

Improvement of Desalination Productivity with Intermittent Impermeable Segments

BY

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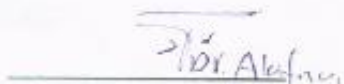
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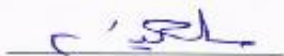
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This thesis is dedicated to my family, whose unwavering support and encouragement have been invaluable throughout this journey.

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LIST OF ABBREVIATIONS

CDFP	Concentration Dependent Fluid Property
CFD	Computational Fluid Dynamic
CP	Concentration Polarization
CPM	CP Modulus
FEM	Finite Element Modelling
MED	Multi Efficient Desalination
MSF	Multi-Stage Flashing
NF	Nano Filtration
PARDISO	Parallel Direct Solver
RO	Reverse Osmosis
TMC	Trimesoyl Chloride
TPMS	Triply Periodic Minimal Surface
TVD	Total Viscous Dissipation
UF	Ultra Filtration

LIST OF NOMENCLATURE

a	Smallest element size	(m)
A	Water permeability coefficient	(m/s/Pa)
A_D, B_D, C_D, D_D	Diffusivity parameters	
A_μ, B_μ	Viscosity parameters	
A_ρ, B_ρ	Density parameters	
B_Π, C_Π, D_Π	Osmotic pressure parameters	
c	Concentration of salt	(wt.%)
c_f	Bulk feed concentration	(wt.%)
c_p	Concentration of permeate	(wt.%)
c_w	Concentration along wall	(wt.%)
CDFP	Concentration dependence of fluid properties	
CPM	Concentration polarization modulus	(-)
$\mathfrak{D}$	Binary diffusivity coefficient of salt in water	(m ² /s)
ΔP	Axial pressure drop	(Pa)
ΔP_T	Transmembrane pressure difference	(Pa)
$\Delta \Pi$	Osmotic pressure difference	(Pa)
e	Unit cost of electricity	(\$/J)
E	Productivity rate	(\$/s)
ER	Element ratio, as defined in COMSOL	(-)
H	Channel height	(m)
J_w	Volumetric flux of permeate	(m/s)
$\bar{J}_w$	Permeate flux averaged over length	(m/s)
L	Length of active segment	(m)
L_{sum}	Total channel length	(m)

$\dot{m}_p$	Mass flowrate of permeate	(kg/s)
n	Number of elements per dimension	(-)
$\mathbf{n}$	Unit vector normal to boundary	
N	Total number of elements in channel	(-)
P	Gauge pressure	(Pa)
P_0	Outlet gauge pressure	(Pa)
P_p	Gauge pressure on permeate side	(Pa)
$\dot{Q}_f$	Volumetric flowrate of feed	(m ³ /s)
$\dot{Q}_p$	Permeate volumetric flowrate	(m ³ /s)
R_{es}	Average mesh resolution in x-direction	(m)
R_i	Intrinsic rejection of membrane	(-)
Re	Channel Reynolds number	(-)
$\mathbf{u}$	Velocity vector of channel flow	(m/s)
U	Velocity magnitude	(m/s)
$\hat{U}_f$	Bulk inlet velocity of feed	(m/s)
w	Permeate value	(\$/kg)
x	x-coordinate	(m)
y	y-coordinate	(m)

Greek letters

β^*	Optimum inactive segment length	(m)
β	Inactive segment length	(m)
μ	Dynamic viscosity	(Pa.s)
ρ	Fluid density	(kg/m ³)

Indices

$*$	Optimum value
D	Diffusivity
f	Feed
i	Intrinsic
0	Outlet
p	Permeate
sum	Sum along channel
T	Transmembrane
w	Wall
$w1, w2$	Wall re. segments 1, or 2
μ	Viscosity
Π	Osmotic
ρ	Density

ABSTRACT

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This study explores a method to enhance productivity in membrane devices by incorporating intermittent impermeable segments within a spacer-free geometry for RO, NF, and UF applications. The model features a hybrid channel structure where a central impermeable segment of length β is seamlessly integrated between two standard membrane segments, each of length L . The productivity parameter, defined herein as economic flow (E , \$/s), balances hydraulic losses with permeate production. Key parameters include segment lengths (L and β), inlet flow speed (Re), and permeate value (w , \$/kg). Results show impact of single impermeable segment use on concentration polarization and permeate flux. Optimal segment lengths (β^*) are dependent upon Re and w . Such lengths increase monotonically with w , reflecting the growing importance of mitigating polarization as permeate value rises. For a channel length of 300 mm, at $Re = 4.8$, the optimal segment length peaked at $\beta^* = 160$ mm for $w > 2 \times 10^{-4}$ \$/kg. Low Re values within the laminar range are necessary for effective use of impermeable segments, while higher values are not justified ($\beta^* = 0$). Compared to a standard staggered-cylinder filament arrangement, our design achieves a 99.9% reduction in axial pressure drop for a 300-mm commercial channel at 0.5 m/s cross-flow, yielding an estimated \$40 in annual savings. This configuration offers an alternative approach for advancing desalination towards greater efficiency and economic sustainability.

ملخص الرسالة

الاسم الكامل: طلال بن ضيف الله بن حمدان العامري الشمراني

عنوان الرسالة: تحسين تحلية المياه بالترشيح الأسموزي المتقطع

التخصص: الهندسة الكيميائية

تاريخ الدرجة العلمية: ذو القعدة ١٤٤٦

في هذه الرسالة يتم مناقشة تحسين تحلية المياه بالترشيح الأسموزي المتقطع وإثبات الجدوى الاقتصادية من ذلك بالمقارنه مع المرشحات الأسموزية التي تعتمد على وجود الفواصل المساحية ضمن تصميمها. تم إجراء الدراسة عن طريق المحاكاة الحاسوبية باستخدام برنامج الفيزياء المتعدده في نموذج ثنائي الابعاد. تم إعداد التصميم بدون اي فواصل مساحية بين مكونات المرشح الأسموزي. حيث يتميز التصميم بوجود وحدة الترشيح الأسموزي على طرفي النموذج عند مدخل و مخرج اللقيم (الماء المالح) , بينما يوجد في منتصف الوحدة جزء لا يسمح بترشيح المياه المالحة. يعطي هذا التصميم بعض المرونه لإعادة مزج الاملاح في اللقيم و خفض إستقطاب التركيز الملحي على سطح المرشح الأسموزي و زياده في كميته المياه المترشحه. يعتمد تحديد الطول المثالي لجزء الغير منفذ للماء . يتميز هذا التصميم المقترح على معيارين رئيسيين هما القيمه الحقيقيه للماء المنقى و قيمة عدد رينولد للقيم بإمكانية إنقاص مقدار إنخفاض الضغط المحوري (على امتداد المرشح الاسموزي بين المدخل و المخرج للمياه المالحة) في المرشحات الإسموزيه التجاريه (طول ٣٠٠ مل و تدفق مقداره ٠,٥ متر/ثانيه) إلى ٩٩,٩% عند مقارنته بالمرشحات الأسموزية التي تحتوي ضمن تصميمها على الفواصل المساحية الاسطوانية المتدرجه. مما يساهم في خفض التكلفة بما يقارب ال ٤٠ دولاراً في السنة ورفع الاستدامة الاقتصادية للترشيح الأسموزي

CHAPTER 1

INTRODUCTION

Water scarcity poses a critical challenge in Saudi Arabia since it has an arid climate. As the demand for freshwater continues to escalate due to population growth and industrial expansion, the need for innovative solutions to boost freshwater resources becomes increasingly urgent. As the kingdom has over depleted its underground water reservoir over the years and is lacking natural water resources. In addition, the country faces irregular and minimal rainfall with an average less than 100 mm per year (Hasanean & Almazroui, 2015). For that, the government has implemented various initiatives to ensure water security. Most important among these initiatives is investing in seawater desalination plants (The U.S.-Saudi Business Council, 2021). Nowadays, Saudi Arabia is a global leader in desalination, with 43 desalination plants along its coastlines with a total daily production of 13.3 million m³ (Saudi Arabia Leads Global Desalinated Water Production Surge, Arab News)



Figure 1: Largest Sea Water Desalination Plant in Ras Al-Khair, Saudi Arabia

There are two main technologies for sea water desalination in Saudi Arabia, Multi-Stage Flashing (MSF) and Reverse Osmosis (RO). MSF is heavily dependent on fossil fuel, requires a lot of maintenance and is not ecofriendly. On the other hand, RO desalination technology is in the sweet spot of efficiency, cost, and operation simplicity. Even though current RO membranes are highly efficient, there is still room for improvement to RO systems to increase efficiency and optimize water productivity in conjunction with increasing production profitability margin.

1.1 Methods of Desalination

The concept of desalination is not new; ancient civilizations like the Greeks and Romans used rudimentary techniques such as simple distillation to for wine production. However, modern desalination technologies have evolved significantly, driven by advancements in science and engineering. The primary methods of desalination in contemporary practice include thermal processes, membrane processes, and hybrid systems. Desalination techniques utilization varies based on location, available energy and quantity of produced (Ghaffour et al., 2013).

1.2 Thermal Processes

Among the various desalination methods, thermal desalination stands out for its simplicity and effectiveness in treating large volumes of water. Thermal desalination processes are based on the principle of phase change, where saline water is heated to produce vapor, which is then condensed to yield fresh water, leaving the salt and other impurities behind (Al-Karaghoul & Kazmerski, 2013a).

The energy required for phase changes can come from various sources, including fossil fuels, solar energy, and waste heat, making thermal desalination adaptable to different contexts. There are several types of thermal desalination processes, each with its specific design and operational characteristics. The most common ones include Multi-Stage Flash (MSF) and Multi-Effect Distillation (MED) (Al-Karaghoul & Kazmerski, 2013).

1.2.1 Multi-Stage Flash Desalination (MSF)

MSF is one of the most widely used thermal desalination methods. It operates by heating seawater to high temperatures and then rapidly reducing the pressure in a series of stages or chambers. This causes the water to "flash" into steam in each stage. The steam is condensed into fresh water, and the process repeats across multiple stages, each at progressively lower pressures and temperatures. MSF is highly efficient for large-scale desalination, capable of producing millions of gallons of fresh water per day.

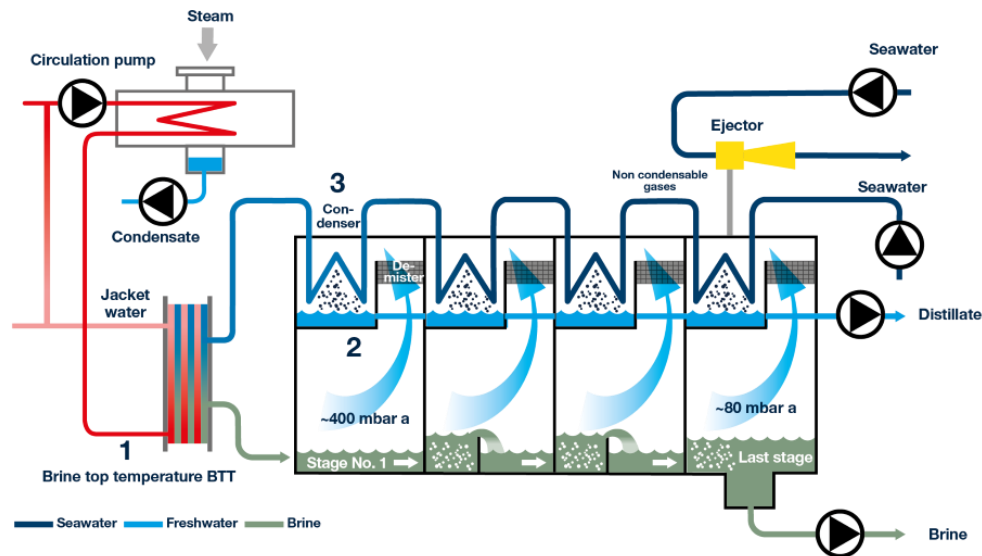


Figure 2: MSF stages (Multistage Flash Evaporator (MSF) - Onboard Desalination of Seawater).

1.2.2 Multi-Effect Desalination (MED)

MED involves a series of evaporator effects, each operating at a lower pressure than the previous one. Seawater is preheated and introduced into the first effect, where it evaporates. The vapor from the first effect serves as the heating source for the second effect, and so on, creating a cascading effect of evaporation and condensation. MED is known for its energy efficiency, as it reuses the heat energy multiple times, making it a popular choice for medium to large-scale desalination plants (Al-Karaghoul & Kazmerski, 2013b).

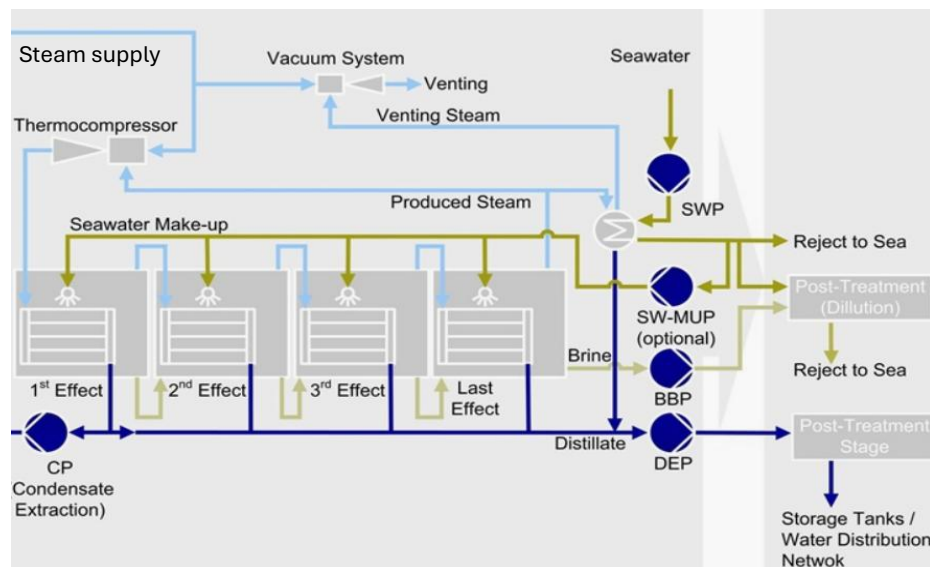


Figure 3: MED stages (Multi Effect Distillation MED, Sulzer)

1.3 Membrane Desalination

Among various desalination technologies, membrane-based processes rely on a filtration medium where permeability and porosity of this medium plays the biggest role in desalinating water. There are many types of membrane desalination such as micro, ultra nano filtration and reverse osmosis membrane. Particularly reverse osmosis membrane (RO) has gained prominence due to its relatively low energy consumption and high produced water quality. RO is the most energy-efficient desalination technology in terms

of overall energy consumption, especially for seawater desalination. As shown in Table 1, RO compared to MSF and MED rely only on electrical energy and that gives it a significant advantage in regions where electricity is more readily available (Elimelech & Phillip, 2011; Semiat, 2008)

Table 1: Comparative analysis of energy consumption for MSF, MED & RO (Elimelech & Phillip, 2011; Imbrogno & Belfort, 2016).

Desalination Technology	Energy Type	Energy Consumption (kWh/m³)
Multi-Stage Flash (MSF)	Thermal + Electrical	50-80 (thermal) + 3-5 (electrical)
Multi-Effect Distillation (MED)	Thermal + Electrical	40-70 (thermal) + 1.5-2.5 (electrical)
Reverse Osmosis (RO)	Electrical	2.5-4.5 (seawater RO)

Although, RO has achieved quite an impressive efficiency in terms of salt rejection and recovery rate, there is still room to optimize its functionality and increase its productivity. Many researchers are trying to achieve the optimization goal by focusing on enhancing RO spacer geometry, optimizing the polyamide chemical structure, or via modification to RO auxiliary systems. Yet not many papers have discussed optimizing membrane efficiency through manipulating the fluid dynamic behavior within the membrane structure. Specifically in RO, there are two governing phenomena that negatively affect its desalination performance. First is pressure drop across the membrane module due to friction loss with membrane surface and water permeation. The second is the buildup of concentration polarization (CP) due to salt accumulating along the membrane surface. For that, there are some optimizations that should be done to enhance water flux and reduce the effect of these two phenomena.

1.4 Thesis Objectives

Conventional reverse osmosis (RO) membranes incorporating feed spacers face notable performance limitations, as they promote hydraulic losses (Song et al., 2002) and localized concentration polarization (CP) at membrane contact points (Song & Ma, 2005). This phenomenon not only promotes scaling and membrane fouling but also contributes to elevated hydraulic losses and reduced water permeation. Collectively, these issues hinder overall system productivity and increase energy demands, thereby compromising economic and operational sustainability.

In response to these challenges, the primary objective of this research is to investigate a novel method for improving the productivity of RO membranes through the strategic integration of intermittent impermeable segments within a spacer-free channel design. The study introduces a hybrid configuration in which non-permeable segments are embedded between conventional, selectively permeable membrane segments. These impermeable segments function as concentration-balancing zones, enabling solute redistribution and reducing CP before feedwater enters a downstream permeable segment. This process essentially "resets" the salt concentration conditions closer to the inlet level, mitigating polarization build-up and enhancing overall system performance.

To achieve this objective, finite element modeling (FEM) will be employed through COMSOL Multiphysics to simulate the fluid dynamics and mass transport processes occurring within the membrane channel. Additionally, to demonstrate the viability of the novel approach, model performance will be assessed using a techno-economic analysis to quantify the productivity rate E (\$/s), which captures the trade-off between permeate production and the energy cost of pressure losses, thus offering a more realistic assessment

of system efficiency. This enables a determination of whether the use of impermeable segments is both technically and economically justifiable.

CHAPTER 2

LITERATURE REVIEW

There are two phenomena that strongly govern RO process efficiency first is pressure drop, which is a resulted from fluid friction against the membrane surface, and second is concentration polarization (CP), an effect arising from salt accumulation on the membrane surface. To mitigate these adverse impacts, one promising approach is utilizing spacer geometry. For example, adopting a Triply Periodic Minimal Surface (TPMS) Structure, as demonstrated by Toh et al. in Fig. 4, can enhance permeate flux by up to 15.5% and reduce pressure drop by 5% (Toh et al., 2020) .

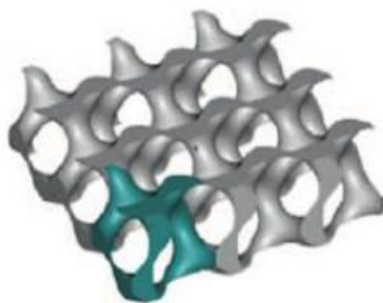


Figure 4: Triply Periodic Minimal Surface (TPMS) Spacer Structure (Toh et al., 2020).

Despite the considerable benefits offered by spacer geometry modification, spacers take up space within the membrane module, slightly reducing the effective membrane surface area available for water filtration. Also, contribute to hydraulic losses (Song et al., 2002) and trap particles, biofilms, and other contaminants, near the contact points with the membrane (Song & Ma, 2005) leading to biofouling, scaling, and particulate fouling. This can reduce membrane efficiency and require more frequent cleaning (Bucs et al., 2014).

RO performance is significantly dependent on membrane selectivity and water diffusion capabilities (Table 2). Therefore, the chemical composition of the active polyamide layer is crucial. As elucidated in a review by Kim et al., the structural variance in amine components can yield polyamide layers with different water flux and salt rejection characteristics (C. K. Kim et al., 2000). For instance, membranes synthesized through the interfacial reaction between Trimesoyl Chloride (TMC) and Poly(m-AminoStyrene) (PmAS) with various diamine monomers have demonstrated water flux permeances ranging from 0.81 to 0.9 L m⁻² h⁻¹ bar⁻¹, and salt rejection percentages between 79% and 95%. However, in polyamide chemical structure there should be a balance in cross-linking density as if it is so high salt rejection will be high, yet it will compromise water permeability. On the other hand, if cross-linking density is low water permeability will increase. Yet, salt rejection will not be optimum (Yoshioka et al., 2018).

Table 2: List of some polyamide types with water permeability and salt rejection percentage.

Polyamide Type	Monomers / Modifications	Water Permeability (L·m⁻²·h⁻¹·bar⁻¹)	Salt Rejection (%)	Reference
Conventional Aromatic Polyamide	m-Phenylenediamine (MPD) + Trimesoyl Chloride (TMC)	1.0 – 2.5	98 – 99.5	(Beydaghi et al., 2014)
Hydrophilic Modified Polyamide	MPD + TMC + Hydrophilic additives (e.g., PEG)	2.5 – 4.0	97 – 99	(Jiang et al., 2023)
Crosslinked Aromatic Polyamide	MPD + TMC + Crosslinkers (e.g., isophthaloyl chloride)	0.8 – 2.0	95-99	(Yu et al., 2021; C. K. Kim et al., 2000)
Aliphatic Polyamide	Aliphatic diamines (e.g., HMD) + Dicarboxylic acids	2.0 – 5.0	90 – 98	(Aristizábal et al., 2023)

In terms of enhancing RO auxiliary systems, the pressure exchanger device represents a significant technological advancement for reducing the energy required to pump feedwater into the membrane inlet. This device, integrated downstream the reject water stream, can capture energy from this stream and redirect it to the membrane's inlet. According to Guirguis, the device can achieve energy efficiencies ranging from 95% and 98% (Guirguis, 2011).

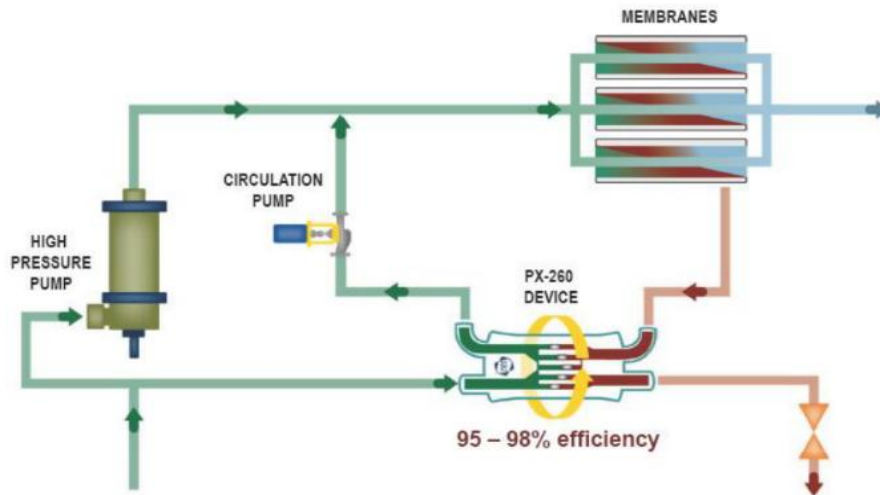


Figure 5: Pressure exchanger device (Multi Effect Distillation MED, Sulzer).

Currently, traditional avenues for membrane enhancement are approaching their limits. Spacer geometry has intrinsic constraints in terms of shape variety, thus capping the potential to have an efficiency gain. Chemically, the formation of densely packed, sterically hindered functional groups for improved salt rejection is challenging. Moreover, advancements in RO auxiliary systems have reached a plateau in terms of efficiency improvements.

In the realm of our study, there were few papers in literature that have acquired similar design to this paper proposed design. The first was back in 1972 by Shaw et al. (Shaw et al., 1972) where they comprised a theoretical model of a hybrid membrane cell with selectively permeable (salt-rejecting), and fully permeable membrane sections designed to disperse the boundary layer. Their findings indicated increased permeate production through the reduction of CP. The convective draw-off of the boundary layer in the fully permeable sections significantly reduced CP. However, such an implementation faces potential issues – notably the tendency for fouling within the pores of the fully permeable membrane sections, which may gradually render them impermeable over time. Additionally, extracting the salt-free permeate would require a more complex setup with separate permeate streams, further increasing costs. This early study did not include optimization or an economic analysis.

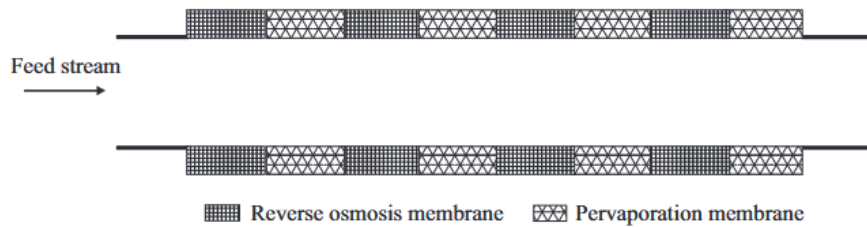


Figure 6: Hybrid membrane design from the study of (Miranda & Campos, 2007).

A similar study by Miranda and Campos in 2007 (Miranda & Campos, 2007) used a numerical finite element model to examine a hybrid membrane cell with semi-permeable (VOC-rejecting) and fully permeable membrane sections as shown in Fig. 6. They also reported increased productivity through CP reduction. While their findings mirrored those of Shaw et. al.(Shaw et al., 1972), they extended the concept by using multiple hybrid sections in series and quantifying the impact of the number of repeat segments on CP. However, like the earlier study, it lacked economic analysis.

CHAPTER 3

MODEL SETUP

3.1 Model Fundamentals

The aim is to simulate accurately the hydrodynamic and mass transfer processes specifically within the feed channel using Galerkin Finite Element Modelling (FEM) via COMSOL Multiphysics Simulation software.

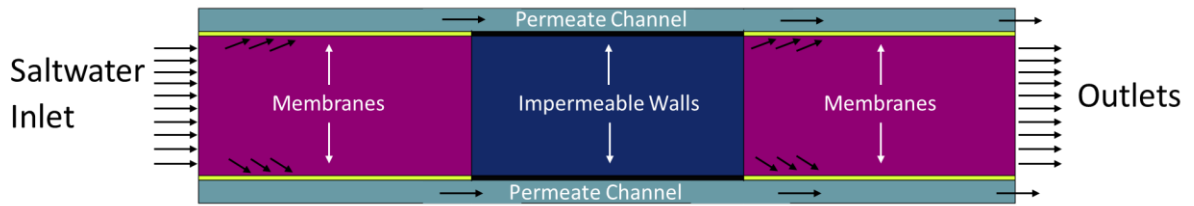


Figure 7: Example channel section featuring active segments (purple) and corresponding selectively permeable membranes (yellow), interceded by an inactive concentration balancing segment (blue). Teal-colored regions denote the permeate collection layers. Numbers represent the boundary IDs used within the model.

The computational domain features a hybrid channel structure, as illustrated in Fig. 7. This structure includes two active segments with selectively permeable membranes, interceded by a single non permeable segment. The active segments, highlighted in purple, allow water permeation through the membranes while completely rejecting salt. Also, no slip conditions are enforced at the top and bottom boundary of the feed side membrane sections to ensure accurate representation of the flow dynamics. Saline feed water enters the left boundary where developed flow is assumed for realistic inlet conditions. The impermeable segment, shown in blue, serves to redistribute solutes and reset the flow conditions for the subsequent active permeable segment, thereby reducing concentration polarization and enhancing overall permeate production. The membranes in this model are designed to reject all salt and pass only water, whereby the intrinsic rejection, $R_i (\equiv 1 - c_p/c_w)$, and water permeability, A , are invariant along the membrane. The teal-colored regions in Fig.

7 denote the permeate collection layers, whose flow is not modelled in this study, thereby simplifying the computational complexity while focusing on the primary objectives at the feed channel.

3.1.1 Fluid Mechanics Governing Equation

The Navier-Stokes equations were employed to model the fluid dynamics within the hybrid channel, operating under steady-state conditions, as given by Eq 1. The 2-D differential operator, $\nabla = (\partial/\partial x, \partial/\partial y)$ is based on x and y being the coordinates in the cross-flow direction and normal to the membrane surface respectively, and $\mathbf{u}$ is the velocity vector of the channel fluid. A two-dimensional setup was adopted for the flow given that channel widths are customarily much deeper than the height. Furthermore, the shallow channel approximation was invoked, where the channel length is two orders of magnitude longer than its height, in which the gravity term was deemed unnecessary and omitted. The primary outputs of this model are the two-dimensional fields of velocity and pressure. Alongside these, the continuity equation was solved by Eq. 2, assuming liquid water to be incompressible, though slight variations in density were incorporated to account for the weak dependence on salt concentration. Additionally, the viscosity was modeled as a function of salt concentration.

$$\rho(\mathbf{u} \cdot \nabla \mathbf{u}) = -\nabla P + \mu \nabla^2 \mathbf{u} \quad (1)$$

$$\rho \nabla \cdot \mathbf{u} = 0$$

Consistent stabilization techniques were applied to the numerical model to ensure accuracy and stability. A linear discretization approach, denoted as P1+P1 in COMSOL, was used for both velocity and pressure variables. This method provided a robust framework for resolving the flow dynamics and capturing the essential features of the system.

3.1.2 Mass Transfer Governing Equation

Mass transport for a single salt within the channel was modeled using the steady-state convection-diffusion equation to determine the concentration field, as described by Eq. 3. This model assumes a single, dilute salt species, and that diffusion is isotropic. Modelling based on a single salt can be justified since seawater is commonly used during reverse osmosis and chiefly consists of NaCl (Li et al., 2017) . The primary output is a two-dimensional concentration field, which is coupled to the Navier-Stokes equations via the velocity field in the convective term. Additionally, the diffusion coefficient was modeled as a function of concentration. To enhance numerical stability, consistent stabilization techniques were applied to the mass transport equations. A quadratic discretization method was used for the concentration variable, ensuring higher accuracy in capturing concentration gradients. For the finite element model, specific solver settings were implemented to ensure convergence. A fully coupled solver approach was used, employing the direct PARDISO solver in COMSOL. These settings facilitated robust and efficient resolution of the coupled fluid dynamics and mass transport equations.

$$\mathcal{D}\nabla^2 c - \mathbf{u} \cdot \nabla c = 0 \quad (3)$$

The permeate flow through the membrane was computed using the solution-diffusion model - described by Wijmans and Baker (Wijmans & Baker, 1995) – as shown in Eq. 4. It should be noted that this is applicable to RO but adaptable for NF or UF with the use of alternative submodules.

In the solution-diffusion model, A represents the water permeability coefficient of the membranes, and ΔP_T is the transmembrane pressure difference ($P(x) - P_p$). In this, ΔP_T is equivalent to the gauge pressure owing to the permeate layer being assumed to be at atmospheric pressure (i.e. $P_p = 0$ barg) - this assumption is reasonable, especially for typical RO operations $P \gg P_p$. The osmotic pressure difference was modeled using Eq. 5 (Lyster & Cohen, 2007), with coefficients provided in Table 4. This relation is applicable for salinities ranging from 0 to 3.5 wt.% NaCl, ensuring accuracy within the specified concentration range.

$$J_w = A(\Delta P_T - \Delta \Pi) \quad (4)$$

$$\Delta \Pi = B_{\Pi} \cdot c + C_{\Pi} \cdot c^2 + D_{\Pi} \cdot c^3 \quad (5)$$

3.1.3 Measuring Productivity

The "Productivity" term, E , introduced in this work, serves as a holistic economic production metric for the system, as defined in Eq. 6. This metric provides a straightforward evaluation by considering both the dissipation losses and the production rate of the permeate. Dissipation losses are characterized by the axial pressure drop along the channel, referred to as the "total viscous dissipation" (TVD), which is calculated as $\Delta P \dot{Q}_f$. The contribution of these losses to E is governed by the unit cost of electricity, e

(\$/J). Conversely, the production rate of the permeate, $\dot{m}_p$, contributes positively to E , and is governed by the intrinsic permeate value, w (\$/kg). The resulting productivity term E has units of (\$/s), representing a net "economic flow", balancing energy costs with permeate production.

$$E = \dot{m}_p \cdot w - \Delta P \dot{Q}_f \cdot e \quad (6)$$

3.2 Model Parameters

For the Navier-Stokes equations, specific boundary conditions are summarized in Table 2. At the inlet, a parabolic velocity profile is assumed, reflecting fully developed flow. This assumption facilitates direct comparison with classical analytical solutions and is justified by the entrance length, calculated to be less than approximately 25 mm, ensuring the flow is fully developed upon entry into the module. The outlet is maintained under a back pressure of 15 or 60 barg. No-slip conditions are enforced along the entire segments. Additionally, the permeate velocity normal to the membranes is determined using Equation 4, based on the solution-diffusion model.

For the mass transport equations, the feed concentration is prescribed at the inlet, establishing the initial condition for salt concentration. Zero diffusive flux is assumed at the outlet owing to convection dominating the transport of salt at this boundary. Furthermore, salt transport is assumed to be negligible along membranes surfaces, indicating impermeable conditions for these surfaces.

Table 3: Boundary conditions.

Boundary ID	Navier-Stokes		Mass transport	
1	$2\hat{U}_f y(H - y)/(H/2)^2$	normal flow	c_f	inlet mass fraction
2	$P = P_0$	back pressure	$\mathbf{n} \cdot \nabla c = 0$	zero diffusive flux
3	$J_{w1}(x)$	SD model, no slip	$\mathbf{n} \cdot (-\mathfrak{D}\nabla c + c\mathbf{u}) = 0$	zero salt transport
4	$\mathbf{u} = 0$	impermeable, no slip	$\mathbf{n} \cdot (-\mathfrak{D}\nabla c + c\mathbf{u}) = 0$	zero salt transport
5	$J_{w2}(x)$	SD model, no slip	$\mathbf{n} \cdot (-\mathfrak{D}\nabla c + c\mathbf{u}) = 0$	zero salt transport

The model parameters for both equipment and flow dynamics are summarized in Tables 3 and 4, respectively. The geometry of the system includes active segment lengths (L) ranging from 10 mm to 150 mm and inactive flow conditioning segment lengths (β) from 0.1 mm to 400 mm. The channel height is maintained at 0.5 mm, reflecting dimensions typical of microfluidic systems. The hydraulic permeability of the membranes is set at 9.8×10^{-12} m/s/Pa ($3.53 \text{ L m}^{-2} \text{ h}^{-1} \text{ bar}^{-1}$), a value consistent with findings from previous studies (S. Kim & Hoek, 2005; Song et al., 2002; Usman et al., 2024).

For the flow parameters, the bulk flow speed at the inlet varies between 0.003 m/s (creeping flow) and 0.5 m/s. The inlet feed consists of a brackish solution with a NaCl concentration of 0.417 wt.% (72 mM). The fluid properties, including density, viscosity, and diffusivity, are modelled as functions of salt concentration using Eqs. 7-9. The intrinsic value of the produced permeate (w) is examined over a range from 0 to 0.5 \$/kg to assess economic viability. Additionally, the energy unit cost (e) is based on typical market rates (Consumer Price Index Average Price Data, US Bureau of Labor Statistics, 2023).

Empirical models were employed to accurately represent the behaviour of sodium chloride (NaCl) solutions in water (Lyster & Cohen, 2007). Such concentration dependent fluid property (CDFP) models provide essential data for the simulation of osmotic pressure and other fluid properties. The sub-model for osmotic pressure, as described by Eq. 5, is applicable for NaCl concentrations ranging from 0 to 3.5 wt.%. This model is essential for predicting the osmotic effects within the membrane system under varying salinity conditions. Additionally, the fluid property sub-model for diffusivity, described by Eq. 9, is specifically applicable for NaCl concentrations between 0 and 0.77 wt.%. This range ensures accurate modelling of the diffusion process at lower salinities. Other fluid properties, including density and viscosity, as described by Eqs. 7 and 8, are applicable for a broader concentration range from 0 to 3.5 wt.% NaCl. These comprehensive sub-models enable precise simulations of the fluid dynamics and mass transport processes within the membrane system.

$$\rho = A_{\rho} + B_{\rho} \cdot c \quad (7)$$

$$\mu = A_{\mu} + B_{\mu} \cdot c \quad (8)$$

$$\mathfrak{D} = \frac{A_D + B_D \cdot c}{1 + C_D \cdot c + D_D \cdot c^2} \quad (9)$$

Table 4: Model parameters.

Parameter	Symbol	Value	Unit
Active segment length	L	10 – 150	mm
Inactive segment length	β	0.1 – 400	mm
Channel height	H	0.5	mm
Membrane permeability	A	9.8×10^{-12}	m/s/Pa
Bulk flow velocity	$\hat{U}_f$	0.003 – 0.5	m/s
Feed salt concentration	c_f	0.417	wt. %
Back pressure	P_o	60 or 15	barg
Density	ρ	$\rho = A_\rho + B_\rho \cdot c$	kg/m ³
Viscosity	μ	$\mu = A_\mu + B_\mu \cdot c$	Pa.s
Diffusivity	$\mathfrak{D}$	$\mathfrak{D} = \frac{A_D + B_D \cdot c}{1 + C_D \cdot c + D_D \cdot c^2}$	m ² /s
Permeate value	w	0 – 1	\$/kg
Energy unit cost	e	3.6×10^{-8}	\$/J

Table 5: Fluid coefficients.

Parameter	Symbol	Value	Unit
Osmotic pressure			
	B_Π	0.04572	m ³ .bar/mol
	C_Π	-1.797×10^{-6}	m ⁶ .bar/mol ²
	D_Π	4.631×10^{-9}	m ⁹ .bar/mol ³
Density			
	A_ρ	999	kg/m ³
	B_ρ	0.04099	Kg/mol

<i>Dynamic viscosity</i>			
	A_μ	0.001004	Pa.s
	B_μ	8.377×10^{-8}	$\text{m}^2.\text{kg}/(\text{mol}.\text{s})$
<i>Diffusivity</i>			
	A_D	1.418×10^{-9}	m^2/s
	B_D	1.159×10^{-10}	$\text{m}^5/(\text{mol}.\text{s})$
	C_D	0.085	m^3/mol
	D_D	1.514×10^{-5}	m^6/mol^2

As the simulated fluid experiences an inlet pressure of 60 or 15 barg far less than the water bulk modulus ($k = 2.25$ GPa) (Bulk Modulus and Fluid Elasticities) and has low flow velocity within the laminar flow range ($Re = 1.33$ to 240), it is reasonable to consider the fluid incompressible. Thus, all governing equations (Navier–Stokes equation, mass transfer equation) and sub-model equations (density, osmotic pressure, and diffusion) are presented in a time-independent format, which is consistent with a steady-state process.

3.3 Mesh Independence

To ensure the numerical model's accuracy and reliability, it is important to verify that the laminar flow and dilute species mass transfer equations yield results consistent with experimental data reported in the literature for similar setups. Our system is modelled using FEM, which requires a well-defined mesh to solve the governing equations accurately.

Figure 8 illustrates an example of a structured rectangular mesh for an active segment length of $L = 10$ mm and an inactive length of $\beta = 20$ mm. The setup shown in Fig. 8 represents the typical length scales used in this study and comprises a total of $N = 12,000$

elements. The example mesh includes $n = 100$ elements in the y -direction and $n = 30$ elements in the x -direction for the active segments, while the inactive segments contain $n = 60$ elements in the x -direction. The element size is distributed non-linearly, with higher element density near the membranes, walls and segment boundaries, following a geometric progression implemented in COMSOL. Table 5 summarizes the mesh characteristics used throughout the study, highlighting the element sizes, particularly in the y dimension where the resolution reaches 230 nm near the membrane surface. The number of elements in the x -direction was scaled with the segment length based on L/R_{es} and β/R_{es} , where ($R_{es} = 0.333$ mm) represents the average element size. The resulting ratio was rounded to the nearest integer, ensuring consistent mesh resolution while studying the effects of varying segment lengths. This approach ensures detailed resolution in critical areas while maintaining computational efficiency. Notably, this structured mesh produces results consistent with those obtained from an unstructured triangular mesh, but with improved computational speed, making it the preferred choice for our simulations.

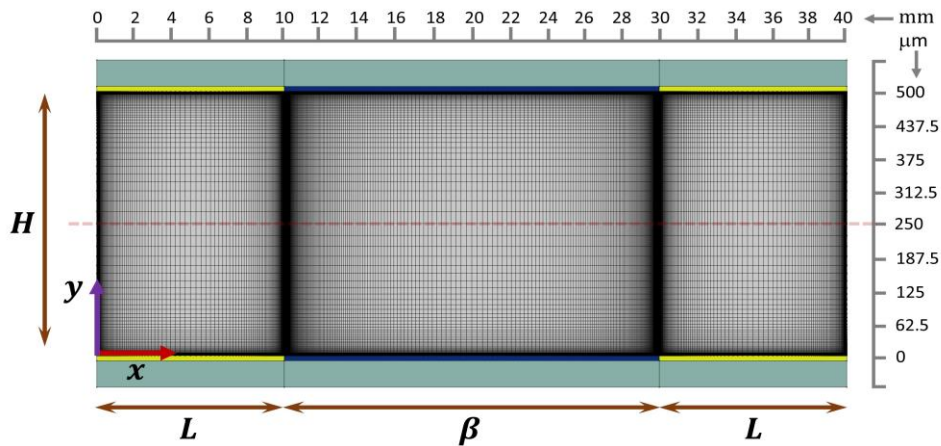


Figure 8: Representation of rectangular mesh.

Table 6: Mesh characteristics.

<i>Dimension</i>	<i>Element ratio (ER)</i>	<i>No. of elements (n)</i>	<i>Smallest element size (a)</i>
H	100	100	0.23 μm
L	30	L/R_{es}	37 μm
β	30	β/R_{es}	37 μm
* R_{es} Average mesh resolution along x-direction, measured in millimetres. H, L, β Dimensions measured in millimetres.			

3.4 Model Validation

One of the fundamental checks before validating against experimental data is ensuring mesh resolution independency. This step confirms that the simulation results are not affected by the chosen mesh size, ensuring accuracy and results reliability. Given that productivity, E , depends on both the overall permeate production, $\dot{Q}_p$, and hydraulic loss terms, $\Delta\dot{P}Q_p = (\text{TVD})$, as described in Eq. 6, it is instructive to verify mesh independency for both components. Figure 9a shows the dependence of total viscous dissipation (TVD, shown in orange) and total volumetric permeate flow ($\dot{Q}_p$, shown in green) on the average element size, defined as the channel cross-sectional area divided by the total element count. The parameter L_{sum} is defined by Eq. 10 for basic model. Both TVD and $\dot{Q}_p$ become independent of mesh size below approximately $250 \times 10^{-4} \text{ mm}^2$, indicating that E is also independent of mesh size under these conditions, with a tolerance of less than 0.01%. This result confirms that the chosen mesh resolution is sufficient for accurate simulations.

$$L_{sum} = \beta + 2L \quad (10)$$

Mesh independence can also be assessed by varying element size and observing its effect on spatial profiles. One critical spatial profile in this study is the permeate flux $J_w(x)$. Figure 9b shows the permeate profile's dependence on resolution, which, similar to the global parameters, becomes invariant for mesh sizes below $300 \times 10^{-4} \text{ mm}^2$. This approach captures mesh resolution dependencies that might be masked when using only global parameter checks. The agreement between both methods is reassuring, confirming that the mesh resolution used in this study is appropriate.

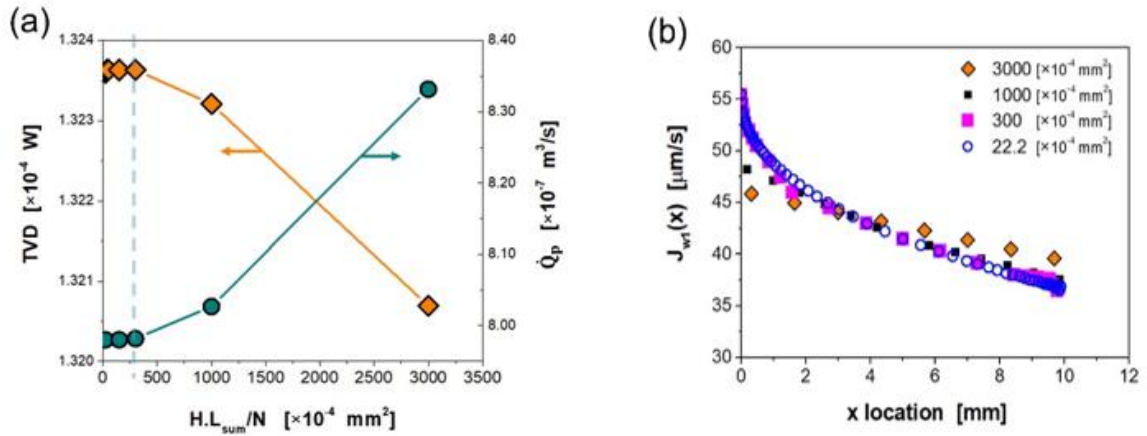


Figure 9: Influence of mesh element size on: (a) total viscous dissipation, TVD , overall permeate flow rate, $\dot{Q}_p$, and (b) permeate flux profile at the membrane surface of the first segment.

3.5 Comparison with Experimental Data

To test the accuracy of the model, it shall be compared to experimental data from the literature. A standard experimental setup by Kim and Hoek (S. Kim & Hoek, 2005) serves as a reference. Their laboratory-scale cross-flow membrane filter, devoid of mesh feed spacers, was configured as a single-sided flow channel. The channel dimensions were a length of 146 mm, width of 95 mm, and height of 1.73 mm, with a membrane water permeability, A , specified as $2.15 \times 10^{-11} \text{ m/s/Pa}$. Figure 10a illustrates a comparison between their experimental data and our simulation results of their configuration. The

experimental data points are represented by single points, while the curves denote our simulation results. This comparison is made for a typical feed flow rate of approximately 0.086 m/s, corresponding to a Reynolds number, $Re = 290$ in their study. Note here that Reynolds numbers for channel flows are often defined by a characteristic length of channel height, H , with fluid properties at inlet conditions. The results are consistent across three different salinities of NaCl, ranging 0.058 wt.% - 0.29 wt.%, and under realistic pressure conditions. The data is expressed as the average permeate flux over the length of the membrane cell, L , in their study, defined by Eq. 11.

$$\bar{J}_w = \frac{1}{L} \int_0^L J_w(x) \, dx \quad (11)$$

It is observed that the average permeates fluxes, of the order of micrometres per second, are typical of RO processes. Our numerical results agree with the experimental data to within 5% marginal error, indicating the reliability of the model. Figure 10b presents a comparison with experimental data from an additional study (Alexiadis et al., 2007), involving a single-sided flow channel followed by an impermeable segment. The membrane, with a permeability of 8.72×10^{-12} m/s/Pa, measures 260 mm in length, while the impermeable segment is 8 mm long. The channel dimensions are 2.5 mm in height and 50 mm in width. The transmembrane pressure for the conditions in Fig. 9b was 1398.7 kPa, corresponding to a low feed salinity of 0.2 wt.% NaCl.

Similarly, the results show agreement with their experimental data within a margin of 5%. This accuracy is expected given the laminar flow regime, which simplifies the modelling by eliminating the need to account for turbulence, along with the well-defined boundary conditions in these two studies. This comparison underscores the robustness of our model in replicating the experimental conditions accurately.

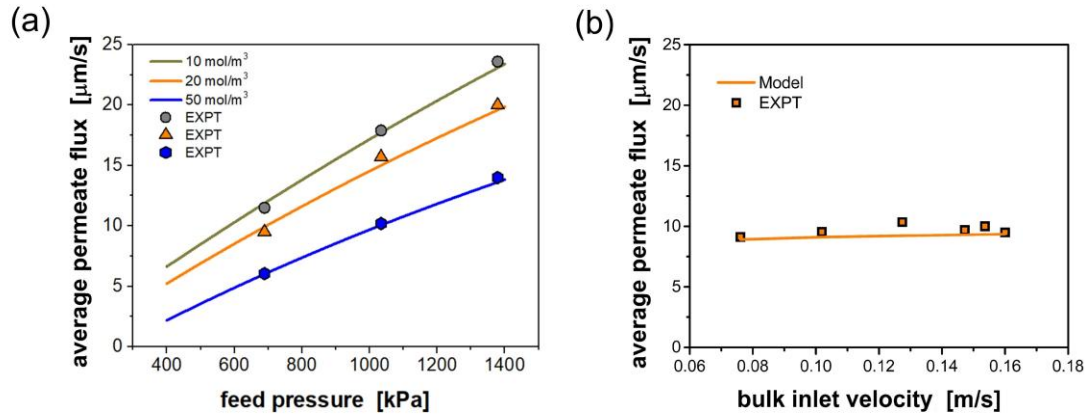


Figure 10: a) Average permeate flux dependence on feed pressure $\bar{J}_w(P)$. Simulated vs. Experimental data for three salinities at $Re=290$. (b) Average permeate flux on inlet velocity. Simulated vs. Experimental data for transmembrane pressure of 1398.7 kPa.

CHAPTER 4

RESULTS AND DISCUSSION

4.1 Single Impermeable Segment

4.1.1 Velocity and Concentration Maps

Figure 11 presents two-dimensional maps of the direct computational FEM output, showcasing the velocity field and salt concentration within the channel. Figure 11a displays the velocity magnitude, $U(x, y)$, as a shaded plot, complemented by arrows indicating the local flow profiles. Figure 11b illustrates the concentration field, $c(x, y)$, as the mass fraction of salt (NaCl), with contour lines included for enhanced clarity.

In this example, the lengths are $L = 10$ mm for the active segments and $\beta = 20$ mm for the inactive segment. The velocity values (U) within the flow field of Fig. 11a range from 0 to 0.02 m/s, corresponding to a bulk inlet velocity of $\hat{U}_f = 0.01$ m/s in this case. A parabolic velocity profile is imposed at the inlet to simulate developed flow conditions accurately, while no-slip conditions are strictly enforced along the channel boundaries. As anticipated, the flow field decreases in the x -direction within the active segments due to permeation through the membrane and remains constant in the inactive segment, where no permeation occurs. For an inlet salinity of $c_f = 0.417$ wt.% NaCl, representative of brackish water, Fig. 11b depicts the evolution of concentration polarization (CP) along the membrane surface in the first active segment. Here, the concentration at the membrane surface rises to approximately 3 wt.%, highlighting a significant CP effect. In the following inactive segment, the concentration recovers close to the inlet value owing to absence of

permeation and the occurrence of back diffusion in towards the centre of the channel. This is followed by a similar CP development in the subsequent active segment, with conditions closely mirroring those at the inlet.

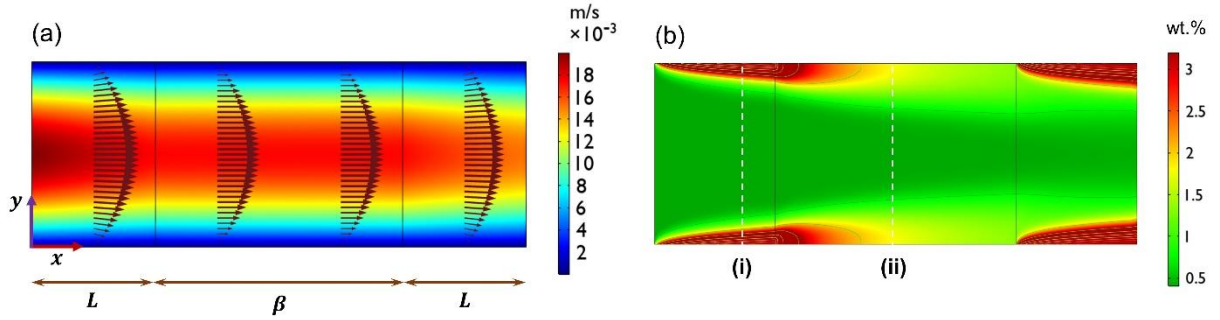


Figure 11: 2-D maps of: (a) velocity magnitude $U(x, y)$, and (b) percentage salt mass fraction $c(x, y)$ for a channel partitioned into segment lengths L and β . Representative salt flux profiles in model are based on cross-sections (i) and (ii).

4.1.2 Water Permeation and Flux Profile

The examination of water permeation and salt flux profiles is important for understanding the performance of the membrane system. A key aspect of this analysis is the velocity field structure near the membranes, as it significantly impacts the system's efficiency.

Figure 12 presents the x -profile of the volumetric flux of permeate, which corresponds to the local flow speed of water through the membrane surface. The flux values are in the micrometer per second range, reflecting a hydraulic permeability of $9.8 \times 10^{-12} \text{ m/s/Pa}$ and a back pressure of 60 barg under brackish water conditions (0.417 wt.% NaCl). Initially, the flux is high at $55.2 \text{ } \mu\text{m/s}$ ($199 \text{ L m}^{-2} \text{ h}^{-1}$), corresponding to the inlet conditions before concentration polarization has developed. The profiles exhibit an exponential decline in magnitude, indicating the influence of the permeation velocity in the y -direction through the membrane boundary. This decline is dominated by the effect of CP, a common

observation in membrane systems, especially in the absence of spacers that induce an axial pressure drop in the x -direction.

The average value of each profile is represented by a pink dashed line, serving as a measure of the total permeate production, which is defined earlier by Eq. 11. This average value is critical for assessing the overall performance and efficiency of the membrane system. As expected, the profiles show a decrease in permeate flux from the first to the second of the active segments, decreasing from $42.4 \mu\text{m/s}$ to $37.3 \mu\text{m/s}$. This decline is primarily due to the overall rise in concentration along the axial direction, which increases the osmotic pressure and reduces the driving force for permeation.

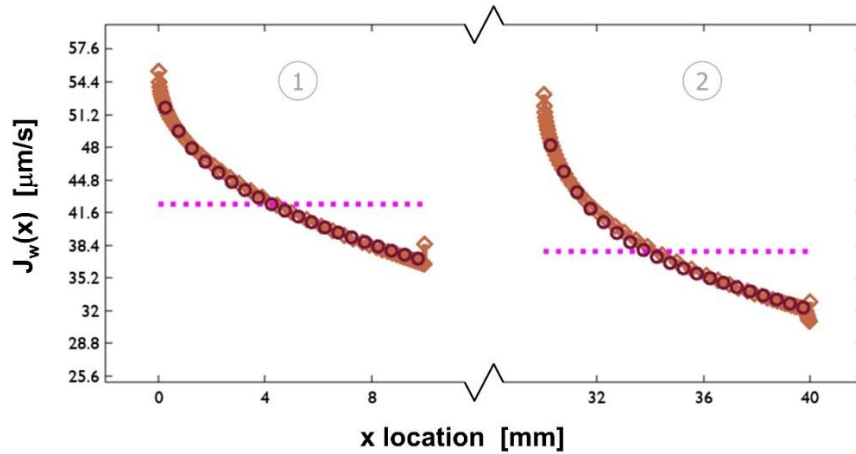


Figure 12: Permeate flux profiles, $J_w(x)$, for the two active segments 1 and 2. Pink dashed lines represent their corresponding average values, $\bar{J}_{w1}$ and $\bar{J}_{w2}$.

The cross-sectional analysis of salt flux is important for understanding the salt transport dynamics within the membrane system. The cross sections (i) and (ii), as marked on Fig. 11b, correspond to the y -profiles of salt flux shown in Figs. 13a and 13b, respectively.

Figure 13 illustrates the contributions to the total salt flux in the y -direction, profiled along the channel height. In Fig. 13a, we observe distinct behaviours of diffusive and convective

flux components. The diffusive flux exhibits a steep decline towards the channel axis in the vicinity of the membrane, indicating a high concentration gradient near the membrane surface. Conversely, the convective flux is generally non-zero and moves in the opposite direction to the diffusive flux. Notably, the convective flux matches perfectly with the diffusive flux right at the membrane surface. The total salt flux, therefore, results in a zero value at the membrane surface, as expected for a membrane exhibiting 100% rejection. This zero net flux at the membrane surface is consistent in Figs. 13a-b, confirming that the convective and diffusive fluxes are in equilibrium at this interface. Away from the axis, the combined effect of the convective and diffusive fluxes results in non-zero values. When combined with the total flux in the x -direction within the boundary layer, this contributes to the cumulative build-up of concentration polarization, thereby maintaining steady-state conditions. Understanding these flux profiles is essential for optimizing membrane performance and mitigating the adverse effects of CP.

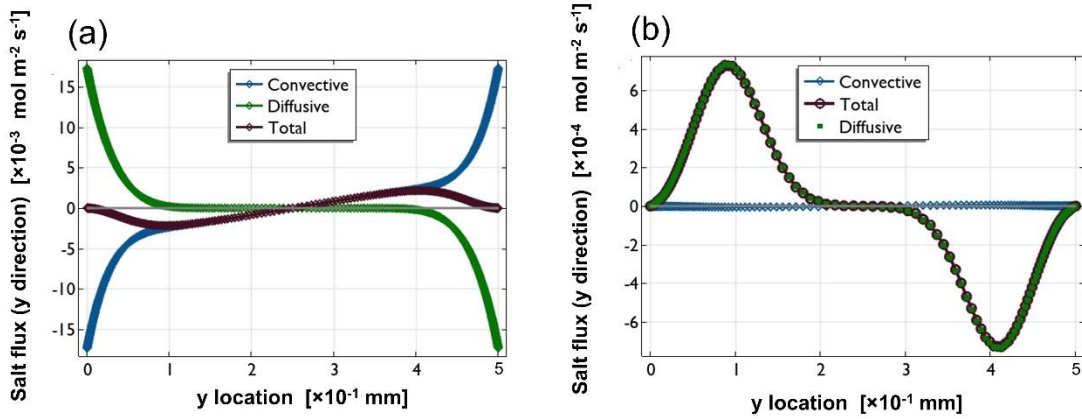


Figure 13: Profiles for y-component of convective and diffusive salt fluxes corresponding to: (a) cross-section (i) at $x=7.5$ mm in active segment, and (b) cross-section (ii) at $x=20$ mm in inactive segment as illustrated in Fig. 11b. Total flux demonstrate zero value at the membrane and wall surfaces.

The balance between diffusive and convective fluxes at the membrane surface plays a critical role in maintaining the efficiency of the desalination process. By analysing these profiles, we can identify key areas for improving membrane design and operational strategies to enhance overall system performance.

4.1.3 Mechanism of Concentration Balancing

One effective approach to mitigating CP involves introducing impermeable segments interspersed between permeable ones. These impermeable segments act as zones for concentration balancing, where solutes are redistributed from the boundary layer back towards the channel axis. This redistribution promotes transverse mixing, effectively disrupting the CP layer and resetting the flow conditions for subsequent permeable segments.

This mechanism is evident in Fig. 13b, where the total salt flux is predominantly driven by diffusion, as indicated by the close alignment of the green and brown curves. Unlike Fig. 13a, where the flux direction is outward due to permeation, the flux within the impermeable segments reverses direction, moving inward toward the channel axis. This inward flux depletes the CP layer, distinguishing it from the active permeable segments where solutes accumulate near the membrane. By alternating between these segments, the CP layer is periodically disrupted, maintaining a more uniform solute distribution and enhancing the overall efficiency of the flow.

The surface plot in Fig. 14a facilitates visualization of the overall CP distribution within the channel. This plot illustrates conditions for a Reynolds number of 4.8 and permeability coefficient of 9.8×10^{-12} m/s/Pa. The domain corresponds to half channel height, based

on axial symmetry, with the axis centred on $y = 0.25$ mm. The grey back surface in the plot denotes the channel boundary, comprising the membranes and impermeable wall. The CP modulus (CPM) in Fig. 14a is defined as $c(x, y)/c_f$.

CPM exhibits variation with x and y in Fig. 14a, adopting an exponential surface topology. The exponential increase in concentration with x in the active segments is consistent with the cumulative salt migration via total flux observed in Fig. 13a, which directs salt towards the membranes. This exponential rise in CPM within the active segments is notable close to the membrane, where CPM peaks at 6.5 and 8 for $0 < x < 10$ mm and $30 < x < 40$ mm, respectively. The maximum CPM of 8 corresponds to 3.3 wt.% NaCl, which is substantially below the saturation limit of 26.5 wt.% at ambient temperature (25°C) (S. Kim & Hoek, 2005). It is critical to include the concentration dependence of fluid properties (CDFP), as the peak CPM in the first active segment was found to be 5.4% higher when CDFP was considered. Modelling CDFP is prudent in general, given that localized high concentrations can occur in regions of flow recirculation, where scaling may initiate. However, this is less critical in our case, where the absence of spacers prevents the formation of recirculation zones. Within the active segments, there is a sharp exponential decrease in concentration away from the membrane surfaces in the y direction towards the channel axis, with a reduction towards a background CPM close to unity. This exponential trend in the y direction persists in the inactive segment (i.e., for $10 < x < 30$ mm); however, an exponential decrease of CPM in the axial direction is also observed. The absence of the convective component of salt, combined with back diffusion towards the channel axis, facilitates concentration reduction - and hence homogenization.

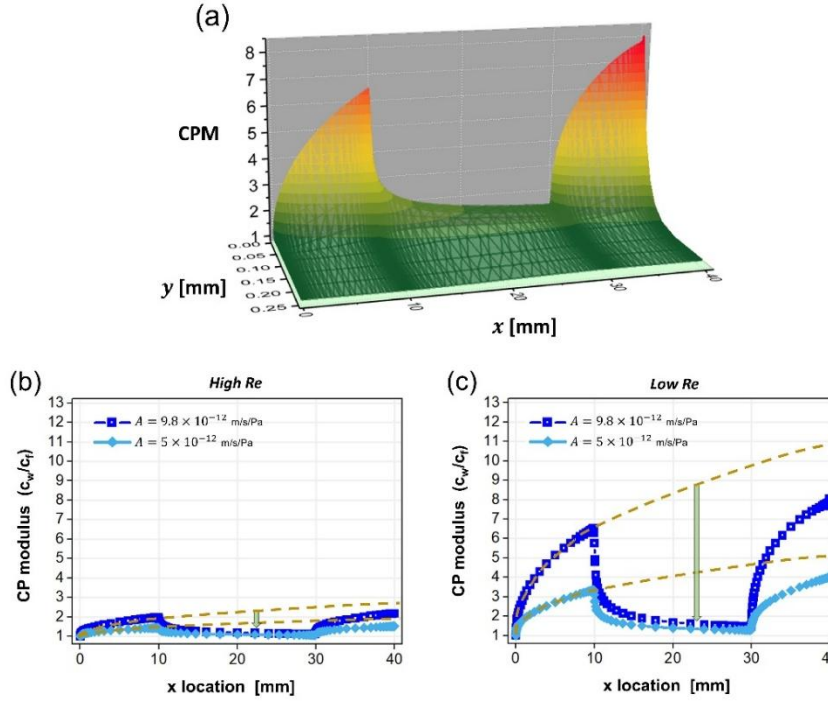


Figure 14: (a) Distribution of salt buildup within the channel, quantified by CPM (x, y), for $Re = 4.8$ and $A = 9.8 \times 10^{-12}$ m/s/Pa. Specific CPM profiles along the wall shown for (b) high Reynolds ($Re = 240$) and (c) low Reynolds ($Re = 4.8$). Dashed lines represent the CPM profiles in the absence of concentration balancing

A more detailed examination can be achieved by plotting CPM at the boundary, as illustrated in Figs. 14b-c. These figures investigate the effects of varying inlet flow (Re) and the permeability coefficient (A) on the CPM profile; here, CPM is defined as c_w/c_f . Regarding permeability, a second value of 5×10^{-12} m/s/Pa was also analysed, representing a lower typical value found in the literature.

As anticipated, a lower permeability decreases CP – in this case by approximately 50%, corresponding to a reduction in A from 9.8×10^{-12} to 5×10^{-12} m/s/Pa. For the effect of flow, comparing Figs. 14b and 14c reveals that a higher Re of 240 results in a smaller difference between CPM profiles of differing A . Additionally, the profiles are lower and flatter with reduced curvature.

The brown dashed lines in Figs. 14b-c depict the CP modulus if the entire boundary were a single membrane without concentration balancing. The presence of the inactive segment evidently results in a lower overall concentration at the wall. This is further highlighted by the comparison of the green arrows, where the longer arrow in Fig. 14c indicates a larger reduction of CP at the low Reynolds flow. The "concentration reset" observed leads to a similar CP modulus profile in the subsequent active segment downstream, which reduces the average osmotic back pressure at the membrane (as indicated in Eq. 4), thereby enhancing the permeate flux.

4.1.4 Segment Length and Flow Velocity Effects

The effects of segment length and flow velocity on system productivity, E , are illustrated in Fig. 15. This figure plots productivity E against the inactive segment length, β , for six different bulk velocities, corresponding to an example fixed active segment length $L = 10$ mm and permeate value w of 2×10^{-7} \$/kg. Black curves denote positive E values on the left axis, while red indicates negative on the right axis.

The plots in Fig. 15 reveal productivity E reaching a maximum value at an optimal β , herein referred to as β^* . As the flow velocity increases, the position of this maximum shifts towards smaller β values. For sufficiently large flow velocities, no stationary point is observed, though a constrained maximum at $\beta = 0$ is noted. In these cases, E often assumes negative values, as indicated on the red-coloured secondary axis. This behaviour suggests that at higher flow velocities, the benefits of optimizing β are diminished, and the system's productivity can decrease, potentially resulting in negative E values. This trend

underscores the importance of carefully balancing segment length and flow velocity to maximize the economic performance of the membrane system.

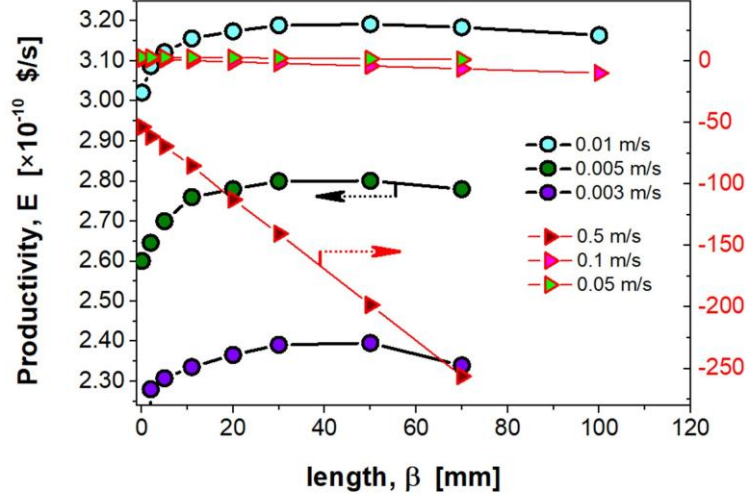


Figure 15: Dependence of productivity, E , on inactive segment length, β , at six different flow velocities, $\hat{U}_f$. Values correspond to active segment length, $L = 10$ mm, for a permeate value, $w = 2 \times 10^{-7}$ \$/kg. Black curves denote positive E values, and red as negative on the secondary axis-denoted by the arrows

Figure 16 provides an overall illustration of how productivity, E , varies with different combinations of both active segment length L and inactive length β . The surface plot corresponds to an inlet flow velocity $\hat{U}_f = 0.01$ m/s and permeate value w of 2×10^{-7} \$/kg. An asymptotic rise in E is apparent with increasing L across various β values. This indicates that while productivity continues to increase with longer active segments, it eventually levels off due to the balancing effects of increased permeate production and rising pressure drop. For this specific case, E increases monotonically with L . However, it exhibits a maximum with respect to β , resulting in a slightly folded surface in Fig. 16. The locus of these maxima at each L - denoted by the blue dotted line - represents the optimal β (denoted as β^*). This line highlights the most effective combination of segment lengths for maximizing productivity under the given conditions.

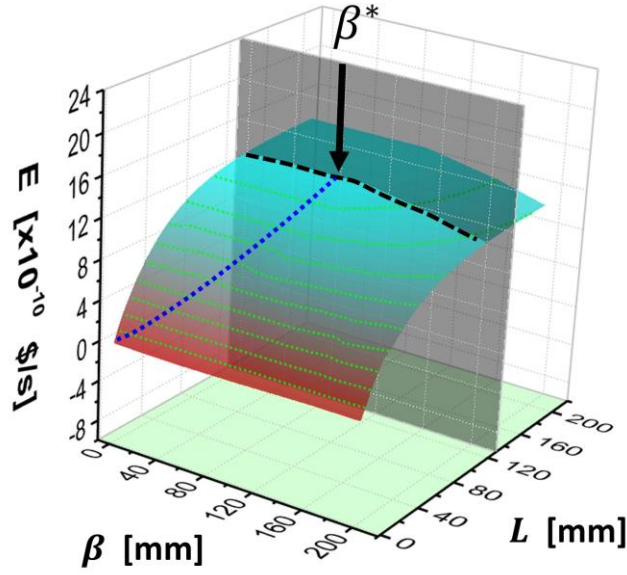


Figure 16: Contoured surface plot illustrating productivity, E , dependence on both β and L for a flow velocity of 0.01 m/s and permeate value, $w = 2 \times 10^{-7} \$/kg$. Cross-section at $L = 130$ mm highlights example intersection curve (black dashed line), whose local maximum E value occurs at β^* . Blue dotted line represents the locus of such local maxima for various lengths, L .

Figure 17a explores in more detail the dependence of the optimal inactive segment length, β^* , on the active segment length, L – for example conditions: inlet velocity of 0.01 m/s, transmembrane pressure of 15 bar, and permeate value $w = 1 \times 10^{-3} \$/kg$. The brown curve in this figure demonstrates a distinct rise and fall in β^* , presenting an explicit quantitative relationship between the two parameters. This behaviour contrasts with the blue dotted curve in Fig. 16, which shows the locus of maxima without a clear rise and fall as L increases. The total channel length at optimum conditions, $L_{sum} = \beta^* + 2L$, is also shown; this is given by the blue curve in Fig. 17a exhibiting a steady increase despite non-monotonic variation in β^* .

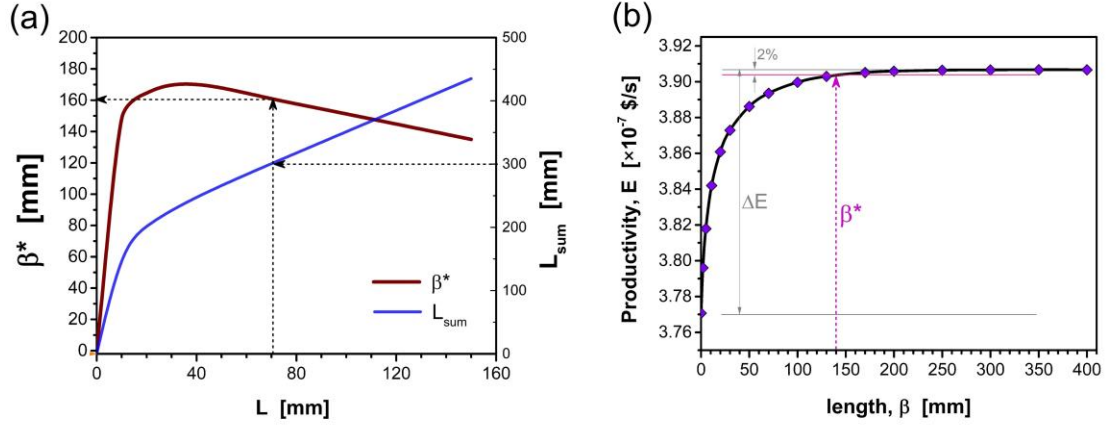


Figure 17:(a) Dependence of optimum inactive segment length, β^* , with active segment length, L , (brown curve), and resulting total channel length, L_{sum} , shown in blue. (b) Example curve showing how β^* is evaluated based on 2% tolerance, corresponding to length of $L = 10$ mm. All curves correspond to conditions: $\bar{U}_f = 0.01$ m/s, $\Delta P_T = 15$ bar, and $w = 1 \times 10^{-3}$ \$/kg

The determination of β^* relies on the characteristics of the E vs. β curves. While clear stationary points – as observed in Fig. 15 – are often used to locate maxima, these points are not always well-defined, particularly at higher permeate values, w . Figure 17b highlights such conditions, where the E vs. β curves exhibit a monotonic exponential character. In these cases, traditional calculus-based methods are insufficient for accurately locating β^* . To address this challenge, the Threshold-Based Maximum Evaluation method is employed. This approach systematically identifies the maximum by determining the point, β^* , corresponding to a small offset just below the curve's plateau. The offset used in this work is defined as 2% of the rise, ΔE , between the curve's intercept and its peak value. This approach ensures reliable evaluation of β^* when stationary points are ambiguous. Notably, as w decreases toward zero, stationary points become more defined, and the results obtained using the Threshold-Based Maximum Evaluation method converge with those of calculus-based methods, as expected.

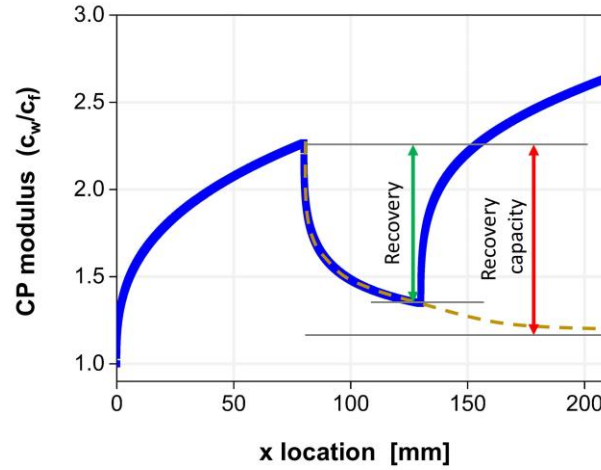


Figure 18: Example CPM profile along the wall for $Re = 4.8$, $L = 80$ mm, $\beta = 50$ mm, and $\Delta P_T = 15$ bar. The dashed line represents the extrapolation of the CPM profile for the inactive segment. Both Recovery and Recovery Capacity are indicated.

The rise and fall of β^* , depicted by the brown curve in Fig. 17a, is governed by the recovery capacity (Chrystie et al., 2025) defined as the difference in solute concentration between the peak at the end of the active segment and the lowest possible value at the end of the inactive segment (as $\beta \rightarrow \infty$). Initially, as L increases from zero, the peak wall concentration at the end of the active segment rises, necessitating longer inactive segments (β^*) to effectively dissipate the solute. However, as L continues to grow, the bulk concentration in the subsequent inactive segment also starts to rise, increasing the overall background salinity. This faster rise in bulk concentration diminishes the recovery capacity, reducing the need for longer inactive segments and thus forces β^* to decrease. It should be noted that while all this is happening there is a consistent background hydraulic loss that increases linearly with segment length. Such losses become more significant with larger L , further discouraging extended inactive segments. In general, the optimal inactive segment length, β^* , is determined by balancing the amount of recovery along with the associated hydraulic loss. This occurs where the wall concentration is sufficiently reduced

over the steepest descending portions of the blue curve in Fig. 18, without incurring excessive hydraulic penalties. These dynamics collectively explain the rise and fall behaviour of β^* observed in Fig. 17a.

4.1.5 Significance of Concentration Balancing

A critical aspect in assessing the economic viability of using impermeable segments for concentration balancing is determining when it maximizes productivity. The useful parameter in this analysis is β^* , which represents the optimal inactive segment length that maximizes E under a set of given conditions. When $\beta = \beta^*$, the system operates at peak efficiency, and the converse is true when $\beta \neq \beta^*$. Therefore, selecting the appropriate inactive segment length is important for achieving maximum economic productivity.

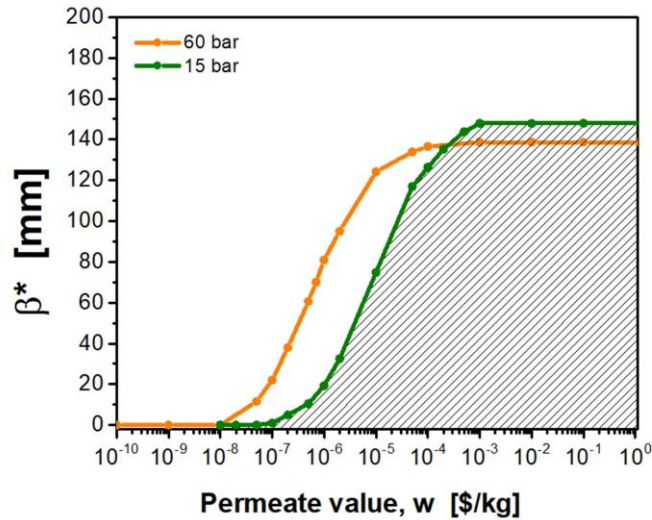


Figure 19: Plots tracking optimum inactive segment length, β^* , with permeate value, w , for two ΔP_T values of 60 and 15 bar corresponding to $L = 10$ mm and $\hat{U}_f = 0.01$ m/s.

The parameter β^* is a function of four key variables: the segment length (L), feed velocity ($\hat{U}_f$), transmembrane pressure drop (ΔP_T) and permeate value (w). Among these, the dependence of β^* on w is of particular interest, as it determines the optimal length of

impermeable segments and evaluates whether such segments are even necessary based on this economic criterion. However, the interdependence with the other parameters, particularly ΔP_T , cannot be overlooked. This interaction is explored in detail by analysing the joint dependence of β^* on w and ΔP_T at fixed values of L and $\hat{U}_f$. Specifically, Fig. 19 illustrates the trends for β^* as a function of w and ΔP_T , with example representative values of $L = 10$ mm and $\hat{U}_f = 0.01$ m/s.

To capture the trends effectively, two pressure values commonly used in commercial applications, 15 bar and 60 bar, are plotted. The figures use a linear scale for β^* and a logarithmic scale for w . For 60 bar, β^* initially remains zero for $w < 10^{-8}$ \$/kg, then rises steeply and plateaus at $\beta^* = 140$ mm. A similar trend is observed at 15 bar, although the curve is shifted toward larger w values and the plateau occurs at a higher value of $\beta^* = 150$ mm. The differences in these trends can be explained by considering the interplay between permeation rate and bulk concentration dynamics. At 60 bar, starting from low values of w near 10^{-8} \$/kg, β^* increases more rapidly than at 15 bar. This is because the higher pressure induces a greater rate of permeation, leading to a more pronounced peak in wall concentration at the downstream end of the active segment. To mitigate this effect of concentration polarization and maintain effective recovery, a larger β^* is required. However, as w increases, β^* eventually reaches a plateau because the inherent recovery capacity limit is reached, beyond which no additional benefit in recovery can be achieved. In contrast, at 15 bar, the rate of permeation is lower, resulting in a less significant increase in bulk concentration downstream. Consequently, the system can achieve a higher peak β^* value before reaching its recovery limit. The plateau at 15 bar thus occurs at a higher β^*

value of 150 mm, reflecting the system's greater recovery capacity at this lower pressure. These insights highlight the nuanced interplay between β^* , w , and ΔP_T , emphasizing the importance of considering all parameters for optimizing system performance.

The dependence $\beta^*(w)$ is further analysed by plotting β^* against w for various fixed lengths of the active segment, L . Figure 20a illustrates this relationship for an inlet flow velocity $\hat{U}_f = 0.01$ m/s and transmembrane pressure $\Delta P_T = 15$ bar, a typical operating pressure for brackish water systems, where the feed salinity is set here at 0.417 wt.%. The figure depicts five black curves corresponding to different L values spanning 10 to 150 mm, showing the co-dependence of β^* on w and L .

For $L = 10$ mm, β^* remains zero at low permeate values ($w < 10^{-8}$ \$/kg), rises sharply beyond this threshold, and eventually plateaus to a maximum value. A similar trend is observed for the other L values, with notable differences between the curves. At low w values, just exceeding 10^{-8} , β^* increases monotonically with L . This increase gradually saturates as L approaches 150 mm, a behaviour attributed to the saturation of wall concentration at the end of the active segment as L increases. At higher w values (e.g., $w = 10^{-3}$ \$/kg), the trend in β^* becomes more complex, exhibiting a non-monotonic rise and fall as a function of L . This behaviour is explicitly captured in Fig. 17a (brown curve), where the underlying physical reasons for this trend have been discussed earlier. The interplay between permeation rate, bulk concentration, and wall concentration dynamics explains the observed variation in β^* , highlighting the nuanced impact of L on system performance across different permeate values.

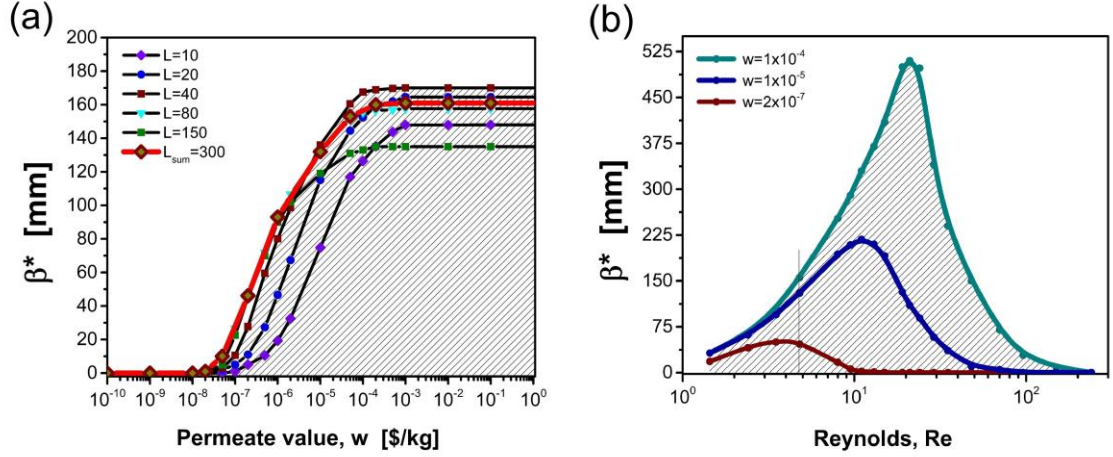


Figure 20: Dependence of optimum length, β^* , on: (a) permeate value, w , and (b) Reynolds number, Re . Panel (a) shows the effect of active length, L , and includes a fixed total channel length plot ($L_{sum} = 300$ mm, red curve) along with corresponding productivity (orange curve). Conditions for (a) are $\bar{U}_f = 0.01$ m/s and $\Delta P_T = 15$ bar. Panel (b) depicts variation of β^* for three permeate values w , with conditions: $L = 80$ mm and 15 bar.

To further analyse the behaviour of β^* with w , we consider the case where the total channel length is constrained to 300 mm. This scenario, represented by the red curve in Fig. 20a, corresponds to typical residential reverse osmosis module lengths, particularly in residential applications. The choice of $L_{sum} = 300$ mm balances the need for practical relevance along with computational feasibility. The red curve in Fig. 20a was calculated with the aid of Fig. 17a, which displays two key relationships for $w = 10^{-3}$ \$/kg: (i) $\beta^*(L)$, and (ii) $L_{sum}(L)$. Using $w = 10^{-3}$ as an example, the process begins by identifying $L_{sum} = 300$ mm on the blue curve. The corresponding L value is then used to determine β^* from the brown curve. This approach is illustrated in Fig. 17a with a series of three dotted lines, maintaining the constraint $\beta^* + 2L = 300$ mm. A similar methodology applies to other curve pairs for different values of w . It can be seen that the red curve in Fig. 20a exhibits a trend consistent with the black curves: β^* rises from zero for $w > 10^{-8}$ \$/kg and eventually saturates, in this case at $\beta^* = 160$ mm.

Figure 20b presents three curves illustrating how β^* varies with channel inlet flow, characterized by the Reynolds number (Re), for the case of $L = 80$ mm, $\Delta P_T = 15$ bar. Collectively, the results reveal a monotonic relationship between β^* and the permeate value w . Notably, at $Re = 4.8$, indicated by a vertical marker, β^* saturates at high w values, a trend consistent with the red curve in Fig. 20a. Each curve in Fig. 20b exhibits a rise and fall in β^* with increasing Re , with the peak β^* shifting toward larger Re as w increases. The peak value of β^* also consistently rises with increasing w . This behaviour is accompanied by an upward shift in the upper limit of Re , where β^* transitions from zero. Specifically, the sequence $Re: 10 \rightarrow 100 \rightarrow 200$ is observed in Fig. 20b as w increases.

The rise and fall behaviour of β^* with Re can be explained by a mechanism analogous to the trend of β^* with L , as illustrated by the brown curve in Fig. 17a. However, this relationship operates in reverse: whereas an increase in L enhances concentration build-up due to polarization, a decrease in inlet flow (Re) produces a similar effect. For instance, using the blue curve ($w = 10^{-5}$ \$/kg) in Fig. 20b as an example, as Re decreases from 100, β^* begins to rise from zero. This increase in β^* continues as Re approaches ~ 10 , driven by elevated concentration polarization at the end of the active segment, induced by the slower cross-flow. Consequently, a longer optimal segment length (β^*) is required to mitigate this effect. However, as the cross-flow slows further ($Re < 10$ in this case), the overall bulk concentration starts to rise. Under these conditions, the recovery capacity diminishes, limiting the maximum recovery achievable. As a result, the need for a high β^* decreases, leading to a decline in β^* with further reductions in Re .

The monotonic relationship between β^* and w , as seen in Fig. 20b, reflects the maximization of recovery capacity, which is consistent with expectations. For example, at $Re = 4.8$ ($U_f = 0.01$ m/s), the results align with the trends observed for $L = 80$ mm in Fig. 20a. Furthermore, as w increases, the influence of recovery capacity becomes more pronounced. This reduces the need for significant reductions in Re to approach full recovery capacity, causing the peak of the curve in Fig. 20b to shift toward higher Re values for larger w . From the representative ranges of w and Re shown here, it is evident that the advantages of concentration balancing during optimization depend critically on the concurrent thresholds of both Re and w . Also, this relationship demonstrate that intermittent membrane desalination is effective only at laminar flow regime since having Re in turbulent flow will diminish the length of β^* ($Re \gg 200$). Equation 12 succinctly captures this relationship, specifying the conditions under which β^* remains positive.

$$\left. \begin{array}{l} Re < 200 \\ w > 10^{-8} \text{ \$}/\text{kg} \end{array} \right\} \beta^* > 0 \quad (12)$$

These threshold values are specific to the example case analysed in this study. Under different conditions, with variations in key parameters, the threshold values would shift accordingly. Nevertheless, the example case serves to illustrate the fundamental concept: both Re and w must simultaneously meet their respective thresholds to ensure optimal performance, as indicated by a positive β^* . This highlights the importance of careful parameter selection and balance in achieving effective system optimization. The analysis reveals that w must be non-zero to enable effective optimization, rendering it a necessary,

but not sufficient, condition. Additionally, the Reynolds number for channel flow must remain within the laminar range to fully exploit the benefits of incorporating impermeable segments.

This study emphasizes the need to consider both Re and w when evaluating the advantages of the proposed novel approach. In particular, w has been treated as a variable throughout this analysis. To optimize a specific system, reliable w values should be determined for contexts and geographical regions, requiring a comprehensive economic analysis. While such a detailed economic investigation is beyond the scope of this paper, the study underscores the critical interplay between economic factors and engineering design. This interplay must be carefully navigated to achieve optimal membrane system performance.

4.2 Dual Impermeable Segments

In this section we will investigate productivity in a model where the number of inactive segments is two. Each segment is positioned between two active segments as shown in Fig. 21. The reason for that is to study if further segmentation may result in further reduction in CP and higher productivity.

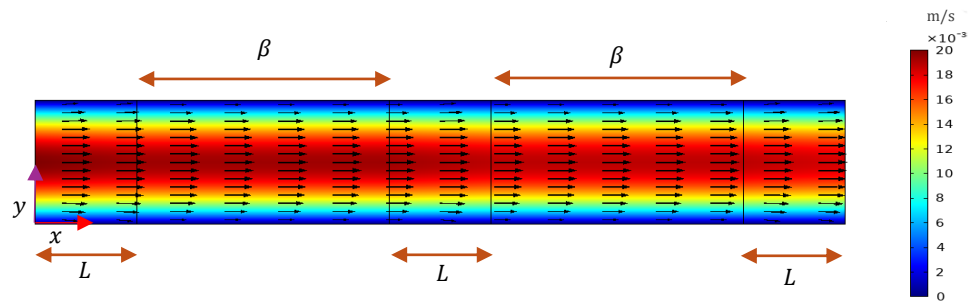


Figure 21: 2-D map showing the dual impermeable segment model as a velocity profile diagram.

4.2.1 Segment Length and Flow Velocity Effects

First, we shall study the behaviour of the productivity rate versus lengths at different feed velocity value. This study will allow us to identify the highest productivity rate at the most optimum length of β which is denoted earlier as β^* . In Fig. 22, the model was tested for six different velocity values at $L = 10$ mm and $\$/\text{kg}$.

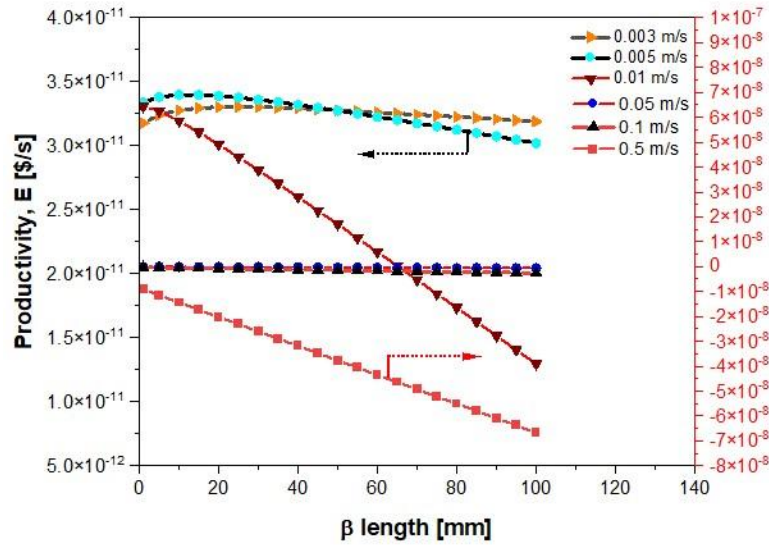


Figure 22: Dependence of productivity, E , on inactive segment length β , at six different flow velocities

Lower velocities (0.003 m/s – 0.005 m/s) show a slight increase in productivity rate at lower β length ($\beta < 20$ mm) after which it decreases. On the other hand, the velocities at (0.01 m/s and 0.5 m/s) exhibit a very sharp decline for productivity rate (negative) represented by the secondary y-scale in red, while moderate velocities (0.05 m/s & 0.1 m/s) remain nearly constant. Overall, the impact of β^* length on productivity varies depending on velocity, indicating a strong relationship between feed flow rate and efficiency.

As shown in the graph, the productivity range has reduced significantly comparing to the single impermeable model (Fig. 15). Also, the graph indicate that the model maximum productivity is at lower velocity of 0.005 m/s comparing to the single impermeable model at 0.01 m/s. This is due to the balancing effect between permeate production and the

pressure drop that takes place within the feed channel. The more impermeable segment to be introduced, the more the hydraulic loss, and less permeation. So, lower velocity is favourable.

4.2.2 Concentration Balancing

Utilizing intermittent flow membrane is not always the best choice. The judgment shall come from studying economic viability first. In Fig. 23, we examine the range of length at permeate value under two different operational pressures to have better understanding on how to achieve the maximum productivity and profitability.

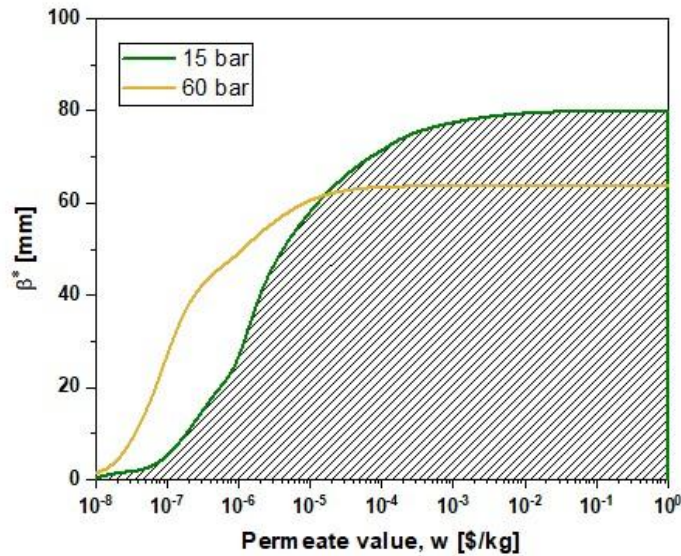


Figure 23: Plots tracking optimum inactive segment length, β^* , with permeate value, w , for two ΔP_T values of 60 and 15 bar corresponding to $L=10$ mm and $\bar{U}_f=0.005$ m/s.

At 15 bar, length increases gradually at low permeate values, then rises sharply before plateauing around 80 mm. At 60 bar, length grows more rapidly initially but stabilizes around 65 mm. This indicates that the model behaviour varies significantly with pressure, particularly at higher permeate values. Still, herein we see that the trend of 15 bar shows higher recovery capacity than 60 bar similar to what was shown for the single model. The

only difference between both models is that at the 15 bar the trend starts rising from zero earlier at (\$/kg). This could be attributed to the introduction of the second impermeable segment creating further hydraulic loss compared to the single model; where it required some length at lower to adjust to maintain effective recovery and maintain the dynamic balance between, and CP.

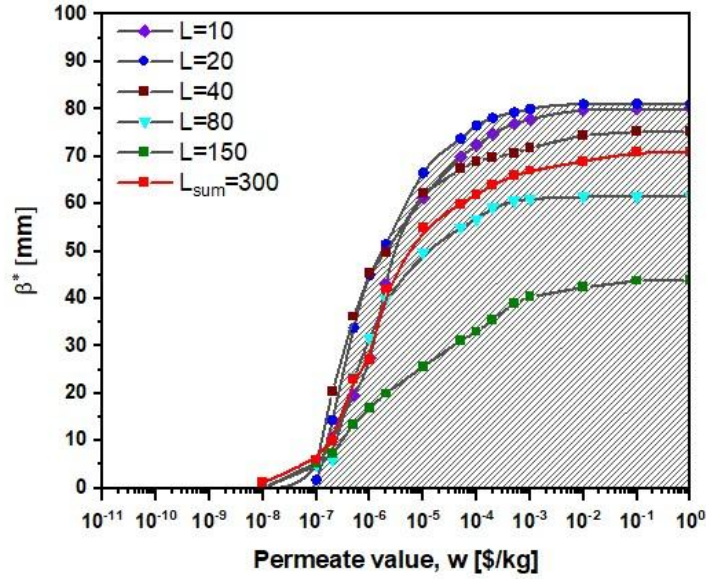


Figure 24: Optimum length of β^* , versus permeate value w , for multiple L , lengths

For Further analysis, (β^*) was plotted against a logarithmic scale of permeate value (w) for six different fixed length of (L), pressure of 15 bar and flow velocity $\hat{U}_f = 0.01$ m/s, feed salinity of 0.417wt.% as shown in Fig. 24. All curves are increasing monotonically until $w = 10^{-3}$ (\$/kg) after the trend in β^* becomes more complex, exhibiting a non-monotonic rise as a function of L . This show case the same behavior as the single impermeable model in terms of the interplay between permeation rate, bulk concentration, and wall concentration dynamics.

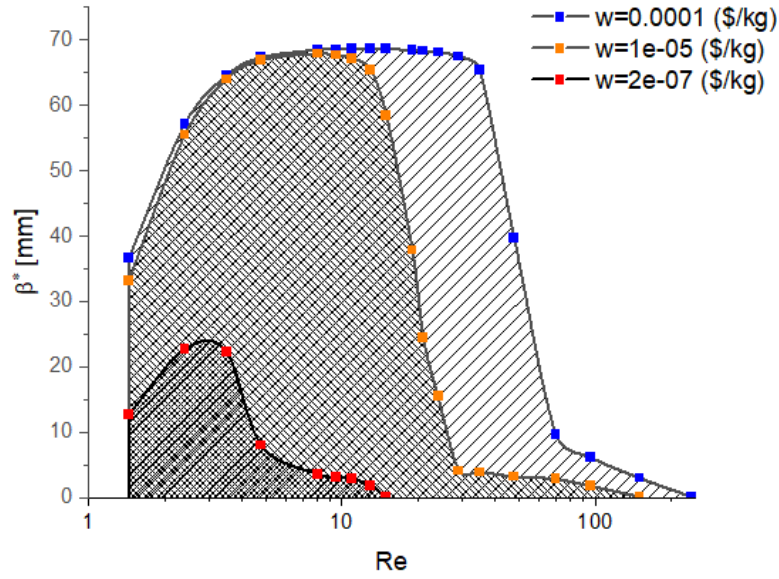


Figure 25: Optimum length of β^* , versus Reynold number Re , for three permeate values w

For the effect of length on feed velocity, Fig. 25 shows that the model plays on a significantly shorter range of Reynold number and length compared to the single impermeable segment model. This is attributed to the increased hydraulic losses taking place by the introduction of the second impermeable segment. From all the results that were obtained, it is still evident that further increment of feed flow velocity to turbulent regime ($Re \gg 200$) will diminish the length of β^* . Also, it is verified that the dual impermeable model does not have any additional advantages over the single impermeable model either in the Reynold number nor permeate value range.

$$\left. \begin{array}{l} Re < 200 \\ w > 10^{-8} \text{ \$/kg} \end{array} \right\} \beta^* > 0 \quad (13)$$

4.3 Economic Savings Potential

The discussion above focuses on optimizing the segmented channel configuration to achieve its best possible operational performance. While the β^* parameter offers valuable insight into whether incorporating a finite inactive segment length is advantageous, it does not directly measure the magnitude of performance enhancement, as evaluated by the productivity E defined in Eq. 6. To address this gap, a direct comparison with conventional feed-spacer-filled channels is instructive for quantifying differences in effective productivity. This comparison is framed by the metric $E_{seg} - E_{spacer}$, where E_{seg} and E_{spacer} represent the productivity of the segmented and spacer-filled configurations, respectively. For the segmented configuration, calculations were performed for the 300-mm-long channel with a feed velocity of 0.01 m/s, a channel height of 0.5 mm, and a width of 1 m. The determination of optimal segment lengths is informed by the red curve shown in Fig. 20a for the 300-mm long channel. With knowledge of permeate value, w , the optimal conditions, and hence rate of permeation, can be used to properly evaluate E_{seg} . For comparative purposes, E_{spacer} computations draw from (Gruber et al., 2016) open-source CFD model, which simulates a standard staggered-cylinder filament arrangement. This reference configuration employs cylindrical filaments with a diameter of 0.25 mm and axial spacing of 1 mm. The overall channel dimensions (length, height, and width) were kept the same as the segmented case – ensuring a fair basis for evaluating performance differences between the two setups.

Figure 26 presents two curves illustrating the productivity difference, $E_{seg} - E_{spacer}$, plotted on both primary and secondary axes for ease of comparison. The first curve (grey) corresponds to the case where the spacer-filled channel operates at a feed velocity of 0.01 m/s, matching the segmented channel. This velocity equates to a Reynolds number of 4.8, remaining below the aforementioned limit of 200. The second curve (green) represents the spacer-filled channel with a feed velocity of 0.5 m/s. Both curves are based on a typical unit energy price of 3.6×10^{-8} \$/J (Consumer Price Index Average Price Data, in, US Bureau of Labor Statistics, 2023).

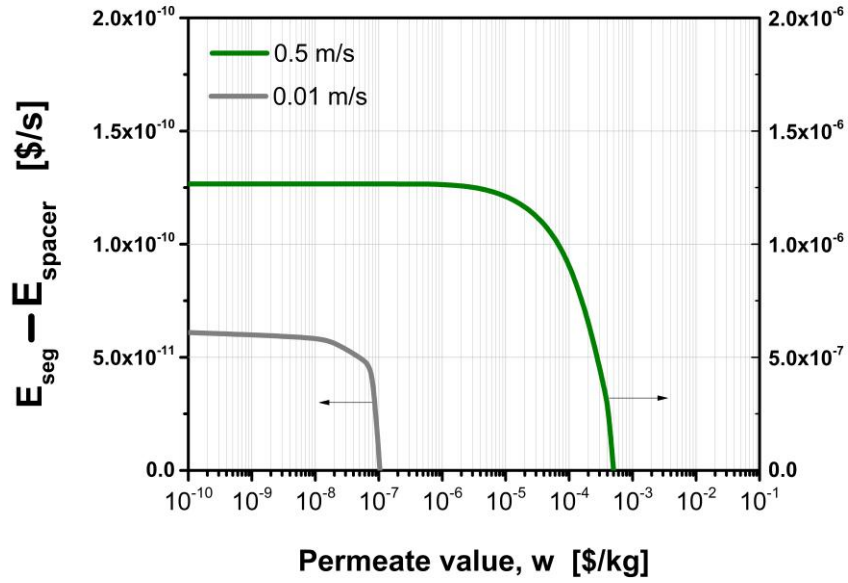


Figure 26: Dependence of productivity difference on permeate value, w , at a segment feed velocity of 0.01 m/s. Each curve represents a corresponding feed velocity in the spacer-filled channel for comparison.

The break-even point, where the segmented and spacer-filled configurations show equal productivity ($E_{seg} = E_{spacer}$), is observed in Fig. 26. For the case where the spacer-filled channel operates at 0.01 m/s, break-even occurs at $w = 1 \times 10^{-7}$ \$/kg, as indicated by the grey curve. At lower w values, the productivity difference saturates to a level of

5.9×10^{-11} \$/s. In contrast, for the 0.5 m/s spacer-filled channel, the break-even point shifts to a higher value of $w = 5 \times 10^{-4}$ \$/kg, with the productivity difference increasing by four orders of magnitude and saturating at 1.26×10^{-6} \$/s.

The comparison with a feed velocity of 0.5 m/s is relevant, as this value represents a typical operating condition for conventional spacer-filled channels. This allows for a fair assessment of the segmented configuration at 0.01 m/s against a commercial spacer-filled system. Under these conditions, the maximum economic savings achieved by the segmented configuration is approximately 1.26×10^{-6} \$/s, which translates to an annual cost reduction of \$40, assuming continuous operation. This saving is largely attributed to the 99.9% reduction in axial pressure drop when comparing the segmented and commercial spacer-filled configurations. It is important to note that this calculation is based on the unit energy price used in this study, and the maximum annual savings figure here can be directly scaled according to alternative energy prices. Furthermore, the estimated savings apply to a single channel with dimensions of 0.5 mm $\times$ 300 mm and width of 1 m. For similar channels of different width, the annual cost savings can also be proportionally adjusted.

CHAPTER 5

CONCLUSION AND RECOMMENDATIONS

5.1 Conclusion

This study introduces an alternative approach for the implementation of reverse osmosis (RO), nanofiltration (NF), or ultrafiltration (UF) – with the aim of eliminating the need for conventional spacers within channels. The concentration balancing approach in this work uses a built-in impermeable segment to reset the concentration boundary layer back towards its initial state. The spacer-free design offers several advantages, including a lower hydraulic loss in the channel, a reduced tendency to accumulate foulants, and easier cleaning. To explore this concept, a first-principles numerical model was developed using the finite element method. The model simulates a two-dimensional channel under laminar flow conditions with dilute species transport, considering brackish water as an example feed. The first configuration featured a central impermeable segment, which was integrated between two standard membrane segments, forming a hybrid channel, while the second configuration featured a dual positioned between two active segments. For the first model, the impermeable segment essentially resets the wall concentration profile for the subsequent standard membrane segment downstream. This has the effect of reducing the level of concentration polarization and hence enhancing the overall permeate rate for the module. However, the use of an impermeable segment introduces additional hydraulic loss, necessitating an optimization trade-off. This trade-off is captured by the parameter E , which represents the productivity rate in \$/s, incorporating both the permeate yield – weighted by its economic value (w in \$/kg), and the hydraulic loss – weighted by the unit

energy cost (e in $\$/J$). A techno-economic analysis reveals that an optimal impermeable segment length exists, influenced by both w and the cross-flow Reynolds number (Re).

The optimal segment length increases monotonically with w , reflecting the growing importance of mitigating concentration polarization as permeate value rises. Conversely, its dependence on Re follows a non-monotonic trend: initially increasing with lower cross-flows due to an increased need for concentration balancing but eventually declining as further reduction in Re leads to higher background concentration levels, diminishing the recovery capacity and making it harder to offset hydraulic losses. This dynamic interplay between w , Re , and concentration balancing governs the feasibility and performance of the spacer-free channel configuration.

On the other hand, for the dual impermeable segment model, even though it exhibits similar behaviour as the single model, the practicality of using the dual model is not preferable due to its lower productivity range and the further reduction imposed by the addition of a second impermeable segment. This creates more hydraulic loss whose effect cannot be overcome by trying to reset salt concentration as close as possible to similar levels at the inlet section.

Optimising the proposed channel configuration of the first model helps determine whether incorporating an inactive segment is inherently beneficial, but it does not directly quantify performance gains. To assess the practical advantages of this approach, we compare it with a conventional feed-spacer-filled channel using the metric $E_{seg} - E_{spacer}$, which evaluates the effective productivity difference. On comparing with a standard staggered-cylinder filament arrangement, we reveal key findings: a 99.9% reduction in axial pressure drop for

a 300-mm long commercial channel operating at 0.5 m/s cross-flow, leading to an estimated annual cost reduction of approximately \$40 under continuous operation. These savings can also be directly scaled based on alternative energy prices and adjusted proportionally for similar channels of different widths. Overall, this method offers an alternative path for more efficient and sustainable desalination, with potential for reduced energy use and operational costs.

5.2 Recommendations for Future Work

The economic feasibility of this approach ultimately depends on the notional permeate value, w , which is determined by geography, scarcity, and governmental policy. A robust economic analysis is needed to quantify w , which can then be used in conjunction with the trends in this study. Future work should refine these estimates to better guide implementation. This study focused on a 300-mm channel, similar to small-scale residential units, and brackish water feed. While these conditions provide a starting point, exploring other scenarios could further explore the approach. In this study, the dual impermeable segments model was examined under the constraint that both segments possess equal lengths ($\beta_1 = \beta_2$). Future studies are recommended to explore configurations with unequal segment lengths ($\beta_1 \neq \beta_2$), as such variations may offer enhanced performance and reveal unobserved advantages associated with non-uniform segment distributions.

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