

Understanding the evolution of organic fouling in membrane distillation through driving force and resistance analysis

Original

Understanding the evolution of organic fouling in membrane distillation through driving force and resistance analysis / Ricceri, F., Blankert, B., Ranieri, L., Piciooreanu, C., Ghaffour, N., Vrouwenvelder, J.S., Tiraferri, A., Fortunato, L.. - In: JOURNAL OF MEMBRANE SCIENCE. - ISSN 0376-7388. - 686:(2023), p. 121993. [10.1016/j.memsci.2023.121993]

Availability:

This version is available at: 11583/2982027 since: 2023-09-12T07:18:06Z

Publisher:

Elsevier

Published

DOI:10.1016/j.memsci.2023.121993

Terms of use:

This article is made available under terms and conditions as specified in the corresponding bibliographic description in the repository

Publisher copyright

Elsevier preprint/submitted version

Preprint (submitted version) of an article published in JOURNAL OF MEMBRANE SCIENCE © 2023,
<http://doi.org/10.1016/j.memsci.2023.121993>

(Article begins on next page)

1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
20
21

**Understanding the evolution of organic fouling in membrane distillation
through driving force and resistance analysis**

¹ Department of Environment, Land and Infrastructure Engineering (DIATI), Politecnico di Torino, Corso Duca degli Abruzzi 24, Turin, 10129, Italy

² CleanWaterCenter@PoliTo, Corso Duca degli Abruzzi 24, Turin, 10129, Italy

³ Water Desalination and Reuse Center (WDRC), Biological & Environmental Science & Engineering Division (BESE), King Abdullah University of Science and Technology (KAUST), Thuwal 23955-6900, Saudi Arabia

22 **Highlights**

- 23 • Flux decline and fouling thickness are linearly correlated.
- 24 • The feed to permeate heat transfer decreases during fouling.
- 25 • OCT allows direct analysis of the fouling layer development.
- 26 • Fouling thickness is governed by the permeate drag force and shear stress
27 increment.
- 28 • The balance between driving force and resistance yields a stationary flux.

29

30 **Abstract**

31 Fouling is one of the main issues hampering the implementation of thermally-driven membrane
32 distillation (MD). While the mutual influence of driving force and fouling deposition has been
33 critically assessed in pressure- and osmotically-driven processes, fouling mechanisms have not
34 been fully understood in MD. Using non-invasive optical coherence tomography, this study
35 describes for the first time the evolution of resistance and driving force evolution during the
36 development of the organic fouling layer in direct contact MD. Foulant layer thickness was
37 found to be strongly and linearly correlated to water flux under different conditions of feed
38 temperature and cross-flow velocity. Experimental and modeling results indicate that this
39 phenomenon is associated to the increase of the overall resistance to water vapor transport.
40 With a clean membrane, heat loss is mainly governed by the permeate flux and by temperature
41 polarization. As fouling evolves over time, temperature polarization increases and affects,
42 together with the additional fouling resistance, the water flux and the heat transfer from feed
43 to permeate. Indeed, foulant accumulation was observed to lead to a gradual reduction of heat
44 transfer from the feed to the permeate side, causing a steady increase of the average nominal
45 driving force, i.e., difference between vapor tension in the feed bulk and in the permeate bulk.
46 The driving force and the resistance evolved together during this dynamic process of fouling
47 development, resulting in the achievement of a near-stable flux value over time.

48

49 **Keywords:** membrane distillation; optical coherence tomography (OCT); organic fouling;
50 fouling evolution; driving force.

51 **1 Introduction**

52 Membrane distillation (MD) is a thermally-driven process for water desalination and for
53 the concentration of challenging wastewater and hypersaline feed [1]. The simplest of the
54 MD configurations is the so-called direct contact membrane distillation (DCMD). In
55 DCMD, the warmer feed solution is separated from the colder distillate product by a semi-
56 permeable hydrophobic membrane that only allows, under ideal conditions, the passage of
57 water vapor [2]. The driving force is linked to the temperature difference between the
58 warmer feed solution and the colder distillate stream. MD has been mainly investigated for
59 desalination purposes, but recent interest has grown also for the treatment of produced
60 waters, surface waters, and groundwaters with the goals of reuse and/or stream
61 concentration [3-7]. These feed solutions are all characterized by the presence of organic
62 substances, often consisting of natural organic matter and humic acids (HA). Srisurichan *et*
63 *al.* showed that when HA is combined with CaCl_2 , a heavy and dense HA foulant layer
64 forms during MD operation and results in a significant flux decline and increase of the heat
65 transfer resistance [8].

66 It is generally accepted that organic fouling in pressure-driven membrane processes is more
67 severe than in MD [9, 10]. On the other hand, fouling in MD is also still believed to be one
68 of the main factors limiting the commercial use of this technology [11]. In direct contact
69 membrane distillation, fouling involves both mass and heat transfer, which are highly
70 interconnected with each other. Here, fouling formation on the membrane surface affects
71 the mass transfer across the membrane, causing a decline in permeate flux. Since the
72 permeate flux is also responsible for transferring latent heat of vaporization from the feed
73 to the permeate side, as fouling evolves, also the heat flow decreases. In particular, the
74 overall heat transfer decreases during fouling in MD both due to the permeate flux decline
75 and for the formation of an additional thermally insulating layer [12]. Moreover, the fouling

76 layer results in temperature polarization, causing large temperature deviations with respect
77 to the nominal gradient [13]. In this complex phenomenon, the degree of resistance can be
78 theoretically calculated based on the characteristics of the fouling layer, such as its
79 thickness and porosity [8, 14]. However, the available data are usually not sufficiently
80 accurate as the fouling layer has been mainly characterized by destructive techniques,
81 limiting the possibility of providing insight into its development over time [15]. Indeed, so
82 far, most studies on fouling in MD have limited their scope to the water vapor productivity
83 along the filtration time or the recovery rate value [16, 17]. In recent years, optical coherence
84 tomography (OCT) has been used to conduct non-invasive dynamic analysis of the fouling
85 development in membrane systems [18, 19]. OCT allows spatial and time monitoring of
86 fouling development on the membrane module with micron resolution, as well as
87 morphological investigation of the deposited layer [20].

88 For the first time in this study, the OCT technology is used in combination with continuous
89 flux and temperature measurements to provide a complete assessment of fouling behavior
90 in direct contact membrane distillation and to identify the main factors dominating mass
91 and thermal transfer during this process. For this purpose, a feed with high load of humic
92 acids is used as of particular interest for MD applications and also due to the homogeneous
93 deposition that this matrix is able to form on the membrane surface, allowing more reliable
94 data analysis through the OCT. Moreover, the contributions of the various resistances to
95 mass transport are simultaneously analyzed and correlated to the loss of water vapor
96 productivity. As in the osmotic and pressure driven membrane processes, fouling thickness
97 evolution and the overall fouling behavior are linked to the variation of the driving force
98 and the fouling resistance. The permeate drag force and the shear stress also increase during
99 distillation showing a counteracting effect on the foulant deposition. Finally, important
100 insight into the fouling mechanism in MD under a wide range of realistic operating

101 conditions is provided and a new mechanism relating fouling to the magnitude of driving
102 force and resistance in DCMD is discussed.

103 **2 Materials and Methods**

104 **2.1 Membrane characteristics and feed solution composition.**

105 A synthetic feed solution with an initial humic acid (HA, Sigma-Aldrich) concentration of 500
106 mg/L was employed in this study. To accelerate the fouling deposition and to enhance high
107 load of HA solubilization, 20 mM of calcium chloride (CaCl₂, Sigma-Aldrich) was also added
108 to the feed solution [8, 25]. These concentrations of organics and salts can be typical of
109 produced water streams [17], for which MD is particularly appealing, but it is important to note
110 that a high load were mainly selected to accelerate fouling and to simulate an overall mass flow
111 of foulants that can be observed under long term operation in up-scaled systems. Initial volumes
112 of 1 L were used for both the feed and the distillate solutions. For the feed, 500 mL of pure
113 water were initially used for each test to evaluate a stable initial water flux. After stabilization,
114 a stock solution was added to reach the desired feed volume and concentrations. The
115 concentrate stock addition indicates the beginning of the fouling test. For all the experiments,
116 a commercially available membrane consisting of a hydrophobic polytetrafluoroethylene
117 active layer with a polypropylene support (PP-PTFE) (Membrane Solutions corp., US) was
118 used. The membrane characteristics are listed in Table 1, with several data provided by the
119 manufacturer. The membrane permeability coefficient was experimentally determined by
120 measuring the water flux and dividing it by the calculated vapor pressure difference across the
121 membrane at the operating temperatures.

122 **Table 1.** Porous PP-PTFE membrane characteristics

Parameter	Value	Units	Source
Thickness	174 - 245	µm	manufacturer
Mean pore size	0.22	µm	manufacturer
Bubble point	16-20.3	psi	manufacturer

Permeability coefficient	144	$\text{kg m}^{-2} \text{h}^{-1} \text{bar}^{-1}$	experiments
--------------------------	-----	--	-------------

123

124 **2.2 Direct contact membrane distillation lab system**

125 All the MD tests were performed in direct contact membrane distillation (DCMD)
126 configuration with a lab-scale flow-cell where the warm feed and the cold permeate were
127 flowed in countercurrent mode. The warm liquid stream is in direct contact with the membrane,
128 which allows vapors to pass and then condense at the distillate side in the cold water stream
129 that is also in contact with the membrane. The DCMD flow-cell was made of polymethyl
130 methacrylate and customized to allow in-situ characterization with OCT (see section 2.4). The
131 flow-cell had an active membrane area of 33 cm^2 with dimensions of $10 \times 3.3 \text{ cm}$ (length \times
132 width). Flow-cell images are reported in Fig S1 of the Supporting Information file (SI) together
133 with the overall setup describe here below. The temperatures in the warm feed and cold
134 permeate inlet streams were maintained constant throughout each experiment using two
135 separate heating circulators keeping the desired water temperature inside an insulated stainless-
136 steel bath(Corio-CD, Julabo, Germany), in which the coil circulating the streams was
137 immersed. The temperature sensors were integrated in the conductivity meters (TetraCon 325,
138 Xylem Analytics, Germany) located just before the inlet of the flow-cell. Before foulant
139 addition, those sensors allowed measurements of the stream temperatures entering the flow-
140 cell and the right settings of the heating circulators to get the desired inlet temperature of the
141 two streams. Two more sensors were used to measure the outlet temperature values of the feed
142 and the permeate streams during operation. Cross-flow velocities of both water streams were
143 measured by digital flow meters (MINI CORI-FLOWTM, Bronkhorst, Netherlands). The flux
144 across the membrane was calculated from the change in weight of the permeate tank over time,
145 measured through a computer-interfaced balance. All the instruments were digitally connected
146 and controlled by Lab View software (National Instruments, United States).

147 **2.3 Operating conditions and protocol of the fouling experiments**

148 In this study, we investigated feed temperatures in the range of 35 - 65 °C and feed cross-flow
149 velocities in the range of 0.2 - 0.4 m/s. Constant temperature of 20 °C and cross-flow velocity
150 of 0.1 m/s were maintained on the permeate side. The central composite design method was
151 used in Design Expert software to identify an efficient set of experiments within the ranges of
152 the selected feed temperatures and flow velocities (see Table S1, SI). For all the experiments,
153 the flux was first stabilized using de-ionized water as feed solution, without organic foulants,
154 thus obtaining a steady-state flux value referred to as J_{w0} . The fouling phase started when an
155 appropriate volume of organic foulant stock solution was added into the feed tank and it was
156 run until a volume concentration factor of 2.5 was achieved. A schematic presentation of the
157 described protocol is showed in Fig. S2. The decrement of flux (J_w) observed during the fouling
158 phase can be largely attributed to foulant deposition as salinity in the feed solution was low.
159 The maximum reduction of the feed vapor tension during tests was 0.02% from the initial value
160 due to CaCl_2 concentration (reduction due to HA concentration was even more negligible,
161 since HA accounted for less than 20% of the total contaminants weight). Therefore, the flux
162 decline ratio (J_w/J_{w0}) decreased during operation due to fouling. This parameter was adopted
163 to allow an easier correlation between flux and the fouling layer thickness analyzed through
164 the OCT, which is described in the next section.

165 **2.4 Direct monitoring of the fouling thickness with OCT**

166 A spectral-domain optical coherence tomography (SD-OCT) system (Ganymede II, Thorlabs
167 GmbH, Germany) was used to assess the fouling deposition on the membrane surface. The
168 instrument was equipped with a scan lens (LSM 03BB). The OCT probe was positioned on top
169 of the middle point along the length of the DCMD flow-cell, to monitor the fouling layer
170 development over time. Two-dimensional cross-section OCT scans were taken over a
171 transversal area of $8.0 \times 2.6 \text{ mm}^2$ (length \times depth). FiJi software was used to process the OCT

172 images by reducing the noise and adjusting contrast and brightness. The fouling layer thickness
 173 was calculated from the OCT images using a custom-made MATLAB code.

174 2.5 Heat transfer analysis

175 Heat balances

176 The heat balances over permeate volume (eq. (1)) and over the whole flow-cell (eq. (2)) can be
 177 written in terms of heat flows (or power), Q (J/s), here with incoming heat streams on the left-
 178 hand side and outgoing on the right-hand side of the balance equations:

$$179 \quad F_{P,in} \rho C_P T_{P,in} + Q_M = (F_{P,in} + F_w) \rho C_P T_{P,out} + Q_{P,loss} \quad (1)$$

$$180 \quad F_{F,in} \rho C_P T_{F,in} + F_{P,in} \rho C_P T_{P,in} = (F_{F,in} - F_w) \rho C_P T_{F,out} + (F_{P,in} + F_w) \rho C_P T_{P,out} + Q_{F,loss} + Q_{P,loss} \quad (2)$$

181 F_F and F_P are the volumetric flow rates (m^3/s) on the feed and permeate side of the membrane;
 182 T_F and T_P are the bulk temperatures (K) of feed and permeate, all at inlet and outlet (denoted
 183 by indices *in* and *out*). It is assumed that the heat capacity of water is approximately constant
 184 between 20 and 60 °C, with $C_P = 4180 \text{ J kg}^{-1}\text{K}^{-1}$, and the water density can also be taken as
 185 constant, $\rho = 998 \text{ kg/m}^{-3}$. The heat balances can be further simplified by assuming that the total
 186 flowrate of water passing the membrane, F_w , is negligibly small compared with the flows of
 187 feed and permeate ($F_w \ll F_{P,in}, F_{F,in}$). Also, the heat loss from the permeate side to exterior,
 188 $Q_{P,loss}$, can be neglected because the permeate temperature is very close to the ambient
 189 temperature, thus heat losses predominately occur at the hot side of the system as $Q_{F,loss}$. By
 190 introducing the notations for heat flows between inlet and outlet, Q_P for permeate side and Q_F
 191 for feed side:

$$192 \quad Q_P = F_P \rho C_P (T_{P,out} - T_{P,in}) \quad (3)$$

$$193 \quad Q_F = F_F \rho C_P (T_{F,in} - T_{F,out}) \quad (4)$$

194 the heat balances over permeate (eq. (5)) and over the flow-cell (eq. (6)) become:

$$195 \quad Q_M = Q_P \quad (5)$$

$$196 \quad Q_{F,loss} = Q_F - Q_P \quad (6)$$

197 Eq. (5) means that the permeate heat gain, Q_P , is a result of the total heat transfer through the
198 membrane $Q_M = Q_w + Q_m$ caused by the condensation of water, Q_w , and by conduction, Q_m .

199 Eq. (6) allows the estimation of heat loss through the feed channel walls to the exterior, as the
200 difference between the power change between inlet and outlet for feed and permeate.

201 *Temperature drop from feed to permeate*

202 In DCMD, heat is transferred from the feed to the permeate side by several mechanisms. First,
203 there is conduction from the warmer feed solution (bulk temperature T_F) across a thermal
204 boundary layer to the surface of the fouling layer (temperature $T_{F,L}$), which links to the
205 temperature polarization on the feed side. Second, if there is a fouling layer, this will induce an
206 additional heat transfer resistance by conduction, usually named cake-enhanced temperature
207 polarization, with a temperature drop from $T_{F,L}$ to $T_{F,M}$ at the membrane surface. Third, several
208 heat transfer mechanisms take place in the membrane: conduction through the membrane
209 polymeric material, conduction through the vapors in the pores (which may be negligible), and
210 an important heat flux due to evaporation/condensation at the feed/permeate interfaces. Finally,
211 there is also temperature polarization on the permeate side, with conduction in the quasi-
212 stagnant water layer adjacent to the membrane driven by the difference between $T_{P,M}$ at the
213 membrane surface at T_P in the bulk permeate.

214 The average temperature value in the feed and permeate side, T_F and T_P , was the arithmetic
215 mean of measured inlet and outlet water temperatures. The difference between the feed and

216 permeate bulk temperature give the total temperature drop between the feed and permeate
 217 across the thermal boundary layers ($\Delta T_F, \Delta T_P$), foulant layer (ΔT_L) and membrane (ΔT_M):

$$218 \quad T_F - T_P = \Delta T_F + \Delta T_L + \Delta T_M + \Delta T_P \quad (7)$$

219 However, the individual temperature differences can also be estimated. Assuming the existence
 220 of a fouling layer, the continuity of total heat flux through the membrane, q_M ($\text{J m}^{-2}\text{s}^{-1}$), implies:

$$221 \quad q_M = h_F (T_F - T_{F,L}) = h_L (T_{F,L} - T_{F,M}) = h_M (T_{F,M} - T_{P,M}) = h_P (T_{P,M} - T_P) \quad (8)$$

222 with the four heat transfer coefficients ($\text{J m}^{-2}\text{s}^{-1}\text{K}^{-1}$) as h_F and h_P for thermal boundary layers
 223 on the feed and permeate side, h_L through the fouling layer, and h_M through the membrane (a
 224 lumped value involving both evaporation/condensation and conduction). With the heat flux as
 225 heat flow $Q_M = Q_P$ divided by membrane area, A_M (0.0033 m^2), one can express the
 226 temperature differences across the different layers:

$$227 \quad \Delta T_F = T_F - T_{F,L} = \frac{Q_P}{h_F A_M}, \quad \Delta T_L = T_{F,L} - T_{F,M} = \frac{Q_P}{h_L A_M}$$

$$228 \quad \Delta T_M = T_{F,M} - T_{P,M} = \frac{Q_P}{h_M A_M}, \quad \Delta T_P = T_{P,M} - T_P = \frac{Q_P}{h_P A_M}$$

229 Average heat transfer coefficients through the thermal boundary layers, h_F and h_P , may be
 230 estimated from correlations for the Nusselt number, $\text{Nu} = hH/k$, function of Reynolds
 231 number, $\text{Re} = uH\rho/\mu$, and Prandtl number, $\text{Pr} = C_p\mu/k$. These involve the physical
 232 properties of water (density ρ , dynamic viscosity μ , specific heat C_p and thermal conductivity
 233 k), as well as the water velocity u and the height H of the specific channel (feed or permeate)
 234 that decreases at the feed side in time due to the growth of the fouling layer. While ρ , C_p and k
 235 ($0.6 \text{ J s}^{-1}\text{m}^{-1}\text{K}^{-1}$) can be assumed constant in the interval of temperatures, viscosity has a
 236 significant change taken as $\mu = 0.497(42.5 + T)^{-1.5}$ with T in $^\circ\text{C}$. Generally, convective heat

237 transfer correlations are developed based on the assumption of small rate of mass transfer. We
 238 adopted the Nusselt relation for heat transfer between two plates, for both feed and permeate
 239 channels, $\text{Nu} = 0.664 \text{Re}^{1/2} \text{Pr}^{1/3}$. This allows estimation of $\Delta T_F = \frac{Q_P}{A_M} \frac{H_F}{k_F \text{Nu}_F}$ and
 240 $\Delta T_P = \frac{Q_P}{A_M} \frac{H_P}{k_P \text{Nu}_P}$. The fitting of experiments before foulant addition also allowed a
 241 determination of the temperature difference across the membrane as a function of the measured
 242 flux (see Fig. S3), $\Delta T_M = f(J_w)$, where the water flux is $J_w = F_w/A_M$. Finally, the
 243 temperature drop over the fouling layer can be computed during each test by:

$$244 \quad \Delta T_L = T_F - T_P - \Delta T_F - \Delta T_M - \Delta T_P = \frac{Q_P}{A_M} \frac{\delta_L}{k_L}$$

245 This assumes heat transfer through the fouling layer occurs predominately by conduction.
 246 Equations were implemented in custom MATLAB code processing OCT and MD data
 247 acquired during each test. The value of k_L was thus retrieved for each test prior assuming this
 248 value as $0.6 \text{ W m}^{-1} \text{K}^{-1}$, i.e., water thermal conductivity. Thus, this value was assumed constant
 249 to retrieve the Nusselt number during fouling evolution for each test.

250 **2.6 Driving force and fouling resistance analysis**

251 In MD, the driving force for the mass transfer is provided by the water vapor pressure
 252 difference between the feed, p_F , and the permeate side, p_P . The vapor pressure p (Pa) for
 253 water can be calculated from the Antoine equation $p = \exp(23.238 - 3841/(T - 45))$ with T
 254 in K. The nominal driving force calculated in this study is based on the average value from the
 255 inlet and outlet bulk temperatures of the feed stream, T_F , and the permeate stream, T_P . The
 256 overall resistance to mass transfer between feed and permeate, R , was calculated as ratio
 257 between the driving force DF and the measured water vapor flux J_w [24]:

258
$$R = \frac{DF}{J_w} = \frac{p_F - p_P}{J_w} \quad (9)$$

259 Changes in the driving force and in the resistance were evaluated during the tests to determine
260 the mechanism of fouling development.

261

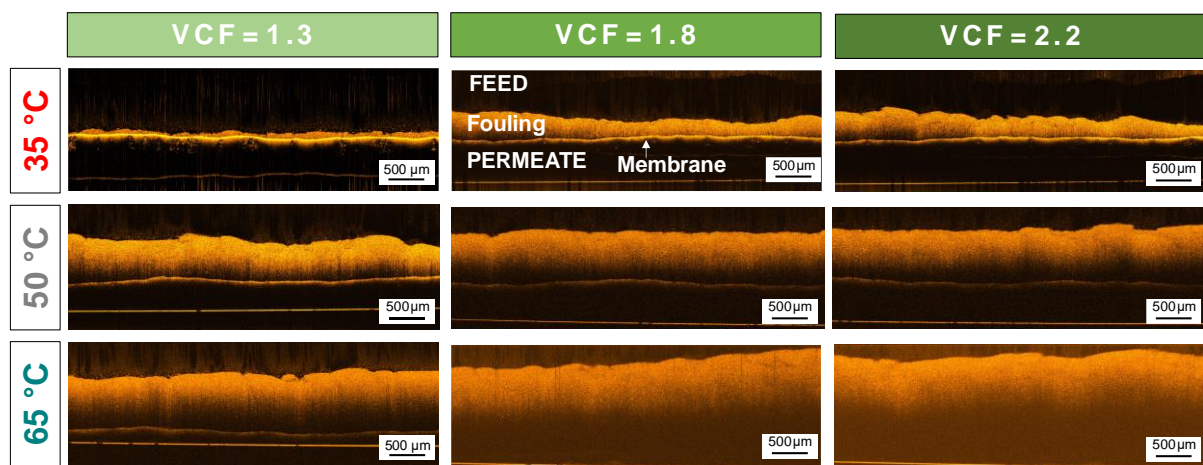
262 **3 Results and discussion**

263 **3.1 Impact of temperature and cross-flow velocity on process performance and fouling** 264 **development**

265 Fouling evolution in membrane distillation was evaluated in this study under different realistic
266 conditions of feed inlet temperature ($T_{F,in}$) and cross-flow velocity (u_F). As expected, the initial
267 flux was strongly correlated to the applied feed inlet temperature. Results reported in Fig. S4
268 suggest that the initial flux can be increased from ~3 to 22.5 kg m⁻²h⁻¹ by increasing $T_{F,in}$ from
269 35 to 65 °C. This is a result of the nature of the MD driving force, i.e., the vapor tension
270 difference between the feed and the permeate [26]. On the other hand, an almost negligible
271 effect on J_w was observed when increasing the cross-flow velocity. Indeed, despite u_F plays a
272 role on the heat transfer coefficient and consequent temperature polarization, this effect was
273 negligible compared to the applied feed temperature. It is also important to note that in this
274 study the membrane housing length was sufficiently small to minimize the temperature profiles
275 along the cross-flow direction.

276 The fouling layer development over time was monitored with OCT microscopy. Fig. 1 presents
277 representative OCT scans obtained for tests performed with applied feed inlet temperatures of
278 35, 50, and 65 °C, and acquired at the three volume concentration factors (VCF) of 1.3, 1.8,
279 and 2.5. Images refer to the tests performed at a cross-flow velocity of 0.3 m/s. The fouling
280 layer always increased during distillation: in details, the measured thickness when operating at

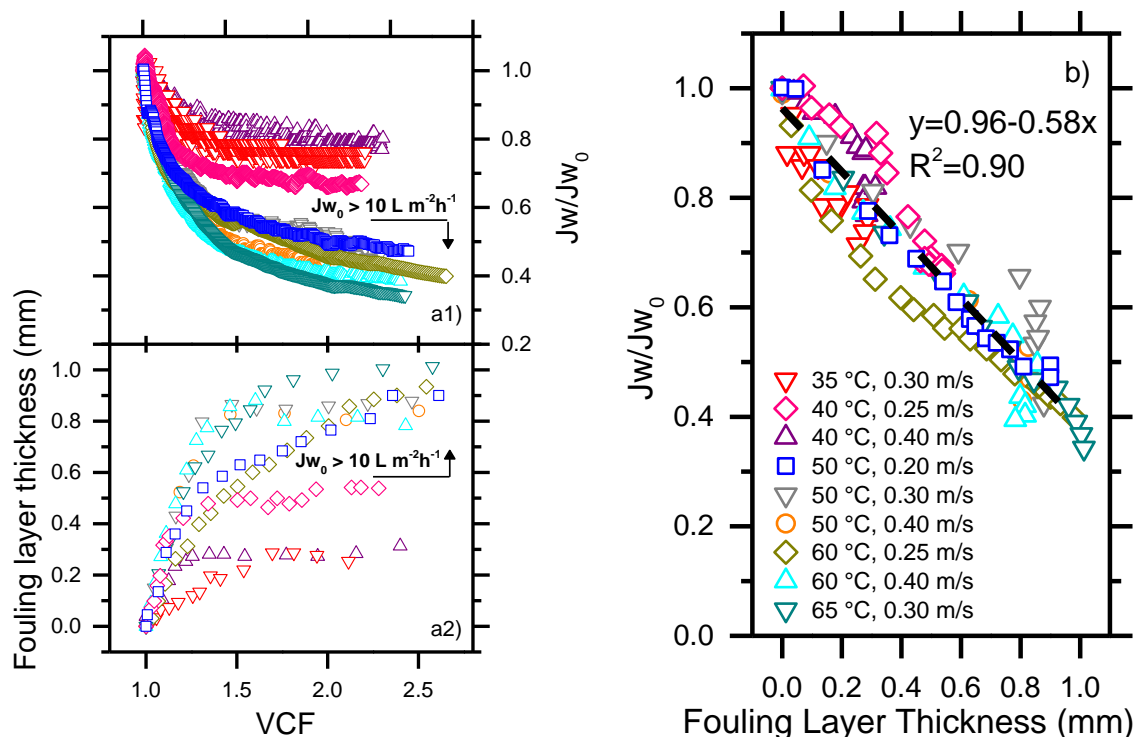
281 65 °C and 50 °C changed from 622 μm to 1013 μm and from 355 μm to 878 μm, respectively,
 282 when the VCF was increased from 1.3 to 2.2. Indeed, for all VCF values reported in Fig 1, a
 283 decrease of fouling layer thickness was observed when the feed inlet temperature was lowered.
 284 Interestingly, considerable lower deposition was observed for the experiment performed at 35
 285 °C, whereby thickness only changed from 196 μm to 273 μm in the same VCF range.
 286 Therefore, the fouling deposition rate in DCMD may be directly linked to the feed temperature:
 287 severe organic fouling is observed at higher temperatures.



288 **Fig. 1.** Cross-sectional OCT scans of the fouling layer deposited on the membrane surface, at
 289 three VCF values, for experiments performed at fixed cross-flow velocity of 0.3 m/s and with
 290 different inlet feed temperatures (35, 50, 65 °C). The OCT scans were acquired at the middle
 291 position of the membrane cell during continuous operation.
 292

293
 294 The dependency of flux decline ratio and fouling thickness on the VCF is shown in Fig. 2. In
 295 general, the flux decline (Fig. 2a1) showed a similar behavior of fouling layer thickness (Fig.
 296 2a2). This evidence highlights how these parameters are similarly influenced by the different
 297 applied conditions of $T_{F,in}$ and u_F . The cross-flow velocity mainly influenced the flux and
 298 fouling deposition at lower inlet feed temperatures: the J_w/J_{w0} varied from 0.8 to 0.6 (higher
 299 flux decline) when decreasing u_F from 0.40 to 0.25 m/s at 40 °C, while negligible influence on
 300 the final value of J_w/J_{w0} was observed by changing the u_F for the experiments performed at 50
 301 and 60 °C. Symmetrically, lower final layer thickness was observed when increasing the cross-

302 flow velocity at 40 °C, from 538 μm to 313 μm, while negligible difference was obtained when
 303 the u_F was increased at a $T_{F,in}$ of 50 °C and 60 °C. This effect could be attributed to the reduced
 304 ability of the cross-flow velocity in counteracting fouling when operating at higher flux. On
 305 the other hand, a strong effect was played by $T_{F,in}$, whose increase generally reduced the value
 306 of J_w/J_{w0} and evidently increased the fouling deposition along the VCF. This result is in good
 307 agreement with previous studies and may be correlated to the role of temperature in increasing
 308 both initial flux (see Fig. S4) and temperature polarization [27-29].



309 **Fig. 2.** (a1) Normalized flux J_w/J_{w0} and (a2) fouling thickness development plotted against
 310 VCF. The experiments were performed with the synthetic feed water in the presence of humic
 311 acid and calcium at different initial permeate flux J_{w0} obtained by varying the feed inlet
 312 temperature $T_{F,in}$ and the cross-flow velocity u_F in DCMD configuration. Fouling layer
 313 thickness was determined from the OCT scans acquired in the middle position of the cell during
 314 continuous operation. (b) The correlation between fouling layer thickness and normalized
 315 permeate flux. The coefficient of determination R^2 is for the line fitting all the experimental
 316 data.

317

318 Interestingly, from the data presented in both Fig. 2a1 and 2a2, it is possible to notice the
 319 presence of a low- and a high- fouling region, respectively, below and above the initial flux of

320 roughly $10 \text{ L m}^{-2}\text{h}^{-1}$. Note that J_{w0} below this value are obtained when experiments are
321 performed at a feed temperature below $50 \text{ }^\circ\text{C}$ (see Fig. S4), confirming previous findings [7,
322 19, 25]. In parallel, below this $T_{F,in}$ value, also considerable lower foulant accumulation was
323 measured (see Fig. 2a1). Previous studies linked the presence of different fouling regions to
324 the existence of a possible threshold flux [30]. The concept of critical or threshold flux has
325 been widely reported in osmotically-and pressure-driven membrane processes [22, 31].

326 Fig. 2b shows the flux decline J_w/J_{w0} as a function of the fouling layer thickness, both measured
327 at different times during the various experiments. Overall, the data reflect the inverse linear
328 dependency between the two parameters, regardless of the experimental conditions ($R^2=0.9$ for
329 the aggregate regression). This result is in agreement with previous studies on wastewater
330 treatment with MD and confirms how, as in the other membrane separation process, the flux
331 decline during long-term operation is directly correlated to the amount of fouling deposited on
332 the membrane surface [32, 33]. A possible explanation of this strong correlation is associated
333 with the magnitude of the driving force for water separation (vapor tension difference between
334 feed and permeate side of the membrane). Any loss of flux may be caused by a corresponding
335 percentage of the driving force lost along the fouling layer thickness, a phenomenon named
336 cake-enhanced temperature polarization [28, 29, 34]. Such simple linear correlation implies the
337 possibility to obtaining a reasonable estimation of the foulant accumulation based on flux data
338 during DCMD operation for a wide range of feed inlet temperatures and cross-flow velocities.
339 From this prospective, there is still a lack of knowledge related to modelling of foulant
340 deposition in DCMD.

341 In summary, this investigation highlights how flux decline rate and fouling layer thickness
342 similarly increased with the feed temperature and closely correlated with initial flux. On the
343 other hand, the value of the cross-flow velocity did not show a clear influence on flux, while it
344 mainly decreased the fouling layer thickness. Interestingly, a low fouling region was identified

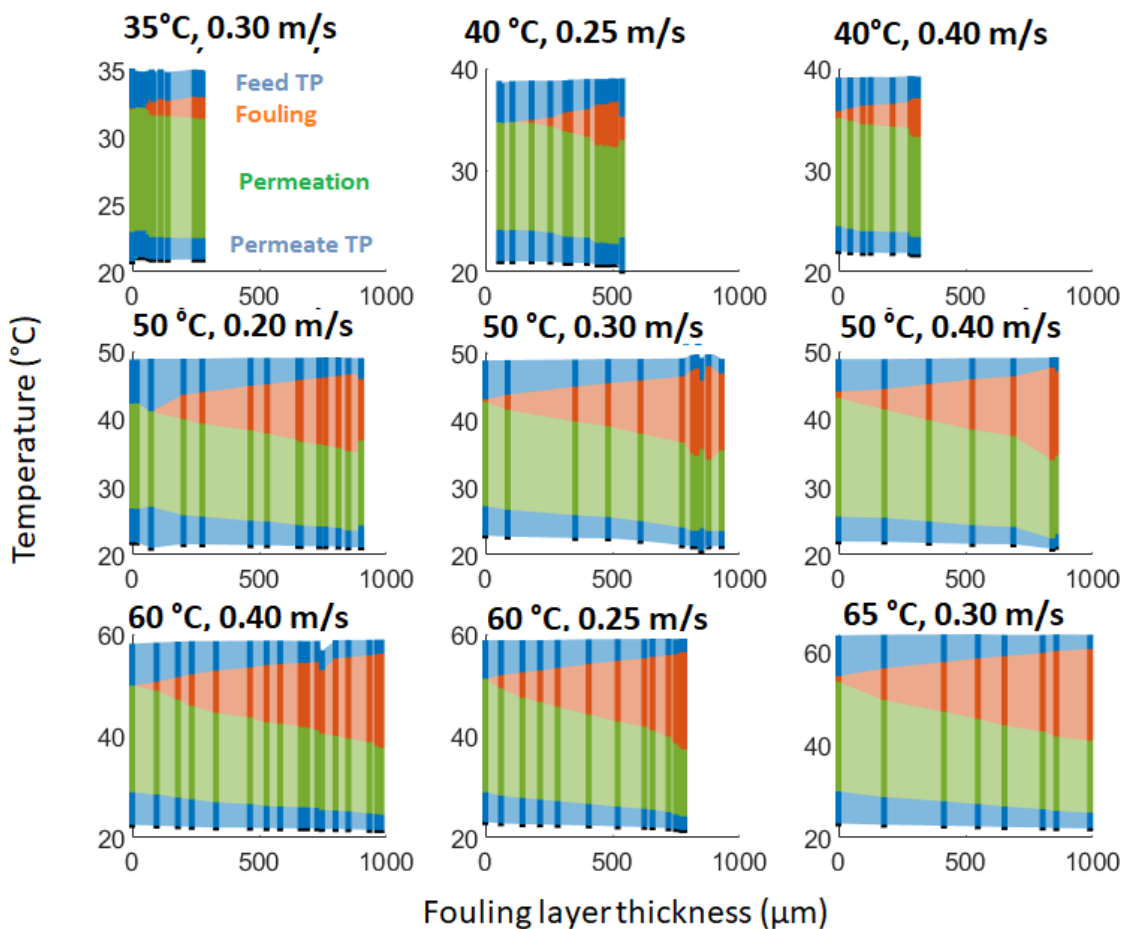
345 when operating below 50°C of inlet feed temperature. Finally, strong linear correlation
346 between the fouling layer development and the vapor permeate flux was observed.

347 **3.2 Analysis of the factors contributing to the temperature evolution during fouling**

348 The membrane, the fouling layer, and temperature polarization layers can be seen as a series
349 of resistances to water flux, as each layer accounts for a certain amount of heat loss from
350 the hot to the cold side. Thus, the nominal driving force (vapor tension difference between
351 bulk feed and permeate) is lowered to a net driving force across the membrane (vapor
352 tension difference between the two membrane interfaces, at the feed and at the permeate
353 side). The temperature profile on both the feed and the permeate side before foulant
354 addition (clean membrane) was retrieved for all the experiments from Eq.7, which also
355 accounts for temperature polarization. The temperature drop across the membrane was then
356 retrieved as a difference between the temperature values as a result of polarization in the
357 two channels. Linear interpolation of these results, shown in Figure S3, allowed estimation
358 of the temperature drop across the membrane for any flux values observed during the
359 experiments. The growth of a fouling layer exacerbated the driving force losses, which
360 added to the effects already present for clean membranes.

361 The separate contribution of each heat transfer resistance from feed to permeate during the
362 fouling process is presented in Fig.3. The temperature profile was plotted against the
363 fouling layer thickness and calculated by Eqs. 1-7. In all tests, the temperature loss due to
364 polarization decreased during fouling (see blue color bands in Fig 3), due to the decrement
365 of the heat flow from feed to permeate (Eq. 3). This is due to the increment of the total heat
366 subtracted by the increasing fouling thickness fouling and by the decrement of permeation
367 (see orange and green color bar in Fig 3, respectively). According to the model, the feed
368 temperature polarization (ΔT_F) reduction was more pronounced than that in the permeate

369 stream (ΔT_p), because the fouling layer growth also decreased the effective height of the
 370 feed channel. This effect can be observed under all the tested conditions in Fig 3, by
 371 comparing the two blue bands with each other. The foulant layer thickness influences the
 372 feed temperature polarization term, due to the heat flow reduction from feed to permeate
 373 and also because the reduction of flow channel height increases the cross-flow velocity,
 374 which thereby decreases the temperature boundary layer thickness. On the other hand, an
 375 enhanced cross-flow velocity leads to more shear stress over the fouling layer, which can
 376 be accountable for the decreasing effective deposition of foulants in time. This mechanism
 377 can explain the gradual approach of a foulant thickness plateau during the last phase of the
 378 experiments (see Fig. 2a2).



379

380 **Fig. 3.** Separate contributions of each heat transfer resistance in the overall temperature loss,
 381 estimated during fouling development for experiments performed in different conditions. The

382 uppermost and lowermost values indicate the bulk temperatures in the feed and permeate
383 during fouling, as average between the measured inlet and outlet temperature. Blue areas:
384 temperature polarization in feed (ΔT_F , top) and permeate (ΔT_P , bottom); Red area:
385 temperature drop through fouling layer (ΔT_L); Green area: temperature drop over the
386 membrane (water vapor flux contribution, ΔT_M).

387

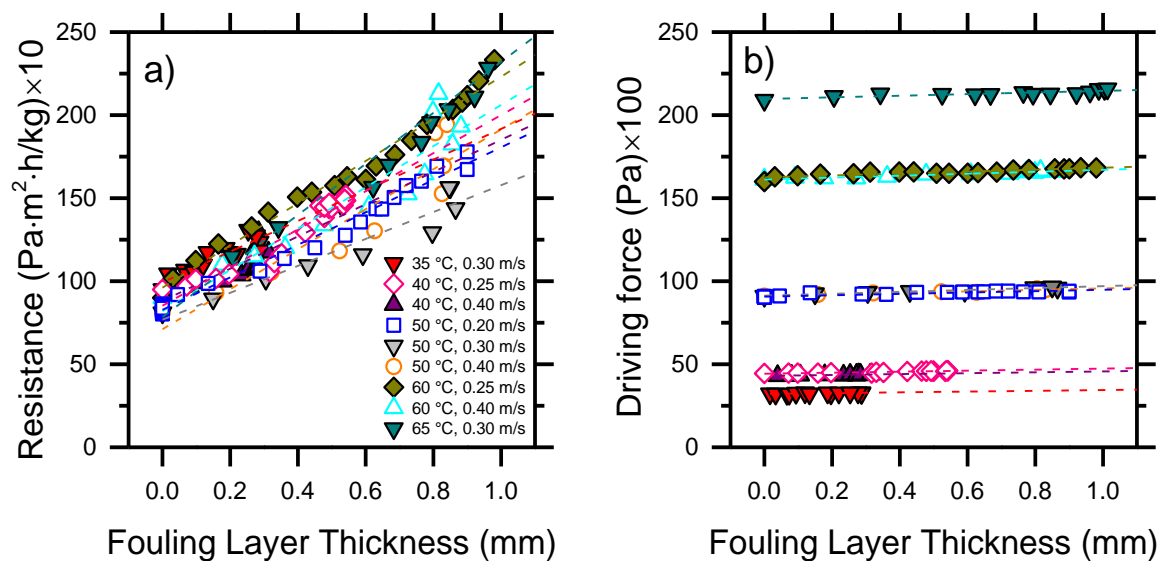
388 In all tests, the water vapor heat flow generally decreased as fouling accumulation occurred
389 (see green band in Fig. 3). Indeed, higher feed temperatures generate more rapid flux
390 decline, which is here reflected by the higher discrepancy in the amount of heat loss due to
391 permeation from the beginning to the end of the tests when increasing the feed inlet
392 temperature. As an example, the test at 35 °C shows a heat loss due to permeation of roughly
393 10 °C during the entire duration of the test, while at 65°C feed bulk temperature, the related
394 heat loss decreased from 30°C to 20 °C throughout the test. In parallel, the analysis allows
395 estimation of the amount of heat subtracted by the growing fouling layer. This portion
396 gradually increased in all the experiments as fouling layer thickness evolved as a result of
397 foulant deposition, while both temperature polarizations and convective heat decreased.
398 Once again, as the fouling thickness is mainly governed by the feed temperature, heat
399 subtracted by the fouling layer increased more for tests performed at higher temperatures.
400 As an example, the highest difference can be observed by comparing again the 35 °C and
401 65 °C test, where the final temperature decline within the fouling thickness was about 3 °C
402 and 20 °C, respectively. In other words, while the loss of driving force is related mainly to
403 temperature polarizations and water vapor flux with a clean membrane, fouling becomes
404 relatively more and more significant in terms of driving force losses compared to the other
405 two phenomena during operation. Interestingly, this behavior evolved similarly regardless of
406 the operational conditions.

407 Another consideration resulting from the analysis is that a small variation of the two average
408 bulk temperature profiles from initial values occurred during operation. Specifically, the

409 average bulk feed temperature was estimated to increase while the average permeate
410 temperature to decrease slightly as organic foulants deposited onto the membranes. This
411 mechanism translated into an overall small increment of the nominal (bulk) driving force, the
412 net effect of a gradually lower amount of heat transferred from the feed to the permeate side
413 due to fouling accumulation and the concomitant flux decline. In other words, the average
414 nominal driving force in the membrane housing increased in time, while the net driving force
415 decreased due to foulant accumulation. Despite the effect on the bulk driving force was not
416 particularly pronounced in this study due to the small size of the membrane, it has important
417 implications on the fouling evolution, discussed in depth below, and it would be much more
418 significant in large-scale systems.

419 **3.3 Overall driving force and resistance analysis during fouling in DCMD**

420 To approach a mechanistic explanation of how fouling evolves in DCMD, the driving force
421 (DF) and the fouling resistance (R) are investigated in this section in the light of the measured
422 fouling layer thickness. Fig. 4 shows the values of the two parameters calculated for
423 experiments performed under different conditions of feed inlet temperature and cross-flow
424 velocity. A linear fit was calculated for each experiment, with intercept and standard error
425 values reported in Table S2 (SI). The overall resistance to the permeate flux was calculated by
426 Eq. 8. All tests started with similar resistance in the range between 700 and 1000 Pa m² h kg⁻¹,
427 as highlighted by the intercept values in Figure 4a. Results imply that fouling deposition
428 inevitably leads to the development of an additional resistance: the good quality of the linear
429 fit implies the role of the fouling thickness on the proportional increment of total resistance,
430 consistent with the discussion of the sections above. Furthermore, the rate of resistance
431 increment was dependent on the operating conditions, possibly due to the different flux decline
432 rates (Fig. 2).



433 **Fig. 4.** (a) Total resistance, R , and (b) nominal driving force, DF , calculated from the data
 434 reported in Fig. 1, as a function of the foulant thickness. The fits are shown by dash line with
 435 the respective colors. The multiplicative factors indicates that the showed y-axes values must
 436 be multiplied for ten and one hundred to get the real R and DF values, respectively.

437

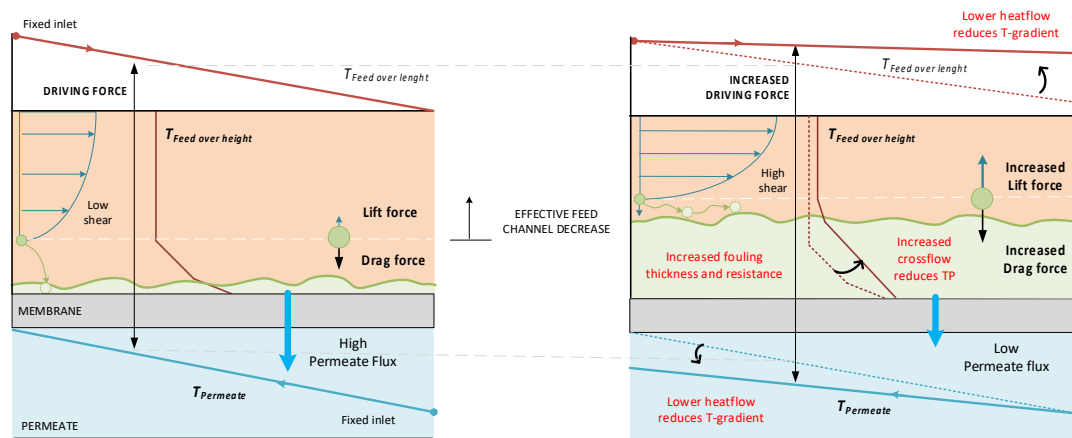
438 The initial values of the total driving force depend solely on the initial operating feed
 439 temperatures. As discussed in the previous section, organic fouling caused an increase in the
 440 average nominal driving force, rationalized with the decrement of the heat transfer from the
 441 feed to the permeate [35]. The slopes reported in Table S2 suggest that also this effect was
 442 somewhat proportional to the initial temperature. In summary, both the total resistance and the
 443 bulk driving force increased during operation. A clearer picture of this mutual increment can
 444 be observed in Fig. S5, reporting results for all the experiments performed at 50 °C. It is
 445 important to note that, while the reduction in polarization phenomena with fouling is present
 446 in all membrane-based filtration processes, the increment of the nominal driving force during
 447 fouling is not a mechanism shared by all filtration processes. For example, in pressure-driven
 448 membrane processes, the nominal driving force is largely independent of foulant deposition, as
 449 the fouling layer has not been found to significantly influence the pressure profiles within the
 450 channels of the membrane housing or module [22, 24]. To summarize, during fouling
 451 deposition both the total resistance and the average nominal driving force increased, translating

452 into a nearly constant net driving force, hence stabilized flux, attained after a certain time. This
453 phenomenon is mainly attributed to a decrement of the convective heat flow from the feed to
454 the permeate as an additional resistance is generated by fouling layer accumulation.

455 **3.4 Proposed fouling evolution mechanism in DCMD**

456 Although fouling is a continuous process, it can be also described as the result of discreet
457 steps to mechanistically depict the evolution of deposition as governed by the mutual
458 increment of the driving force and mass transfer resistance (Fig. 5). This type of
459 mechanistic model has already been used in osmotically- and pressure-driven processes,
460 since it is not dependent on the nature of the driving force [21]. According to the results
461 obtained in this study, the increase of the fouling layer thickness on the membrane surface
462 linearly increased the total resistance to the water vapor flux. This phenomenon was found
463 to be analogous under a wide range of investigated conditions. The increase in resistance
464 translated into a heat transfer reduction from the feed to the permeate side and a consequent
465 increment of the average nominal driving force, i.e., the difference between the average
466 bulk temperature of the feed and of the permeate stream. This increase in driving force and
467 the reduction of polarization outside the fouling layer inevitably caused the permeate drag
468 force to increase, likely causing more foulant deposition. The feedback between driving
469 force and resistance continued to evolve as fouling thickness increased in a framework whereby
470 also hydrodynamic conditions were influenced by the growing layer. In the schematic of Fig.
471 5, it can be observed how in the initial stage of fouling, the permeate drag should dominate
472 fouling deposition as only partially counterbalanced by the shear stress. A strong decrement of
473 the permeate flux is observed as result of fouling accumulation (Fig. 5a). However, with layer
474 growth, the feed channel would tend to narrow with a consequent increment of the effective
475 cross-flow velocity (Fig. 5b, feed side). The shear stress thus thwarts the further deposition of
476 foulants and promotes foulant back-transport, therefore preventing further growth of the

477 fouling layer [36-38]. During the late stage of filtration, the fouling layer thickness approached
 478 a near-stable value as the lift and drag forces reached equilibrium in the feed channel (see Fig
 479 2a2). In parallel, also the flux approached a near-stable value (see Fig 2a1). Therefore, as
 480 overall effect, the mutual increase of both the average nominal driving force and of the overall
 481 resistance should lead to a self-compensation phenomenon during which the fouling thickness,
 482 the effective cross-flow velocity and the cake-enhanced temperature polarization reach near
 483 equilibrium and productivity reaches a near steady-state condition.



484
 485 **Fig. 5.** Micro to macroscale analysis of the mechanisms occurring upon organic foulant
 486 deposition and fouling layer formation in direct contact membrane distillation, including
 487 influence on the hydrodynamic parameters, effects on temperature profiles and on the driving
 488 force. Left panel refers to conditions of clean membrane, right panel to conditions after cake
 489 layer formation. The experiment were performed with a feed solution consisting of 500 mg/L
 490 of humic acid and 20 mM of CaCl₂.

491

492 4 Conclusion

493 This study presented an analysis of organic fouling in membrane distillation under a wide range
 494 of temperature and cross-flow velocity conditions. During operation, the analysis of the fouling
 495 layer with OCT highlighted the linear correlation between layer thickness increase and flux
 496 reduction. An investigation of the main resistances to water vapor flux confirmed that the
 497 heat transfer was reduced due to fouling accumulation. In particular, the amount of heat

498 loss within the fouling layer grew at the expenses of lower heat loss due to convective heat
499 (i.e., heat transported by water vapor flux) and to the two temperature polarizations at the
500 feed and permeate side. Also, the average nominal driving force increased while the overall
501 resistance also increased, overall reducing the water vapor flux. Results suggest that fouling
502 resistance and the driving force evolved together and governed the fouling evolution dynamics
503 over time. Fouling was found to be a dynamic phenomenon, whereby governing factors
504 evolved together resulting in a final near steady-state productivity value as net result. This
505 continuous process may be discretized in steps as follows:

- 506 (i) Fouling deposition increases the overall resistance to the water vapor flux.
- 507 (ii) The fouling layer and the related water flux decrement reduce the heat transferred
508 from the feed to the permeate stream with the overall effect of an increase in
509 average nominal driving force, i.e., bulk temperature difference between the
510 feed and the permeate side.
- 511 (iii) As in the other membrane process, the driving force increase inevitably leads to
512 an increment of the permeate drag force, which thereby promotes foulant transport
513 and accumulation onto the membrane surface.
- 514 (iv) The mutual increments of the driving force and fouling resistance generate a self-
515 compensation phenomenon which is responsible for the near-stable flux gradually
516 approached during fouling development.
- 517 (v) The plateau is reached for both flux and fouling thickness as the gradual fouling
518 accumulation also leads to an increase of the shear force in the feed channel that
519 thwarts foulant deposition and counterbalances the increasing permeate drag force
520 of foulant toward the membrane.

521

522 **Acknowledgements**

523 The research reported in this paper was supported by funding from King Abdullah University
524 of Science and Technology (KAUST), Saudi Arabia and by Politecnico di Torino, Italy.
525 Francesco Ricceri acknowledges funding from the CleanWaterCenter@PoliTo for his Ph.D.
526 scholarship (01_TRIN_CI_CWC).

527
528
529
530
531
532
533
534
535
536
537
538
539
540
541
542
543
544
545
546
547
548
549
550
551

Supplementary Material

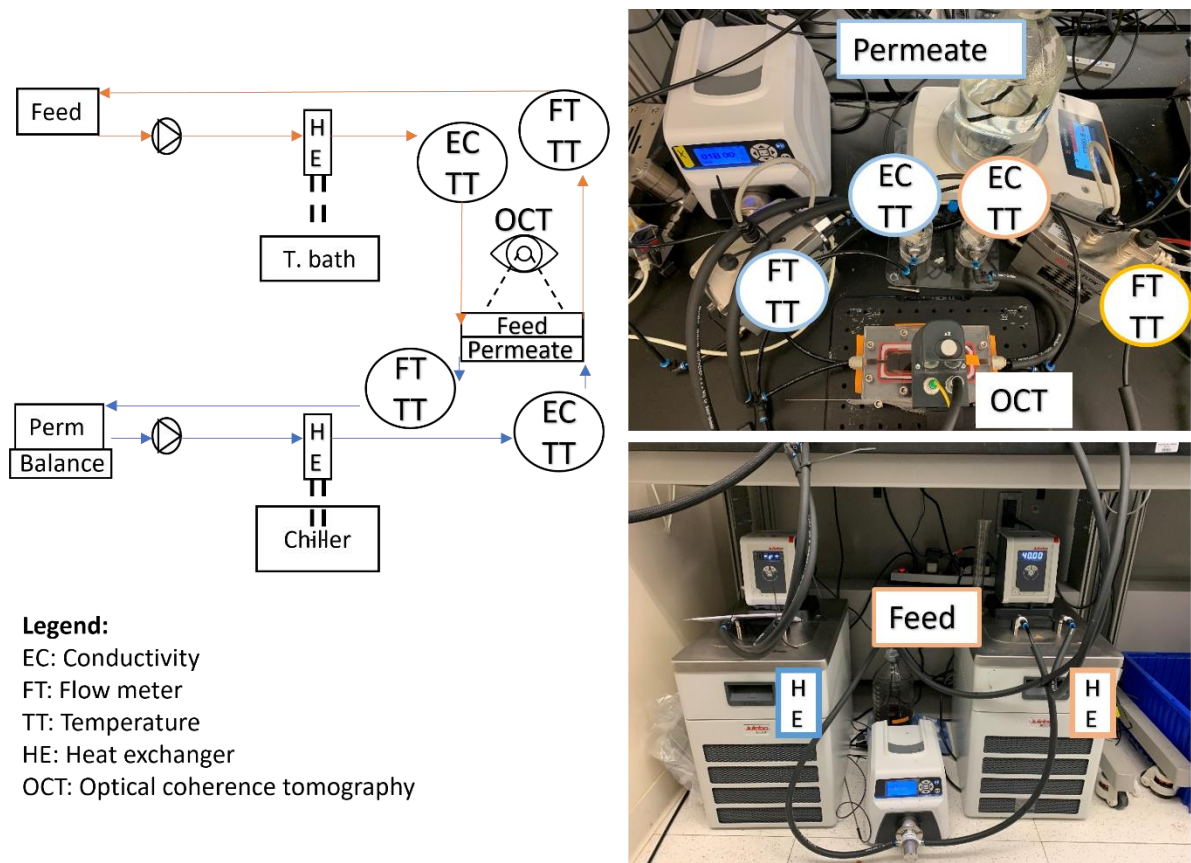
Understanding organic fouling evolution in membrane distillation through driving force and resistance analysis

Francesco Ricceri^{1,2},

1: Department of Environment, Land and Infrastructure Engineering, Politecnico di Torino, Corso Duca degli Abruzzi, 24 – 10129 Torino (Italy)

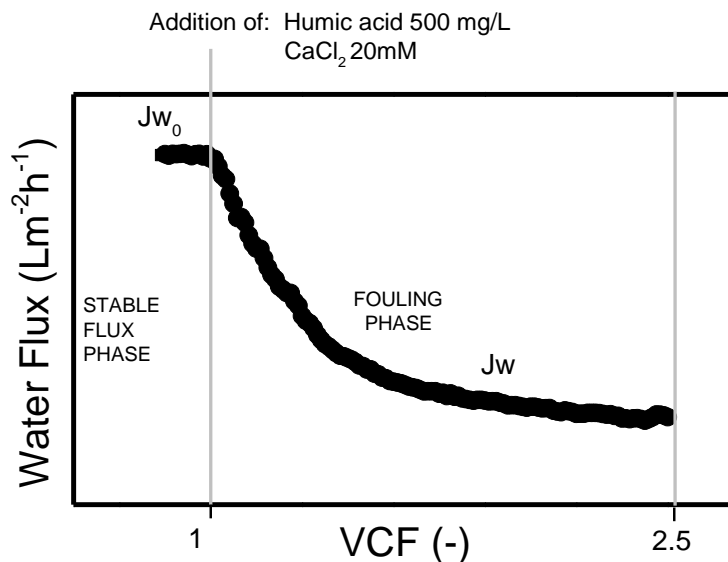
2: CleanWaterCenter@PoliTo, Corso Duca degli Abruzzi, 24 – 10129 Torino (Italy), web:

<http://cleanwater.polito.it/>



552

553 **Fig. S1.** Schematic representation and images of the setup used for this study. The orange
 554 and blue lines are used for the warm feed and cold permeate respectively, the legend indicates
 555 the acronyms used for each component. Regarding the OCT, only the camera positioned on
 556 the transparent cell is shown in this picture.

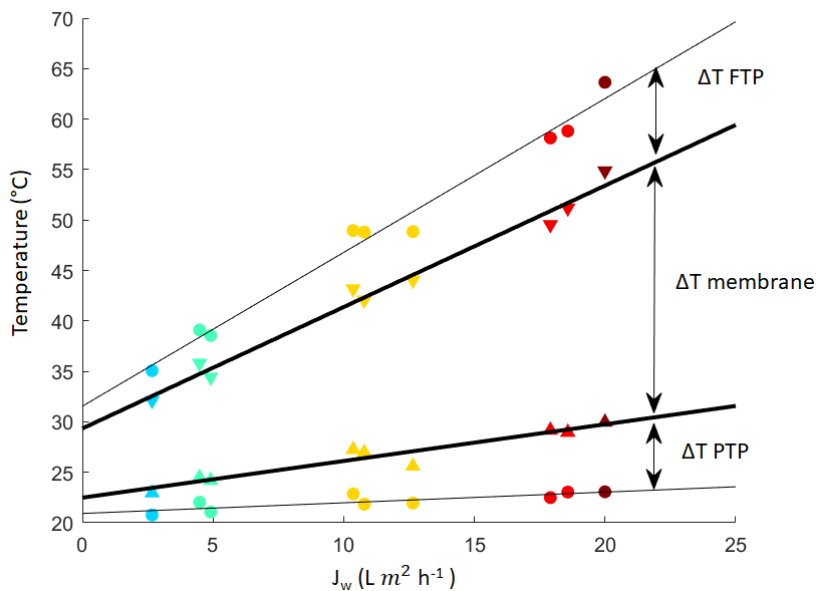


557

558 **Fig. S2.** Representative experimental determination of water flux through the MD membrane.
 559 The protocol consisted of two stages: (i) the initial stabilization of flux, J_{w0} , using deionized
 560 water, for roughly 30 min; (ii) the fouling phase started at the volume concentration (VCF)

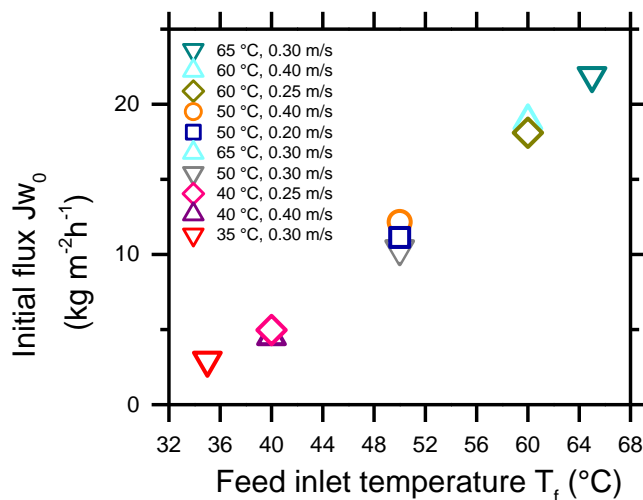
561 equal to 1 and was carried out for a volume concentration factor of 2.5. Here, the flux is
 562 indicated by J_w .

563



564

565 **Fig. S3.** Temperature profile between the feed and permeate bulk solutions before fouling
 566 (clean membrane). The profile was retrieved by implementing Eq. 3, related to the temperature
 567 profile in the presence of temperature polarization, for all the performed experiments (Table
 568 S1). The temperature drop across the membrane was retrieved as difference between the
 569 temperature values at the membrane interface. The fitting was performed to assess the
 570 temperature polarization and temperature difference across the membrane for any measured
 571 flux even during fouling.



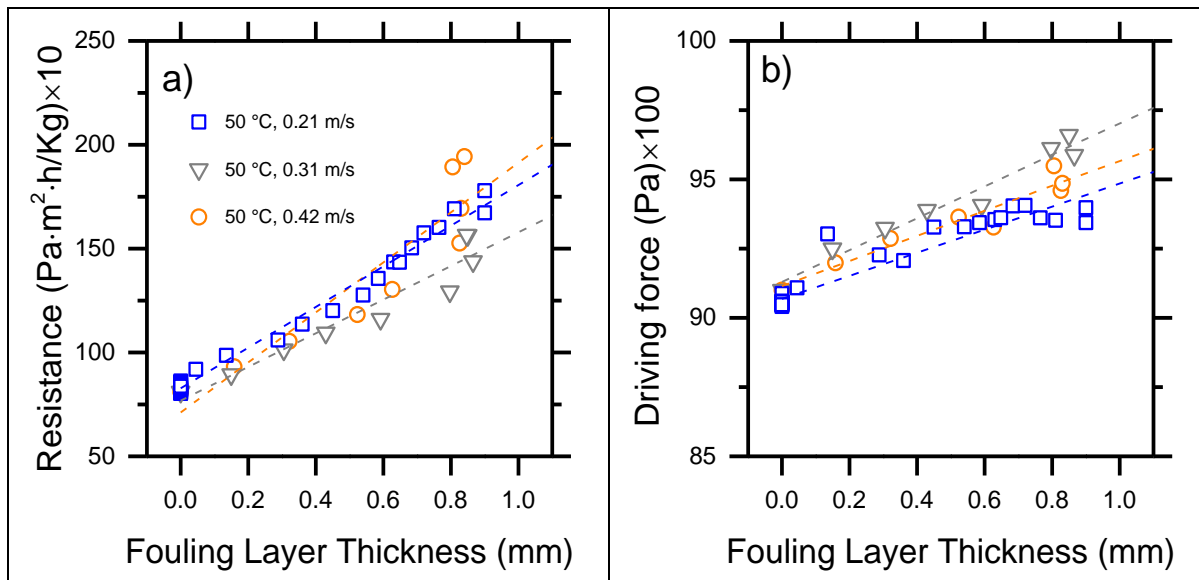
572

573 **Fig. S4.** Initial flux (J_w0) as a function of the feed inlet temperature prior to foulant addition
 574 for the nine different tests reported in legend, each run with a different combination of
 575 temperature and cross-flow velocity. The list of experiment was designed by central composite

576 design method through Design Expert software, from a selected temperature range of 35-65 °C
 577 and a range of cross-flow velocity of 0.20-0.40 m/s.

578

579



580 **Fig. S5.** Evolution of the (a) overall resistance and (b) average nominal driving force during
 581 fouling for the experiments performed at 50 °C. Results are included in Fig. 4 and are here
 582 reported by narrowing the y-axes range to better assess the increasing trends.

583

584 **Table S1.** List of experiments determined by a central composite design method in Design
 585 Expert software, a tool used to get efficient experimental protocols.

Experiment	
Feed temperature (°C)	Feed cross-flow velocity (m/s)
35	0.30
40	0.25
40	0.40
50	0.20
50	0.30
50	0.40
60	0.40
60	0.25
65	0.30

586

587

588 **Table S2.** Intercept, slopes and related standard error values of best lines fitting the average
 589 nominal driving force and foulant resistance when each is plotted against fouling thickness
 590 (dash lines showed in Fig. 4 of the main manuscript)

Experiment			Resistance (Pam2h/Kg)		Driving force (Pa)	
Feed Temperature (°C)	Cross-flow velocity (m/s)		Value	St.Error	Value	St.Error
35	0.30	Intercept	993.30	24.26	3211.02	14.3
		Slope	0.92	0.13	0.23	0.07
40	0.25	Intercept	852.61	35.50	4430.06	10.73
		Slope	1.14785	0.08	0.30	0.02
40	0.40	Intercept	881.13	46.97	4300.45	13.58
		Slope	0.97	0.19	0.27	0.05
50	0.20	Intercept	827.48	6.11	9068.42	9.53
		Slope	0.97	0.01	0.41	0.02
50	0.30	Intercept	768.28	55.62	9129.78	27.86
		Slope	0.81	0.08	0.57	0.04
50	0.40	Intercept	711.14	113.89	9114.81	31.03
		Slope	1.20	0.18	0.45	0.05
60	0.25	Intercept	961.08	26.19	16250.98	40.6
		Slope	1.27	0.04	0.58	0.06
60	0.40	Intercept	819.94	96.47	16121.9	27.25
		Slope	1.24	0.15	0.58	0.04
65	0.30	Intercept	784.43	97.04	20972.12	60.9
		Slope	1.53	0.13	0.48	0.08

591

592 **References**

- 593 [1] Y. Choi, G. Naidu, L.D. Nghiem, S. Lee, S. Vigneswaran, Membrane distillation crystallization for
594 brine mining and zero liquid discharge: opportunities, challenges, and recent progress,
595 *Environmental Science: Water Research & Technology*, 5 (2019) 1202-1221.
- 596 [2] B. Ashoor, S. Mansour, A. Giwa, V. Dufour, S. Hasan, Principles and applications of direct contact
597 membrane distillation (DCMD): a comprehensive review, *Desalination*, 398 (2016) 222-246.
- 598 [3] E. Curcio, E. Drioli, Membrane distillation and related operations—a review, *Separation and*
599 *Purification Reviews*, 34 (2005) 35-86.
- 600 [4] L.M. Camacho, L. Dumée, J. Zhang, J.-d. Li, M. Duke, J. Gomez, S. Gray, Advances in membrane
601 distillation for water desalination and purification applications, *Water*, 5 (2013) 94-196.
- 602 [5] M. Gryta, M. Tomaszewska, K. Karakulski, Wastewater treatment by membrane distillation,
603 *Desalination*, 198 (2006) 67-73.
- 604 [6] M.M. A Shirazi, A. Kargari, A review on applications of membrane distillation (MD) process for
605 wastewater treatment, *Journal of Membrane Science and Research*, 1 (2015) 101-112.
- 606 [7] D. Hou, J. Wang, C. Zhao, B. Wang, Z. Luan, X. Sun, Fluoride removal from brackish groundwater
607 by direct contact membrane distillation, *Journal of Environmental Sciences*, 22 (2010) 1860-1867.
- 608 [8] S. Srisurichan, R. Jiratananon, A. Fane, Humic acid fouling in the membrane distillation process,
609 *Desalination*, 174 (2005) 63-72.
- 610 [9] K.W. Lawson, D.R. Lloyd, Membrane distillation, *Journal of membrane Science*, 124 (1997) 1-25.
- 611 [10] L.N. Nthunya, M.F. Bopape, O.T. Mahlangu, B.B. Mamba, B. Van der Bruggen, C.A. Quist-Jensen,
612 H. Richards, Fouling, performance and cost analysis of membrane-based water desalination
613 technologies: A critical review, *Journal of Environmental Management*, 301 (2022) 113922.
- 614 [11] F. He, K.K. Sirkar, J. Gilron, Studies on scaling of membranes in desalination by direct contact
615 membrane distillation: CaCO₃ and mixed CaCO₃/CaSO₄ systems, *Chemical Engineering Science*, 64
616 (2009) 1844-1859.
- 617 [12] L.D. Tijing, Y.C. Woo, J.-S. Choi, S. Lee, S.-H. Kim, H.K. Shon, Fouling and its control in membrane
618 distillation—A review, *Journal of Membrane Science*, 475 (2015) 215-244.
- 619 [13] L. Martínez-Díez, M.I. Vazquez-Gonzalez, Temperature and concentration polarization in
620 membrane distillation of aqueous salt solutions, *Journal of membrane science*, 156 (1999) 265-273.
- 621 [14] P. Wang, M.M. Teoh, T.-S. Chung, Morphological architecture of dual-layer hollow fiber for
622 membrane distillation with higher desalination performance, *water research*, 45 (2011) 5489-5500.
- 623 [15] E. Guillen-Burrieza, A. Ruiz-Aguirre, G. Zaragoza, H.A. Arafat, Membrane fouling and cleaning in
624 long term plant-scale membrane distillation operations, *Journal of membrane science*, 468 (2014)
625 360-372.
- 626 [16] M. Gryta, M. Tomaszewska, J. Grzechulska, A. Morawski, Membrane distillation of NaCl solution
627 containing natural organic matter, *Journal of Membrane Science*, 181 (2001) 279-287.
- 628 [17] F. Ricceri, M. Giagnorio, G. Farinelli, G. Blandini, M. Minella, D. Vione, A. Tiraferri, Desalination
629 of produced water by membrane distillation: Effect of the feed components and of a pre-treatment
630 by fenton oxidation, *Scientific reports*, 9 (2019) 1-12.
- 631 [18] J. Guo, L. Fortunato, B.J. Deka, S. Jeong, A.K. An, Elucidating the fouling mechanism in
632 pharmaceutical wastewater treatment by membrane distillation, *Desalination*, 475 (2020) 114148.
- 633 [19] L. Fortunato, H. Elcik, B. Blankert, N. Ghaffour, J. Vrouwenvelder, Textile dye wastewater
634 treatment by direct contact membrane distillation: Membrane performance and detailed fouling
635 analysis, *Journal of Membrane Science*, 636 (2021) 119552.
- 636 [20] L. Fortunato, Y. Jang, J.-G. Lee, S. Jeong, S. Lee, T. Leiknes, N. Ghaffour, Fouling development in
637 direct contact membrane distillation: Non-invasive monitoring and destructive analysis, *Water*
638 *research*, 132 (2018) 34-41.
- 639 [21] F.A. Siddiqui, Q. She, A.G. Fane, R.W. Field, Exploring the differences between forward osmosis
640 and reverse osmosis fouling, *Journal of Membrane Science*, 565 (2018) 241-253.

641 [22] F. Ricceri, M. Giagnorio, K.R. Zodrow, A. Tiraferri, Organic fouling in forward osmosis: Governing
642 factors and a direct comparison with membrane filtration driven by hydraulic pressure, *Journal of*
643 *Membrane Science*, 619 (2021) 118759.

644 [23] Q. She, R. Wang, A.G. Fane, C.Y. Tang, Membrane fouling in osmotically driven membrane
645 processes: A review, *Journal of Membrane Science*, 499 (2016) 201-233.

646 [24] S. Srisurichan, R. Jiratananon, A. Fane, Mass transfer mechanisms and transport resistances in
647 direct contact membrane distillation process, *Journal of membrane science*, 277 (2006) 186-194.

648 [25] F. Ricceri, B. Blankert, N. Ghaffour, J.S. Vrouwenvelder, A. Tiraferri, L. Fortunato, Unraveling the
649 role of feed temperature and cross-flow velocity on organic fouling in membrane distillation using
650 response surface methodology, *Desalination*, 540 (2022) 115971.

651 [26] L. Mariah, C.A. Buckley, C.J. Brouckaert, E. Curcio, E. Drioli, D. Jaganyi, D. Ramjugernath,
652 Membrane distillation of concentrated brines—Role of water activities in the evaluation of driving
653 force, *Journal of Membrane Science*, 280 (2006) 937-947.

654 [27] A.S. Alsaadi, A. Alpatova, J.-G. Lee, L. Francis, N. Ghaffour, Flashed-feed VMD configuration as a
655 novel method for eliminating temperature polarization effect and enhancing water vapor flux,
656 *Journal of Membrane Science*, 563 (2018) 175-182.

657 [28] A. Hausmann, P. Sancio, T. Vasiljevic, U. Kulozik, M. Duke, Performance assessment of
658 membrane distillation for skim milk and whey processing, *Journal of dairy science*, 97 (2014) 56-71.

659 [29] M. Laqbaqbi, J.A. Sanmartino, M. Khayet, C. García-Payo, M. Chaouch, Fouling in membrane
660 distillation, osmotic distillation and osmotic membrane distillation, *Applied Sciences*, 7 (2017) 334.

661 [30] R.W. Field, G.K. Pearce, Critical, sustainable and threshold fluxes for membrane filtration with
662 water industry applications, *Advances in colloid and interface science*, 164 (2011) 38-44.

663 [31] T.-T. Nguyen, C. Lee, R.W. Field, I.S. Kim, Insight into organic fouling behavior in polyamide thin-
664 film composite forward osmosis membrane: Critical flux and its impact on the economics of water
665 reclamation, *Journal of Membrane Science*, 606 (2020) 118118.

666 [32] X.-y. Li, X.-m. Wang, Modelling of membrane fouling in a submerged membrane bioreactor,
667 *Journal of Membrane Science*, 278 (2006) 151-161.

668 [33] L. Fortunato, M. Li, T. Cheng, Z.U. Rehman, W. Heidrich, T. Leiknes, Cake layer characterization
669 in activated sludge membrane bioreactors: real-time analysis, *Journal of membrane science*, 578
670 (2019) 163-171.

671 [34] Z. Ding, L. Liu, Z. Liu, R. Ma, Fouling resistance in concentrating TCM extract by direct contact
672 membrane distillation, *Journal of Membrane Science*, 362 (2010) 317-325.

673 [35] A. Alkhatib, M.A. Ayari, A.H. Hawari, Fouling mitigation strategies for different foulants in
674 membrane distillation, *Chemical Engineering and Processing-Process Intensification*, 167 (2021)
675 108517.

676 [36] S. Goh, Q. Zhang, J. Zhang, D. McDougald, W.B. Krantz, Y. Liu, A.G. Fane, Impact of a biofouling
677 layer on the vapor pressure driving force and performance of a membrane distillation process,
678 *Journal of Membrane Science*, 438 (2013) 140-152.

679 [37] R.W. Field, J.J. Wu, On boundary layers and the attenuation of driving forces in forward osmosis
680 and other membrane processes, *Desalination*, 429 (2018) 167-174.

681 [38] R. Schofield, A. Fane, C. Fell, Gas and vapour transport through microporous membranes. II.
682 Membrane distillation, *Journal of Membrane Science*, 53 (1990) 173-185.

683