

Ceramic vs polymeric membrane implementation for potable water treatment

P. Jarvis^a, I. Carra^a, M. Jafari^b, S.J. Judd^{a,*}

^a Cranfield Water Science Institute, Cranfield University, Beds, UK

^b PWNT, Velsbroek, the Netherlands

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ABSTRACT

The continued technological developments and decreased purchase costs of ceramic membranes have seen increased recent interest in the technology as an alternative to the more widely used polymeric membranes. This paper assesses the relative technical, practical and economic merits of the two membrane materials in the context of potable water production from surface water sources.

The work focuses on phenomena of direct technoeconomic significance, namely cleaning efficacy (manifested as permeability recovery), membrane integrity and incurred labour effort. Topics reviewed thus comprise: (a) practical comparison of the two technologies challenged with the same feedwater, (b) comparative technoeconomic analyses, (c) membrane integrity studies of polymeric membranes - incorporating aged samples extracted from operating installations, (d) sludging incidents, and (e) pilot and full-scale data.

Available relevant data reveal:

(a) bench-scale comparative tests do not indicate a consistent significant difference in the net permeability between the two membranes;

(b) polymeric membranes are subject to a decline in both mechanical strength and permeability from the loss of the hydrophilic agent over a period of years from the action of hypochlorite used for cleaning;

(c) the decreased mechanical strength with age of polymeric membranes increases the manual repair requirement and shortens membrane life, respectively impacting on labour and membrane replacement costs where the latter is also determined by the permeability;

(d) the chemical and mechanical robustness of ceramic membranes permits more aggressive chemical cleaning, which then affects the chemicals consumption cost; and

(e) anecdotal evidence suggests that polymeric membranes challenged with pre-coagulated surface waters may be subject to sludging, the agglomeration of solids in the membrane channels, which may also be age-related.

Notwithstanding the above, data from published comparative technoeconomic studies indicate a linear relationship between the overall cost benefit and the membrane module cost ratio mitigated by the relative membrane life and operating flux.

1. Introduction

1.1. Background and rationale

The relative benefits of ceramic and polymeric membranes for municipal water applications has attracted increasing interest over the past five years-or-so. This is in part evidenced by the number of ceramic membrane-based reviews published, with at least six produced since the start of 2020 on topics ranging from pathogen rejection (Goswami and Pugazhenth, 2020) and surface fouling (Hakami et al., 2020; Li et al.,

2020; Arumugham et al., 2021; Asif and Zhang, 2021) to fabrication impacts on material characteristics (Liang et al., 2021).

Whilst there has been a strong focus on fouling and its amelioration by the academic community, in practice membrane fouling is rarely the sole limitation on membrane process implementation and operation in municipal applications. Recent studies suggest that a substantial part of the operating expenditure (OPEX) can be attributed to the labour effort incurred (Qiblawey and Judd, 2019; Judd and Carra, 2021), particularly for small installations. This in turn is strongly associated with process challenges such as clogging (or sludging) of the membrane interstices

* Corresponding author.

E-mail address: simon@juddwater.com (S.J. Judd).

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with solids (Buzatu et al., 2012) and/or, in the case of potable water treatment, membrane repair from integrity breaches (Ferrer et al., 2015; Lee et al., 2019).

Given the technical and commercial significance of the different facets of the two materials, it is instructive to examine papers where either:

- (a) direct comparison in performance of ceramic and polymeric membranes has been made, or
- (b) key aspects of each technology studied where differences in performance and/or technical characteristics generally can be expected.

This paper aims to review available data pertaining to (a) and (b) above which reflect on the operability and cost of the two technologies primarily for municipal potable applications at full-scale, viz:

- i side-by-side studies of ceramic and polymeric membranes challenged with the same feedwater,
- ii comparative cost or technoeconomic studies,
- iii membrane integrity studies,
- iv membrane channel clogging (or sludging),
- v operability of pilot and full-scale installations.

Potable water applications considered relate to surface water feeds. Surface waters, with seasonally or otherwise intermittently high levels of natural organic matter (NOM), are widely recognised as being more challenging than borehole water due to the NOM fouling propensity.

1.2. A brief history

Low-pressure membrane technology for generating potable water from low-salinity inland raw waters has predominantly been based on polymeric hollow-fibre (HF) ultrafiltration/ microfiltration (UF/MF) membranes (AWWA, 2005; Goswami and Pugazhenth, 2020; Judd and Carra, 2021). Implementation for potable waters in Europe began in 1988 with the 0.16 MLD Aquasource plant in Amoncourt, France (Letourneur and Mehra, 1993); the first >10 MLD capacity system in the US was the Memcor plant installed at Saratoga five years later (AWWA, 2005).

Ceramic membranes were first implemented for this duty in 1996 at Miyazaki, Japan with a small (0.55 MLD) installation based on the original Metawater ceramic membrane (Ceramac). By mid-2009 there were 87 WTWs in Japan based on the Metawater system delivering >400 MLD of drinking water (Hattori, 2010). This system is based on vertically-orientated, cylindrical monoliths with multiple tubular channels (hence MT modules). Whilst other MT ceramic products are available, the potable water market appears to be dominated by the Metawater technology.

The mid-late noughties saw the implementation of the flat sheet (FS) ceramic module (Binkle et al., 2005; Kaplan and Kaschek, 2009), with its subsequent implementation for industrial process water and effluent and municipal potable water and wastewater applications. One of the earliest wastewater installations, the 0.5 MLD ItN Nanovation (now part of Saffon) membrane bioreactor (MBR) plant at Petersberg which started up in 2007 (Kaplan and Kaschek, 2009) and still operates with the original membranes. The ceramic FS technology is currently offered by at least five providers internationally: Cembrane/Ovivo (Denmark/USA), Ceraflo (Singapore), ItN/Saffon (Germany/China), Meidensha (Japan) and Liqtech (Denmark). Ceramic MT modules have also found use for niche industrial effluent treatment applications, dating back to the 1990s or earlier (Lahiere and Goodboy, 1993; Chang et al., 1994).

Whilst the data available is constrained by the limited age of the majority of the installations, evidence suggests that the membrane life of 15–20 years generally warranted by most ceramic membrane product providers is not unreasonable - and possibly even conservative for the

most recent products.

2. Analysis

2.1. Side-by-side studies of net permeability

Side-by-side comparisons of the performance of polymeric and ceramic membranes, where the ceramic and polymeric membranes have been challenged with the same feedwater, date back 30 years (Moulin et al., 1991). A number of other comparative studies, predominantly at bench-scale, have subsequently been reported (Bodzek and Konieczny, 1998; Yuasa, 1998; Klomfas and Konieczny, 2004; Hofs et al., 2011; Lee and Kim, 2014; Oligny et al., 2016; Alresheedi et al., 2019). Outcomes from these studies (Table 1) have been conflicting regarding the relative sustainable permeability values for the two material types. Some studies, principally those conducted prior to ~2010 (Moulin et al., 1991; Bodzek and Konieczny, 1998; Klomfas and Konieczny, 2004), have reported a higher overall permeability for the polymeric membrane over the ceramic material when challenged with the same feedwater. This may reflect the intrinsically lower clean water permeability of a ceramic membrane compared with an equivalent polymeric one, due to the greater membrane layer thickness and lower bulk porosity. However, these studies appear not to have been conducted under the same controlled hydrodynamic conditions for the two membrane materials; it is well known that the membrane module volume to membrane surface area ratio (V/A) significantly affects fouling, ostensibly through the Reynolds number (Re), and hence permeability (Hofs et al., 2011; Alresheedi et al., 2019).

More recent studies (Hofs et al., 2011; Lee and Kim, 2014; Oligny et al., 2016; Kenari and Barbeau, 2016; Alresheedi et al., 2019), where the hydrodynamic impacts are acknowledged and/or accounted for, have tended to demonstrate either comparable (Oligny et al., 2016) or slightly lower (Hofs et al., 2011) overall fouling propensity or fouling rate for the ceramic material compared with the polymeric one. The work of Lee and Kim (2014) indicated the relative fouling propensity (with reference to Savannah River humic acid) to be strongly influenced by the presence of calcium, with the ceramic membrane more resistant to irreversible fouling at calcium concentrations of 10 mM. The most recent published comparative study (Alresheedi et al., 2019) demonstrated increased backwash efficiency for the ceramic membrane, but little difference in chemical cleaning efficiency between the two materials. Recorded differences in overall net permeability appear to be marginal, whether measured as the sustainable net permeability in LMH/bar, the overall fouling resistance (Lee and Kim, 2014), or the physical and chemical cleaning efficiency (Alresheedi et al., 2020).

In summary, comparative tests of polymeric and ceramic membrane materials challenged with the same feedwater and operated under similar hydrodynamic conditions suggest that there is little difference in fouling propensity when representative backwashing cleaning is applied. This confirms an earlier observation by Guerra & Pellegrino (2013), based on the analysis of 209 published permeability data from studies conducted between 1997 and 2011. Against this, hydrodynamic conditions for full-scale plants based on commercial membrane products are likely to differ between the two materials due to commensurately differing V/A ratios - and associated packing densities. Moreover, much of the work in this area has focused on fouling characterisation in terms of the foulant physicochemistry, rather than the study of hydrodynamics and membrane configuration impacts.

2.2. Side-by-side cost or technoeconomic analyses

A few comparative economic assessments of polymeric and ceramic membranes have been presented in the past decade (Guerra and Pellegrino, 2013; Park et al., 2015; Kurth et al., 2018). The outcomes of these studies (Table 2) have tended to demonstrate the expected sensitivity of the cost benefit or penalty to the sustainable flux and the membrane

Table 1
Outcomes of comparative studies of polymeric and ceramic membrane, potable water production.

Source	Type	Material	d _p /MWCO	Config	Parameter(s)	Parameter values(s)/ranges	Comment
Moulin et al, 1991	Polym.	PS	0.003 μm	0.5 mm HF	Flux, LMH: <i>Permeability, LMH/bar</i>	180–80: 260–114	Significantly differing TMP and Re values and high CFV used. 1 h test period.
	Ceram.	Al ₂ O ₃	0.2 μm	4 mm MT		420–120: 210–60	
Bodzek & Konieczny, 1998	Polym.	PP	0.2 μm	1.8 mm*	Flux, LMH: <i>Permeability, LMH/bar</i>	169–198: 169–198	Higher permeability + greater permeability decline for polymeric membrane. 5 h test period.
	Ceram.	Ti/Zr oxide	0.1–0.2 μm	MT MF		122: 61	
	Ceram.	Ti/Zr oxide	15–300 kDa	MT UF		79–94: 40–47	
Komflas & Konieczny, 2004	Polym.	PP	0.2 μm	1.8 mm*	Flux, LMH: <i>Permeability, LMH/bar</i>	122–89: 122–89	Impact of 100 mg/L dose PAC studied: final flux values higher for PAC-pretreated PP MF membrane than for ceramic membranes
	Polym.	PES	UF	1.5 mm*		109–84: 109–84	
	Ceram.	Ti/Zr oxide	0.1 μm	2.5 mm MT		112–81: 112–81	
	Ceram.	Ti/Zr oxide	0.1 μm	4.5 mm MT		83–54: 83–54	
Hofs et al, 2011	Polym.	PES/PVP	0.14 μm	CT	Irreversible fouling rate, μbar.m ² /L (or μbar/h per unit flux in LMH)	200	Significantly lower irreversible fouling rate for all ceramic membranes cf. polymeric membrane partly attributed to wider membrane channel and subsequent hydrodynamic impacts.
	Ceram.	Al ₂ O ₃	0.1 μm	Single tube		20–60	
	Ceram.	TiO ₂	0.1 μm	Single tube		20–60	
	Ceram.	ZrO ₂	0.1 μm	Single tube		20–60	
	Ceram.	SiC	0.04 μm	Single tube		1–3	
Lee & Kim, 2014	Ceram.	ZrO ₂	50 kDa	Disc	Irreversible fouling resistance, 10 ⁻¹⁴ m ⁻¹ , SRHA: SRHA/Ca	0.4; 0.8	Rotating disc used to control hydrodynamics. Irreversible fouling rate significantly influenced by presence of Ca ²⁺ at 10 mM concentration.
	Ceram.	ZrO ₂	150 kDa	Disc		-	
	Polym.	PES	100 kDa	Disc		<0.1: 3	
	Polym.	Reg. cell.	100 kDa	Disc		<0.1: 2	
Oligny et al, 2016	Polym.	PES/PVP	0.025 μm	CT	Flux range, LMH: <i>Total fouling coefficient</i>	20–140: 0.58	Higher permeability & greater permeability decline for polymeric membrane
	Ceram.	TiO ₂	0.1 μm	CT		20–140: 0.56	
Alresheedi et al, 2019	Polym.	PES/PVP	0.01 μm	CT	Backwash efficiency: chemical cleaning eff. <i>alkaline hypo: caustic</i>	0.2–0.63: 0.5–0.72: 0.5–0.72	Difference in backwash efficiency greater than that of chemical cleaning efficiency
	Ceram.	TiO ₂	0.01 μm	MT		0.48–0.82: 0.58–0.8: 0.86–0.94	

MWCO Molecular weight cut-off in kDa (kiloDaltons); PS polysulphone; PES polyethylsulphone; PP polypropylene; Reg. cell. regenerated cellulose; PVP polyvinylpyrrolidone; *Denoted "capillary"; CT capillary tube; MT multitube; SRHA Savannah River humic acid; CFV crossflow velocity; PAC powdered activated carbon.

cost. An early cost comparison (Guerra and Pellegrino, 2013) demonstrated a 24–54% increase in the total water production cost (TPC) for the ceramic membrane compared with the polymeric one for a ten-fold higher membrane cost and a 25% higher net flux (125 LMH for the ceramic compared with 100 LMH for the polymeric). In the most recent study (Kurth et al., 2018) a net flux of up to 381 LMH was assumed (based on the outcomes of a pilot trial) along with an implied cost of ~\$175 USD per m² for the ceramic membrane, compared with \$35 USD m⁻² for the polymeric one operating at a net flux of 68 LMH. Based on these assumptions the authors calculated a net cost benefit of 7–15% for the ceramic membrane system.

The data from these three studies can be depicted as a correlation of cost penalty - or negative cost benefit - expressed as life cycle or total production cost (LCC or TPC respectively) against the membrane cost-flux-life (CFL) ratio (Fig. 1). For the membrane CFL ratio, the higher cost of the ceramic membrane is mitigated by its longer life and higher sustainable flux:

$$\text{Cost penalty} = \frac{\text{Total cost parameter for ceramic membrane installation}}{\text{Total cost parameter for polymeric membrane installation}} - 100\%$$

$$\text{CFL} = \frac{\text{Ceram. cost}}{\text{Polym. cost}} \times \frac{\text{Polym. flux}}{\text{Ceram. flux}} \times \frac{\text{Polym. life}}{\text{Ceram. life}}$$

Fig. 1 indicates a strong correlation ($R^2 = 0.991$) between the cost

penalty and membrane cost-flux-life ratio for life values of 5–8 y for the polymeric membrane and 20 y for the ceramic one. This trend arises despite significant differences across the three studies (Table 2) in (a) the respective characteristics of the polymeric and ceramic membranes and modules, (b) the interest rate value selected, (c) the installation flow capacity, (d) cleaning protocols and frequencies, (e) assumptions regarding labour costs and other OPEX-related parameters, and (f) both the absolute CAPEX (capital expenditure) and the CAPEX per unit flow. Whilst assumptions around discounting and inflation are questionable the outputs for the ceramic membranes have been normalised against the polymeric ones determined for the same set of assumptions. As such both sets of data are subject to similar absolute errors pertaining to assumed discounting and inflation; the relative values should nonetheless be reasonably representative.

Regarding (f) above, the data of Park et al. (2015) and Kurth et al. (2018) respectively imply a normalised CAPEX of \$257 and \$156 per m³/d flow for the polymeric membrane technology. This compares to a rather lower value of \$137–147 per m³/d flow determined from 14 full-scale UK membrane installations (Judd and Cara, 2021).

The chemical resistance of ceramic membranes permits more aggressive conditions for cleaning (d) than polymeric membranes. Increasing cleaning reagent strength, combined with pre-coagulant dosing, has been shown to more than double the chemicals consumption cost compared with polymeric membranes to \$29 per m³/d according to Kurt et al. (2018). However, this study suggests that this cost

Table 2

Outcomes of comparative economic assessments, potable water production.

Parameter or component	Guerra and Pellegrino (2013)		Park et al. (2015)		Kurth et al. (2018)	
Membrane	PES	Al ₂ O ₃	Polymeric	Ceramic	PES/PVDF	Al ₂ O ₃
Feed	100 mg/L bentonite		Not specified		River water, 3–10 NTU, <4 mg/L TOC	
Capacity, MLD	18.9		30		7.57	
Pore size/MWCO	10 kD	0.01 µm	Not specified		0.02 µm*	0.03 µm
Membrane config., size	CT, 1.02 mm ID	monolith, 1.05 mm ID	Not specified		CT, 0.9 mm ID	Monolith, 2.4 mm ID
Membrane cost, USD/m ²	50	300–500	120	600	35	175
Net flux, LMH	100	125	42	96	68	196–381
Maximum TMP, bar	1.4	10.3	Not specified		Not specified	0.8–1.8
Membrane life, y	5	20	8 (assumed)	20	6	20
Plant life, y	40		20		20	
Interest rate	4%		6%		Not specified	
Membrane CAPEX metric	6%, \$0.06/m ³ ^a	38%, \$0.59/m ³ ^a	13%, \$0.017/m ³ ^a \$256 per m ³ /d ^{b,c}	24%, \$0.036/m ³ ^a \$381 per m ³ /d ^{b,c}	\$156 per m ³ /d ^b	\$145–172 per m ³ /d ^b
• Ratio, ceram/polym ^d	+983%		+49–85%		+10% to –6%	
TPC or LCC, \$/m ³	\$0.122/m ³	\$0.151–0.188/m ³	\$0.27/m ³	\$0.28/m ³	\$280 per m ³ /d ^b	\$238–261 per m ³ /d ^b
• Ratio, ceram/polym ^e	+24–54%		+4%		–7 to –15%	

MLD megalitres per day; MWCO molecular weight cut off; LMH litres per m² per hour; TPC total production cost; LCC life cycle cost; PES polyethylsulphone; PVDF polyvinylidene fluoride; kD kiloDaltons; ID internal diameter; CT capillary tube (in-to-out flow); CAPEX capital expenditure.

^a Membrane cost as% of total CAPEX, and as absolute cost (either amortised or not amortised) per volume permeate;

^b Cost per unit flow;

^c Based on membrane + skid costs only;

^d refers to membrane CAPEX metric;

^e refers to TPC or LCC metric.

* assumed to be based on Pentair XIGA 46.

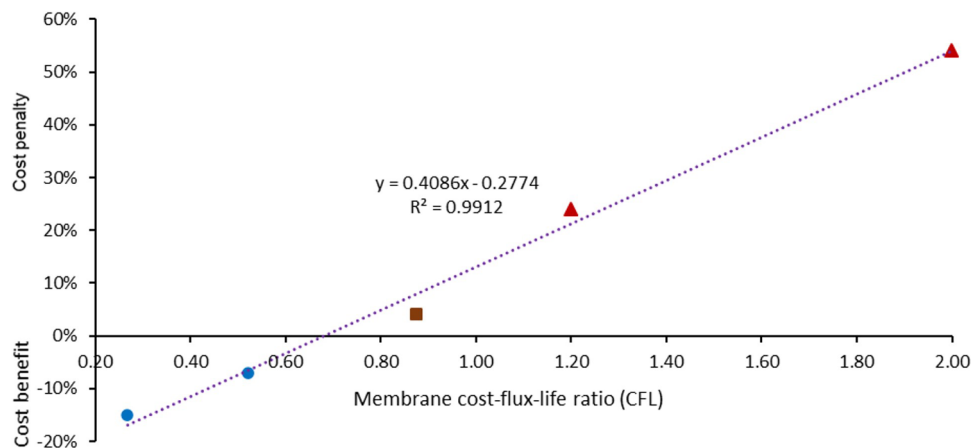


Fig. 1. Cost penalty vs. Membrane cost-flux ratio, based on published data from three studies: ● Kurth et al. (2018), ■ Park et al. (2015), and ▲ Guerra and Pellegrino (2013).

penalty is surpassed by the cost benefit associated with the resultant doubling in the net flux. The trials conducted by these authors and those of Alresheedi et al. (2019) (Table 1) suggest that there are clear cost benefits in fully optimising the backwash and cleaning protocols.

The impact of labour requirements (e) for polymeric membrane-based installations can differ substantially from those for ceramic ones if significant membrane integrity loss takes place (Section 2.3). A labour effort of 2.5 FTE (full-time equivalent) was assumed by Guerra and Pellegrino (2013) for operation of a 18.9 MLD plant, with ~0.5 FTE assumed to be assigned to integrity monitoring and fibre repair of 24 polymeric membranes filaments per year (thus 1.27 repairs per MLD p. a.). Neither Park et al. (2015) nor Kurth et al. (2018) identified labour effort specifically in their respective reported analyses.

Notwithstanding the above factors, Fig. 1 implies a cost benefit of implementing ceramic over polymeric membranes when the membrane

CFL ratio falls below ~0.68. However, the slope of the relationship must necessarily depend on OPEX, and specifically the key parameters of energy and chemicals consumption, and of labour effort. Energy consumption relates mainly to membrane permeability and hydraulic losses and chemical demand to membrane cleaning. Labour effort is strongly correlated with the maintenance incurred by membrane breaching.

2.3. Membrane integrity studies

There are far fewer studies of membrane integrity than there are of membrane fouling. However, a number of studies have nonetheless demonstrated age-related impaired integrity of polymeric membranes, with an accompanying concomitant decrease in permeability and membrane mechanical strength.

For the more extensively-studied PES membrane a few potable water

studies have indicated increased fouling propensity (Touffet et al., 2015; Cui et al., 2017) and reduced integrity, generally quantified as the elongation at break ϵ_R (Causserand et al., 2015; Cui et al., 2017) (Table 3). These changes have been associated with the loss of polyvinylpyrrolidone (PVP), a widely used hydrophilic additive in membrane fabrication. Loss of PVP has been assumed to be promoted by hypochlorite (OCl^-) oxidation (Touffet et al., 2015; Ravesreau et al., 2016; Fu and Zhang, 2019). Authors have observed reasonable correlation of PVP and mechanical strength loss with hypochlorite dose expressed as Ct , the product of the concentration ($\text{g Cl}_2/\text{L}$) and exposure time (h), and its action is attenuated to an extent by the pH (Regula et al., 2014). Ct is estimated to be around 2000 g.h/L over the life of a polymeric membrane potable water plant based on standard chemical cleaning protocols (Causserand et al., 2015; Robinson and Bérubé, 2020).

In a recent study in this area (Robinson and Bérubé, 2020) samples of a specific PVDF membrane product were harvested from 8 full-scale installations across an age range of 1–8 years. Results indicated a deterioration in permeability and a decline in mechanical strength at around 5 years of use, associated with the decreased hydrophilic additive bulk and surface concentration. Similar observations of PVP loss have been made with MBR PVDF HF membranes (Mavukkandy et al., 2016; Marbelia et al., 2019).

Integrity of ceramic membranes is largely unreported. It is generally accepted that the ceramic membrane material itself is not susceptible to breaching pertaining to oxidative degradation, though hypochlorite-promoted failure of the seals has recently been demonstrated (Kramer et al., 2019). For the purposes of this analysis, it is assumed that whilst

integrity monitoring is required for ceramic membrane installations, no repair associated with integrity loss is incurred.

In practice membrane integrity is routinely assessed using established off-line methods, most usually using the pressure decay test (PDT). Whilst PDTs are automated, examination and repair (“pinning”) of polymeric HF membranes failing the PDT is laborious and incurs significant labour effort and associated operating costs (Guerra and Pellegrino, 2013; Judd and Carra, 2021). The proportion of the total OPEX that this cost component contributes increases with decreasing flow capacity (Judd and Carra, 2021). Moreover, the impact the breakage of a single fibre has on the treated water quality also increases with decreasing size, since the proportion of the total number of fibres also increases (Gijbetsen-Abrahamse et al., 2006).

A crucially important aspect is thus the integrity loss manifested as required membrane filament repairs over the life of the installation. Fibre breakage rate data is rarely reported in the peer-reviewed literature. The small amount of available data (Fig. 2), taken from Brehant et al. (2015) and two unpublished data sets (“PES 1 & 2”), suggests wide ranging breakage rates – from <10 to ~80 per MLD plant capacity per year – which may increase exponentially or linearly with time. These values are substantially higher than the estimate of 24 p.a. for an 18.9 MLD capacity installation assumed by Guerra and Pellegrino (2013).

Evidence across a number of sites suggests that polymer integrity loss is more strongly associated with feedwater quality than membrane material. Specifically, both permeability and integrity diminish at full-scale polymeric membrane plants when feedwater turbidity increases (Brehant et al., 2015; Ferrer et al., 2016).

Translating breakage rates into cost is challenged by the differing protocols used across different sites for identifying and isolating the module with the integrity breach, removing the membrane element, identifying and repairing the breach, and reassembling the module/skid. However, if an individual breach is assumed to incur 1.5 h of effort (i.e. ~\$75 assuming a gross cost of \$50/h, and thus comparable to the value of \$70 for a single fibre breakage determined by Freeman, 2012) then, using the average values from the data depicted in Fig. 2, this corresponds to a range of 0.011 to 0.028 FTE per MLD. Assuming 1600 working hours per year, i.e. \$80,000/FTE, this equates to \$0.0046–0.012 per m^3 .

Calculated overall OPEX values vary widely according to assumptions made (Table 4). The greatest sensitivity is to the electrical energy tariff, followed by the membrane replacement and the labour effort. The overall OPEX calculated by Brehant et al. (2015) is around a quarter of that determined by Guerra and Pellegrino (2013) due to the lower energy tariff and labour effort assumed Judd and Carra, 2021., using data taken from full-scale UK installations, determined the normalised labour effort (FTE per MLD or m^3) to increase with decreasing flow capacity. Based on such assumptions, fibre breakage may contribute up to 43% of the OPEX in labour costs alone, and more still if membrane life is subsequently reduced.

2.4. Clogging (sludging) determination

Membrane channel clogging (or sludging), the agglomeration of solids in the membrane interstices, has been widely recognised by MBR practitioners in municipal wastewater treatment for at least 14 years (Itokawa et al., 2008; Frechen et al., 2008). Sludging has been observed in both aerobic (Stone and Livingston, 2008; Ewert, 2014) and anaerobic MBRs (Giménez-Lorang et al., 2021).

It has recently become evident that the challenge also extends to potable water treatment where polymeric HF membranes are downstream of a coagulation step, so-called “direct filtration”. Thus far no research studies have been conducted on potable water membrane channel sludging. However, practitioner experience of operating installations in the UK and North America suggests that sludging presents a significant challenge on direct filtration of surface waters. This slow accumulation of solids within the fibre bundle is manifested as an only

Table 3
Integrity studies and outputs.

Source	Permeability (K) & fouling rate (FR)	Integrity	Other
Causserand et al. (2015)	Doubling of K independent of OCl^- concentration	CEBs with 1.05 g/L NaOCl for 144 h pH 8 (151 g.h/L): ϵ_R decreases 70% PVP content decreases linearly with decreasing ϵ_R	Logarithmic decline of both ϵ_R and PVP content with NaOCl Ct . Lifetime Ct estimated at 200 g.h/L ~30% loss of PVP from ClO-exposed membranes
Touffet et al. (2015)	1.44 g.h/L NaOCl @ pH 8 trebles FR. 15 g.h/L Cl_2 @ pH 8 increases K by 240%	-	-
Ravesreau et al. (2016)	-	Soaking in pH 11.5 4 g/L NaOCl @ 40°C: - 50d: longitudinal shrinkage of ~4% - 135d: elongation at break ϵ_R decreases ~25%	Ct equates to 4800 g.h/L @ 50d and 12,960 g.h/L at 135d
Cui et al. (2017)	Between 0.8 and 1.8 y: K decreases ~25%	Between 0.8 and 1.8 y: - Porosity decreases by ~9% - Contact angle decreases by ~13% - Breaking strength decreases by ~5%	-
Robinson and Bérubé (2020)	Between 5 and 8 years: - K halves - FR trebles FR increases six-fold when surface HA decreases five-fold	Between 5 and 8 years: - Maximum stress decreases by ~21% - Young's modulus decreases by ~27% - Surface HA content decreases by ~44%	OCl^- load estimated at 143 and 197 g.h/L at 5 & 8 y respectively. 2000 g.h/L Ct reduced HA surface content by 35% in accelerated test.

HA Hydrophilic agent (assumed to be PVP, polyvinylpyrrolidone); ϵ_R elongation at break; Ct hypochlorite (OCl^-) concentration as $\text{g Cl}_2/\text{L}$ x exposure time in h;.

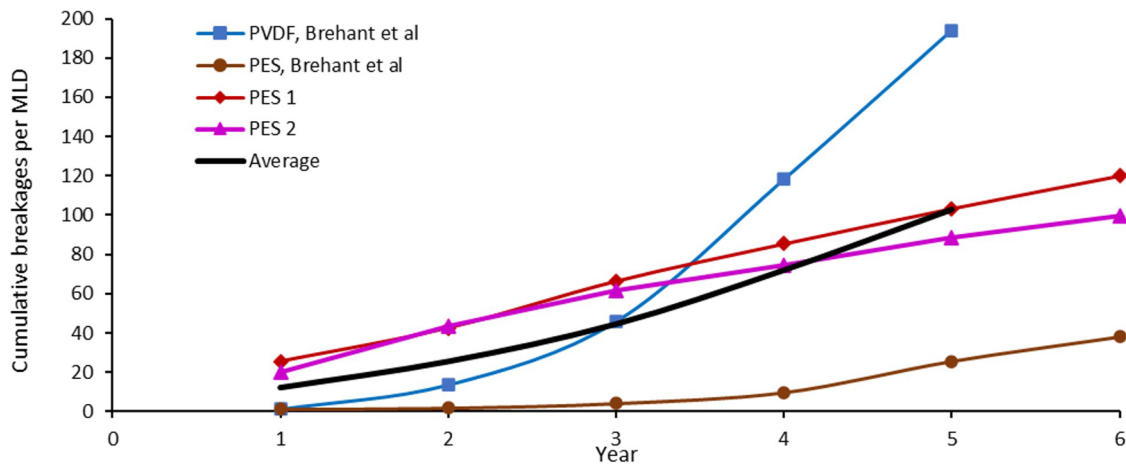


Fig. 2. Recorded normalised fibre breakage rate for four full-scale plants (partly extracted from Brehant et al., 2015).

Table 4
Values of key OPEX contributors.

Parameter	Judd and Carra (2021) (adjusted)	Guerra and Pellegrino (2013)	Brehant et al. (2015)	
			PES	PVDF
Energy cost, \$/kWh	0.1	0.1	0.062	
Specific energy cost, \$/m ³	0.0147	n.a.	0.0064	0.0031
Assumed membrane cost, \$/m ²	40	50	24	35
Membrane life, y	7	5	7	5
Net flux, LMH	70	100	111	49
Membrane replacement cost, \$/m ³	0.0125	0.022	0.0063	0.0132
Specific chemicals cost, \$/m ³	0.00030	7 ^a	0.0029	0.0025
Labour, staff cost, \$/y	80,000	80,000	n.a.	n.a.
Labour effort, FTE/MLD	0.008–0.063	0.13	n.a.	n.a.
Labour cost, \$/m ³	0.002–0.017 ^b	n.a.	0.0004	0.0033
Waste disposal cost, \$/m ³	0.0015	n.a.	n.a.	n.a.
TOTAL, \$/m³	0.031–0.046^b	0.074	0.016	0.022

^a \$/(m².y);

^b refers to flow range of 5–150 MLD. n.a. not available.

fleetingly-effective CIP, and has been quantified as an increase in the weight of the wetted membrane module at full-scale potable water installations (Booker, 2019) and pilot-scale MBR studies (Zsirai et al., 2014).

2.5. Full-scale experience and pilot trial outcomes

In practice membrane plants tend to operate at fluxes well below the design fluxes, which are themselves also below the values tested and/or

assumed in some of the published studies (Kurth et al., 2018). An example is the installation in the Netherlands (Fig 3), which employs suspended ion exchange (SIX) upstream of the membrane skid for improved dissolved organic carbon removal. This plant has a design peak flux of 110 LMH (Table 5), though in practice the flux is in the region of 65–90 LMH since the flow is demand driven. Cleaning is entirely through chemically-enhanced backwashing (CEB): no recovery CIP (clean in place) is currently installed at this site (Table 6).

Whilst operating under very conservative conditions, experience at this site nonetheless indicates:

- (a) no integrity loss: integrity tests (pressure decay measurement) provide no evidence of damage to the membrane affecting its perm-selectivity

Table 5
Summary of key design and operational parameters.

<i>Flux</i>		
Design, peak	110	LMH
Actual, max	90	LMH
Actual, min	65	LMH
<i>Membrane maintenance</i>		
Transmembrane pressure (TMP)	0.2–1.8	bar
Maximum permitted TMP	2	bar
Backwash (BW) frequency	>36	min
BW duration	3	min
BW maximum pressure	3.5	bar
<i>Chemically enhanced backwash (CEB)</i>		
Reagent	100 mg/L NaOCl	100 mg/L H ₂ O ₂ , pH3
Frequency	per 11th BW	per 55th BW
Duration, minutes	10	10
Temperature	ambient	ambient
Net chemical consumption rate,	1.3 gNaOCl/m ³	0.3 ¹ gH ₂ O ₂ /m ³
<i>Performance</i>		
Recovery	92	%
Projected membrane life	20	years

¹ refers to chemical use in CEB only.

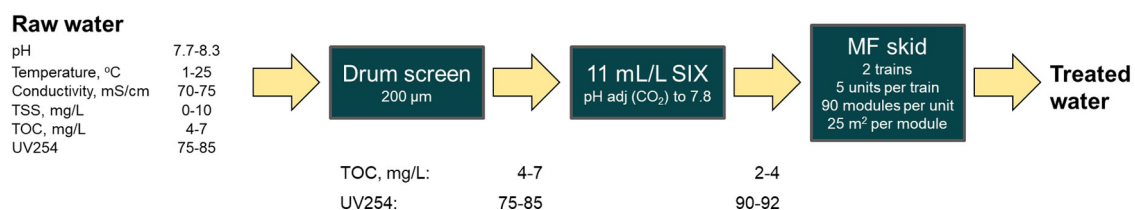


Fig. 3. Schematic with outline water quality determinants of surface water treatment plant, the Netherlands.

Table 6
Fluxes sustained for recent pilot-scale studies of ceramic membranes.

Source	Flux, LMH	Permeability, LMH/bar	Pre-treatment	Feed
Metcalfe et al. (2016)	185	600–900	PACl (in-line)	Surface w
Ng et al. (2018)	300	300–420	Coag, ozonation	2ndary Ww
Ng et al. (2018)	200	300–400	Coag	2ndary Ww
Spencer et al. (2019)	150	300–430	PACl, ozonation	2ndary Ww
Noguchi et al. (2019)	160	400–600	Alum	Surface w
Campinas et al. (2021)	154–174	220–240	PAC, alum	Surface w
Lam et al. (2021)	118–183	>1200	Alum, ozonation	Surface w
Shorney-Darby et al. (2021)	200–250	200–600	PACl (SCC)	Surface w
	250	800–1000	PACl (in-line)	Surface w

PACl polyaluminium chloride; PAC powdered activated carbon; Ww wastewater; SCC solids contact clarifier.

- (b) manageable permeability decline: the permeability from Year 2 onwards has plateaued at 180–200 LMH/bar during operation without an installed CIP system. Trials with CIP indicate that the permeability can be increased to as much as 400 LMH/bar.

There have been no “events” associated with integrity loss and permeability decline, the latter being associated with sludging at some polymeric membrane plants. The FTE for the site is 0.6, i.e. 0.005 FTE/MLD, which is somewhat lower than the value of 0.009 computed by Judd and Carra (2021) for a plant of similar size. Against this, the chemicals consumption is higher than that reported for polymeric plants and so contributes a greater proportion of the OPEX at this site (~\$0.010/m³ if a cost of \$1000 per m³ of bulk reagent is assumed for both NaOCl and H₂O₂).

Available information from recent pilot-scale studies, based on at least one full-scale membrane module, suggest sustainable operating flux values generally between 150 and 250 LMH. This range refers to optimised systems, and specifically to optimised pre-treatment. Whilst optimisation of pre-treatment relies on foreknowledge of the feedwater organic matter characteristics (Peters et al., 2021), evidence suggests that combined ozonation and coagulation presents a robust means of mitigating ceramic membrane fouling (Fan et al., 2014; Ng et al., 2018; Spencer et al., 2019; Lam et al., 2021; Shorney-Darby et al., 2021).

4. Discussion

A review of the available information relating to membrane filtration employed for potable water production suggests the following:

- 1 Bench scale trials have not conclusively demonstrated a significant difference in overall net permeability between the membranes, especially for tests based on pristine materials (Oligny et al., 2016; Alresheedi et al., 2019).
- 2 Polymeric membranes are subject to irreversible impairment of both mechanical strength and permeability due to loss of the hydrophilic agent polyvinylpyrrolidone (PVP) over a period of many years. The loss of PVP roughly correlates with the total hypochlorite dose *Ct*, the product of the reagent concentration and total exposure time. In practice the total *Ct* for a polymeric membrane installation at the end of its life has been reported as being in the region of 200 gh/L (Causserand et al., 2015; Robinson and Bérubé, 2020).
- 3 Since polymeric membranes decrease in mechanical strength with age it follows that pinning frequency must correspondingly increase,

incurring higher labour costs with time until the membranes are replaced. This supplementary labour cost represents a significant additional component of the OPEX which does not arise in ceramic membrane installations.

- 4 The irreversible loss of permeability also determines the membrane life. As with membrane integrity, long-term permeability loss does not appear to arise for ceramic membranes.
- 5 The chemical and mechanical robustness of ceramic membranes allows them to be exposed to high *Ct* levels. This then implies a higher chemical consumption than that associated with polymeric membranes, such that it may become the primary OPEX component.
- 6 There is additionally increasing anecdotal evidence, unpublished and based only on practitioner experience, to suggest that polymeric membranes challenged with pre-coagulated NOM-laden surface waters may be subject to clogging (or sludging). As with permeability decline and integrity, sludging appears to be age-related, and also incurs significant staff effort (and thus labour cost) for its amelioration (Judd and Carra, 2021).

The above means that the actual OPEX for a polymeric membrane plant challenged with a surface water feed is likely to resemble the trend shown in Fig. 4. The ceramic membrane OPEX (the dashed line in Fig. 4) is generally higher than the polymeric one due to the increased chemical consumption required to maintain membrane permeability, but is independent of time. Lastly, costs will be affected by investment in equipment to:

- (a) suppress polymeric membrane sludging through installing clarification technologies such as dissolved air flotation, employed upstream of the membrane filter and downstream of coagulant dosing at some plants,
- (b) suppress organic fouling of ceramic membranes using pre-coagulation, SIX, as at the Netherlands site (Section 2.5), and/or powdered activated carbon, tested at pilot scale by Campinas et al. (2021), and
- (c) enhance oxidative membrane cleaning using ozonation, such as at the Choa Chu Kang Waterworks (CCKW) in Singapore (Galjaard et al., 2011, 2013; Lam et al., 2021) as an alternative to hydrogen peroxide.

Despite the envisaged significance of the above OPEX elements, evidence suggests that an indication of the relative cost benefit or penalty of implementing a ceramic membrane system over a polymeric one is through the membrane cost-flux-life (CFL) ratio:

$$\text{Cost penalty} = 0.41 \times \text{CFL} - 0.277$$

where

$$\text{Cost penalty} = \frac{\text{Total cost parameter for ceramic membrane installation}}{\text{Total cost parameter for polymeric membrane installation}} - 100\%$$

and

$$\text{CFL} = \frac{\text{Ceram. cost}}{\text{Polym. cost}} \times \frac{\text{Polym. flux}}{\text{Ceram. flux}} \times \frac{\text{Polym. life}}{\text{Ceram. life}}$$

However, both the coefficient and intercept values in this equation must necessarily depend on factors impacting on the OPEX used to determine the overall cost penalty, including the assumed energy and chemicals consumption and the labour effort.

5. Conclusions

A review of available information pertinent to the key practical concern of the comparative operability and total/life cycle cost of polymeric and ceramic membrane-based installations for treating

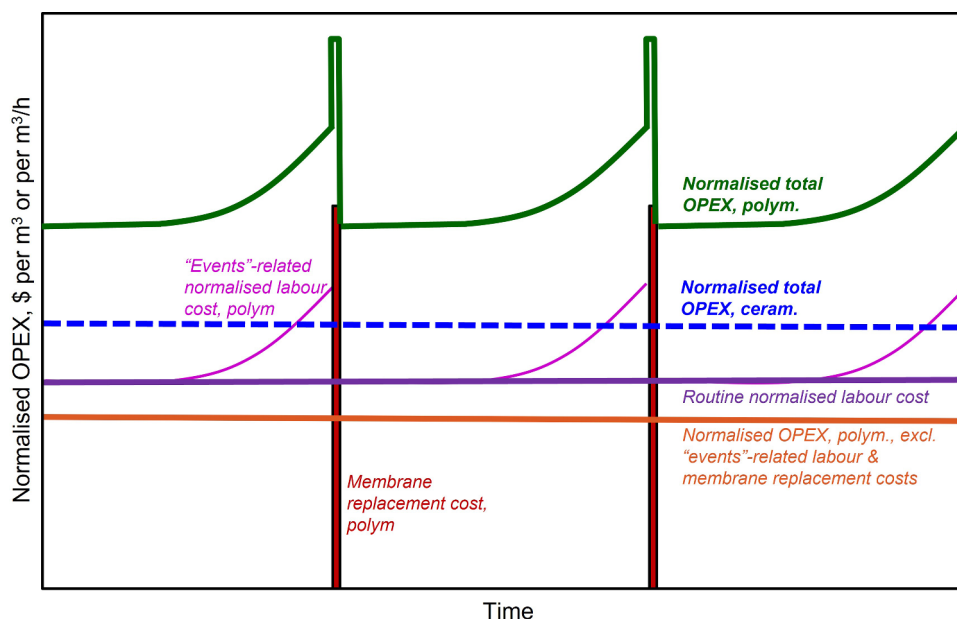


Fig. 4. Notional trend in OPEX for polymeric and ceramic membrane plants.

surface waters for potable water production has been conducted. The following conclusions can be drawn.

- 1 Short-term bench-scale tests provide conflicting conclusions regarding the relative sustainable permeability of the two materials, in part because of the differing hydrodynamic conditions employed. Bench-scale trials also cannot precisely replicate the cleaning protocols of full-scale plants or, unless based on aged membranes sampled from full-scale plants, account for the reduced polymeric membrane hydrophilicity with age.
- 2 There is a measurable correlation between impaired polymeric membrane performance, both with respect to permeability and mechanical strength, and the total exposure to hypochlorite over the life of a plant. The impact of the hypochlorite is on the PVP content of the membrane, which is denaturised and/or leached from the membrane material bulk or surface through the action of the hypochlorite, hypochlorite being widely used to recover membrane permeability. When challenged with waters having a significant organic matter content (such as surface waters), the exposure to hypochlorite during the cleaning cycle appears to limit the membrane life to ~6–8 years.
- 3 Contrary to the observations made regarding polymeric membranes, evidence from existing full-scale ceramic membrane installations suggests that ceramic membranes are sufficiently robust to allow aggressive permeability recovery measures without impairment of the warranted membrane life of 15–20 years or more. The increased intensity of the cleaning protocol, which can extend to ozonation at some installations, increases the contribution of this component to the OPEX.
- 4 Labour costs associated with maintaining polymeric membrane-based installations depend on the rate and extent of the decline in permeability and mechanical strength. There may also be additional costs associated with “sludging”, the accumulation of solids within the polymeric hollow fibre bundle which incurs both significantly increased labour effort and chemicals consumption for cleaning as well as further reducing the membrane life.
- 5 The cost benefit offered by a ceramic membrane system can be inferred from the ratio of the ceramic and polymer membrane cost mitigated by the flux and life ratios. However, this simple ratio excludes key operating parameters such as energy and chemicals consumption and, in particular, the labour costs.

Ceramic membrane systems are associated with higher CAPEX, due to the higher-cost membrane material, but lower OPEX if both the membrane replacement and labour costs are accounted for (Fig. 4). Procurement processes which (a) favour or permit offsetting of CAPEX and/or (b) rely on accurate prediction of OPEX may therefore favour ceramic over polymeric membrane systems.

Finally, it should be acknowledged that the fabrication of commercial membranes is continuously evolving. Whilst the cost of ceramic membranes has decreased over the past 15–20 years, the same is also true – though perhaps to a lesser extent – of polymeric membranes. At the same time, improvements in membrane materials and membrane module design have generally increased their robustness, which challenges accurate predictions of the comparative costs of technologies based on the two different materials.

Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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Ceramic vs polymeric membrane implementation for potable water treatment

Jarvis, Peter; Carra, Irene; Jafari, M.; Judd, Simon J.

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