to the enhancement in the permeate flux due to the combined effect of the exponential relationship between temperature and vapor pressure (Eqs. 4 and 5). According to this equation, when the feed temperature is low, the effect of temperature on vapor pressure is low; however, when the feed temperature is high, the effect becomes more significant. An increase in the vapor pressure difference between the membrane's two sides is caused by a rise in the temperature difference across it.

Similarly, AD system produce more freshwater when the feed (seawater) temperature is increased, because this results in rising the heating water temperature inlet to the desorption bed which results in higher desorbed water vapor amounts, hence the performance of the AD system is enhanced (Figure 22, b). However, at $T_{f,DCMD} \geq 80\text{ }^{\circ}\text{C}$ is almost constant. Configuration 3 resulted in excellent overall freshwater amounts from the hybrid system. E.g. the total freshwater amount reaches 18.3 and 28 kg/h at $T_{f,DCMD}$ of 80 and 90 $^{\circ}\text{C}$, respectively. Freshwater trends show clearly that almost the same total freshwater amounts are produced at all permeate temperatures between 15 and 30 $^{\circ}\text{C}$.

Figure 23 depicts the GOR values of DCMD, AD, and AD-DCMD hybrid systems as a function of seawater inlet temperature at different DCMD cold permeate temperature values. Total GOR increases with increasing the temperature of the inlet (feed) is raised with a high contribution from the DCMD system. GOR of DCMD range between 1.6 and 1.95 is obtained for feed temperatures ranging from 60 $^{\circ}\text{C}$ to 90 $^{\circ}\text{C}$ (Figure 23, c). Similarly, GOR of the AD system is greater than 0.7 with a maximum of 0.725 at $T_{f,DCMD}$ feed temperatures of 80 $^{\circ}\text{C}$ (Figure 23, b). The integrated system's overall GOR reaches 2.69 which is higher than those of configurations 1 and 2.

As previously discussed, the flux increases as the temperature difference increases. However, a little deviation is noticed in Figure 23, c specially at $T_{f,DCMD} \geq 80\text{ }^{\circ}\text{C}$. This may be due to the fact that the low temperature of the cold permeate stream causes high conduction heat losses across the membrane and from the module itself to the surroundings. According to the result analysis of configuration 3 $T_{f,DCMD}$ of 80 $^{\circ}\text{C}$ results in an optimal performance.



[Figure 22 Effect of inlet feed temperature of DCMD system on the freshwater production at different DCMD permeate temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30$ °C, $T_{cond,in} = 30$ °C, $T_{chw,in} = 30$ °C, $\dot{m}_{hw,in} = \dot{m}_{f,DCMD} = 0.08$ kg/s, $\dot{m}_{cw,in} = 0.3$ kg/s, $\dot{m}_{chw,in} = 0.5$ kg/s, $\dot{m}_{cond,in} = 0.3$ kg/s, $\dot{m}_{bp} = 0.06$ kg/s, and half-cycle time= 700 s for configuration 3.]

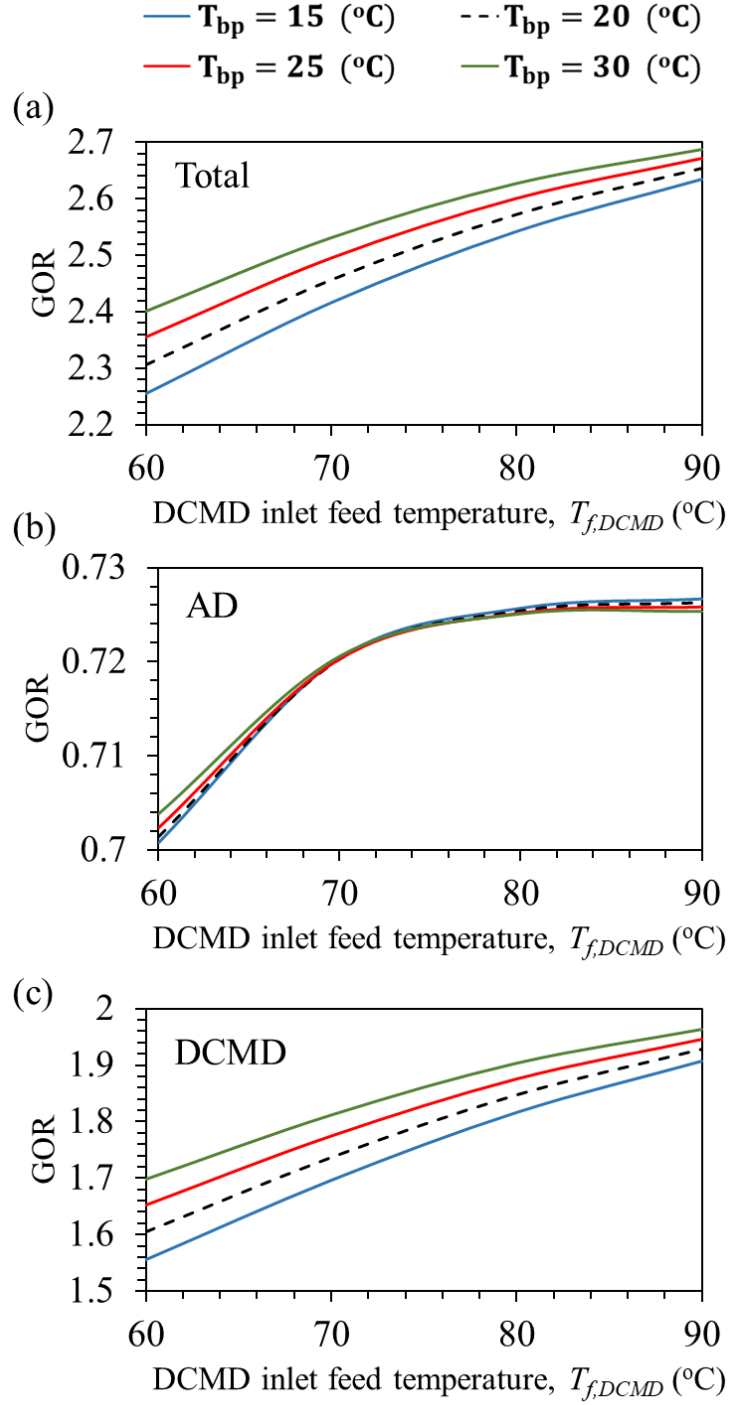


Figure 23 Effect of inlet feed temperature of DCMD system on the system GOR at different DCMD permeate temperatures for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30$ °C, $T_{cond,in} = 30$ °C, $T_{chw,in} = 30$ °C, $\dot{m}_{chw,in} = 0.5$ kg/s, $\dot{m}_{hw,in} = \dot{m}_{f,DCMD} = 0.08$ kg/s, $\dot{m}_{cw,in} = 0.3$ kg/s, $\dot{m}_{cond,in} = 0.3$ kg/s, $\dot{m}_{bp} = 0.06$ kg/s, and half-cycle time= 700 s for configuration 3.

The effect of DCMD hot seawater feed flow rate ($\dot{m}_{f,DCMD}$) and cold permeate mass flow rate ($\dot{m}_{bp}$) values on the hybrid system's performance is also investigated. Freshwater production of DCMD system is enhanced by increasing the mass flow rate of seawater inlet (Figure 24). This can be attributed to the fact that the turbulence level in the DCMD feed channel raises as the flow rate increases because the channel flow's Reynolds number increases. As a direct result, the boundary layer mixing beside the membrane surface is enhanced, hence increasing the heat and mass transfer coefficient. Because both the feed and permeate streams are in direct contact with the membrane in DCMD module, it is expected that the mass flow rates on both membrane sides will be effective in the permeation process, and therefore freshwater production is enhanced.

Figure 24, c depicts the relationship between feed mass flow rate and system freshwater production. The rise in freshwater amount is significantly noticeable when the feed flow rate is increased. For instance, at a fixed feed temperature of 80 °C, increasing the mass flow rate from 0.03 to 0.2 kg/s results in a freshwater rate increase from 8.6 to 21.5 kg/h, which is considered a significant jump in system's freshwater production. For the selected conditions of feed mass flow rate, it was observed that the channel flow Reynolds number on the feed side is turbulent (> 2300) beyond mass flow rate of 0.07 kg/s. The more the mass flow rate, the higher the turbulence; therefore, the less water amounts settle down in the feed channel, hence the mass transfer of water vapor across the membrane is enhanced. AD system's freshwater production is also enhanced by increasing the mass flow rate of seawater inlet (DCMD feed); which results in increasing the heating water amounts, leading to better adsorption/desorption performance (Figure 24, b).

GOR values of the hybrid system are also influenced by the DCMD feed mass flow rate (Figure 25, a). Total GOR values reach 2.45, 2.65 and 2.7 at $\dot{m}_{f,DCMD}$ of 0.03, 0.08 and 0.2 kg/s, respectively. It is noticed that increasing $\dot{m}_{f,DCMD}$ above 0.08 kg/s results with a low enhancement in the hybrid system's performance. Thus, 0.08 is the optimum $\dot{m}_{f,DCMD}$ for an efficient performance.

It is clear that, unlike feed flow rate, permeate flow rate doesn't have a noticeable impact on the overall freshwater production and the GOR of the hybrid system. The feed side is more effective in enhancing the performance of AD-DCMD hybrid system, because it is the source of vaporization and the controller of the permeation process for DCMD process, as well as, it is the supplier of the heating water for AD process.

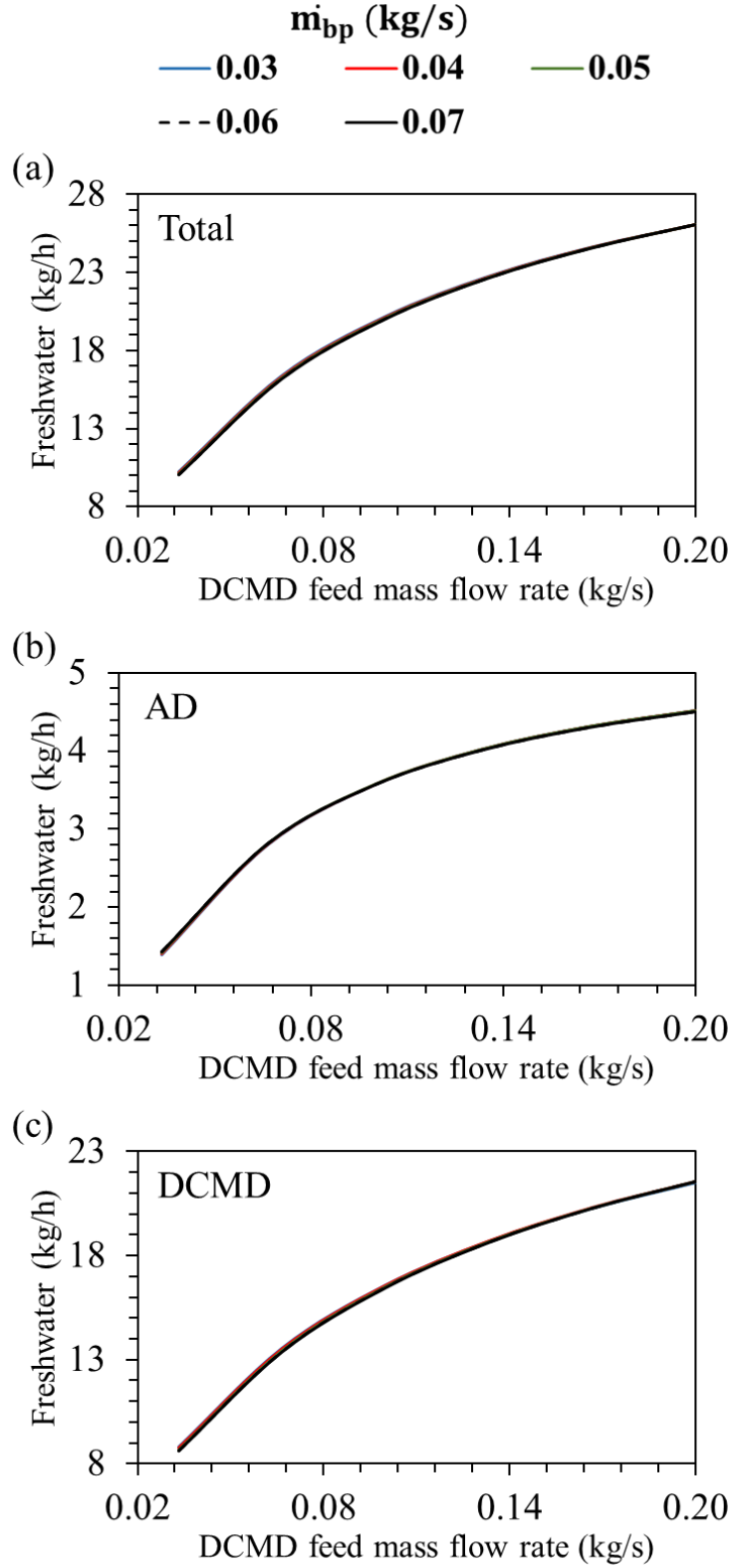


Figure 24 Effect of inlet feed flow rate of DCMD system ($\dot{m}_{f,DCMD} = \dot{m}_{hw,in}$) on the freshwater production at different DCMD permeate mass flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30^\circ\text{C}$, $T_{cond,in} = 30^\circ\text{C}$, $T_{chw,in} = 30^\circ\text{C}$, $\dot{m}_{cw,in} = 0.3\text{ kg/s}$, $\dot{m}_{chw,in} = 0.5\text{ kg/s}$, $\dot{m}_{cond,in} = 0.3\text{ kg/s}$, $T_{bp} = 20^\circ\text{C}$, $T_{bf} = 80^\circ\text{C}$, and half-cycle time = 700 s for configuration 3.

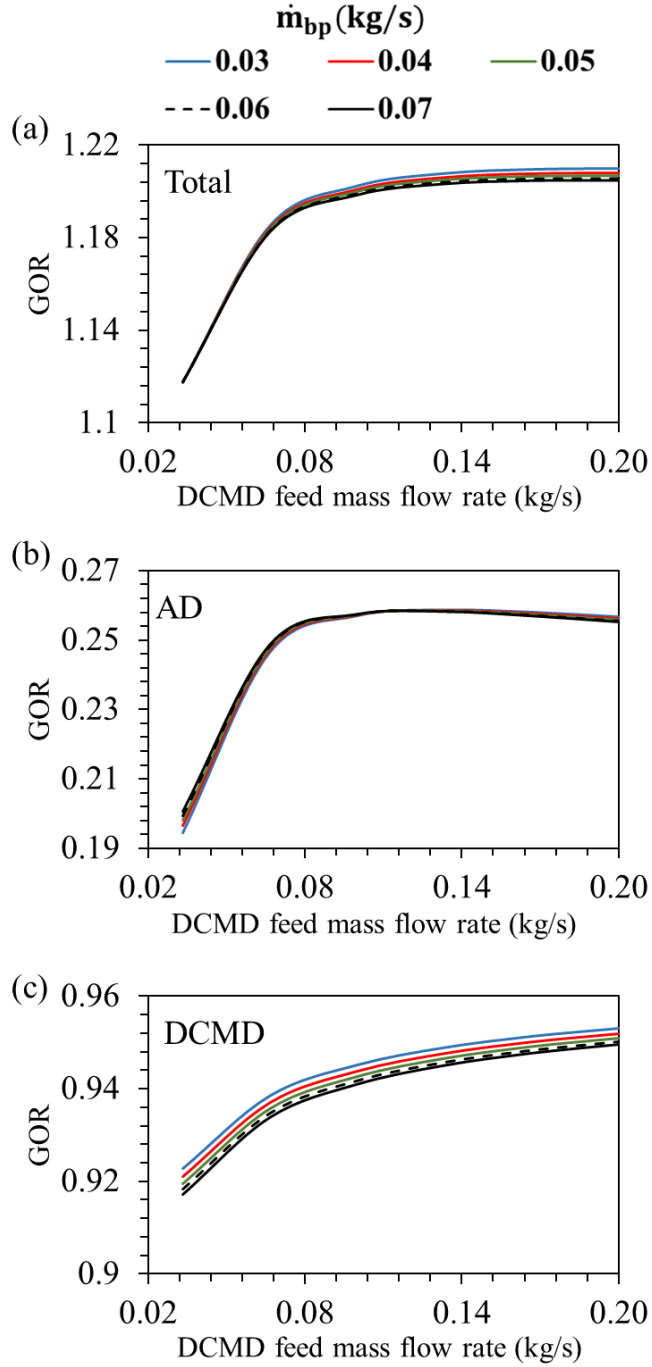


Figure 25 Effect of inlet feed flow rate of DCMD system on the system GOR at different DCMD permeate mass flow rates for (a) AD-DCMD (total), (b) AD, and (c) DCMD under $T_{cw,in} = 30\text{ }^{\circ}\text{C}$, $T_{cond,in} = 30\text{ }^{\circ}\text{C}$, $T_{chw,in} = 30\text{ }^{\circ}\text{C}$, $\dot{m}_{chw,in} = 0.5\text{ kg/s}$, $\dot{m}_{cw,in} = 0.3\text{ kg/s}$, $\dot{m}_{cond,in} = 0.3\text{ kg/s}$, $T_{bp} = 20\text{ }^{\circ}\text{C}$, $T_{bf} = 80\text{ }^{\circ}\text{C}$, and half-cycle time= 700 s for configuration 3.

Figure 26 shows how the proposed system's fresh water production is affected by the temperature of the heat source. For configurations 1 and 2, with an increase in the heat source temperature, an improvement in adsorption uptake occurs throughout the adsorption process, which in turn improves the water vapor desorption. Moreover, the temperature of the DCMD feed (seawater) is raised either by the waste heat inside the AD condenser (as in configuration 1) or inside the AD adsorption bed (as in configuration 2). Therefore, total freshwater production of the hybrid system is enhanced. Overall freshwater production values are found to be 26 and 53 m³/ton.day for configurations 1 and 2, respectively. Freshwater production of configuration 2 is remarkably high. This can be attributed to the appropriate arrangement of configuration 2 in which the feed stream of the DCMD system enters with a relatively higher temperature than that in configuration 1 resulting in a larger amount of freshwater.

Regarding configuration 3, overall freshwater production values are 25, 41, 62, and 93 m³/ton.day for the heating source temperatures of 60, 70, 80 and 90 °C, respectively. This can be attributed to the appropriate arrangement of the hybrid system in which the feed stream of the DCMD system enters with a relatively high temperature ≥ 60 °C resulting in a large amount of freshwater, and this also results with an outlet temperature of the DCMD brine stream (Figure 21) which is high enough to operate the AD system.

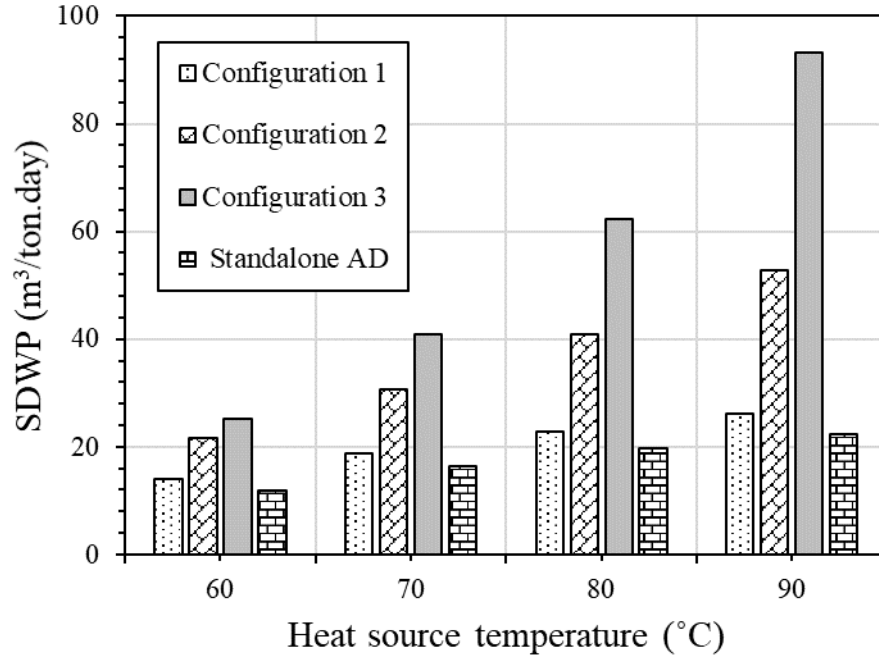


Figure 26 SDWP of the proposed configurations at different heat source temperatures.

As a by-product, the configurations studied generate a cooling effect which can be used for air conditioning. Figure 27 depicts how COP of the hybrid system is influenced by the heat source temperature. COP of configurations 1 and 2 are 0.76 and 0.7, respectively at heat source temperature of 60°C which is higher than some standalone AD refrigeration systems [98,99] at similar operating conditions. COP is noticed to decrease with increasing the heating source temperature and this is explained by Eqn. 49. The heat removed in the evaporator is almost constant while increasing the heat source temperature results in increasing the inlet heat of the AD system, hence decreasing the COP.

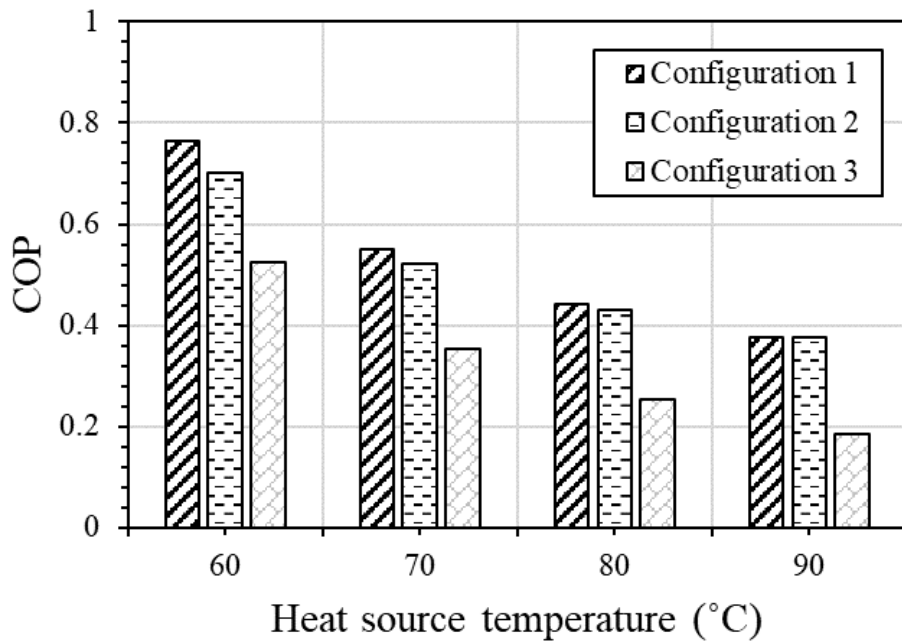


Figure 27 COP of the proposed configurations at different heat source temperatures.

Capital investment costs for the AD-DCMD hybrid system are estimated using the model reported in Section “ Economic model” of this thesis. Figure 28 shows the costs of freshwater produced for the optimal performance of the two configurations being emphasized compared with the cost of freshwater produced by standalone AD system . Freshwater costs of 0.049 and 0.026 (\$/L) for configurations 1 and 2, respectively were obtained at heat source temperature of 90 °C. Freshwater cost of configuration 2 is reduced by 58% compared with the cost of that produced by standalone AD system.

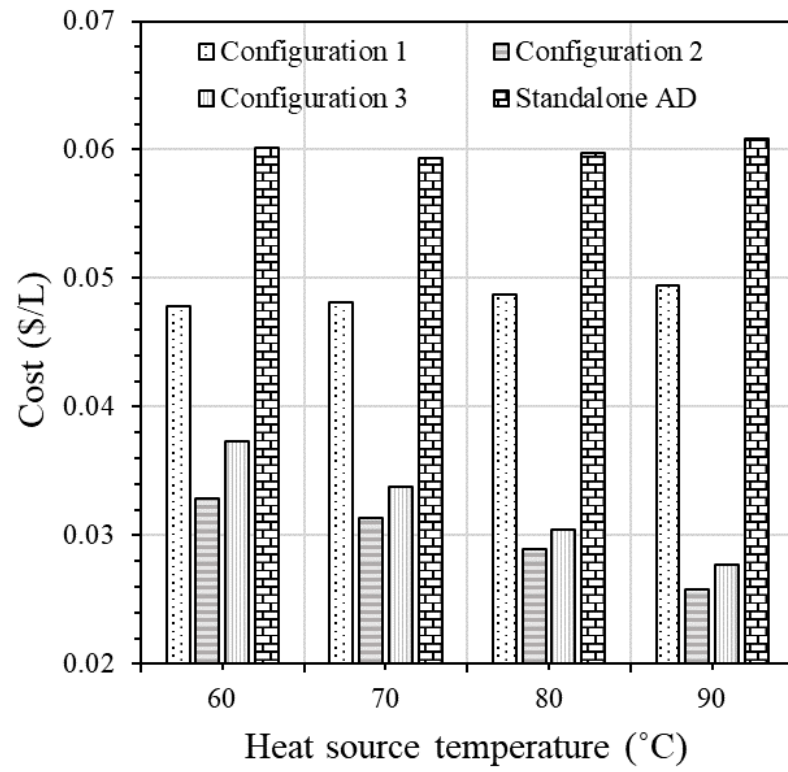


Figure 28 Freshwater costs at different heat source temperatures.

CHAPTER 5

CONCLUSION

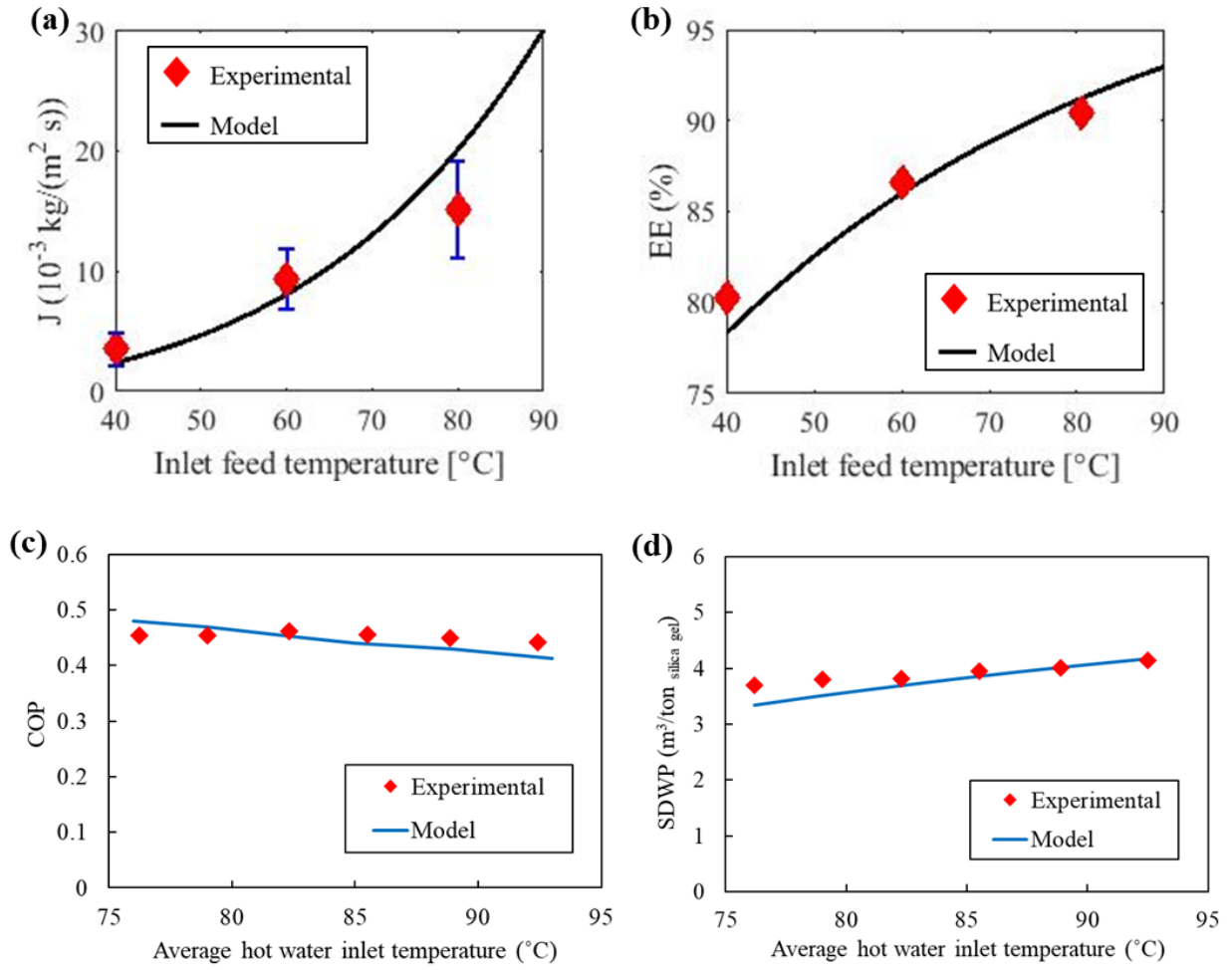
DCMD-AD hybrid desalination system has been designed, and its performance has been evaluated using a robust mathematical model in which each component of the system is taken into account. Two possible configurations have been described, which all have been demonstrated to be feasible and practical. Experimentally-validated mathematical models are employed in the simulation. GOR, freshwater production rate, SDWP, COP, and percent reduction in freshwater cost were all evaluated and investigated to obtain optimum performance for each configuration. For configurations 1 and 2, the GOR values reach 1.48 and 1.6, respectively, which is almost double that of a standalone AD system. Whereas GOR of configuration 3 is always less than 1.25, which means it is not as energy efficient as configurations 1 and 2; however, its water production is the highest. At a source temperature of 90 °C, the freshwater production of the DCMD-AD hybrid system attains SDWP values of 26, 53 and 93 m³/ton. day for configurations 1, 2, and 3, respectively. Moreover, each of the three configurations has proved a capability of producing a cooling effect as a by-product, with COP values reaching 0.76 and 0.70, and 0.52 for configurations 1, 2, and 3, respectively. Furthermore, the percent reduction in the freshwater cost produced by configuration 2 (which reaches 58%) demonstrates a significant improvement in the configuration's design.

Appendix 1: Model Validation

The hybrid AD-DCMD system that was investigated in this article has never been studied by researchers. In order to validate the current mathematical model, the AD and DCMD are validated individually using experimental data from the literature.

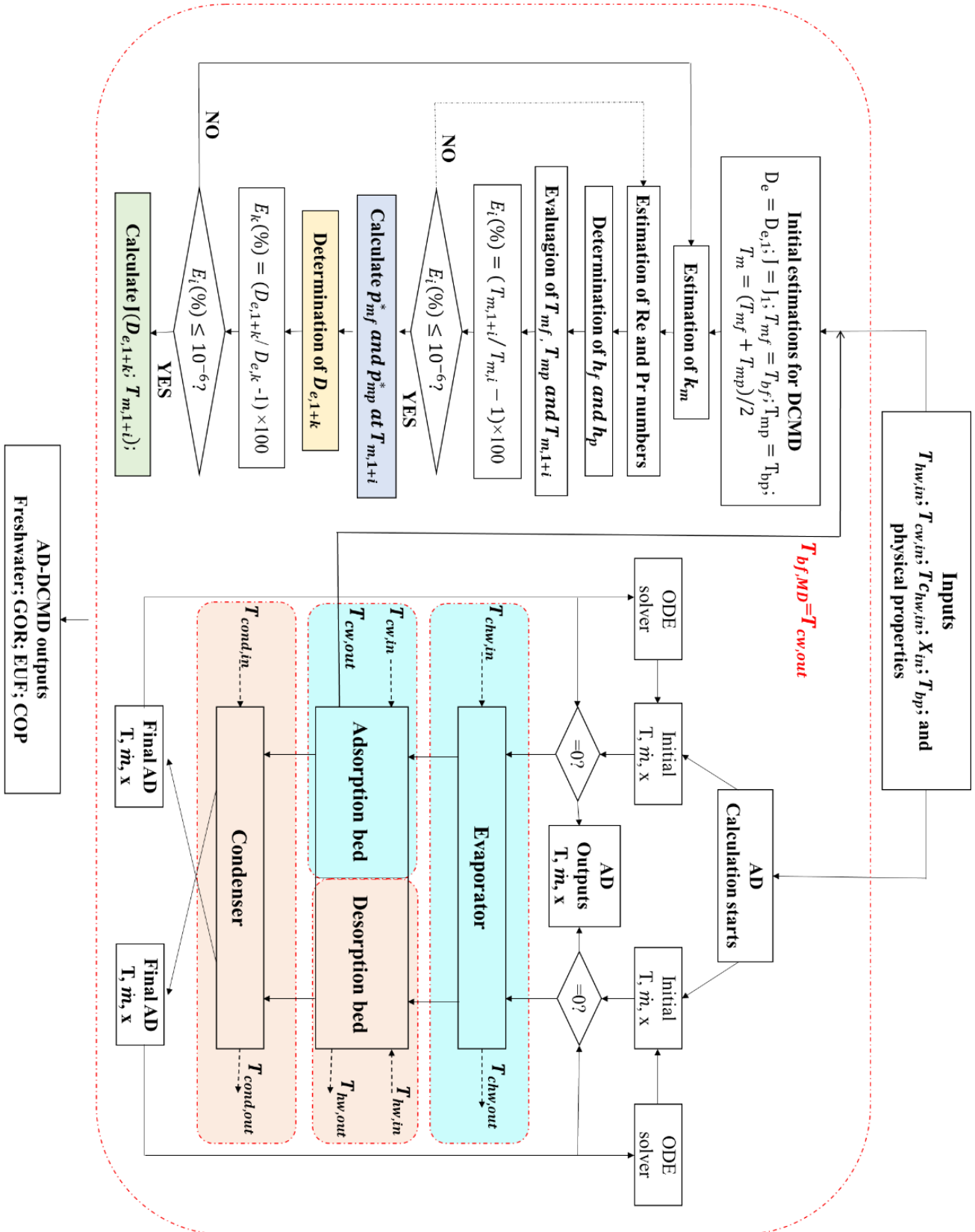
The DCMD model was validated using the experimental work of Essalhi and Khayet [69] with the same constructive and operating conditions. Figures 29, a and 29, b compare the permeate flux (J) and the evaporative efficiency expected by the proposed DCMD model to the experimental data (Essalhi and Khayet, 2013) and a good agreement was realized with an error of 0.4% – 10% and 0.6% 2.4% for permeate flux and evaporative efficiency, respectively.

The experimental data of Alsaman et al. [84] was used to validate the AD model because they used silica-gel and water as adsorption pair. In addition to the operational and constructive parameters indicated in Table 4, the heating water mass flowrate and the chilled water mass flowrates were 0.2 kg/s and 0.1 kg/s, respectively. Figures 29 (c and d) show an excellent agreement between the simulated and experimental COP and SDWP values with a percent error ranges between 5.8 - 6.4%.



[Figure 29 A comparison of experimental results values and those of the present simulation.]

Appendix 2: AD-DCMD Model Algorithm



References

- [1] H.T. El-Dessouky, H.M. Ettouney, Fundamentals of salt water desalination, (2002) 670.
- [2] M.I. Zubair, THERMAL AND ECONOMICAL ANALYSIS OF HDH SYSTEMS DRIVEN BY SOLAR THERMAL ENERGY WITH A STORAGE OPTION - KFUPM ePrints, KFUPM, 2015. <https://eprints.kfupm.edu.sa/id/eprint/139813/> (accessed May 9, 2022).
- [3] Ridhish Kumar, Shubham LanjewarShubham Lanjewar, Anupam Mukherjee, Anirban RoyAnirban Roy, Seawater desalination process by osmosis, in: Chem. Eng. World, Jasubhai Media Pvt. Ltd, 2018. https://www.researchgate.net/publication/329115755_Seawater_desalination_process_by_osmosis (accessed May 9, 2022).
- [4] T. Luo, R. Young, P. Reig, Technical Note: Aqueduct Projected Water Stress Country Rankings | World Resources Institute, (2015). <https://www.wri.org/research/aqueduct-projected-water-stress-country-rankings> (accessed May 9, 2022).
- [5] W.K. Al-Zubari, Sustainable water resources management in the Gulf Cooperation Council countries. in: WWAP (United Nations World Water Assessment Programme), 2015.
- [6] Saudi Arabia - Water Resources Equipment (WRE) and Technologies Market | export.gov, (n.d.). <https://legacy.export.gov/article?id=Saudi-Arabia-Water-Resources-Equipment-WRE-and-Technologies-Market> (accessed May 9, 2022).

- [7] A.J. Beaudoin, Analysis of Domestic and International Desalination to Outline the Decision Making Landscape for Implementation and Operation of Desalination Plants in the United States, (2015).
- [8] A.H. Hassabou, Experimental and Numerical Analysis of a PCM-Supported Humidification-Dehumidification Solar Desalination System, (2011) 224. <http://www.td.mw.tum.de/tum-td/de/forschung/dissertationen/index.html/>.
- [9] Water in Saudi Arabia: Desalination, Wastewater, and Privatization - USSBC, (n.d.). <https://ussaudi.org/water-in-saudi-arabia-desalination-wastewater-and-privatization/> (accessed May 9, 2022).
- [10] O.K.M. Ouda, Y. Khalid, A.H. Ajbar, M. Rehan, K. Shahzad, I. Wazeer, A.S. Nizami, Long-term desalinated water demand and investment requirements: a case study of Riyadh, J. Water Reuse Desalin. 8 (2018) 432–446. <https://doi.org/10.2166/WRD.2017.107>.
- [11] E. Zaidan, M. Al-Saidi, S.H. Hammad, Sustainable development in the Arab world – is the Gulf Cooperation Council (GCC) region fit for the challenge?, <https://doi.org/10.1080/09614524.2019.1628922>. 29 (2019) 670–681. <https://doi.org/10.1080/09614524.2019.1628922>.
- [12] O.K.M. Ouda, Water demand versus supply in Saudi Arabia: current and future challenges, <http://dx.doi.org/10.1080/07900627.2013.837363>. 30 (2014) 335–344. <https://doi.org/10.1080/07900627.2013.837363>.
- [13] E. Jones, M. Qadir, M.T.H. van Vliet, V. Smakhtin, S. mu Kang, The state of desalination and brine production: A global outlook, Sci. Total Environ. 657 (2019) 1343–1356. <https://doi.org/10.1016/J.SCITOTENV.2018.12.076>.

- [14] H.S. SON, Sustainability Evaluation of Hybrid Desalination Systems: Multi Effect Distillation – Adsorption (MED-AD) and Forward Osmosis – Membrane Distillation (FO-MD), (2020).
- [15] K. Elsaid, M. Kamil, E.T. Sayed, M.A. Abdelkareem, T. Wilberforce, A. Olabi, Environmental impact of desalination technologies: A review, *Sci. Total Environ.* 748 (2020) 141528. <https://doi.org/10.1016/j.scitotenv.2020.141528>.
- [16] M. Elimelech, W.A. Phillip, The future of seawater desalination: Energy, technology, and the environment, *Science* (80-.). 333 (2011) 712–717. <https://doi.org/10.1126/SCIENCE.1200488>.
- [17] N. Ghaffour, J. Bundschuh, H. Mahmoudi, M.F.A. Goosen, Renewable energy-driven desalination technologies: A comprehensive review on challenges and potential applications of integrated systems, *Desalination.* 356 (2015) 94–114. <https://doi.org/10.1016/J.DESAL.2014.10.024>.
- [18] M. Qasim, M. Badrelzaman, N.N. Darwish, N.A. Darwish, N. Hilal, Reverse osmosis desalination: A state-of-the-art review, *Desalination.* 459 (2019) 59–104. <https://doi.org/10.1016/J.DESAL.2019.02.008>.
- [19] B. Van der Bruggen, C. Vandecasteele, Distillation vs. membrane filtration: overview of process evolutions in seawater desalination, *Desalination.* 143 (2002) 207–218. [https://doi.org/10.1016/S0011-9164\(02\)00259-X](https://doi.org/10.1016/S0011-9164(02)00259-X).
- [20] S. Bhojwani, K. Topolski, R. Mukherjee, D. Sengupta, M.M. El-Halwagi, Technology review and data analysis for cost assessment of water treatment systems, *Sci. Total Environ.* 651 (2019) 2749–2761. <https://doi.org/10.1016/J.SCITOTENV.2018.09.363>.

- [21] Desalination of saline and brackish water is becoming more affordable, *Membr. Technol.* 2009 (2009) 8–10. [https://doi.org/10.1016/S0958-2118\(09\)70142-9](https://doi.org/10.1016/S0958-2118(09)70142-9).
- [22] L.F. Greenlee, D.F. Lawler, B.D. Freeman, B. Marrot, P. Moulin, Reverse osmosis desalination: water sources, technology, and today's challenges, *Water Res.* 43 (2009) 2317–2348. <https://doi.org/10.1016/J.WATRES.2009.03.010>.
- [23] K.P. Lee, T.C. Arnot, D. Mattia, A review of reverse osmosis membrane materials for desalination—Development to date and future potential, *J. Memb. Sci.* 370 (2011) 1–22. <https://doi.org/10.1016/J.MEMSCI.2010.12.036>.
- [24] C. Fritzmann, J. Löwenberg, T. Wintgens, T. Melin, State-of-the-art of reverse osmosis desalination, *Desalination.* 216 (2007) 1–76. <https://doi.org/10.1016/J.DESAL.2006.12.009>.
- [25] L. Malaeb, G.M. Ayoub, Reverse osmosis technology for water treatment: State of the art review, *Desalination.* 267 (2011) 1–8. <https://doi.org/10.1016/J.DESAL.2010.09.001>.
- [26] D.L. Oatley-Radcliffe, M. Walters, T.J. Ainscough, P.M. Williams, A.W. Mohammad, N. Hilal, Nanofiltration membranes and processes: A review of research trends over the past decade, *J. Water Process Eng.* 19 (2017) 164–171. <https://doi.org/10.1016/J.JWPE.2017.07.026>.
- [27] J.E. Miller, Review of Water Resources and Desalination Technologies, *Desalination.* 93 (2003) 119–125. [https://doi.org/10.1016/0011-9164\(93\)80099-9](https://doi.org/10.1016/0011-9164(93)80099-9).
- [28] A.D. Khawaji, I.K. Kutubkhanah, J.M. Wie, Advances in seawater desalination technologies, *Desalination.* 221 (2008) 47–69. <https://doi.org/10.1016/J.DESAL.2007.01.067>.

- [29] M. Al-Shammiri, M. Safar, Multi-effect distillation plants: state of the art, *Desalination*. 126 (1999) 45–59. [https://doi.org/10.1016/S0011-9164\(99\)00154-X](https://doi.org/10.1016/S0011-9164(99)00154-X).
- [30] A. Ophir, F. Lokiec, Advanced MED process for most economical sea water desalination, *Desalination*. 182 (2005) 187–198. <https://doi.org/10.1016/J.DESAL.2005.02.026>.
- [31] A.M. Abusorrah, F. Mebarek-Oudina, A. Ahmadian, D. Baleanu, Modeling of a MED-TVC desalination system by considering the effects of nanoparticles: energetic and exergetic analysis, *J. Therm. Anal. Calorim.* 144 (2021) 2675–2687. <https://doi.org/10.1007/S10973-020-10524-1/FIGURES/14>.
- [32] M. Elimelech, W.A. Phillip, The future of seawater desalination: Energy, technology, and the environment, *Science* (80-.). 333 (2011) 712–717. https://doi.org/10.1126/SCIENCE.1200488/SUPPL_FILE/ELIMELECH.SOM.PDF.
- [33] R. Schwantes, K. Chavan, D. Winter, C. Felsmann, J. Pfafferott, Techno-economic comparison of membrane distillation and MVC in a zero liquid discharge application, *Desalination*. 428 (2018) 50–68. <https://doi.org/10.1016/J.DESAL.2017.11.026>.
- [34] A.A. Mabrouk, A.S. Nafey, H.E.S. Fath, Analysis of a new design of a multi-stage flash–mechanical vapor compression desalination process, *Desalination*. 204 (2007) 482–500. <https://doi.org/10.1016/J.DESAL.2006.02.046>.
- [35] H.T. El-Dessouky, H.M. Ettouney, F. Al-Juwayhel, Multiple Effect Evaporation—Vapour Compression Desalination Processes, *Chem. Eng. Res. Des.* 78 (2000) 662–676. <https://doi.org/10.1205/026387600527626>.

- [36] I. Sutzkover-Gutman, D. Hasson, Feed water pretreatment for desalination plants, *Desalination*. 264 (2010) 289–296. <https://doi.org/10.1016/J.DESAL.2010.07.014>.
- [37] M.W. Shahzad, M. Burhan, L. Ang, K.C. Ng, Energy-water-environment nexus underpinning future desalination sustainability, *Desalination*. 413 (2017) 52–64. <https://doi.org/10.1016/J.DESAL.2017.03.009>.
- [38] K. Elsaid, E.T. Sayed, M.A. Abdelkareem, M.S. Mahmoud, M. Ramadan, A.G. Olabi, Environmental impact of emerging desalination technologies: A preliminary evaluation, *J. Environ. Chem. Eng.* 8 (2020) 104099. <https://doi.org/10.1016/J.JECE.2020.104099>.
- [39] T. Serrano-Tovar, B. Peñate Suárez, A. Musicki, J.A. de la Fuente Bencomo, V. Cabello, M. Giampietro, Structuring an integrated water-energy-food nexus assessment of a local wind energy desalination system for irrigation, *Sci. Total Environ.* 689 (2019) 945–957. <https://doi.org/10.1016/J.SCITOTENV.2019.06.422>.
- [40] M.W. Shahzad, M. Burhan, K.C. Ng, Pushing desalination recovery to the maximum limit: Membrane and thermal processes integration, *Desalination*. 416 (2017) 54–64. <https://doi.org/10.1016/J.DESAL.2017.04.024>.
- [41] A. Subramani, J.G. Jacangelo, Emerging desalination technologies for water treatment: A critical review, *Water Res.* 75 (2015) 164–187. <https://doi.org/10.1016/J.WATRES.2015.02.032>.
- [42] A.G. Olabi, T. Wilberforce, E.T. Sayed, K. Elsaid, H. Rezk, M.A. Abdelkareem, Recent progress of graphene based nanomaterials in bioelectrochemical systems, *Sci. Total Environ.* 749 (2020) 141225. <https://doi.org/10.1016/J.SCITOTENV.2020.141225>.

- [43] E.T. Sayed, N. Shehata, M.A. Abdelkareem, M.A. Atieh, Recent progress in environmentally friendly bio-electrochemical devices for simultaneous water desalination and wastewater treatment, *Sci. Total Environ.* 748 (2020) 141046. <https://doi.org/10.1016/J.SCITOTENV.2020.141046>.
- [44] Y.I. Aristov, G. Restuccia, G. Cacciola, V.N. Parmon, A family of new working materials for solid sorption air conditioning systems, *Appl. Therm. Eng.* 22 (2002) 191–204. [https://doi.org/10.1016/S1359-4311\(01\)00072-2](https://doi.org/10.1016/S1359-4311(01)00072-2).
- [45] B.B. Saha, A. Chakraborty, S. Koyama, Y.I. Aristov, A new generation cooling device employing CaCl₂-in-silica gel-water system, *Int. J. Heat Mass Transf.* 52 (2009) 516–524. <https://doi.org/10.1016/J.IJHEATMASSTRANSFER.2008.06.018>.
- [46] US4881376A - Adsorption refrigeration system - Google Patents, (n.d.). <https://patents.google.com/patent/US4881376A/en> (accessed May 9, 2022).
- [47] K.C. Ng, K. Thu, B.B. Saha, A. Chakraborty, Study on a waste heat-driven adsorption cooling cum desalination cycle, *Int. J. Refrig.* 35 (2012) 685–693. <https://doi.org/10.1016/J.IJREFRIG.2011.01.008>.
- [48] O.R.F. Cevallos, Adsorption Characteristics of Water and Silica Gel System for Desalination Cycle, (2012) 72. http://repository.kaust.edu.sa/kaust/bitstream/10754/273097/1/OscarFonseca113292_Thesis_final.pdf.
- [49] R.E. Critoph, R. Vogel, Possible adsorption pairs for use in solar cooling, [Http://Dx.Doi.Org/10.1080/01430750.1986.9675500](http://dx.doi.org/10.1080/01430750.1986.9675500). 7 (2011) 183–190. <https://doi.org/10.1080/01430750.1986.9675500>.

- [50] Y.D. Kim, K. Thu, K.C. Ng, G.L. Amy, N. Ghaffour, A novel integrated thermal-/membrane-based solar energy-driven hybrid desalination system: Concept description and simulation results, *Water Res.* 100 (2016) 7–19. <https://doi.org/10.1016/J.WATRES.2016.05.002>.
- [51] K. Thu, K.C. Ng, B.B. Saha, A. Chakraborty, S. Koyama, Operational strategy of adsorption desalination systems, *Int. J. Heat Mass Transf.* 52 (2009) 1811–1816. <https://doi.org/10.1016/J.IJHEATMASSTRANSFER.2008.10.012>.
- [52] J.W. WU, A Study of Silica Gel Adsorption Desalination System, 2012.
- [53] X. Wang, W. Zimmermann, K.C. Ng, A. Chakraborty, J.U. Keller, Investigation on the isotherm of silica gel+water systems, *J. Therm. Anal. Calorim.* 2004 762. 76 (2004) 659–669. <https://doi.org/10.1023/B:JTAN.0000028045.96239.7E>.
- [54] K.C. Ng, H.T. Chua, C.Y. Chung, C.H. Loke, T. Kashiwagi, A. Akisawa, B.B. Saha, Experimental investigation of the silica gel-water adsorption isotherm characteristics, *Appl. Therm. Eng.* 21 (2001) 1631–1642. [https://doi.org/10.1016/S1359-4311\(01\)00039-4](https://doi.org/10.1016/S1359-4311(01)00039-4).
- [55] A. Alkhudhiri, N. Darwish, N. Hilal, Membrane distillation: A comprehensive review, *Desalination.* 287 (2012) 2–18. <https://doi.org/10.1016/J.DESAL.2011.08.027>.
- [56] G. Naidu, L. Tijning, M.A.H. Johir, H. Shon, S. Vigneswaran, Hybrid membrane distillation: Resource, nutrient and energy recovery, *J. Memb. Sci.* 599 (2020) 117832. <https://doi.org/10.1016/j.memsci.2020.117832>.
- [57] R.K.M. Ferreira, H. Ramlow, C. Marangoni, R.A.F. Machado, A review on the manufacturing techniques of porous hydrophobic ceramic membranes applied to

- direct contact membrane distillation, *Adv. Appl. Ceram.* 120 (2021) 336–357.
<https://doi.org/10.1080/17436753.2021.1981749>.
- [58] N.A. Mohammad Ameen, S.S. Ibrahim, Q.F. Alsalhy, A. Figoli, Highly Saline Water Desalination Using Direct Contact Membrane Distillation (DCMD): Experimental and Simulation Study, *Water* 2020, Vol. 12, Page 1575. 12 (2020) 1575. <https://doi.org/10.3390/W12061575>.
- [59] A. Hagedorn, G. Fieg, D. Winter, J. Koschikowski, A. Grabowski, T. Mann, Membrane and spacer evaluation with respect to future module design in membrane distillation, *Desalination.* 413 (2017) 154–167.
<https://doi.org/10.1016/J.DESAL.2017.03.016>.
- [60] J. Zhang, S. Gray, J. De Li, Predicting the influence of operating conditions on DCMD flux and thermal efficiency for incompressible and compressible membrane systems, *Desalination.* 323 (2013) 142–149.
<https://doi.org/10.1016/J.DESAL.2013.04.002>.
- [61] L. Eykens, T. Reyns, K. De Sitter, C. Dotremont, L. Pinoy, B. Van der Bruggen, How to select a membrane distillation configuration? Process conditions and membrane influence unraveled, *Desalination.* 399 (2016) 105–115.
<https://doi.org/10.1016/J.DESAL.2016.08.019>.
- [62] M. Khayet, Membranes and theoretical modeling of membrane distillation: A review, *Adv. Colloid Interface Sci.* 164 (2011) 56–88.
<https://doi.org/10.1016/J.CIS.2010.09.005>.
- [63] A.M. Helal, Hybridization — a new trend in desalination, *New Pub Balaban.* 3 (2012) 120–135. <https://doi.org/10.5004/DWT.2009.263>.

- [64] A. Farsi, I. Dincer, Development and evaluation of an integrated MED/membrane desalination system, *Desalination*. 463 (2019) 55–68. <https://doi.org/10.1016/J.DESAL.2019.02.015>.
- [65] E. Ali, J. Orfi, H. AlAnsary, J.G. Lee, A. Alpatova, N. Ghaffour, Integration of multi effect evaporation and membrane distillation desalination processes for enhanced performance and recovery ratios, *Desalination*. 493 (2020) 114619. <https://doi.org/10.1016/j.desal.2020.114619>.
- [66] N.A.A. Qasem, S.M. Zubair, Performance evaluation of a novel hybrid humidification-dehumidification (air-heated) system with an adsorption desalination system, *Desalination*. 461 (2019) 37–54. <https://doi.org/10.1016/J.DESAL.2019.03.011>.
- [67] C. Ghenai, D. Kabakebji, I. Douba, A. Yassin, Performance analysis and optimization of hybrid multi-effect distillation adsorption desalination system powered with solar thermal energy for high salinity sea water, *Energy*. 215 (2021) 119212. <https://doi.org/10.1016/J.ENERGY.2020.119212>.
- [68] M.W. Shahzad, K.C. Ng, K. Thu, B.B. Saha, W.G. Chun, Multi effect desalination and adsorption desalination (MEDAD): A hybrid desalination method, *Appl. Therm. Eng.* 72 (2014) 289–297. <https://doi.org/10.1016/J.APPLTHERMALENG.2014.03.064>.
- [69] M. Essalhi, M. Khayet, Self-sustained webs of polyvinylidene fluoride electrospun nanofibers at different electrospinning times: 2. Theoretical analysis, polarization effects and thermal efficiency, *J. Memb. Sci.* 433 (2013) 180–191. <https://doi.org/10.1016/J.MEMSCI.2013.01.024>.

- [70] Z.W. Song, L.Y. Jiang, Optimization of morphology and performance of PVDF hollow fiber for direct contact membrane distillation using experimental design, *Chem. Eng. Sci.* 101 (2013) 130–143. <https://doi.org/10.1016/J.CES.2013.06.006>.
- [71] K.W. Lawson, D.R. Lloyd, Membrane distillation, *J. Memb. Sci.* 124 (1997) 1–25. [https://doi.org/10.1016/S0376-7388\(96\)00236-0](https://doi.org/10.1016/S0376-7388(96)00236-0).
- [72] T.C. Chen, C.D. Ho, H.M. Yeh, Theoretical modeling and experimental analysis of direct contact membrane distillation, *J. Memb. Sci.* 330 (2009) 279–287. <https://doi.org/10.1016/J.MEMSCI.2008.12.063>.
- [73] J. Phattaranawik, R. Jiraratananon, A.G. Fane, Effect of pore size distribution and air flux on mass transport in direct contact membrane distillation, *J. Memb. Sci.* 215 (2003) 75–85. [https://doi.org/10.1016/S0376-7388\(02\)00603-8](https://doi.org/10.1016/S0376-7388(02)00603-8).
- [74] M.S. Khayet, T. Matsuura, *Membrane distillation: principles and applications*, (2011) 477. https://books.google.com/books/about/Membrane_Distillation.html?id=VVY4AvSIgpwC (accessed April 2, 2022).
- [75] J.A. Sanmartino, M. Khayet, M.C. García-Payo, Desalination by Membrane Distillation, *Emerg. Membr. Technol. Sustain. Water Treat.* (2016) 77–109. <https://doi.org/10.1016/B978-0-444-63312-5.00004-8>.
- [76] A. Khalifa, H. Ahmad, M. Antar, T. Laoui, M. Khayet, Experimental and theoretical investigations on water desalination using direct contact membrane distillation, *Desalination*. 404 (2017) 22–34. <https://doi.org/10.1016/J.DESAL.2016.10.009>.
- [77] S.B. Iversen, V.K. Bhatia, K. Dam-Johansen, G. Jonsson, Characterization of microporous membranes for use in membrane contactors, *J. Memb. Sci.* 130 (1997)

- 205–217. [https://doi.org/10.1016/S0376-7388\(97\)00026-4](https://doi.org/10.1016/S0376-7388(97)00026-4).
- [78] I. Janajreh, D. Suwwan, R. Hashaikeh, Assessment of direct contact membrane distillation under different configurations, velocities and membrane properties, *Appl. Energy*. 185 (2017) 2058–2073. <https://doi.org/10.1016/J.APENERGY.2016.05.020>.
- [79] J. Phattaranawik, R. Jiraratananon, A.G. Fane, Heat transport and membrane distillation coefficients in direct contact membrane distillation, *J. Memb. Sci.* 212 (2003) 177–193. [https://doi.org/10.1016/S0376-7388\(02\)00498-2](https://doi.org/10.1016/S0376-7388(02)00498-2).
- [80] L. Francis, N. Ghaffour, A.A. Alsaadi, G.L. Amy, Material gap membrane distillation: A new design for water vapor flux enhancement, *J. Memb. Sci.* 448 (2013) 240–247. <https://doi.org/10.1016/J.MEMSCI.2013.08.013>.
- [81] L. Martínez-Díez, M.I. Vázquez-González, Temperature and concentration polarization in membrane distillation of aqueous salt solutions, *J. Memb. Sci.* 156 (1999) 265–273. [https://doi.org/10.1016/S0376-7388\(98\)00349-4](https://doi.org/10.1016/S0376-7388(98)00349-4).
- [82] S. Srisurichan, R. Jiraratananon, A.G. Fane, Mass transfer mechanisms and transport resistances in direct contact membrane distillation process, *J. Memb. Sci.* 277 (2006) 186–194. <https://doi.org/10.1016/J.MEMSCI.2005.10.028>.
- [83] Y.M. Manawi, M. Khraisheh, A.K. Fard, F. Benyahia, S. Adham, Effect of operational parameters on distillate flux in direct contact membrane distillation (DCMD): Comparison between experimental and model predicted performance, *Desalination*. 336 (2014) 110–120. <https://doi.org/10.1016/J.DESAL.2014.01.003>.
- [84] A.S. Alsaman, A.A. Askalany, K. Harby, M.S. Ahmed, Performance evaluation of a solar-driven adsorption desalination-cooling system, *Energy*. 128 (2017) 196–207.

- <https://doi.org/10.1016/J.ENERGY.2017.04.010>.
- [85] N.A.A. Qasem, R. Ben-Mansour, Adsorption breakthrough and cycling stability of carbon dioxide separation from CO₂/N₂/H₂O mixture under ambient conditions using 13X and Mg-MOF-74, *Appl. Energy*. 230 (2018) 1093–1107. <https://doi.org/10.1016/J.APENERGY.2018.09.069>.
- [86] N.A.A. Qasem, R. Ben-Mansour, M.A. Habib, An efficient CO₂ adsorptive storage using MOF-5 and MOF-177, *Appl. Energy*. 210 (2018) 317–326. <https://doi.org/10.1016/J.APENERGY.2017.11.011>.
- [87] N.A.A. Qasem, R. Ben-Mansour, Energy and productivity efficient vacuum pressure swing adsorption process to separate CO₂ from CO₂/N₂ mixture using Mg-MOF-74: A CFD simulation, *Appl. Energy*. 209 (2018) 190–202. <https://doi.org/10.1016/J.APENERGY.2017.10.098>.
- [88] R. Ben-Mansour, N.A.A. Qasem, M.A. Antar, Carbon dioxide adsorption separation from dry and humid CO₂/N₂ mixture, *Comput. Chem. Eng.* 117 (2018) 221–235. <https://doi.org/10.1016/J.COMPCHEMENG.2018.06.016>.
- [89] D.M. Ruthven, Fundamentals of Adsorption Equilibrium and Kinetics in Microporous Solids, *Mol. Sieves - Sci. Technol.* 7 (2006) 1–43. https://doi.org/10.1007/3829_007.
- [90] REFPROP | NIST, (n.d.). <https://www.nist.gov/srd/refprop> (accessed April 2, 2022).
- [91] H. El-Dessouky, H. Ettouney, Fundamentals of salt water desalination 2002, . (n.d.). https://books.google.com/books?hl=ar&lr=&id=eqssS2EMBU4C&oi=fnd&pg=PP1&ots=6hw7BHHZ0n&sig=bn6isl8HAXY_GbzDa098XszugYk (accessed April 2, 2022).

- [92] J. Gilron, L. Song, K.K. Sirkar, Design for Cascade of Crossflow Direct Contact Membrane Distillation, *Ind. Eng. Chem. Res.* 46 (2007) 2324–2334. <https://doi.org/10.1021/IE060999K>.
- [93] Y. Wang, N. Lior, Thermoeconomic analysis of a low-temperature multi-effect thermal desalination system coupled with an absorption heat pump, *Energy*. 36 (2011) 3878–3887. <https://doi.org/10.1016/J.ENERGY.2010.09.028>.
- [94] A.E. Kabeel, T.A. Elmaaty, E.M.S. El-Said, Economic analysis of a small-scale hybrid air HDH–SSF (humidification and dehumidification–water flashing evaporation) desalination plant, *Energy*. 53 (2013) 306–311. <https://doi.org/10.1016/J.ENERGY.2013.02.042>.
- [95] M.I. Zubair, F.A. Al-Sulaiman, M.A. Antar, S.A. Al-Dini, N.I. Ibrahim, Performance and cost assessment of solar driven humidification dehumidification desalination system, *Energy Convers. Manag.* 132 (2017) 28–39. <https://doi.org/10.1016/J.ENCONMAN.2016.10.005>.
- [96] M.A. Jamil, S.M. Zubair, On thermoeconomic analysis of a single-effect mechanical vapor compression desalination system, *Desalination*. 420 (2017) 292–307. <https://doi.org/10.1016/J.DESAL.2017.07.024>.
- [97] E.S. Ali, R.H. Mohammed, N.A.A. Qasem, S.M. Zubair, A. Askalany, Solar-powered ejector-based adsorption desalination system integrated with a humidification-dehumidification system, *Energy Convers. Manag.* 238 (2021) 114113. <https://doi.org/10.1016/J.ENCONMAN.2021.114113>.
- [98] F.A. Lattieff, M.A. Atiya, J.M. Mahdi, H.S. Majdi, P. Talebizadehsardari, W. Yaïci, Performance Analysis of a Solar Cooling System with Equal and Unequal

Adsorption/Desorption Operating Time, *Energies* 2021, Vol. 14, Page 6749. 14
(2021) 6749. <https://doi.org/10.3390/EN14206749>.

- [99] Q. Pan, H. Shan, Z. Tamainot-Telto, R. Wang, Heat Recovery for Adsorption Refrigeration System via Pinch Technology, *J. Therm. Sci.* 2022 312. 31 (2022) 379–389. <https://doi.org/10.1007/S11630-022-1535-7>.

|

Vitae

Name : Ayman Abdulhak Al-Azab |
Nationality : Yemeni |
Date of Birth : 3/3/1994 |
Email : aymen.alazab2017@gmail.com |
Address : Street no. 60 , Sana'a, Yemen |

Academic Background | **M. Sc. (Chemical Engineering), May, 2022**
King Fahd University of Petroleum and minerals
Dhahran, Saudi Arabia
B. Sc. (Chemical Engineering), December, 2018
King Fahd University of Petroleum and minerals
Dhahran, Saudi Arabia |

Publications:

1. “Geospatial Analysis of Desalination in Saudi Arabia: An Interactive Model for Socio-economic Evaluations and Decision Support”, Journal paper under writing.

2. “A solar-driven integrated adsorption desalination and direct contact membrane distillation system for a net-zero fuel consumption”, Journal paper under writing.
3. “Geospatial Analysis of Desalination in Saudi Arabia: An Interactive Model for Socio-economic Evaluations and Decision Support”, Journal paper under writing.
4. “An Innovative Hybridization of Adsorption Desalination with Direct Contact Membrane Distillation”, Patent to be submitted.