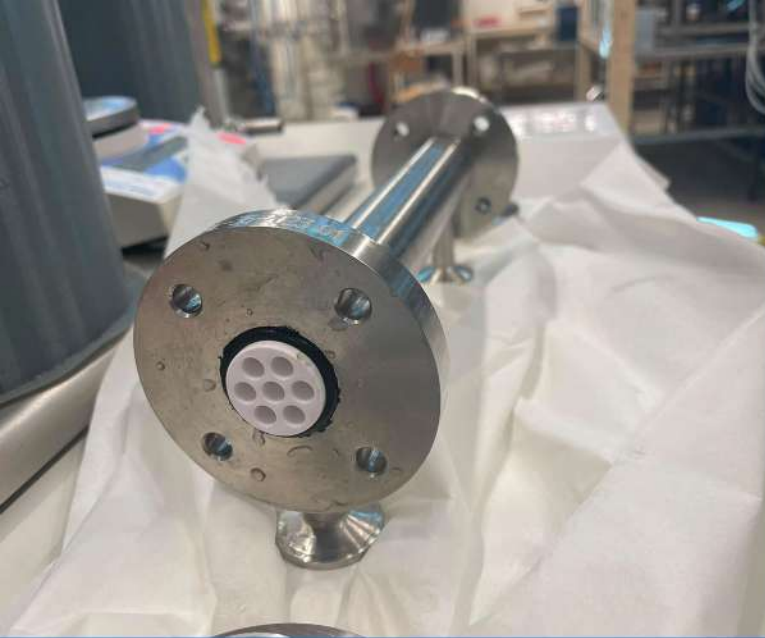




Thin Beet Juice Concentration With Membrane Distillation

DEPARTMENT OF PROCESS AND LIFE SCIENCE ENGINEERING | LUND UNIVERSITY
DAVID ETIDO AWAK | MASTERS THESIS 2026



Master in Membrane Engineering for Sustainable Development (MESD)
Erasmus Mundus Joint Master



Master Thesis

**Vacuum Membrane Distillation for
concentration of Thin Beet SugarJuice
in Swedish Beet Sugar Refining**

by

Awak David Etido

Division of Chemical Engineering
Department of Process and Life Science Engineering
In collaboration with Nordic Sugar AB Ortofta, Member of Nordzucker Group
Lund University, Sweden

June, 2026

Supervisor: Mikael Sjölin and Hallbeck Johan (Production/Process Engineer)

Co-supervisor: Omid Arbabzadeh

Examiner: Frank Lipnizki

Postal address

P.O. Box 124

SE-221 00 Lund, Sweden

Web address

<http://www.ple.lth.se>

Visiting address

Kemicentrum

Naturvetarvägen 22

Telephone

+46-222 82 85

+46 46-222 00 00

223 62 Lund, Sweden

Program Funding and Affiliation

This master thesis project has been conducted as part of the Erasmus Mundus Joint Master Degree in *Master in Membrane Engineering for Sustainable Development*, funded by the European Union.

The author thanks the European Commission-Education, Audiovisual and Culture Executive Agency (EACEA) for the Erasmus Mundus scholarship under the program: Master in Membrane Engineering for Sustainable Development, (MESD), Project number: 101081874, Call: ERASMUS-EDU-2022-PEX-EMJM-MOB

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Preface

This thesis represents the culmination of two years of advanced study and reflects the knowledge, practical skills, and problem-solving abilities developed through dedication, perseverance, and continuous learning. I am deeply grateful for the opportunity to have studied in a stimulating academic environment where impactful research is conducted daily, guided by inspiring scholars and supported by a culture that fosters curiosity and innovation. While this work marks the conclusion of one chapter, it also signifies the beginning of another, one in which I aspire to contribute meaningfully to society through knowledge, inquiry, and innovation. My academic journey in the Erasmus Mundus Master's Programme in Membrane Engineering for Sustainable Development (MESD), together with all partner institutions, reaches its conclusion with this thesis. I would like to express my sincere appreciation to Johan Hallbeck and the entire team at Nordic Sugar AB, Örtofta, Sweden, for their collaboration. This research was conducted in the Pilot Hall at Kemicentrum, Lund University. I would like to extend my heartfelt gratitude to my supervisor, Mikael Sjölin, for his expert guidance and mentorship throughout this project. I am also sincerely thankful to my co-supervisor, Omid Arbabzadeh, for his valuable input, constructive feedback, and active involvement throughout the research process.

My sincere appreciation also goes to Helene von Wachenfeldt, MESD International Coordinator at Lund University, for her dedicated support throughout the programme. I am grateful to my fellow MESD colleagues and friends for their encouragement and shared experiences during this academic journey. To my beloved family, Mr. Etido Awak, Mrs. Comfort Awak, Ekemini Awak, Ndiuwem Awak, Uzoma Ibeanu, and Ofonmbuk Awak thank you for your unwavering love, prayers, and belief in me. Above all, I give thanks to GOD Almighty, whose grace, guidance, and faithfulness have sustained me through every challenge and every achievement along this journey.

Generative Artificial Intelligence has also been used to a limited degree in the making of this thesis to improve language, formulations and structure. In a few instances it has been used to suggest sources of information to further an argument, however this has always been accompanied by scrutiny of other literature and the final text by the authors own interpretation of the source.



Popular Scientific Summary

Beet juice contains a lot of water that must be removed during sugar production. This step requires large amounts of energy and is one of the most energy-intensive stages in the process. However, high temperatures are typically used to evaporate the water, which can also affect the juice quality.

Although the Swedish sugar industry is already energy efficient, further reductions in energy use are required to achieve climate-neutral sugar production by 2050. To reduce the energy demand, researchers are investigating membrane technologies as alternatives to conventional evaporation. While some membrane methods have been successful in other food industries, they have shown limited performance for beet juice due to sugar losses, membrane fouling, and high pressure requirements.

One promising alternative is vacuum membrane distillation (VMD). In this process, the juice is gently heated and water vapour passes through a water-repellent barrier under vacuum conditions. This removes water while retaining sugar and other valuable compounds.

A collaborative study between Nordic Sugar AB and Lund University tested a VMD systems using thin beet juice from the Ortofta sugar mill. The results showed that VMD can efficiently concentrate beet juice while preserving its valuable components. Both experimental and theoretical studies suggest that the technology can reduce energy use and lower the heat requirement for sugar production.

VMD has the potential to improve energy efficiency and support more sustainable sugar production. It could also enable the use of low-grade or waste heat within sugar factories. However, the process still has relatively low water-removal rates, and further optimization and industrial-scale testing are needed before large-scale implementation. Despite these challenges, VMD shows strong potential to contribute to sustainable sugar production.

Summary

This thesis investigates vacuum membrane distillation (VMD) as an energy-efficient method for concentrating thin beet sugar juice. It evaluates the effects of operating conditions on concentration performance, membrane selectivity, product quality, and process stability.

Pilot-scale experiments were conducted using 5000 mL of thin beet sugar juice at each batch. The system was operated at varying temperatures of 70, 75, and 80 °C, vacuum pressures of 27, 95, and 105 mbar, and feed flow rates of 6.7, 8.4, and 12.8 L/min. Two ceramic membrane configurations were evaluated: a 7-channel membrane (0.156 m²) and a 19-channel membrane (0.222 m²).

The 7-channel membrane was limited to 70 °C because of severe concentration polarization. Under these conditions, the juice concentration reached 9.10 °Brix after 3 hours at a 6 hour run. The 19-channel membrane showed better performance under optimal hydrodynamic conditions of 75 °C, 95 mbar, and 8.4 L/min. At this condition, thin juice was concentrated from 11.80 to 15.54 °Brix after 3 hours. The juice quality was maintained above 90 %, with acceptable colour levels in the range of 2707 to 3027 ICUMSA. The improved performance attributes to a higher membrane surface-area to volume ratio at a Reynolds number of 1795 within the laminar flow regime.

The membrane maintained high selectivity throughout the experiments achieving low permeate conductivity, an average flux of 1.41 L/m²·h, and 96.11% ion retention. These results demonstrate efficient water removal and stable process performance. Despite fouling and scaling, the cleaning protocol successfully restored 98% of the membrane performance. The recovery on performance of the membrane after cleaning highlights the robustness of ceramic membranes for VMD applications.

Furthermore, the 19-channel membrane provided the best overall performance under the tested conditions. This highlights the potential of VMD to effectively concentrate thin beet sugar juice. These findings indicate that VMD has potential for integration into sugar factories, particularly when combined with low-grade waste heat. However, further process optimization, long-term testing, and improved membrane and module designs are required before industrial-scale implementation.

Keywords: Vacuum Membrane Distillation (VMD), Ceramic Membranes, Thin Beet Sugar Juice, Concentration Polarization, Membrane Fouling, Internal Pore Scaling, Sugar Concentration, Process Optimization.

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1. Introduction

1.1. Background

Sugar is one of the world's most important agricultural commodities, with approximately 188 million metric tons produced annually (Abadam, 2025). Global sugar production is derived primarily from sugarcane, which accounts for nearly 80 % of total production, while the remaining 20 % originates from sugar beets (A. Shirazi and Kargari, 2019; Criscuoli and Drioli, 2020; Lipnizki and Figueroa, 2021).

Despite its economic importance, beet sugar production remains highly energy-intensive due to the high water removal rate during juice concentration with conventional steam-driven evaporation processes (Misra et al., 2022; Muir, 2022). Consequently, there is increasing interest in developing alternative concentration technologies that can reduce high energy demands and environmental impacts while preserving sugar quality (Criscuoli and Drioli, 2019).

Among these emerging alternatives, membrane distillation (MD) has gained considerable attention as an energy-efficient separation technology. In recent years, MD has been successfully investigated for a variety of applications across the food and water industries (Wang and Chung, 2015). This technology combines thermal evaporation with membrane-based separation, allowing water vapour to pass through a hydrophobic membrane while retaining dissolved solutes. Unlike pressure-driven processes such as reverse osmosis, VMD uses a vapour pressure difference generated by heat and vacuum to achieve separation (Do et al., 2025; Francis et al., 2022; Omar et al., 2022).

Vacuum Membrane Distillation (VMD) operates by applying a vacuum on the permeate side of the membrane, which is maintained below the saturation pressure of the volatile species in the feed solution (Kariman et al., 2026). This creates a large vapour pressure gradient across the membrane, providing a stronger driving force for mass transfer than Direct Contact Membrane Distillation (DCMD), Air Gap Membrane Distillation (AGMD), and Sweep Gas Membrane Distillation (SGMD) (Mericq et al., 2010). As a result, VMD generally achieves higher vapour fluxes and enhanced separation performance. Consequently, it is widely regarded as the most promising membrane distillation configuration for the concentration and purification of aqueous solutions (Boubakri et al., 2024; EL-Bourawi et al., 2007).

Furthermore, VMD achieves excellent solute rejection and improved thermal efficiency while operating at relatively low temperatures and hydrostatic pressures (Alsebaei and Ahmad, 2020; Chen et al., 2014; Kim et al., 2021). Additionally, the absence of a liquid phase on the permeate side minimizes conductive heat losses (Chiam and Sarbatly, 2013).

During VMD operation, volatile components absorb latent heat and undergo a phase transition, and the resulting vapour is transported through the membrane pores via a combination of Knudsen diffusion, viscous (Poiseuille) flow, and convective heat transfer before condensing in the permeate stream (Gryta, 2012; Ko et al., 2018). Simultaneously, the hydrophobic

characteristics of the membrane material and capillary forces prevent liquid penetration into the membrane pores (Alsaadi et al., 2018; R      et al., 2014). This resistance to pore wetting is quantified by the Liquid Entry Pressure (LEP), as described by the Laplace–Young equation from Equation 1, which defines the minimum pressure required to maintain selective vapour-phase transport (Khayet, 2011):

$$P_{f(\text{liquid})} - P_{p(\text{vapour})} = \frac{-2 \cdot \gamma \cdot \cos \theta}{r_{\max}} = \Delta P_{(\text{entry})} \quad (1)$$

Where γ represents the liquid surface tension, θ is the contact angle, and $r_{\max}$ is the maximum pore radius.

The performance of VMD is often constrained by temperature polarization, a phenomenon arising from the formation of thermal boundary layers that creates resistance to heat transfer (Suleman et al., 2021). This limits the effective temperature difference across the membrane interface, consequently reducing the vapour pressure gradient (Do et al., 2025; Francis et al., 2022).

Historically, hydrophobic polymeric membranes, particularly fluoropolymers such as polytetrafluoroethylene (PTFE), polyvinylidene fluoride (PVDF), and polypropylene (PP) have dominated both commercial and laboratory-scale MD applications (Drioli et al., 2015). Their widespread use is attributed to their inherently low surface energy, excellent hydrophobicity, and high permeate fluxes, which facilitate the design of compact and energy-efficient membrane systems (Reza Shirzad Kebria and Rahimpour, 2020).

Membrane distillation (MD) has traditionally relied on hydrophobic membranes because their high water repellency promote efficient vapour transport. More recently, ceramic membranes have emerged as attractive alternatives due to their superior chemical, thermal, and mechanical stability (Alawad et al., 2023; Al-Jariry, 2025; Samavati et al., 2025). In addition to MD, ceramic membranes are widely used in microfiltration, ultrafiltration, and nanofiltration applications and are commercially available in both single-channel and multi-channel configurations (Hakami et al., 2020).

However, unlike inherently hydrophobic polymers, ceramic membranes are hydrophilic and interact strongly with water molecules due to surface hydroxyl groups. Through chemical surface modification with fluorinated silane-based compounds, these ceramic surfaces can be made hydrophobic (Alessandro et al., 2023). Furthermore, ceramic membranes can be manufactured in different geometries, offering varying surface-area to volume ratios and permeance characteristics that can be tailored to specific process requirements (Abadi et al., 2011; Das and Maiti, 2009; Xu et al., 2023). Although ceramic membranes generally exhibit lower permeate fluxes than polymeric alternatives, this behaviour has largely been attributed to the higher mass-transfer resistance of the porous support layer. (Chen et al., 2018; Hubadillah et al., 2019).

As solutes accumulate at the membrane interface, the feed-side vapour pressure decreases, reducing the effective driving force for mass transfer and consequently lowering process

performance (Chen et al., 2020). To overcome these limitations, VMD performance must be optimized through appropriate operating conditions, including feed temperature, cross-flow velocity, and transmembrane pressure. These parameters influence vapour transport, heat transfer, and concentration polarization effects (Tai et al., 2020).

In addition, membrane module design such as tubular, hollow-fibre, spiral-wound, and plate-and-frame modules directly affect system hydrodynamics and overall separation efficiency (Alessandro and Macedonio, 2025). The development of advanced ceramic membrane systems for VMD continues to attract significant research interest, particularly since promising results have been reported for the concentration of fruit and sugarcane juices (Doherty et al., 2010; Sarbatly et al., 2023).

The largely unexplored application of VMD to thin beet sugar juice highlights a critical research gap requiring further systematic investigation.

1.2. Beet sugar Production

In conventional beet sugar manufacturing, the concentration of clarified thin juice with approximately 15 % dissolved solids to thick juice with a dry substance of approximately 70 % relies on multi-effect evaporators operating at elevated temperatures (Lipnizki and Figueroa, 2021) .

This conventional arrangement represents the primary energy and quality bottleneck of the sugar factory, driven concurrently by excessive energy consumption. (Asadi, 2006; Clarke, 1987.). The sugar-beet process involves washing, slicing, and thermally extracting the raw juice. It is then purified using lime and CO₂, heavily concentrated via high-energy multi-effect evaporation with steam. Finally, it is subjected to further evaporation for crystallization and centrifugation (Rafik et al., 2015). Specifically, prolonged exposure to elevated temperatures at the evaporator heating surfaces promotes destructive Maillard-type reactions, sucrose inversion, and caramelization (Misra et al., 2022; Van Der Poel et al., 1998). This dual thermodynamic and chemical constrain results in severe juice colour formation that ultimately impairs downstream crystallization performance and necessitates chemical mitigation using sulfur dioxide (SO₂) (Muir, 2022).

To reduce energy consumption and minimize thermal degradation, vacuum membrane distillation (VMD) has emerged as a promising alternative for thin juice concentration. VMD operates at temperatures below 100 °C and uses a vapour pressure gradient under vacuum conditions to separate water from the thin juice. This approach reduces the reliance on high-pressure steam and enables the utilization of low-grade industrial waste heat. Consequently, VMD offers the potential to lower the thermal footprint of sugar production while preserving juice quality and limiting colour formation

Table 1.1: Typical ranges of colour seen in sugar beet processing of different sucrose solution as represented with the ICUMSA Unit (IU)(Asadi, 2006)

Samples	Colour (IU)
Thin juice	1500 – 3000
Thick juice	2000 – 3500
Molasses	40000 – 70000
Sugar	15 – 45

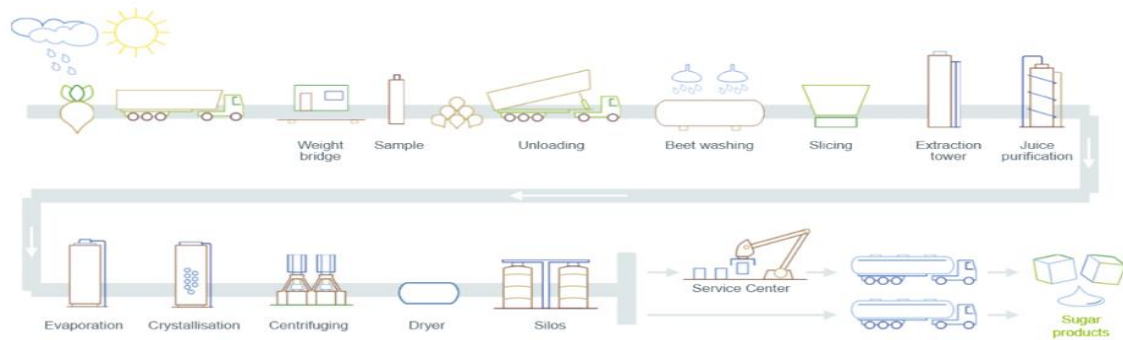


Figure 1.1. Flow diagram of sugar production with permission from Nordzucker company presentation.,2025.

1.3. Project Description

While membrane distillation has demonstrated considerable potential as an energy-efficient concentration technology, its application to thin beet sugar juice remains largely unexplored. This lack of application-specific research, particularly for vacuum membrane distillation (VMD), constrains the development of innovative process solutions aimed at reducing energy consumption in the sugar industry. To address this gap, a collaborative master's thesis between Lund University and Nordic Sugar AB evaluates and optimizes VMD for concentrating thin beet sugar juice from 16 to 35 wt.% dry substance as an alternative to multi-effect evaporation. This study will integrate theoretical analysis with pilot-scale experiments on industrial thin beet sugar juice samples to evaluate operational behaviour, scalability, and process feasibility. This will mitigate previously reported limitations which includes temperature polarization, membrane wetting, module-induced flux constraints, and fouling-related flux decline. This study evaluates operating conditions and ceramic membrane configurations to maximize permeate flux and sucrose retention while minimizing energy consumption, supporting the industrial implementation of VMD in beet sugar factories.

2. Aim and Objectives

This project aims to develop and evaluate an optimized vacuum membrane distillation (VMD) system for concentrating thin beet sugar juice in ranges of 30 to 35 wt.% dry substance, corresponding to a volume concentration factor greater than 2. The goal is to achieve high and stable permeate flux while providing an energy-efficient alternative or complement to conventional evaporation.

The study will address key VMD challenges, including temperature polarization, membrane wetting, flux decline, fouling, and module limitations. This will be achieved through optimization of operating conditions, evaluation of multi-channel ceramic membranes, and assessment of fouling and cleaning strategies to support low-grade heat integration while maintaining sucrose retention and juice quality.

2.1. Research Questions

- i. How does VMD influence permeate flux, fouling behaviour, and juice quality during thin beet sugar juice concentration?
- ii. How do ceramic membrane configurations affect flux, selectivity, and fouling resistance?
- iii. What are the optimal operating conditions for maximizing concentration performance and permeate flux?
- iv. What degree of juice concentration and sucrose retention can be achieved during extended VMD operation?

2.2. Limitations

This study evaluates the feasibility of vacuum membrane distillation (VMD) for concentrating thin beet sugar juice in the sugar industry. The experimental work was limited to industrial thin juice samples collected from the Örtöfta sugar factory during the 2025 production campaign. Furthermore, this study does not include full techno-economic analysis or complete process integration assessments which are limited to qualitative discussions.

3. Literature Review

3.1. Influence of Membrane Material on Performance.

Hydrophobic membrane materials are central to the performance of vacuum membrane distillation (VMD), as they determine vapour transport, LEP, fouling resistance, and long-term operational stability (Hubadillah et al., 2019). Polymeric membranes such as PTFE, PVDF, and PP are widely used in membrane distillation due to their low surface energy and strong hydrophobicity (Defor and Chou, 2024; Kim et al., 2021). Consistent with these properties, Zhang *et al.*, (2016) demonstrated that PTFE membranes delivered high water fluxes and retained more than 99.9 % of non-volatile solutes, producing near-pure water permeate. However, polymeric membranes are susceptible to pore wetting, fouling, and thermal degradation under harsh operating conditions (Xu et al., 2023).

To overcome these limitations, increasing attention has been directed towards hydrophobic modified ceramic membranes based on alumina (Al_2O_3), titania (TiO_2), and zirconia (ZrO_2) (Cong et al., 2019; Ramlow et al., 2019). Unlike polymeric membranes, ceramic membranes are sought for their mechanical, chemical and thermal stability and can tolerate operation in harsh conditions, making them attractive for industrial applications involving concentrated sucrose solutions (Liu et al., 2023). Schnittger *et al.*, (2021) and Ko *et al.*, (2018) demonstrated that fluoroalkyl-silane (FAS) modified ceramic membranes achieved high fluxes while maintaining excellent salt rejection and high thermal stability. Nevertheless, support-layer resistance, thermal conductivity, and fabrication cost remain important challenges for large scale implementation (Qi et al., 2025).

3.2. Effect of Feed Temperature on Permeate flux

Feed temperature is widely recognized as the dominant operating parameter influencing VMD performance because vapour pressure increases exponentially with temperature according to the Antoine equation (Suleman et al., 2020). Studies according to Al-Jariry *et al.*, (2026) reported fluxes above 193 $\text{kg/m}^2 \text{ hr.}$ at 60 °C using PTFE membranes, while Lee *et al.*, (2025) observed fluxes exceeding 200 $\text{kg/m}^2 \text{ hr.}$ at 110 °C using hydrophobic modified ceramic membranes.

Similarly, Ricceri *et al.*, (2022) demonstrated that increasing the temperature from 35 to 65 °C increased flux approximately from 3 $\text{kg/m}^2 \text{ hr.}$ to 22.5 $\text{kg/m}^2 \text{ hr.}$ Despite these benefits, increasing temperature also intensified temperature polarization (TP), fouling, and membrane wetting risks (Rezaei et al., 2018). This was further reported by Persson, (2025) that higher temperatures produced less permeate flux due to severe TP near the membrane surface. Furthermore, increasing the feed temperatures significantly increased permeate flux but studies by Al-Asheh *et al.*, (2006) highlighted that high temperatures can accelerate organic deposition and reduce long-term process stability.

Therefore, while higher temperature generally improves mass transfer and fluxes, optimization is required to balance productivity, fouling control, and membrane performance (Boubakri et al., 2024).

3.3. Effect of Cross-flow Velocity and Hydrodynamics

Cross-flow velocity is a key operating parameter in VMD because it influences heat and mass transfer at the membrane surface (Y. S. Chang et al., 2021). Higher velocities enhance fluid mixing and shear forces, reducing temperature polarization (TP) and concentration polarization (CP), which are major limitations to permeate flux and process efficiency (Alawad et al., 2023). As a result, optimized hydrodynamic conditions can improve flux, reduce fouling, and enhance system stability (Abu-Zeid et al., 2015).

Schnittger *et al.*, (2021) demonstrated that increasing feed velocity from 0.14 to 1.08 m/s significantly improved permeate flux in hydrophobic alumina (Al_2O_3) and titania (TiO_2) tubular ceramic membranes, particularly at high salt concentrations. Similarly, Pan *et al.*, (2022) reported that turbulent flow improves mixing near the membrane surface and reduces polarization effects. However, TP and CP remained significant limitations even under turbulent conditions, indicating that hydrodynamic optimization alone cannot fully eliminate heat and mass transfer resistances (Alawad and Khalifa, 2023).

Similar findings were reported by Ricceri *et al.*, (2022), who observed that increasing the cross-flow velocity from 0.21 to 0.42 m/s reduced boundary-layer thickness and delayed cake layer formation in humic acid solutions. Because fouling layers introduce additional thermal resistance, their reduction can also mitigate temperature polarization (Ricceri et al., 2023). These findings are particularly relevant to thin beet sugar juice concentration, where organic compounds, colloids, and colour precursors can accumulate at the membrane interface and promote fouling.

The influence of cross-flow velocity on concentration polarization was further demonstrated by Liu *et al.*, (2022), who investigated the regeneration of concentrated lithium chloride solutions using PTFE membranes. Increasing the flow velocity from 0.02 to 0.06 m/s improved bulk mixing and reduced boundary-layer resistance. However, excessively high hydrodynamic conditions can promote supersaturation near the membrane surface, leading to crystallization, scaling, and an increased risk of wetting (Liu et al., 2025; Wang et al., 2023). These findings suggest that an optimal cross-flow velocity is required to minimize polarization effects without accelerating scale formation (Rezaei et al., 2018). This consideration is particularly relevant for thin beet sugar juice concentration, where increasing solute concentration and viscosity can intensify polarization effects and fouling.

The beneficial effects of cross-flow velocity are closely related to its influence on Reynolds number and boundary-layer development (Ricceri, 2022). At low velocities, laminar flow promotes the formation of thick thermal and concentration boundary layers, intensifying TP and CP, increasing resistance to heat and mass transfer (Kuang et al., 2019; Yu et al., 2011). As velocity increases, enhanced mixing reduces boundary-layer thickness, bringing the

membrane surface conditions closer to those of the bulk feed and increasing the effective vapour-pressure gradient across the membrane (Dong et al., 2021; Sun et al., 2015). This improves permeate flux and thermal efficiency. In addition, higher shear stresses limit foulant accumulation at the membrane surface, reducing cake formation and the risk of membrane wetting (Horseman et al., 2021; Roslee AB. Jamal and Alias, 2025).

Despite these benefits, excessively high cross-flow velocities present practical limitations. Higher flow rates increase pumping energy requirements, which can reduce the overall energy efficiency of VMD (Alawad et al., 2024). In addition, excessive shear may promote concentration effects, crystallization, scaling, and mechanical stress within membrane modules (Do et al., 2025). Therefore, cross-flow velocity must be optimized to balance mass transfer enhancement with energy consumption and fouling control (Alkhudhiri et al., 2012). For thin beet sugar juice concentration, an optimal velocity is required to sustain high permeate flux while minimizing polarization, fouling, and long-term membrane deterioration.

3.4. Effect of Vacuum Pressure and Transmembrane Pressure on VMD performance

Vacuum pressure is a key operating parameter in VMD because it determines the vapour-pressure gradient driving mass transfer (Yang et al., 2019). Higher vacuum levels generally increase permeate flux by enhancing evaporation. However, excessive pressure differences can increase the risk of membrane wetting by approaching the liquid entry pressure (LEP) (RÁCZ et al., 2015). Therefore, vacuum conditions must be optimized to maximize flux while preserving membrane integrity. Chen *et al.*, (2018) investigated sugar solution concentration using VMD and found that lower permeate pressures increased water flux by enhancing the vapour-pressure driving force, consistent with the findings of Sun *et al.*, (2014).

However, Pyo *et al.*, (2024) reported that the increase in flux was lower than theoretically expected because of temperature polarization (TP) near the membrane surface. These findings suggest that although vacuum pressure is a critical operating parameter, its benefits can be limited by heat-transfer resistances (Bindels et al., 2021). Similar findings were reported by Kim *et al.*, (2021) who observed that reducing permeate-side pressure increased transmembrane mass transfer and permeate flux in PTFE membranes. This improvement was attributed to enhanced water vapour removal from the membrane pores, which maintained a higher vapour-pressure driving force during operation (Pei et al., 2021). However, excessively strong vacuum conditions may increase the risk of membrane wetting, particularly when processing feeds containing surfactants or other surface-active compounds (H. Chang et al., 2021; Liao et al., 2021).

Likewise, Lee *et al.*, (2025) reported that increasing vacuum levels improved permeate flux in hydrophobic modified alumina ceramic membranes by enhancing vapour transport through the membrane pores. The superior thermal and mechanical stability of ceramic membranes also enables operation under more demanding vacuum conditions (Zhou et al., 2026).

Nevertheless, maintaining sufficient membrane hydrophobicity remains essential to prevent liquid penetration and ensure long-term process stability. More recently, Al-Jariry *et al.*, (2026) reported high water fluxes and excellent solute rejection at a vacuum pressure of approximately 50 mbar, demonstrating the effectiveness of low permeate pressures in enhancing mass transfer and water recovery. However, membrane performance depends on maintaining operating pressures below the liquid entry pressure (LEP) to prevent pore wetting and loss of separation efficiency (Guillen-Burrieza *et al.*, 2016; Rácz *et al.*, 2014)

The LEP is not a fixed membrane property and can be significantly influenced by operating temperature (Varela-Corredor and Bandini, 2020). Increasing feed temperature reduces the liquid surface tension, which may decrease the LEP and increase the likelihood of membrane wetting under a given transmembrane pressure difference (Khayet and Matsuura, 2011). Therefore, feed temperature and operating pressure should be jointly optimized to maximize permeate flux while maintaining membrane stability and separation performance.

Exceeding the LEP can cause liquid penetration into membrane pores, leading to partial wetting, reduced selectivity, and deterioration of membrane performance (Ye *et al.*, 2024). Consequently, the LEP is considered a critical design parameter for ensuring long-term membrane stability. Similarly, Xie *et al.*, (2025) reported that excessive pressure differences can destabilize the liquid–vapour interface and increase wetting, particularly when processing concentrated solutions. This effect further reduces separation efficiency and accelerates performance decline during prolonged operation (Guillen-Burrieza *et al.*, 2016). Collectively, these studies demonstrate that vacuum and transmembrane pressures must be carefully optimized to maximize permeate flux while maintaining a sufficient safety margin below the membrane’s LEP.

3.5. Temperature Polarization and Concentration Polarization

Temperature polarization (TP) and concentration polarization (CP) are major transport limitations in VMD because they reduce the effective vapour pressure gradient across the membrane (Ni *et al.*, 2022). TP occurs when heat loss by evaporation lowers the membrane surface temperature relative to the bulk feed, whereas CP results from solute accumulation at the membrane interface (Suleman *et al.*, 2021). Together, these phenomena limit mass transfer, reducing permeate flux and overall process efficiency, particularly in large-scale VMD applications (Boey *et al.*, 2025).

Liu *et al.*, (2024) demonstrated that increasing feed flow rate and optimizing vacuum pressure enhanced mass transfer by maintaining a higher vapour-pressure gradient across the membrane. Similarly, Ricceri *et al.*, (2022) reported that improved hydrodynamic conditions reduced concentration build-up and fouling at the membrane surface, helping to sustain permeate flux over extended operation. These findings indicate that concentration polarization not only reduces the driving force for vapour transport but also promotes fouling and scaling, further compromising membrane performance (Horseman *et al.*, 2021).

Temperature and cross-flow velocity are key operating parameters affecting TP and CP. Higher feed temperatures increase permeate flux but increase TP, whereas higher cross-flow velocities reduce polarization by enhancing heat and mass transfer (Alawad et al., 2024; Eykens et al., 2017; Politano et al., 2019). Although increasing vacuum pressure enhances the vapour-pressure driving force, excessive vacuum may intensify heat-transfer limitations at high fluxes (Anqi, 2026; Onsekizoglu, 2012). Consequently, operating conditions must be optimized to balance flux and polarization effects.

3.6. Fouling, Wetting, and Cleaning Strategies

Fouling and membrane wetting are major challenges in VMD, leading to flux decline and reduced membrane lifespan (Kayvani Fard et al., 2016). Fouling results from the accumulation of contaminants on the membrane surface, which increases mass-transfer resistance, whereas wetting occurs when liquid penetrates the membrane pores, compromising hydrophobicity and separation performance (Pal, 2015; Wang et al., 2025).

Studies have shown that wetting and fouling are strongly influenced by operating conditions, pressure gradients, membrane hydrophobicity, and membrane material properties (Pyo et al., 2024; Xie et al., 2025). Consequently, selecting highly hydrophobic membrane materials is critical for improving resistance to fouling, minimizing wetting, and maintaining long-term VMD performance (El Charif et al., 2025; Xu et al., 2023; Zhao et al., 2018)

Several studies have investigated strategies to restore membrane performance and mitigate fouling in VMD. Popović *et al.*, (2009) reported that alkaline cleaning effectively removed organic foulants and restored flux, while Gruskevica *et al.*, (2021) demonstrated that sequential alkaline, oxidative, and surfactant cleaning improved membrane recovery and operational stability. In addition, Chamani *et al.*, (2021) and Lv *et al.*, (2023) showed that drying and solvent-based regeneration techniques help maintain membrane hydrophobicity and prolong membrane lifespan. Despite these advances, fouling and membrane wetting remain the primary causes of long-term flux decline in VMD systems (Do et al., 2025). Therefore, sustainable operation requires a combination of hydrophobic membrane modification, optimized hydrodynamic conditions to limit concentration polarization, and periodic cleaning protocols to restore membrane performance and minimize wetting risks (Shin et al., 2017; Zhao et al., 2018). Furthermore, restoring flux and tackling complex scaling requires rigorous, sequential maintenance protocols, including periodic cleaning, ethanol regeneration, alkaline treatments, and oxidative-surfactant cleaning sequences (Chen et al., 2008; Gruskevica and Mezule, 2021; Gul et al., 2022).

4. Materials and Methods

4.1. Thin Beet Sugar Solution

At the end of the 2025 beet campaign at the Örtöfta sugar processing plant, thin beet sugar juice samples were supplied by Nordic Sugar AB. The thin beet sugar juice was collected just prior to evaporation from a sampling site located after alkali addition and pH adjustment. To soften the juice, it was immediately treated with Amberlite IRA 200C (Na⁺), a strong acid cation-exchange resin in the sodium form. For experimental use, the thin beet sugar juice was frozen and later thawed as required. The remaining unthawed juice was stored in a refrigerator and utilized strictly for the duration of the experiments.

4.2. Membrane Material

The pilot trial employed Fraunhofer ceramic membranes in a VMD system. The membranes comprised an alumina (Al₂O₃) substrate modified with a fluorinated alkylsilane coating to provide hydrophobicity, allowing selective vapour transport while preventing liquid penetration.

Table 4.1: Ceramic membrane material specifications.

Configuration	7-Channel	19-channel
Pore size	100-200 nm	100-200 nm
Outer diameter	25 mm	25 mm
Inner diameter	6 mm	3.1 mm
Filtration Area	0.156 m ²	0.222 m ²
Total length	1.2 m	1.2 m

The membranes exhibit a robust thermal stability and sustained a stable operation up to 80 °C under liquid operating conditions and likely higher in dry states, which aligned with the thermal demands of the VMD processes. The system operating limits were further defined by the stainless-steel module housing, which is rated up to 100 °C at 25 bars.

The LEP of the modified ceramic material exceeds 7 bar at room temperature and remains approximately 6 bar at 80 °C. This high structural resistance to liquid penetration ensures that pore wetting is effectively minimized under typical VMD vacuum conditions.

Standard cleaning protocols involve rinsing the membrane surface thoroughly with de-ionized (DI) water, followed by thermal drying at approximately 120 °C. In the event of pore wetting, a recovery protocol is initiated where DI water is flushed directly through the membrane wall before the drying stage is repeated.

4.3. Experimental Setup and Module Configuration

As shown in Figure 4.1, the ceramic membrane was installed in a tubular module within a pilot-scale vacuum membrane distillation (VMD) system supplied by MembranProTec (Hultgren Miljö & HälsoDesign AB). The pilot plant was equipped with a multistage feed pump, an electric in-line heater, a vacuum pump, a condenser, and a permeate collection tank. On-line sensors monitored feed temperature, conductivity, system pressure, and permeate tank level. Prior to each experimental run, the pump flow rate was calibrated using a measuring cylinder and timer with the membrane module installed. To regenerate fouled or wetted membranes, the following Clean-in-Place (CIP) and drying protocol was applied:

- i. The system was rinsed with deionized (DI) water for 1 hour.
- ii. The membrane was cleaned with Ultrasil-110 (Ecolab AB), a non-chlorinated alkaline cleaner, at 60 °C for 2 hours.
- iii. The system was subsequently flushed with DI water until the permeate conductivity decreased below 5 $\mu\text{S}/\text{cm}$.
- iv. The membrane was removed from the module and submerged in 99 % (v/v) ethanol for 3 hours to remove inorganic, organic, and biological foulants.
- v. After ethanol treatment, the membrane was drained vertically for 15 minutes and oven-dried at 45 °C for 3 hours.
- vi. The membrane was then reinstalled in the module and rinsed with DI water before reuse.
- vii. Finally, the baseline pure water flux (PWF) was measured to verify membrane permeability and re-establish clean baseline performance.

Prior to introducing the thin beet sugar juice, the system was pre-conditioned and fractions of the raw thin beet sugar juice was recirculated through the system for 30 minutes and later discarded. This purged the residual DI water from the dead volumes and prevented feed dilution. After emptying the purge volume, a fixed 5000 mL batch of thin beet sugar juice feed was introduced into the tank for a standard 7-hr concentration run.

During the experimental runs, key process parameters were recorded at fixed intervals to monitor system stability and separation performance. At every 30 minutes, feed temperature, outlet temperature, system pressure, stream conductivities, and refractometric dry substance (RDS) for both the permeate and return streams were logged in. Afterwards, the colour, product purity, and pH of the collected sample from the return stream were analysed hourly.

The experimental investigation was conducted in a three-step optimization matrix to evaluate the membrane performance and identify the optimal operating conditions for concentration. Firstly, a thermal baseline and wetting susceptibility assessment was performed by gradually increasing the feed temperature from 70 °C, 75 °C, and 80 °C while maintaining a constant volumetric flow rate, and vacuum pressure.

This phase aimed to determine the membrane's resistance to pore wetting and establish a suitable operating temperature range. Next, a hydrodynamic optimization study was conducted

using the optimal temperature identified in the first phase. The flow rate was systematically varied at 6.7 L/min, 8.44 L/min, and 12.8 L/min respectively to determine the flow conditions that maximize mass transfer while minimizing concentration polarization effects. Finally, a critical vacuum pressure assessment was carried out using the pre-optimized temperature and volumetric flow rate conditions. In this phase, vacuum pressures of 27, 95, and 105 mbar were systematically varied to evaluate their influence on membrane performance and establish operating conditions that avoided significant scaling and membrane wetting



Figure 4.1. VMD configurations installed in tubular module with Human-Machine interface (HMI) screen (operator interface) for the pilot plant.

4.4. Refractive Dry Substance Measurement

A temperature-compensated Hanna Instruments HI-96801 sucrose refractometer was used to determine the refractive dry substance (RDS) of permeate and retentate samples. Instrument calibration was verified with deionized (DI) water according to ICUMSA guidelines.

4.5. Sucrose Purity and Polarimetry

Polarimetric sucrose content and juice purity were determined according to ICUMSA methods using an Automatic Polarimeter AA-100 (Optical Activity Ltd.). Thin juice samples of 13 grams were diluted to 100 mL with deionized (DI) water, filtered through a cellulose filter, and analyzed at a wavelength of 589 nm. For highly coloured samples, additional dilution was applied to ensure acceptable light transmittance. The instrument was calibrated using a standard sucrose solution and calibrated with DI water during each analytical session. Polarimetric sucrose values ($^{\circ}\text{Z}$) were converted to sucrose purity (%) by relating the measured sucrose content to the sample refractive dry substance (RDS).

4.6. Colour Spectrophotometry

Colour development was evaluated by measuring absorbance at 420 nm using a 1 cm cuvette, with deionized (DI) water as the reference. Samples with absorbance values greater than 1 were diluted 1:1 (v/v), and the measurements were corrected for dilution. Thin juice samples were first adjusted to pH 7.0 using 0.1 M hydrochloric acid and filtered through a 0.45 μm syringe filter prior to analysis.

$$IU = \frac{100000 \cdot Abs_{420}}{\%RDS \cdot (0.9857 + 0.00488 \cdot \%RDS)} \quad (2)$$

Colour intensity was expressed as volume-adjusted absorbance (Abs·mL) and converted to ICUMSA Units (IU) based on the refractive dry substance (RDS) of the filtered sample.

5. Results and Discussion

5.1. Evaluation and Optimization of Feed operating Temperature

To determine the optimum thermal driving force for the vacuum membrane distillation (VMD) system, trials were carried out across three distinct feed temperatures (70, 75, and 80 °C) for 7-channel and 19-channel ceramic membrane configurations.

5.1.1. Temperature Impact on Transmembrane Permeate flux

Increasing the feed temperature to 80 °C enhanced permeate flux which records at 2.79 L/m².hr and 14.2 °Brix as seen in Figure 5.1, whereas the lowest flux of 0.72 L/m².hr was obtained at 70 °C and 12.5 °Brix. At 15.0 °Brix, increasing the operating temperature from 70 °C to 80 °C increased the permeate flux from 1.26 to 2.43 L/m².hr, representing a 93 % increase. Similarly, at 15.5 °Brix, flux increased from 1.03 to 2.39 L/m².hr, as temperature increased. At 12.5 °Brix, the permeate flux at 80 °C (1.53 L/m².hr) was twice in amount when compared at 70 °C (0.72 L/m².hr).

The higher permeate flux observed at elevated temperatures is consistent with membrane transport theory, where increased temperature reduces feed viscosity, enhances solute diffusivity, and lowers mass-transfer resistance, thereby improving permeate transport (Ricceri et al., 2022).

An exception was observed at 12.1 °Brix, where the flux at 75 °C (1.76 L/m².hr) exceeded that at 80 °C (1.26 L/m².hr). This deviation may be attributed to membrane fouling, and concentration polarization (Persson, 2025; Rezaei et al., 2018). Nevertheless, the overall results demonstrate that operating at 80 °C enhanced permeate flux and improved the separation performance of the VMD system.

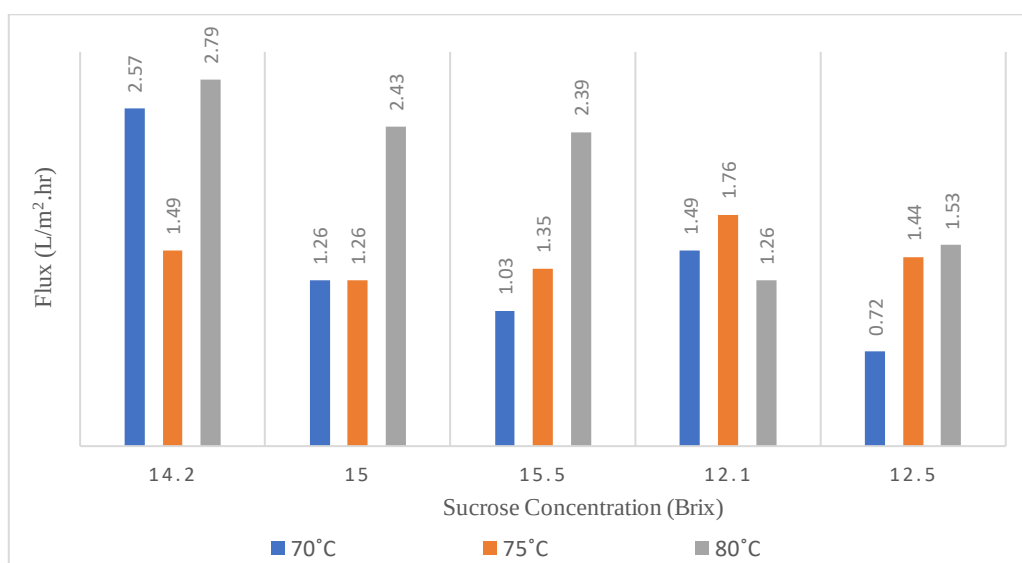


Fig 5.1. Permeate flux and concentration over time at varying feed temperatures and fixed flow rate of 8.44 L/min in the 19-channel configuration (vacuum pressure of 95 mbar).

5.1.2. Temperature Impact on Retention and Sucrose concentration

The gradual decline in permeate flux from 2.57 L/m².hr to 1.03 L/m².hr at 3 hr as shown in Figure 5.2 represented an approximate 60 % reduction. This indicated that resistance to vapour transport increased as concentration progressed. Continuous water removal increased sucrose concentration and solution viscosity. This promoted concentration polarization and reduced the effective vapour-pressure difference across the membrane. Consequently, the membrane system experienced progressively lower permeate flux to 0.72 L/m².hr and slightly increased to 0.98 L/m².hr at 6 hr despite operating under constant process conditions. Similar behaviour has been reported during membrane concentration of highly concentrated sugar solutions, where increasing boundary-layer resistance becomes the dominant limitation to mass transfer (Boey et al., 2025).

In contrast, sucrose retention increased steadily during the experiment from approximately 90 to 99 %. This demonstrated that the ceramic membrane maintained excellent selectivity despite the reduction in productivity. A marked increase was observed between 2 and 4 hr, where retention increased from 93 to 98.5 %, before reaching a plateau near 99 %. The consistently high rejection indicates that transport remained predominantly in the vapour phase, suggesting that membrane wetting was minimal under these operating conditions.

Overall, the results demonstrate an inverse relationship between permeate flux and retention. While flux declined because of increasing mass-transfer resistance, retention improved from 90 to 99 %, confirming the effective sucrose separation and concentration performance of the VMD system.

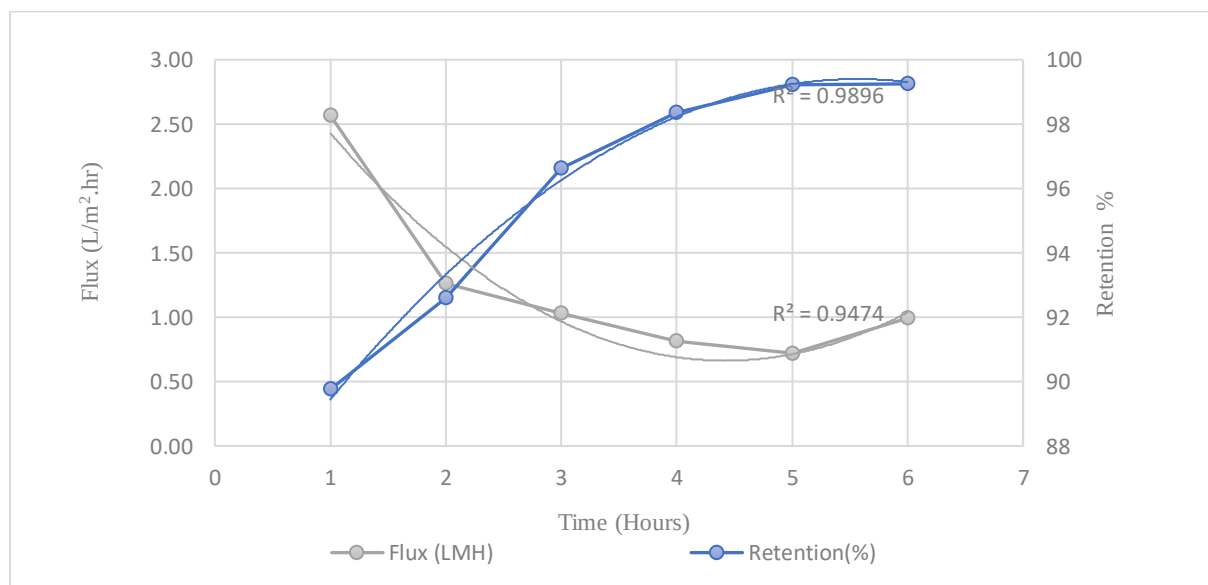


Figure 5.2. Permeate flux and ion rejection at 70 °C in the 19-channel configuration.

At a period of 2 hr, sucrose concentration increased from 12.47 °Brix to 13.26 °Brix indicating a 6.3 % increase in RDS of the retentate stream as shown in Figure 5.3. As concentration proceeded, sucrose concentration slightly declined to 11.68 °Brix at 3 hr and gradually increased to 13.17 °Brix at the end of the experiment. This represented an overall increase of 5.6 % relative to the initial concentration.

In contrast, the permeate was maintained consistently at low sucrose concentrations, decreasing from 1.17 °Brix to 0.18 °Brix at 6 hr. This decline indicated minimal sucrose transport through the membrane and confirms its high rejection efficiency throughout the concentration process. The RDS analysis of the retentate further confirmed the quality of the concentrated product, with purity level approximately 81.6 % and colour variation at 2703 ICUMSA,

The concentration difference between the retentate and permeate agrees with the separation mechanism in which only water vapour permeates the hydrophobic membrane while non-volatile solutes are retained in the feed (Persson, 2025.). This further shows the effectiveness of vacuum membrane distillation for concentrating sucrose solutions.

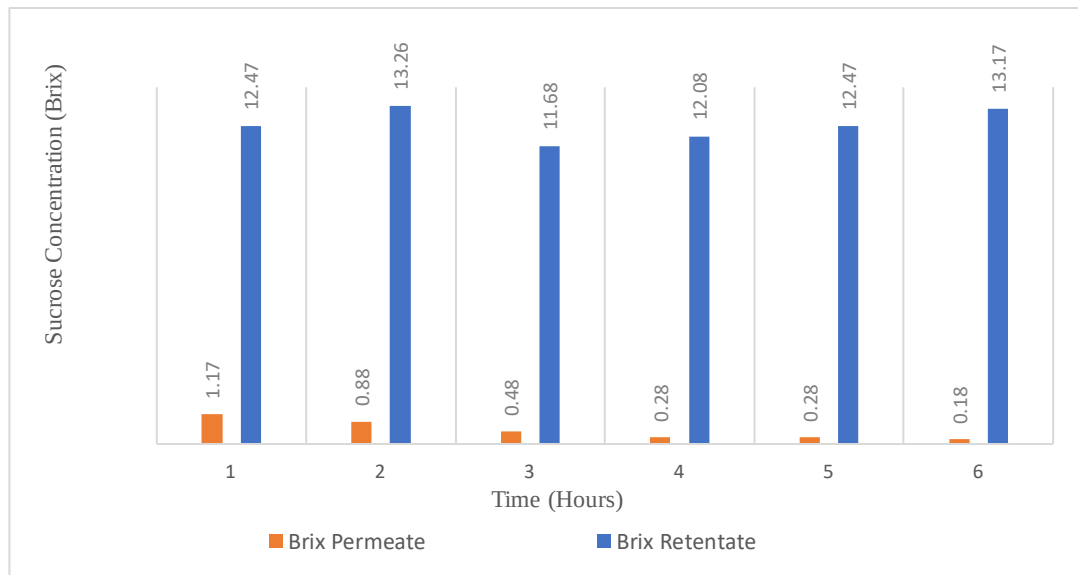


Figure 5.3. Sucrose concentration over time showing RDS of retentate and permeate stream in the 19-channel ceramic membrane at 70 °C

From Figure 5.4 below, it shows the increase in permeate flux from approximately 1.15 L/m².hr to a maximum of 1.60 L/m².hr at 3 hr. The permeate flux further declined to 1.12 L/m².hr at 6 hr. This suggests increase in resistance to water vapour transport with development of concentration polarization and membrane fouling (Al-Asheh et al., 2006).

In addition, retention increases from approximately 89 % to a maximum of 98 % at 2 hr, then further declined to 95 % at 3 hr. As concentration progresses, retention further declines over time to 63 % as observed between 4 hr and 5 hr.

Overall, the 7-channel membrane configuration showed greater fluctuations in both permeate flux and retention than the 19-channel configuration, indicating lower operational stability.

Although sucrose rejection was maintained, the variability in retention suggests that the 7-channel configuration was more susceptible to mass-transfer limitations.

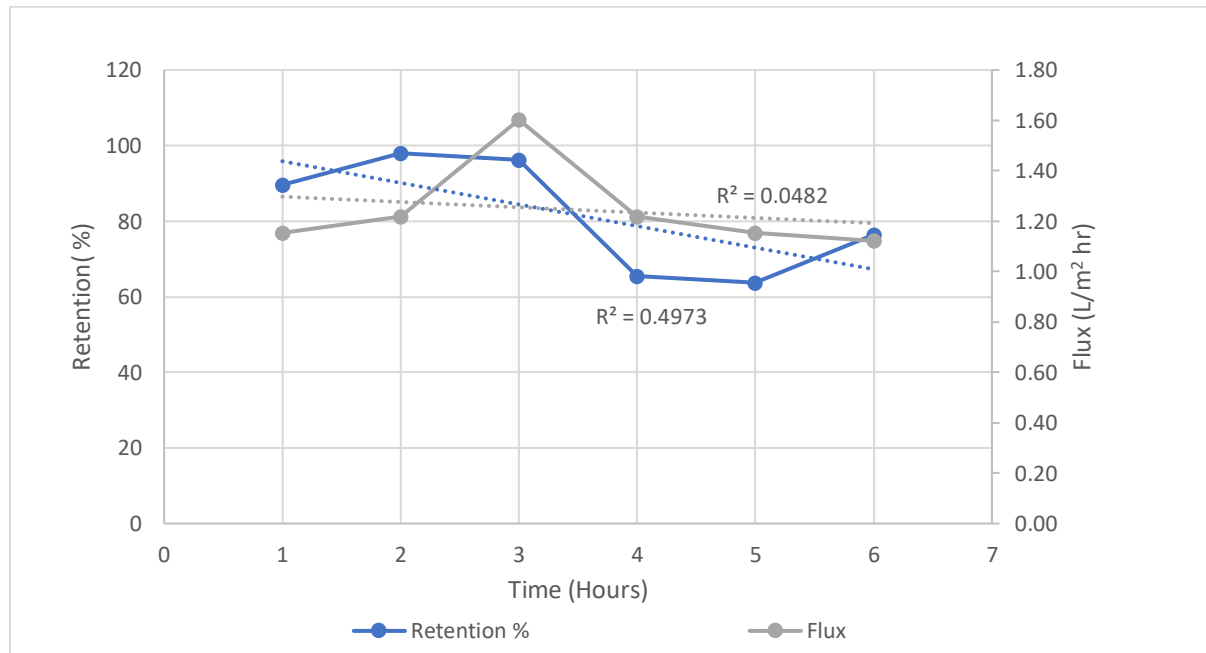


Figure 5.4. Permeate flux and ion retention at 70 °C in the 7-channel configuration

An increase in retentate concentration from 9.50 °Brix to 10.09 °Brix was observed during the first 3 hr of operation as shown in Figure 5.5. This demonstrated an effective concentration of the feed solution. Thereafter, the concentration declined to 8.51 °Brix at 4 hr before stabilizing at 9.10 °Brix. This indicated that the concentration process became less efficient as operation continued, possibly due to increased mass-transfer resistance.

In contrast, the permeate is maintained relatively at low sucrose concentrations. The low permeate concentration at 0.08 °Brix indicates high sucrose rejection, whereas at 3.16 °Brix at 5 hr suggests increased sucrose transport through the membrane pores. The RDS analysis of the final retentate further confirmed the quality of the concentrated product, with purity levels at approximately 91.96 % and colour variation at 2884 ICUMSA after 6 hr run.

The consistent concentration difference between the retentate and permeate confirmed high sucrose retention while allowing water vapour to permeate through the membrane. However, the increase in permeate concentration after 4 hr may be associated to changes in transport behaviour during prolonged operation due to membrane fouling, scaling, and partial pore wetting (Horseman et al., 2021).

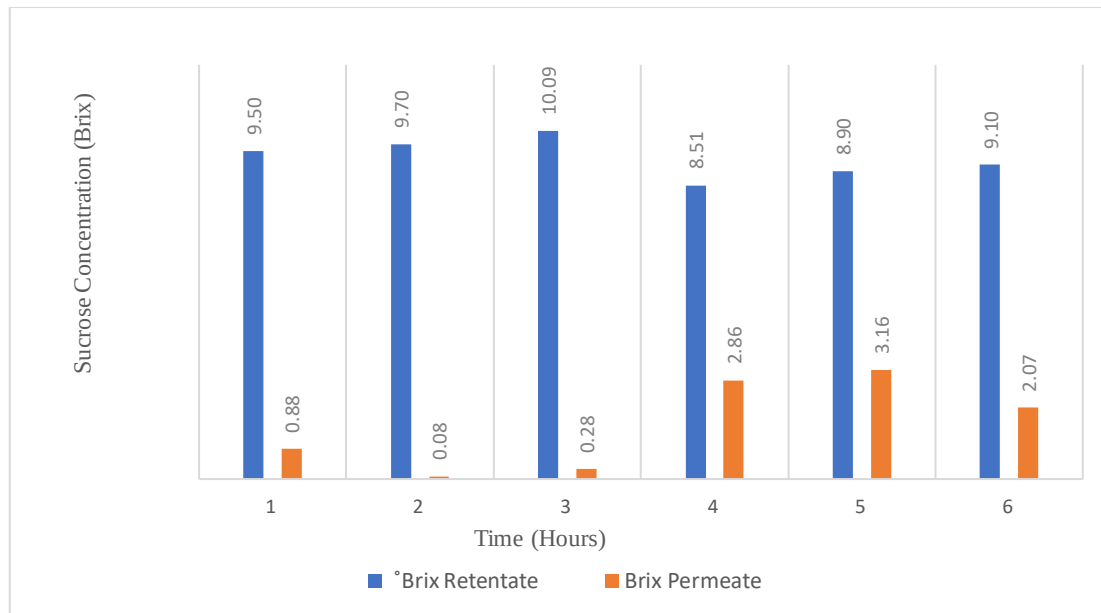


Figure 5.5. Sucrose concentration over time showing RDS of retentate and permeate stream in the 7-channel configuration at 70 °C.

Table 5.1. Sugar analysis results of the thin juice samples at constant Reynolds and vacuum pressure in the 7-channel configuration

Thin juice	70 °C
RDS (°Brix)	9.5
Colour (IU)	2884.24
Purity (%)	91.96
pH	9.0
Conductivity(μS/cm)	4.0

As shown in Figure 5.6, retention increased from 98.0 to 98.6 % during the first 2 hr of operation, before progressively declining to 94.0 % by 6 hr. This reflects a gradual loss in membrane selectivity over time. Permeate flux followed a similar trend, increasing from 1.45 L/m².hr to a peak of 1.65 L/m².hr at 3 hr, and then further declines to 1.20 L/m².hr by the end of the run. The concurrent decline in both retention and flux beyond 3 hr suggest the onset of concentration polarization and membrane fouling, as solute accumulate at the membrane surface increasingly limits permeate transport while simultaneously promoting sucrose leakage through the membrane (Al-Asheh et al., 2006). The steady downward trend in flux, paired with reduced retention, indicate a progressive increase in mass transfer resistance, pointing to a decline in overall membrane separation performance with prolonged operation.

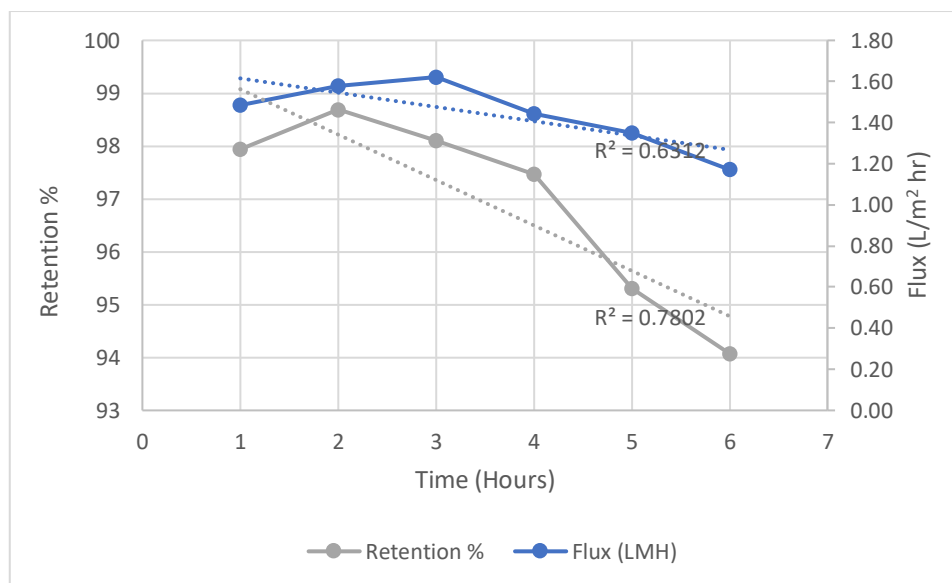


Figure 5.6. Permeate flux and ion rejection at 75 °C in the 19-channel configuration.

An increase in retentate concentration from 14.26 Brix to a maximum of 15.54 Brix was observed during the initial 3hr of operations, reflecting efficient concentration of the feed solution as seen in Figure 5.7. The subsequent decrease to 12.17 Brix at 4 hr, followed by a slight increase to 13.26 Brix at 6 hr, may indicate the influence of concentration polarization and membrane fouling. In contrast, permeate concentration is maintained at low sucrose concentrations, with the lowest value of 0.08 Brix observed at 2 hr, and the highest value of 0.88 Brix at 4 hr. This indicates minimal sucrose transport through the membrane

The large concentration difference between the retentate and permeate indicates high sucrose rejection throughout the concentration process. Although the permeate concentration increased slightly at 4 hr, the membrane continued to retain sucrose which indicates stable separation performance. This behaviour is consistent with the separation mechanism, in which the hydrophobic membrane selectively transports water vapour while retaining non-volatile solutes (Gulied et al., 2023).

The RDS analysis of the retentate at 6 hr further confirmed the quality of the concentrated product, with a purity of approximately 93.14 % and colour variation of 2725 ICUMSA. This indicates that the membrane concentration process produced a relatively high-purity sugar solution while maintaining acceptable colour quality.

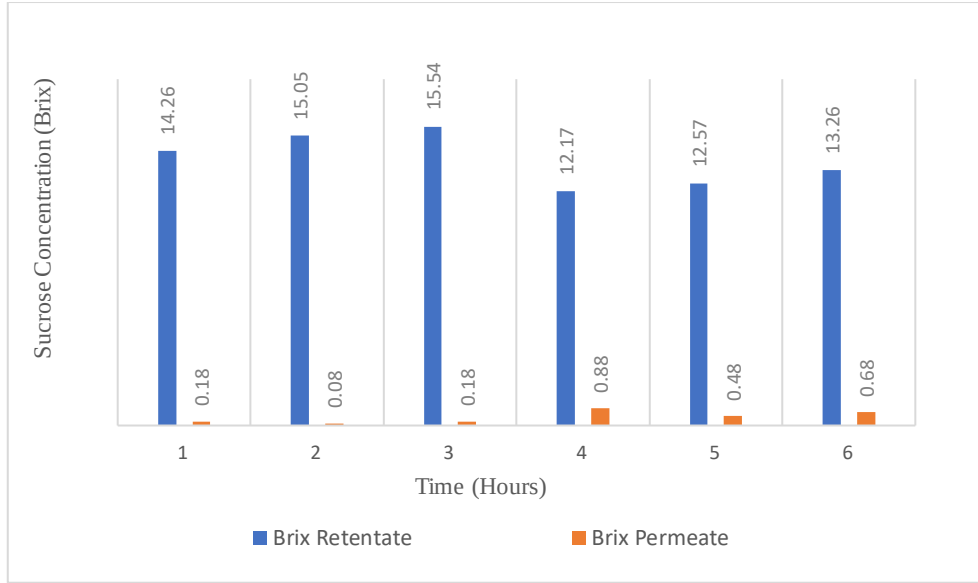


Figure 5.7. Sucrose concentration over time showing RDS of retentate and permeate stream in the 19-channel configuration at 75 °C.

Sugar retention showed an initial increase from 90.9 to a maximum of 95.1 % at 2 hr, before decreasing to 90.3 % at 3 hr and further increased to 94.2 % at 5 hr as seen in Figure 5.8.

In contrast, permeate flux declined steadily from 2.80 L/m² hr to 1.50 L/m² hr at 5 hr, representing an overall reduction of approximately 46 %.

The gradual decline in permeate flux indicates increasing mass-transfer resistance during concentration, which is likely caused by membrane fouling and concentration polarization (Do et al., 2025).

Despite the reduction in flux, sucrose retention remained above 90 % after 5 hr run. The slight decrease in retention at 4 hr may be attributed to scaling, concentration polarization, or partial pore wetting, while the low permeate conductivity of 2.53 µS/cm confirms high membrane selectivity to effectively reject dissolved solutes.

Operating at 80 °C, enhanced molecular diffusivity, solvent mobility, and mass-transfer rates. However, prolonged exposure to elevated temperature may reduce the surface tension of the feed and promote localized evaporation within the membrane pores (Persson, 2025.). These conditions can facilitate partial pore wetting or the transport of trace volatile compounds, leading to the reduction in sucrose retention as observed during the experiment.

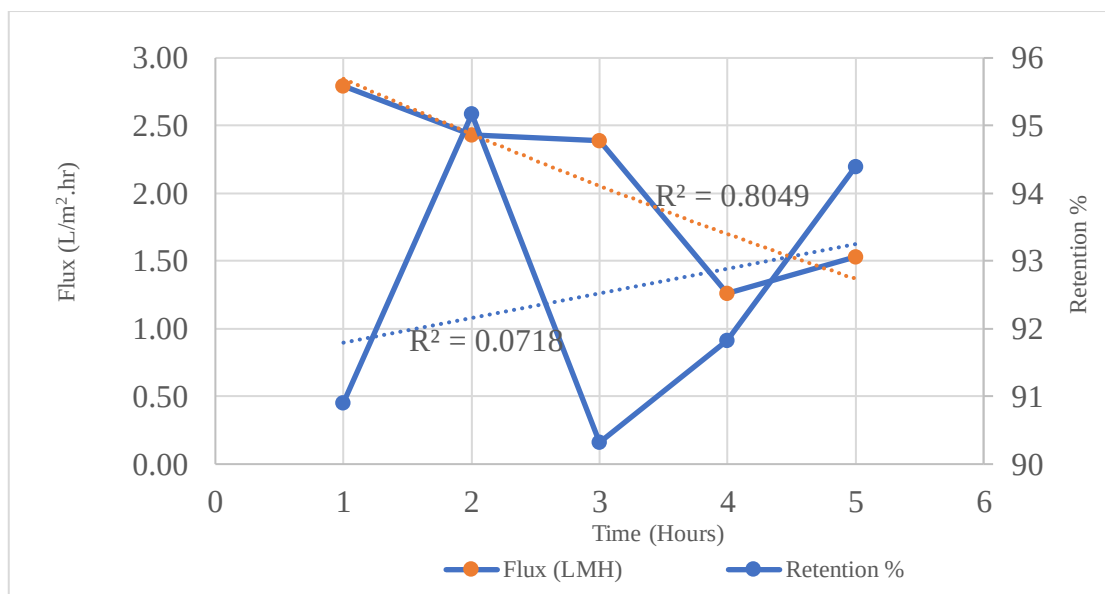


Figure 5.8. Permeate flux and ion rejection at 80 °C in 19-channel configuration

During the first 2 hr of operation, the retentate sucrose concentration increased from 11.88 °Brix to 14.26 °Brix as seen in Figure 5.9, reflecting efficient water removal. This indicates effective concentration during the initial stage of membrane distillation. Thereafter, the retentate concentration decreased to 9.10 °Brix at 3 hr, before increasing slightly to 10.49 °Brix at 5 hr.

In contrast, permeate concentration is maintained at low sucrose concentrations, decreasing from 0.98 °Brix at 1 hr to 0.48 °Brix at 5 hr, confirming effective sucrose concentration by the membrane.

The decline in retentate concentration after 2 hr is not typical of membrane distillation, where concentration generally increases with continued water removal (Aquino et al., 2023). This fluctuation may be attributed to sampling variability or inadequate mixing of the recirculating feed before sample collection.

In addition, prolonged operation at 80 °C may have promoted concentration polarization, scale formation, partial membrane pore wetting, all of which could influence the measured °Brix values (Al-Asheh *et al.*, 2006). In addition, operations at 80 °C results in a significant increase in colour development at 3027 ICUMSA and gave an average purity level of 89.31 % compared with the more stable color and purity profile observed at 70 °C and 75 °C respectively.

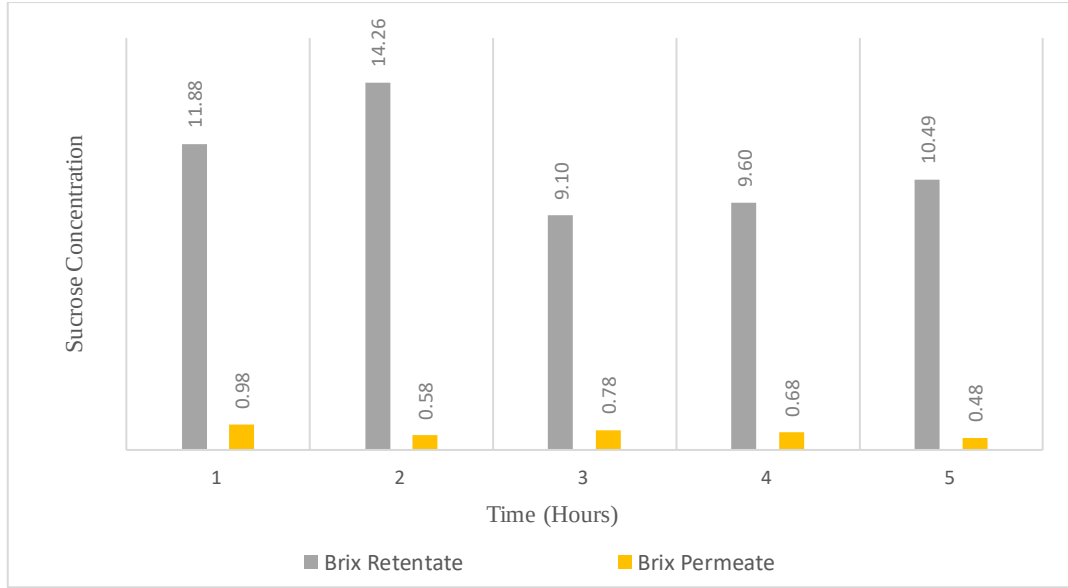


Figure 5.9. Sucrose concentration over time showing RDS of retentate and permeate stream in the 19-channel configuration at 80 °C

Table 5.2: Sugar analysis of thin juice samples at constant Reynolds number and vacuum pressure in the 19-channel configuration

Thin juice	70 °C	75 °C	80 °C
RDS (°Brix)	13.16	13.26	10.48
Colour (IU)	2703.65	2724.80	3027.0
Purity %	81.60	93.13	89.31
pH	9.0	9.0	9.1
Conductivity (µS/cm)	3.02	3.59	2.53

Table 5.2 shows the average values from sugar analysis of the thin beet sugar juice at varying feed temperature, at constant flow rate of 8.4 L/min ($Re = 1795$) in the 19-channel configuration.

Table 5.3: Comparative results of average flux and ion rejection of the 19-channel ceramic membrane at varying feed temperatures.

Temperature (° C)	Ion rejection %	Average Flux (L/m ² hr)
70	95.62	1.34
75	96.11	1.41
80	91.79	2.08

From table 5.3, the 19-channel configuration at 75 °C is observed to achieve the highest average ion retention at 96.11 %, while 80 °C showed the highest average flux validating that increasing temperature improves membrane performance and permeate flux but may facilitate polarization effects (Politano et al., 2019).

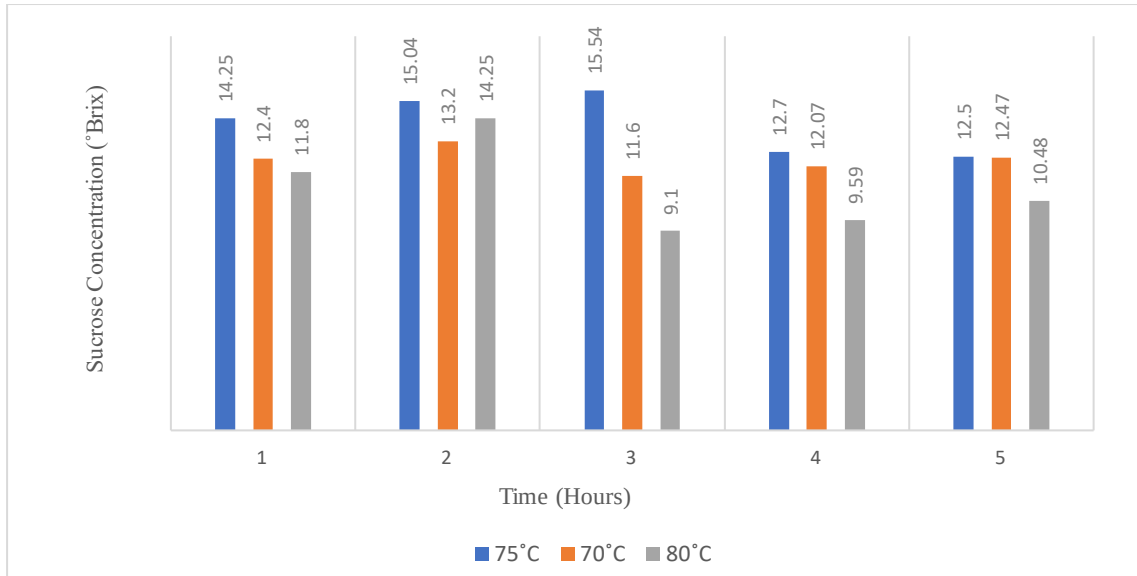


Figure 5.10. Sucrose concentration over time showing RDS of retentate stream in the 19-channel configuration at varying feed temperatures

From figure 5.10, it is observed that the 19-channel configuration at 75 °C shows the highest degree of sucrose concentrations with peak value of 15.54 °Brix over time and validates its optimum condition.

5.2. Membrane Filtration Area and Structural Flux Dynamics

The 7-channel configuration, with a filtration area of 0.156 m², was feasible only at 70 °C. As illustrated in Figure 5.11, the retentate reached a final sucrose concentration of 9.10 °Brix after 6 hr of operation, according to RDS measurements. This variation in concentration is attributed to a lower filtration area-to-volume ratio and water removal rate.

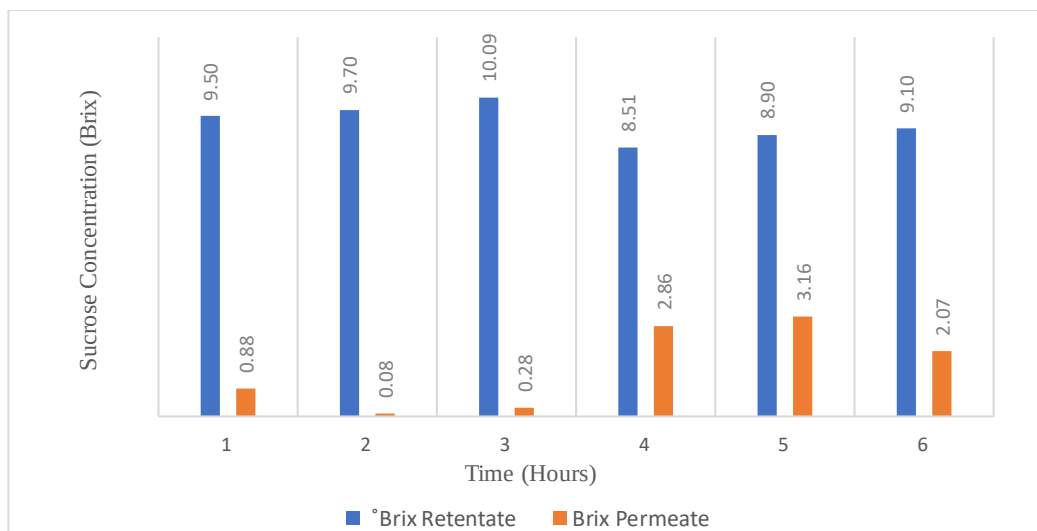


Figure 5.11. Sucrose concentration over time showing RDS of retentate stream in the 7-channel configuration at 70 °C.

The expanded surface area at 0.222 m² of the 19-channel configuration yielded a higher volumetric water removal rate and a higher degree of concentration over time, reaching 13.26 °Brix after 6 hr as seen from figure 5.12. This enhanced performance is attributed to the larger surface-area-to-volume ratio and the narrower dimensions of the individual channels.

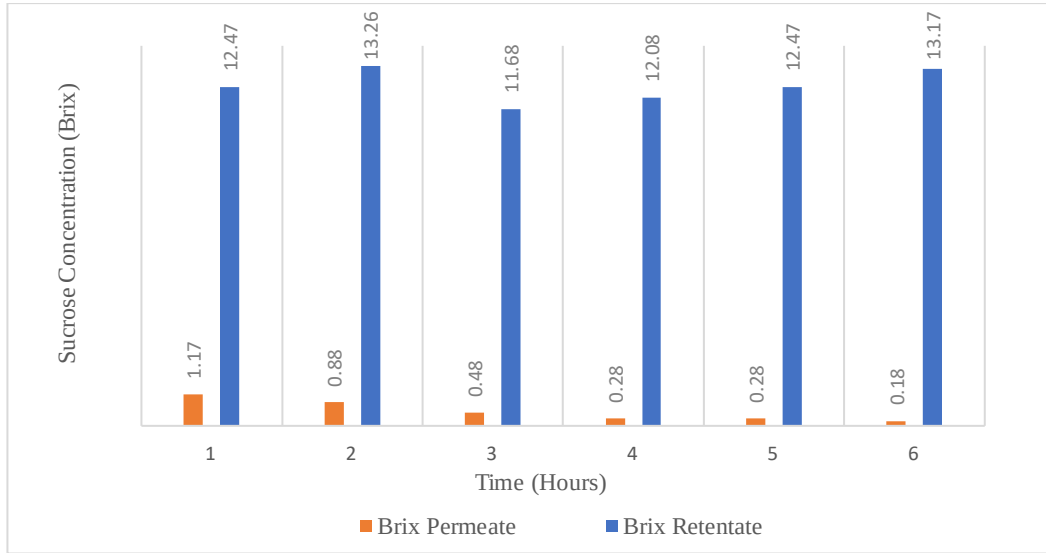


Figure 5.12. Sucrose concentration over time showing RDS of retentate stream in the 19-channel configuration at 70 °C

5.3. Evaluation and Optimization of Feed Flow-Rate

The highest sucrose concentration of 15.54 °Brix was achieved at a flow rate of 8.4 L/min after 3 hr as seen from Figure 5.13. At 6.7 L/min, the sucrose concentration reached a peak of 14.05 °Brix at 1 hr before gradually declining over time. Similarly, at 12.8 L/min, the highest sucrose concentration of 11.48 °Brix was recorded at 2 hr, followed by a slight decrease during the remaining operational period. The superior performance at 8.4 L/min indicates an optimal balance between feed circulation and residence time, thereby enhancing mass transfer while minimizing concentration polarization.

In contrast, the lower flow rate at 6.7 L/min promotes concentration polarization due to insufficient mixing, whereas the higher flow rate (12.8 L/min) reduced the feed residence time, limiting water removal and sucrose concentration (Y.-R. Chen et al., 2018).

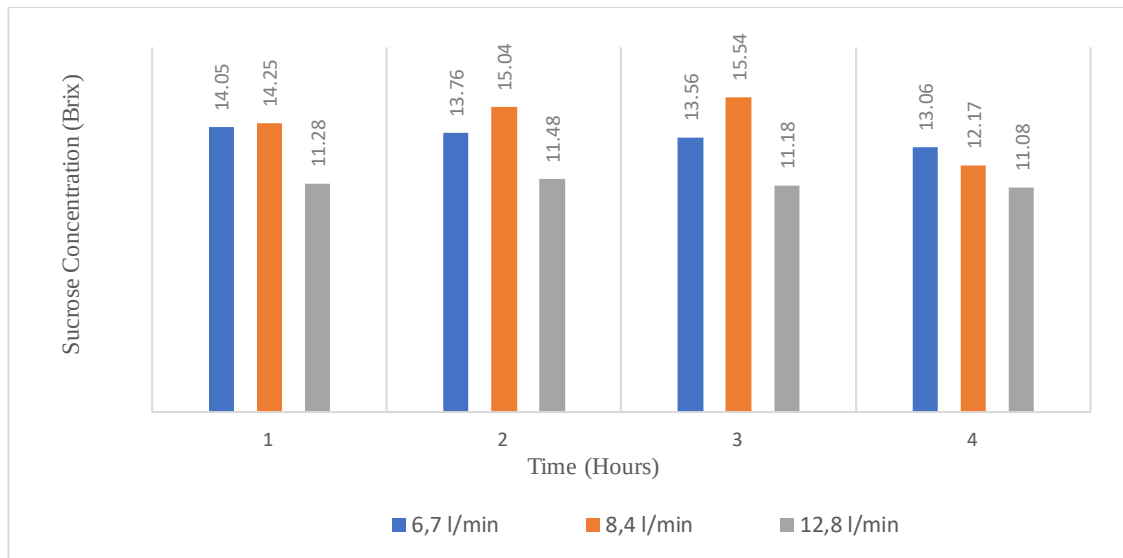


Figure 5.13. Sucrose concentration with respect to time at varying flowrates.

From Figure 5.14, at flow rate of 6.7 L/min, permeate flux remained relatively low as sucrose concentration increased. This indicated a greater mass-transfer resistance due to concentration polarization. Increasing the flow rate to 8.4 L/min increased permeate flux from 1.26 L/m².hr to 1.76 L/m².hr. This suggests that the higher feed velocity reduced the concentration boundary layer and enhanced water vapour transport across the membrane (Ricceri et al., 2022).

The highest permeate flux of 5.41 L/m².hr was consistently obtained at 12.8 L/min, demonstrating that increasing the flow rate enhanced membrane performance by improving hydrodynamic conditions at the membrane surface (Abu-Zeid et al., 2015). The corresponding Reynolds numbers increased from 1431 at 6.7 L/min to 1795 at 8.4 L/min, and then achieves 2736 at 12.8 L/min. While the flow at 6.7 and 8.4 L/min remained within the laminar regime, the higher Reynolds number at 12.8 L/min approached the transitional regime, generating greater turbulence that reduced concentration polarization and increased mass transfer.

Although 12.8 L/min produced the highest flux, 8.4 L/min provided a more balanced operating condition by maintaining relatively high permeate flux while achieving superior sucrose concentration, indicating an optimal compromise between membrane performance and concentration efficiency.

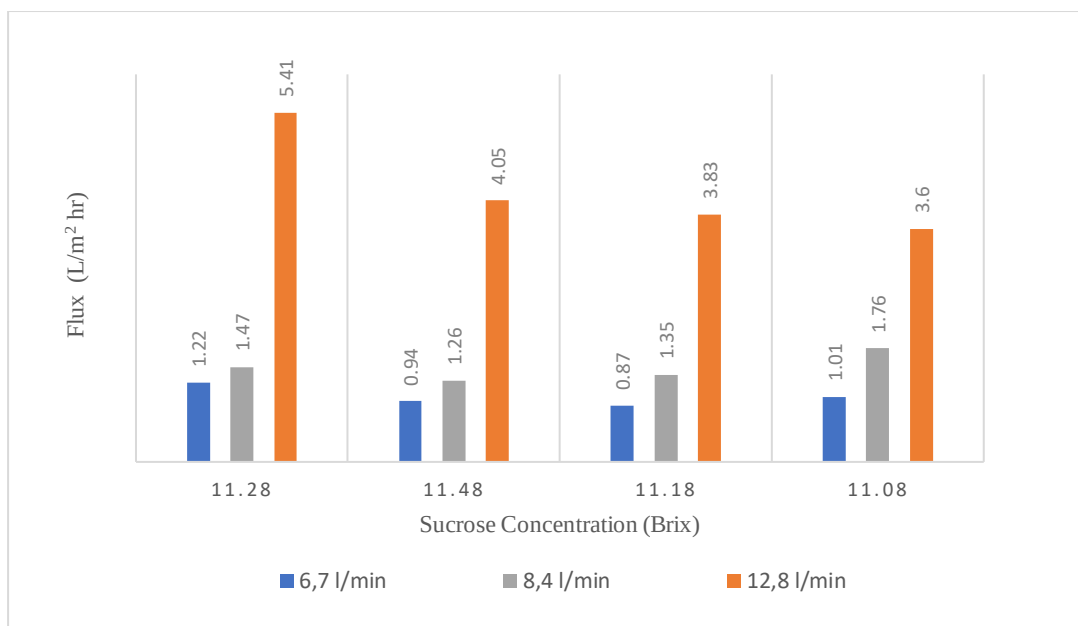


Figure 5.14. Permeate flux with respect to sucrose concentration over time at varying flowrates.

5.4. Effects of Vacuum Pressure on the Permeate Flux.

At a vacuum pressure of 27 mbar, stable baseline water removal rate was observed. The retentate °Brix profile increased slowly and steadily over time achieving sucrose concentration of 15.74 °Brix after 4-hr of run. Its operational assessment was limited to dewatering throughput after 4 hr. At that period, the permeate conductivity increased to 404.94 $\mu\text{S}/\text{cm}$ indicating partial wetting which was mitigated by a sequential clean in place protocol. At a vacuum pressure of 105 mbar, the concentration increases at the first two hours with fluctuations in concentrations and then declines to 12.2 °Brix, lower than the vacuum pressure at 95 mbar.

At 95 mbar vacuum pressure, sugar concentration follows a continuous, nearly linear upward trend. It steadily progresses from an initial feed baseline of 14.25 °Brix, higher than vacuum at 105 mbar, to a peak of 15.54 Brix after 3 hr. in a 6-hr operational time. This balances mass transfer kinetic, allowing maximum concentration to 13.2 °Brix higher than the concentration with vacuum pressure at 105 mbar as seen in Figure 5.15.

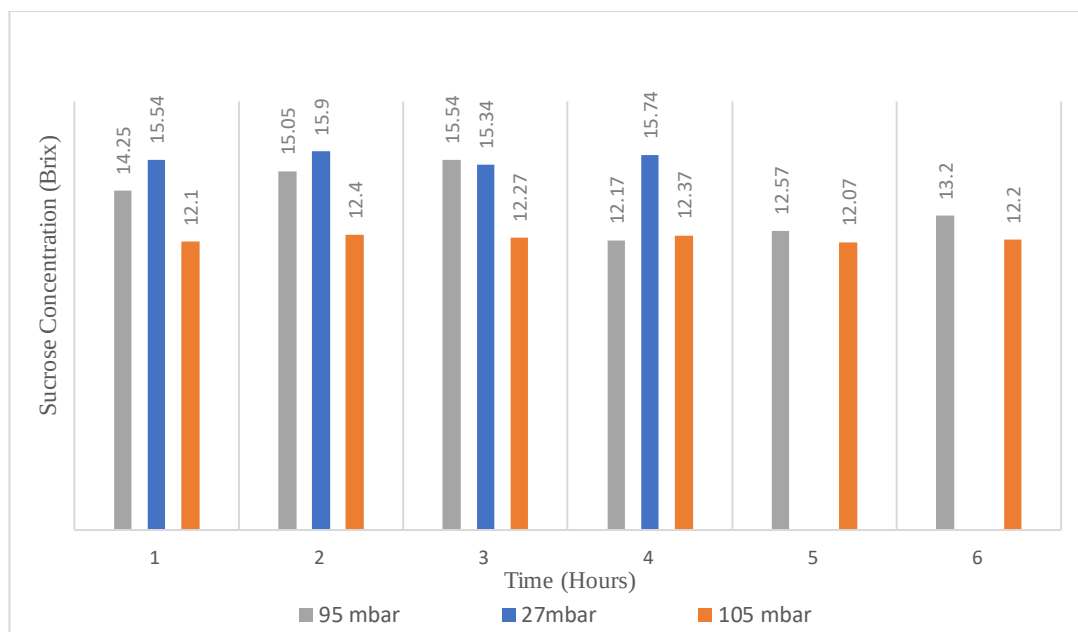


Figure 5.15. Sucrose concentration with respect to time at varying vacuum pressures.

6. Conclusion

The results demonstrated that the 19-channel ceramic membrane outperformed the 7-channel configuration. This is due to its larger membrane area and improved hydrodynamic conditions, which enhanced mass transfer and reduced concentration polarization. Consequently, the 19-channel membrane achieved higher sucrose concentrations and more stable separation performance throughout the concentration process.

Among the operating conditions investigated, the optimum performance was obtained using the 19-channel membrane at 75 °C, 95 mbar vacuum pressure, and a feed flow rate of 8.44 L/min ($Re = 1795$). These conditions provided the best balance between permeate flux, sucrose retention, and concentration efficiency while maintaining process stability. Although operation at 80 °C produced the highest initial permeate flux, prolonged exposure to high temperature accelerated concentration polarization and fouling, limiting the final sucrose concentration. Similarly, increasing the feed flow rate improved permeate flux by enhancing fluid mixing and reducing boundary-layer resistance. However, excessive flow reduced feed residence time and consequently lowered concentration efficiency. Vacuum pressure also influenced membrane performance, with 95 mbar providing a favourable balance between water removal and stable operation.

Product quality was maintained throughout the concentration process, with juice purity ranging from 80 to 93.14% and colour values remaining within the acceptable industrial range of 2703 to 3027 ICUMSA. The ceramic membranes also demonstrated excellent thermal stability at temperatures up to 80 °C, indicating their suitability for high-temperature membrane distillation.

Long-term operation revealed that concentration polarization, sucrose accumulation, and inorganic scaling were the primary fouling mechanisms limiting membrane performance. Nevertheless, low permeate conductivity confirmed that membrane selectivity was maintained throughout the experiments. The fouling was largely reversible, as the clean-in-place protocol restored more than 98% of the original pure water flux. These findings demonstrate the durability and reusability of ceramic vacuum membrane distillation systems for thin beet juice concentration while highlighting the need to optimize operating conditions to minimize fouling during large-scale application.

7. Future Work

The findings of this study highlight several areas that require further investigation to improve the long-term performance and industrial application of ceramic membranes in vacuum membrane distillation (VMD) for thin beet juice concentration.

Future studies should investigate environmentally friendly hydrophobic modification agents, including hexadecyltrimethoxysilane (C16), methyltrichlorosilane (MTS), and monoethanolamine (MEA), as alternatives to fluorinated silanes. These materials may provide stable hydrophobicity while reducing environmental concerns associated with the degradation and leaching of perfluoroalkyl compounds. In addition, non-fluorinated coatings may reduce mass-transfer resistance while maintaining adequate membrane performance.

The development of superhydrophobic and omniphobic ceramic membranes should also be explored to improve resistance to membrane wetting and fouling. Furthermore, membrane surface roughness should be investigated because hierarchical roughness and re-entrant structures can trap air, reduce liquid–solid contact, and improve water repellency during membrane distillation. Feed pretreatment strategies should be evaluated to minimise flux decline caused by scaling and fouling during prolonged operation.

Long-term membrane distillation experiments are recommended to evaluate membrane stability under prolonged operation. Extended operation would provide a more representative assessment of membrane durability and process performance under industrial conditions. The 6-hr experimental period used in this study may not fully capture the gradual development of membrane wetting, fouling, concentration polarization, scaling, and membrane compaction.

Comprehensive membrane characterization should be conducted before and after membrane operation to evaluate the suitability of alumina ceramic membranes for sugar concentration. Techniques such as Fourier Transform Infrared Spectroscopy (FTIR), Scanning Electron Microscopy (SEM), X-ray Photoelectron Spectroscopy (XPS), and contact angle measurements should be used to assess changes in surface chemistry, morphology, hydrophobicity, and the effectiveness of membrane surface modification.

Finally, future research should investigate membrane configurations with higher packing densities to improve process productivity. In particular, hollow-fibre ceramic membranes offer a higher surface-area-to-volume ratio with potential to reduce mass transfer resistance while increasing permeate production. This makes them promising for industrial-scale sugar concentration.

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9. Abbreviations and Explanation

Abbreviation	Explanation
Abs	Absorbance
AGMD	Air-Gap Membrane Distillation
Al ₂ O ₃	Aluminium Oxide
CP	Concentration Polarization
DCMD	Direct Contact Membrane Distillation
DI	De-ionised
EDTA	Ethylenediaminetetraacetic acid
FAS	Fluoroalkyl-silanes
Hr.	Hour
ICUMSA	International commission for uniform methods of sugar
IU	ICUMSA Units
LEP	Liquid Entry Pressure
MD	Membrane Distillation
PP	Polypropylene
P _f (liquid)	Feed pressure
P _p (vapour)	Permeate pressure
PTFE	Polytetrafluoroethylene
PVDF	Polyvinylidene
PFAS	Perfluoroalkyl-silanes
RDS	Refractive Dry Substance
r _{max}	Maximum pore radius
RO	Reverse Osmosis
SGMD	Sweep-Gas Membrane distillation
TiO ₂	Titanium Oxide
TP	Temperature Polarization
VMD	Vacuum Membrane Distillation
ZrO ₂	Zirconia

Appendices

Appendix I

Recorded Data I

Pure water flux evaluation of 7-channel ceramic membrane at 70 °C at 8.44 L/min

Temperature (°C)	TMP (bar)	Flow rate (l/minutes)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hours)
60	2.68	10.8	20135.6	0.8	95	2.67	0.577	90	1
60	2.68	10.8	20207.9	1.02	95	2.67	0.833	130	2
60	2.68	10.8	20207.9	0.55	95	2.67	1.03	160	3
60	2.68	10.8	20244.1	1.09	95	2.67	1.28	200	4
60	2.68	10.8	20232.4	0.9	95	2.67	1.47	230	5

Concentration at 70 °C of 7-channel ceramic membrane at 8.44 L/min

TMP (bar)	Vacuum pressure (mbar)	Feed pressure (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Flux (LMH)	Time (Hours)	Colour	Purity (%)	Permeate volume (ml)
2.71	95	2.74	177406	2.03	1.15	1	2815.01	98.25	180
2.71	95	2.74	179987	6.48	1.22	2	2853.32	91	190
2.71	95	2.74	182084	3.37	1.60	3	2841.19	90	198
2.71	95	2.74	159698	3.33	1.22	4	2916.46	91.6	750
2.71	95	2.74	162699	4.47	1.15	5	2924.49	91.51	260
2.71	95	2.74	165050	4.34	1.12	6	2954.99	89.43	260

% RDS at 70 °C of 7-channel ceramic membrane at 8.44L/min

Permeate	Retentate	Time (Hours)	Outlet Temperature (°C)
0.88	9.50	1	16.4
0.08	9.70	2	16.3
0.28	10.09	3	16.3
2.86	8.51	4	16.5
3.16	8.90	5	17.6
2.07	9.10	6	17.5

Record Data II

Pure water flux evaluation of 7-channel ceramic membrane at 75 °C at 8.44 L/min

Temperature (°C)	TMP (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hours)
60	2.68	20569.6	1.23	95	2.66	0.8	130	1
60	2.68	20641.9	0.58	95	2.66	0.9	145	2
60	2.68	20641.9	0.84	95	2.66	1.0	160	3
60	2.68	20678.1	1.09	95	2.66	0.8	125	4
60	2.68	20656.2	0.96	95	2.66	0.8	125	5

Concentration at 75 °C of 7-channel ceramic membrane 8.44 L/min

TMP (bar)	Cross-flow (m/s)	Vacuum pressure (mbar)	Feed pressure (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Time (Hours)	Permeate volume (ml)
2.71	1	95	2.74	194406	3.86	1	180
2.71	1	95	2.74	195987	5.41	2	190
2.71	1	95	2.74	196084	4.97	3	198

% RDS at 75 °C of 7-channel ceramic membrane at 8.44 L/min

Permeate	Retentate	Overall concentration	variation	Time (Hours)	Outlet Temperature(°C)
8.3	6.8	0	0	0	16.2

Appendix II

Record Data III

Pure water flux evaluation of 7-channel ceramic membrane at 80 °C at 8.44 L/min

Temperature (°C)	TMP (bar)	Flow rate (l/minutes)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hours)
60	2.68	10.8	21112.1	0.58	95	2.66	1.06	165	1
60	2.68	10.8	21292.2	0.58	95	2.66	1.22	190	2
60	2.68	10.8	21943.9	0.73	95	2.66	1.35	210	3
60	2.68	10.8	22052.4	0.58	95	2.66	1.35	210	4
60	2.68	10.8	22154.6	0.67	95	2.66	1.03	160	5

Concentration at 80 °C of 7-channel ceramic membrane at 8.44 L/min

Permeate	Retentate	Overall concentration	variation	Time (Hours)	Outlet Temperature(°C)
11.4	5.8	0	0	1	15.7

Record Data IV

Pure water flux evaluation of 7-channel ceramic membrane at 8.44 L/min

Temperature (°C)	TMP (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hours)
60	2.68	15795.4	50.49	91	2.71	7.05	1100	1
60	2.68	16248.9	49.16	91	2.71	7.05	1100	2
60	2.68	16229.8	46.84	91	2.71	7.69	1200	3
60	2.68	16338.2	47.02	91	2.71	9.42	1470	4

Concentration of 7-channel ceramic membrane at 8.44 L/min

Temperature (°C)	TMP (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (hours)
70	2.86	183350	2.57	91	2.96	2.31	360	1
70	2.86	184218	1.38	91	2.96	2.63	410	2
70	2.86	184652	3.04	91	2.96	2.31	360	3
70	2.86	184765	2.98	91	2.96	2.21	345	4

% RDS at 8.44 L/min of 7-channel ceramic membrane

Permeate	Retentate	Time (Hours)	Outlet Temperature (°C)
8.4	12.1	1	15.5
11.1	11.9	2	15.5
11.7	11.6	3	15.8
11.5	11.5	4	15.6

Record Data V

Pure water flux evaluation of 19-channel ceramic membrane at 70 °C at 8.44 L/min

Temperature (°C)	TMP (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hour)
70	2.37	21112.1	0.58	96	2.35	0.383	85	1
70	2.37	21292.2	0.58	96	2.35	0.450	100	2
70	2.37	21943.9	0.73	96	2.35	0.541	120	3
70	2.37	22052.4	0.58	96	2.35	0.608	135	4
70	2.37	22154.6	0.67	96	2.35	0.676	150	5

Appendix III

Concentration at 70 °C of 19-channel ceramic membrane at 8.44 L/min

TMP (bar)	Vacuum pressure (mbar)	Feed pressure (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Flux (LMH)	Time (Hours)	Colour	Purity (%)	Permeate volume (ml)
2.37	95	2.35	195755	4.42	2.57	1	2721	96.42	580
2.37	95	2.35	201143	4.13	1.26	2	2671.5	91.13	280
2.37	95	2.35	186098	2.07	1.03	3	2729.3	92.7	229
2.37	95	2.35	190077	2.39	0.81	4	2674.11	87.3	330
2.37	95	2.35	198911	2.79	0.72	5	2743	64.58	160
2.37	95	2.35	199726	3.32	0.99	6	2683	47.8	220

% RDS at 70 °C of 19-channel ceramic membrane at 8.4 L/min

Permeate	Retentate	Time (Hours)	Outlet Temperature (°C)
1.17	12.47	1	15.4
0.88	13.26	2	15.8
0.48	11.68	3	15.4
0.28	12.08	4	15.5
0.28	12.47	5	15.2
0.18	13.17	6	15.3

Record Data VI

Pure water flux evaluation of 19-channel ceramic membrane at 75 °C at 8.44 L/min

Temperature (°C)	TMP (bar)	Flow rate (l/minutes)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hour)
75	2.36	8.44	95	2.35	0.428	95	1
75	2.36	8.44	95	2.35	0.495	110	2
75	2.36	8.44	95	2.35	0.563	125	3
75	2.36	8.44	95	2.35	0.581	129	4
75	2.36	8.44	95	2.35	0.586	130	5

Concentration at 75 °C of 19-channel ceramic membrane at 8.44 L/min

TMP (bar)	Vacuum pressure (mbar)	Feed pressure (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Flux ((LMH)	Time (Hours)	Colour	Purity (%)	Permeate volume (ml)
2.35	95	2.35	212644	3.58	1.49	1	2788.6	81.26	330
2.35	95	2.35	22336	2.57	1.58	2	2643.87	95.13	350
2.35	95	2.35	228195	4.6	1.62	3	2527.67	93.95	360
2.35	95	2.35	194055	3.15	1.44	4	2828	97.12	320
2.35	95	2.35	200709	4.31	1.35	5	2772	95.21	300
2.35	95	2.35	205772	3.33	1.17	6	2789	96.14	260

Record Data VII

% RDS at 75 °C of 19-channel ceramic membrane at 8.4 L/min

Permeate	Retentate	Time (Hours)	Outlet Temperature (°C)
0.18	14.25	1	15.4
0.08	15.04	2	15.5
0.18	15.54	3	15.6
0.87	12.17	4	15.4
0.48	12.57	5	15.2
0.67	13.26	6	15.3

Appendix IV

Record Data VIII

Pure water flux evaluation of 19-channel ceramic membrane at 80 °C at 8.4 L/min

Temperature (°C)	TMP (bar)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hour)	Cross-flow velocity (m/s)
75	2.34	95	2.35	1.13	250	1	1
75	2.34	95	2.35	1.19	265	2	1
75	2.34	95	2.35	1.15	255	3	1
75	2.34	95	2.35	1.13	250	4	1
75	2.34	95	2.35	0.991	220	5	1

Concentration at 80 °C of 19-channel ceramic membrane at 8.4 L/min

TMP (bar)	Vacuum pressure (mbar)	Feed pressure (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Flux (LMH)	Time (Hours)	Colour	Purity (%)	Permeate volume (ml)
2.35	95	2.34	190836	1.16	2.79	1	2827	84.14	620
2.35	95	2.34	199733	4.78	2.43	2	2925	89.90	540
2.35	95	2.34	162374	2.54	2.39	3	3141	85.08	530
2.35	95	2.34	169462	3.01	1.26	4	3050.7	88.23	280
2.35	95	2.34	162772	1.2	1.53	5	3191.3	89.67	340

% RDS at 80 °C of 19-channel ceramic membrane at 8.4 L/min

Permeate	Retentate	Time (Hours)	Outlet Temperature (°C)
0.97	11.87	1	15.3
0.58	14.25	2	16.8
0.77	9.10	3	15.7
0.68	9.57	4	15.4
0.48	10.49	5	15.4

Record Data IX

Pure water flux evaluation of 19-channel ceramic membrane at 6.7 L/min

Temperature (°C)	TMP (bar)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hour)
60	1.85	94	1.85	1.53	340	1
60	1.85	94	1.85	2.12	470	2
60	1.85	94	1.85	2.34	520	3
60	1.85	94	1.85	2.52	560	4
60	1.85	94	1.85	2.52	560	5

Concentration at 6.7 L/min of 19-channel ceramic membrane

TMP (bar)	Vacuum pressure (mbar)	Feed pressure (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Flux (LMH)	Time (Hours)	Permeate volume (ml)	Feed Temperature (°C)
2.35	95	2.35	211957	6.62	1.22	1	270	70
2.35	95	2.35	211595	1.23	0.95	2	210	70
2.35	95	2.35	211776	1.27	0.88	3	195	70
2.35	95	2.35	211723	1.24	1.01	4	195	70

% RDS at 6.7 L/min of 19-channel ceramic membrane

Permeate	Retentate	Time (Hours)	Outlet Temperature (°C)
1.17	14.05	1	15.2
6.22	13.76	2	16.2
11.38	13.56	3	16.4
11.57	13.06	4	16.2

Appendix V

Record Data X

Pure water flux evaluation of 19-channel ceramic membrane at 12.8 L/min

Temperature (°C)	TMP (bar)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hour)	Cross-flow velocity (m/s)
60	2.24	93	2.25	1.76	390	1	1.2
60	2.24	93	2.25	2.25	500	2	1.2
60	2.24	93	2.25	2.43	540	3	1.2
60	2.24	93	2.25	2.16	480	4	1.2
60	2.24	93	2.25	2.12	470	5	1.2
60	2.24	93	2.25	2.12	470	6	1.2

Concentration at 12.8 L/min of 19-channel ceramic membrane

TMP (bar)	Vacuum pressure (mbar)	Feed pressure (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Flux (LMH)	Time (Hours)	Permeate volume (ml)	Outlet Temperature (°C)	Outlet Temperature (°C)
2.37	95	2.37	179661	1.13	5.41	1	1200	37.1	70
2.37	95	2.37	181144	1.13	4.05	2	900	34.3	70
2.37	95	2.37	182120	0.91	3.83	3	850	37.2	70
2.37	95	2.37	182916	0.84	3.60	4	800	33	70

% RDS at 12.8 L/min of 19-channel ceramic membrane

Permeate	Retentate	Time (Hours)	Outlet Temperature (°C)
9.79	11.28	1	37.1
10.68	11.48	2	34.3
11.08	11.18	3	37.2
11.18	11.08	4	33

Record Data XI

Pure water flux evaluation of 19-channel ceramic membrane at 27 mbar vacuum pressure

Temperature (°C)	TMP (bar)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hour)
60	2.15	84	2.25	5.86	1300	1
60	2.15	84	2.25	6.76	1500	2
60	2.15	84	2.25	7.21	1600	3
60	2.15	84	2.25	7.21	1600	4
60	2.15	84	2.25	7.21	1600	5

Concentration at 27 mbar vacuum pressure of 19-channel ceramic membrane

TMP (bar)	Vacuum pressure (mbar)	Feed pressure (bar)	Feed conductivity ($\mu S/cm$)	Permeate conductivity ($\mu S/cm$)	Flux (LMH)	Time (Hours)	Permeate volume (ml)	Feed Temperature (°C)
1.5	27	2.17	142375	406.94	6.67	1	1200	70
1.5	27	2.17	137854	406.65	4.50	2	900	70
1.5	27	2.17	142013	405.41	4.50	3	850	70
1.5	27	2.17	142158	404.94	4.50	4	800	70

% RDS at 27 mbar vacuum pressure of 19-channel ceramic membrane

Permeate	Retentate	Time (Hours)	Outlet Temperature (°C)
6.6	15.54	1	15.4
4.5	15.94	2	21.6
4.5	15.34	3	28.9
4.5	15.74	4	28.5

Appendix VI

Record Data XII

Pure water flux evaluation of 19-channel ceramic membrane at 105 mbar vacuum pressure

Temperature (°C)	TMP (bar)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hour)
60	3.00	88	3.06	0.450	1450	1
60	3.00	86	3.06	0.541	1510	2
60	3.00	86	3.06	0.608	1550	3
60	3.00	86	3.06	0.74	1620	4
60	3.00	86	3.06	1.04	1600	5

Concentration of 19-channel ceramic membrane at 105 mbar vacuum pressure

Temperature (°C)	TMP (bar)	Vacuum Pressure (mbar)	Feed pressure (bar)	Flux (LMH)	Permeate volume (ml)	Time (Hour)	Cross-flow velocity (m/s)	Permeate Volume (ml)	Feed conductivity
70	3.01	89	3.06	1.85	410	1	1	410	123171
70	3.01	89	3.06	2.16	480	2	1	480	123315
70	3.01	89	3.06	2.43	540	3	1	540	123786
70	3.01	89	3.06	1.85	410	4	1	410	124220
70	3.01	89	3.06	2.16	480	5	1	480	125305
70	3.01	89	3.06	2.07	460	6	1	460	125196

% RDS at 105 mbar vacuum pressure of 19-channel ceramic membrane

Permeate	Retentate	Time (Hours)	Outlet Temperature (°C)	Permeate conductivity ($\mu S/cm$)
1.85	12.1	1	30.1	0.98
2.16	12.4	2	34.4	0.87
2.43	12.27	3	42.4	4.05
1.85	12.37	4	41.8	1.02
1.44	12.07	5	30.5	0.91
1.17	12.20	6	27.8	0.76

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