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Two indigenous high sheared membrane modules' performance expatiation for the ultrafiltration of polyethylene glycol

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Performance of the ultrafiltration (UF) with anti-thixotropic polyethylene glycol (PEG) 6000 was investigated in the present study using two forward-looking high sheared membrane modules called radial flow membrane module (RFMM) and turbine flow membrane module (TFMM). In addition, the present study has been extended to investigate the effect of both polyether sulfone (PES) and poly sulfone (PSf) membranes equipped within the individual module during ultrafiltration of hydrophilic PEG 6000. Moreover, applicability of the membrane-module combination was judged along with the membrane reusability and the power consumed by these modules. TFMM equipped with PSf membrane was found to be more effective in this respect at 20 kg m^{-3} initial PEG 6000 concentration and 0.2 MPa in creating maximum shear rate over the membrane attributing to moderate flux with low pump energy consumption. Around 97% water flux regains after two consecutive experiments with maximum initial PEG 6000 concentration in order to conclude the reusability of the post-washed membrane even the membrane was exposed to the highest level of concentration.

1 Introduction

Membrane separation nowadays is one of the mostly explored technologies that are frequently used in several industrial sectors for their various purposes. Dairy, beverages, pharmaceuticals, paper-pulp, desalination plants *etc.* are such genres of industries, where membrane separation processes are one of the coveted applications in the downstream processes, either to separate valuable components or to treat the effluent in order to meet the legislation for waste disposal by the local pollution control authority.[1] However, one of the intricate issues here, that causes most of the concerns, is to explore several interactive parameters that enact during the separation process and control the process performance in terms of permeation as well as separation extent of the membrane.[2] Solute-membrane interaction, solute deposition on the membrane surface, blocking of pores because of impingement of the solutes are such controlling factors, collectively called 'fouling'. Such factors need to be understood primarily to design a membrane separation process scheme with utmost efficiency and also to elongate the membrane reusability period. Hence, selection of membrane with proper molecular weight cut off (MWCO) and fabrication of so-called '*high sheared membrane module*' are considered as the key solutions to reduce the possibilities of fouling of the membrane.[3,4] Moreover, the primal issues with such high sheared devices in the present days are its design in order to reduce its operating cost at the expense of substantial break-through in gaining the reusability of the membrane with increased rejection.

Several researches have already been published on such cost effective shear enhanced membrane module to ensure increased separation by reducing the foulants because of deposition/adsorption over the membrane surface. 'Rotating cylindrical membranes' was considered as the most primal shear enhanced membrane device, that was commercialised in mid-80's to collect plasma from donors' blood.[5] Here one of the membranes rotates against a fixed membrane to induce Taylor vortices in the fluid entrapped within the annular space created by the housing of the two membranes and thereby, enhance the extent of separation. However, one of the limitations in this module is its fabrication complexity and low effective membrane area. In 1992, another high shear membrane module, called vibratory shear enhanced processing (VSEP) was primarily proposed by Armando *et al.*[6] where the membranes are stacked along a centrally mounted shaft, that was induced by azimuthal oscillations at resonance frequency by a vibrating base. However, commercial application of VSEP is primarily restricted to nanofiltration (NF) and reverse osmosis (RO) as the system can withstand upto 15 bar,[7] *i.e.* required for the two pre-mentioned membrane

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separation process. Complying with the concept of VSEP, Gomma[8] proposed an oscillating flat 0.22 μm nylon microfiltration (MF) membrane, equipped with turbulence promoters for the separation of yeast cells. In 1995, Lee *et al.*[9] proposed a rotating multi-disk system, where disks were mounted on a single shaft and rotating between fixed circular membranes. Extending the concept of rotating disk module, Ding *et al.*[10] developed a rotating multi-shaft disk membrane, where the membranes were rotating with an overlapping region between them. Unlike rotating disk system, Sarkar *et al.*[11] introduced the concept of rotating disk membrane module (RDMM), where the membrane rotates to create enough turbulence over its surface leading to a reduction in concentration polarization. However, with all such high sheared modules, one of the most intricate issues is the cost of expensed energy in operating the device. Sen *et al.*[1] had investigated on the extent of separation with bovine serum albumin (BSA) and glucose, a simulated solution resembling dairy effluent, using a polyether sulphone (PES) membrane fitted in a vane equipped rotating disk membrane module (RDMM). Vanes fitted within the module, either rotate with the membrane or will be stationary with respect to the membrane. According to them, the attachment of vanes acts as a turbulence promoter within the RDMM to develop high shear environment over the membrane surface, a typical feature manifesting shear-enhanced membrane device, at the expense of low energy consumption compared to a normal RDMM in order to combat with the fouling. However, their application was limited to high MWCO UF range. Luo *et al.*[12] proposed a similar approach with vane attached rotating disc system to treat detergent wastewater, but with MWCO as low as in NF process. Furthermore, among the genre of such indigenous fabricated high shear membrane modules, cross-flow membrane filtration unit is the simplest form that is widely used in industry. Here the fluid flowing across the membrane surface imparts a sweeping action over the membrane to remove the deposited macromolecules from the membrane surface, which attributes to the reduced concentration polarization effect along with an achievement of moderate permeate flux through the membrane. Among several applications with a cross-flow membrane module, separation of whey proteins,[13] industrial oily wastewater treatment for the reduction hydrocarbons prior to its disposal,[14] recovering lignosulfonates from black liquor[15] *etc.* are some typical applications that are being adopted by many industries. However, apart from its several advantages, it attributes to a high capital cost due to its large footprint area and a constant abrasion effect on the membrane surface,[16] that might reduce the reusability of the membrane for a substantial period, especially when the feed to the module is highly viscous. Moreover, the modules described apart from cross-flow filtration module, might contribute to high operational cost in terms of energy consumed per permeate volume collection.

Hence, present work focuses on devising two indigenous membrane modules, named radial flow membrane module (RFMM) and turbine flow membrane module (TFMM), for concentrating polyethylene glycol (PEG) 6000 solution using PES and polysulphone (PSf) UF membrane. One of the critical attempts in fabricating two modules was to create enough shear rates over the membrane without aiding any external arrangement, in order to assert a steady flux from the membrane with enhanced membrane life. According to the modules, described in the subsequent Sections 2.3 and 2.4, the flow pattern of the introduced fluid was controlled in a way, so that it will be effective to create either enough shears or guide the shear enhancement arrangements within the module. Thus, elimination of external aid to create high shear will be expected to reduce the energy consumption per volume of permeate collection from the membrane and henceforth, the operating cost.

2 Experimental

2.1 Materials and methods

PEG 6000 was procured from Merck, India, (CAS-no. 25322-68-3) to prepare 0.004 m^3 (4 L) of feedstock solution with varying concentrations of 10 kg m^{-3}, 20 kg m^{-3} and 30 kg m^{-3} in the present study. Ultra pure deionised water, used to prepare the feed solution at different concentrations, was collected from the Arium RO unit followed by Arium 611DI ultra pure water system (Sartorius, Göttingen, Germany). Prior to the experimental runs, both PES and PSf membranes were subjected to water compaction[1] at a trans-membrane pressure (TMP) (ΔP) of 0.6 MPa, higher than that of experimental TMPs (0.2 MPa, 0.3 MPa and 0.4 MPa). Compaction of the membrane was ensured after 3600 s, when the water flux became steady for a substantial period. Sodium hypochlorite (NaOCl), sodium hydroxide (NaOH) (used for membrane washing) and ethanol (used for membrane storage) were purchased from Merck (Mumbai, India). In each of the experimental runs a volume concentration factor (VCF)[13] (eqn (1)) of 2.0 was maintained.

$$\mathrm{VCF} = \frac{\text{Volume of the feed}}{\text{Volume of the retentate}} \tag{1}$$

2.2 Membrane modules and the flow arrangement within the modules

TFMM and RFMM (fabricated by Concept International Ltd, India), both the modules were made up of SS316. A plunger pump (Maker: American Sparing & Press Workshop Pvt. Ltd., Mumbai, 3 H.P.; Maximum Pressure: 400 psi) was installed with the module to withdraw PEG 6000 feed solution from a jacketed feed tank (Capacity: 0.005 m^3 or 5 L) and fed it to the membrane module. All the experiments had been carried out at 298 K (25 °C). The pressure difference across the membrane was controlled using a flow restrictor, called "back pressure regulator valve (BPRV)", fitted in the retentate line. Fig. 1 shows a schematic diagram of the experimental setup used for experiment. During the experiments, both the modules were equipped individually with 5 kg mol^{-1} PES and PSf membranes (0.041 m effective diameter), procured from Koch Membrane, USA. The concentration of PEG 6000 in the feed, retentate and permeate, had been measured using a calibrated digital Brix refractometer (RFM712; Model no. BX11040). Details of individual modules, TFMM and RFMM, are narrated in the following sub Sections 2.3 and 2.4.

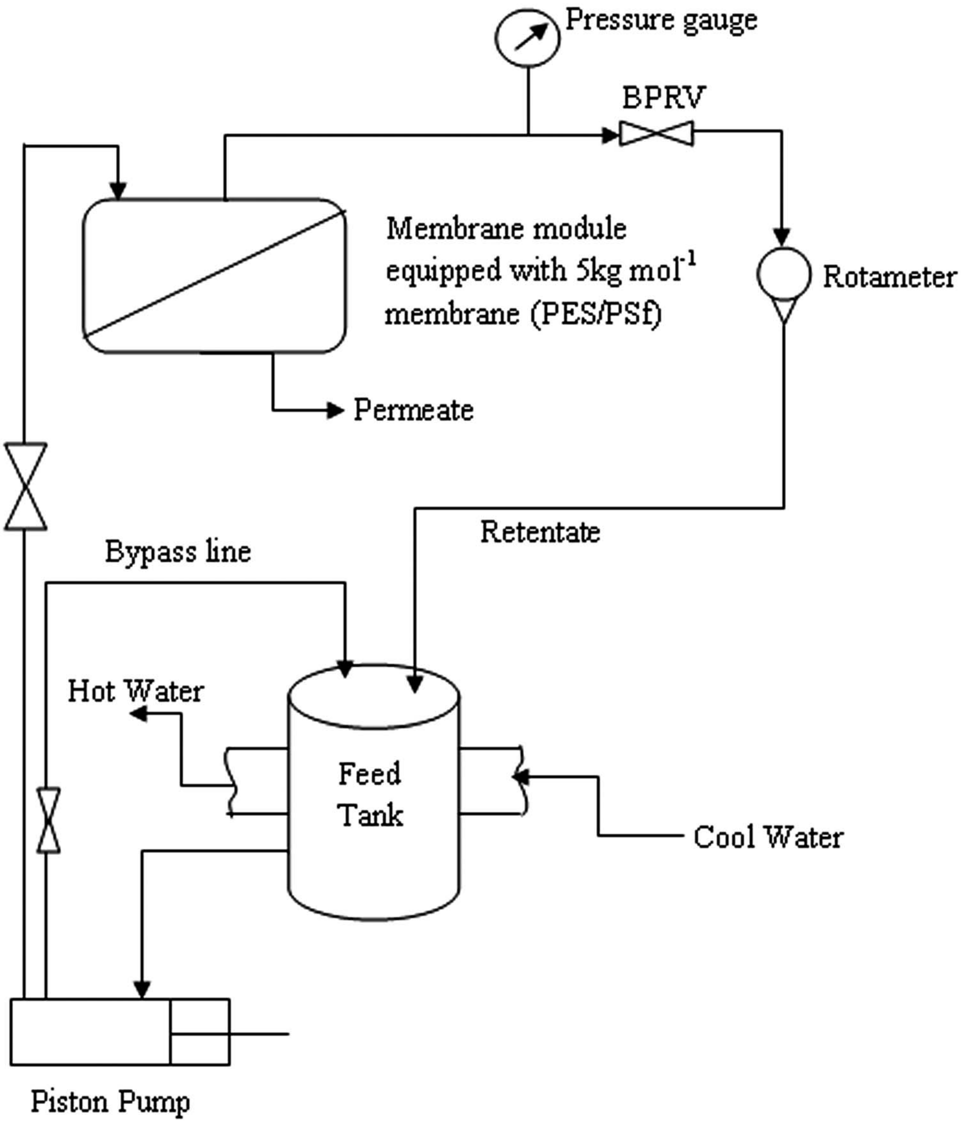

Fig. 1 Schematic diagram of the membrane module set-up.

A series of control experiment has been carried out using Amicon® stirred UF cell model 8050 (Millipore Corporation, Bedford, MA 01730 USA) with an effective membrane diameter of 0.040 m.

2.3 Turbine flow membrane module (TFMM)

Fig. 2(a) shows the original snapshots of TFMM, while Fig. 2(b) demonstrates the schematic representation or may be called the free body diagram in order to understand the forces acting over the blades. The module is made up of SS316, while the blades are made up of PVC. Within the module, the membrane is placed over a perforated sheet and the top casing, with which the blades are bolted, are placed over the membrane using a rubber gasket. The diameters of the inlet and outlet holes within the module are 0.004 m each. Turbine arrangement comprises of four blades rotating around a central axis because of the dynamic head developed by the inlet fluid against the back pressure exerted on the blades by BPRV. Each of the blades are having dimensions of $0.011 \times 0.01 \times 0.002$ m. Blades are having a clearance of 0.001 m with the membrane and with the top casing surface of the module. Hence, as from the Fig. 2(b), the area over which the dynamic head will act is 0.011×0.01 m^2.

2.4 Radial flow membrane module (RFMM)

Fig. 3 shows the original snapshot of RFMM respectively. The material of construction is same as TFMM, while the differences lie only in the inlet and outlet arrangement of the fluid used in the experiment. The feed solution was fed to the module through a channel of 0.0005 m, fabricated by a frontally fixed plate of diameter 0.016 m. The plate has a clearance of 0.001 m with the introduction point near the feed line. The outlet of the retentate from the module has been made through nine holes (shown in the Fig. 3) of diameter 0.005 m. The clearance of the

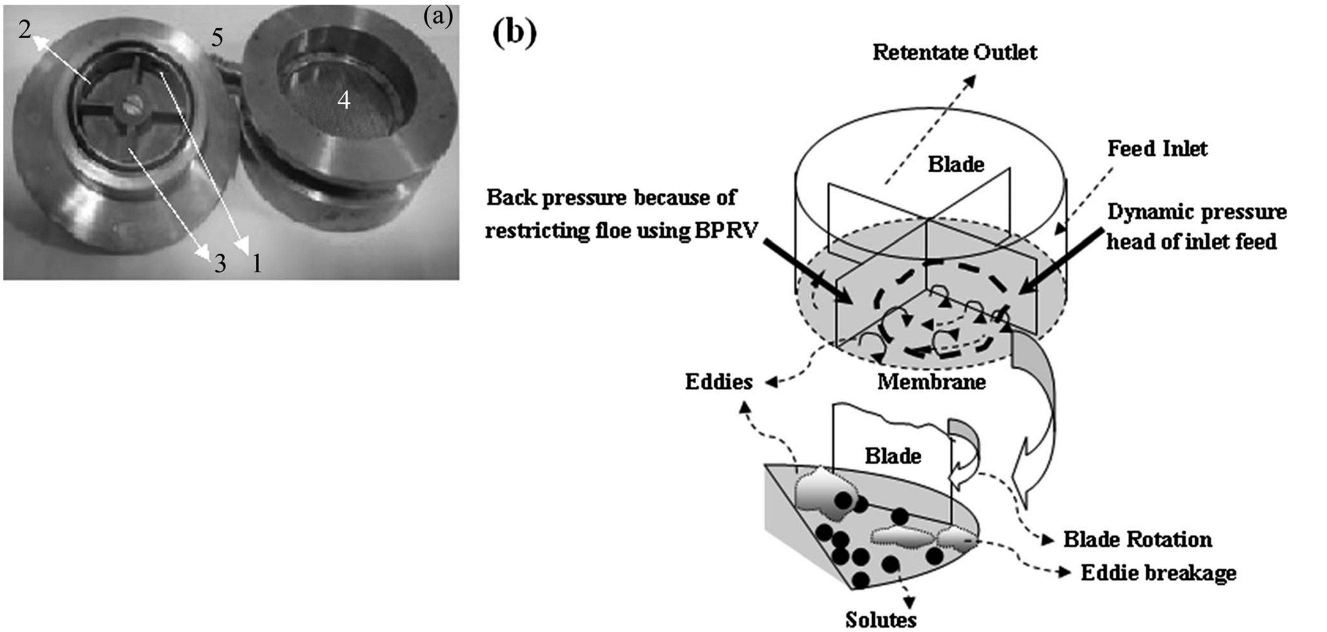

Fig. 2 (a) Turbine flow membrane module (1: feed inlet; 2: retentate outlet; 3: blade; 4: perforated plate; 5: permeate collector). (b) schematic representation of the possible fluid flow within the module.

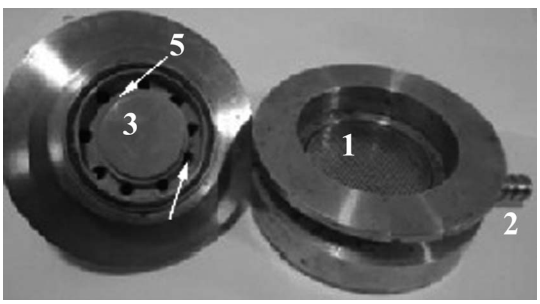

Fig. 3 Radial flow membrane module (1: perforated plate; 2: permeate collector; 3: frontier fixed plate mounted at the feed inlet channel; 4: retentate outlet; 5: feed inlet channel).

holes and the plate (distributor) with the membranes are 0.014 m and 0.007 m respectively.

3 Results and discussion

3.1 Effect of membrane and module on the permeate flux during ultrafiltration of PEG 6000

Fig. 4 through 7 presents the variation of time weighted average flux (TWAF) ($\langle J \rangle$), as given by eqn (2), with the applied trans-membrane pressure TMP and the initial feed concentration for both of the membrane modules, individually equipped with both PES and PSf membranes. The figures represents two-dimensional contour plot, which is the *z*-axis projection of three-dimensional surface plot, where the *x*-axis shows the initial feed concentration, *y*-axis shows the applied TMP and *z*-axis, given as the projected contour lines here, shows the TWAF. So far, the utility of TWAF over the normal average flux measurement is to include the effect of time, might be in an

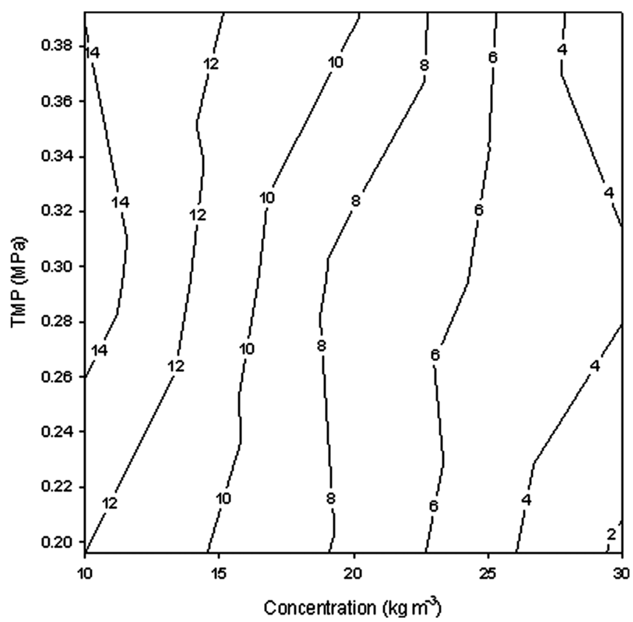

Fig. 4 Variation of TWAF ($\langle J \rangle \times 10^6$ m^3 m^{-2} s^{-1}) from RFMM equipped with PES membrane with varying initial feed concentration and TMP.

implicit way, on the permeate flux that either continuously decreases or becomes steady with the run time of the experiment. Hence, the concept of TWAF becomes advantageous with an inclusion of time over mean flux, when the variation of the time dependent flux ($J(t)$) was investigated with the variation of both primitive conditions, applied TMP and initial feed concentration in a single contour plot. Consequently, the contour plots effectively assist to locate the optimum parametric condition at which the modules can be operated for having a moderate permeate flux during ultrafiltration of PEG 6000.

$$\langle J\rangle = \frac{\sum_{t=0}^{t_f} J(t)t}{\sum_{t=0}^{t_f} t} \tag{2}$$

where, 't' represents the time at any instant and 't_f' is the final run time.

On investigating Fig. 4 and 6, which are representing PES equipped RFMM and TFMM respectively, there are not many differences depicted in the TWAF profile with varied feed concentration and applied TMP. While, on the contrary, with the PSf membrane equipped modules RFMM (Fig. 5) and TFMM (Fig. 7), TWAF is more or less 60–90% lower in RFMM compared to TFMM in all possible conditions. This incites to infer on a considerable effect of the membrane along with the module on the ultrafiltration of PEG 6000. Such low TWAF might ascribe to the hydrophobic nature of the PSf membrane compared to PES[17] that shows a likely disadvantage with the RFMM because of its hydrodynamics. At low feed concentration, because of the high water content, PSf repels the water enriched low concentrated solution of PEG. Therefore, the flux becomes 88–90% lower in RFMM compared to TFMM at low concentration of the PEG solution and at all TMPs. On the contrary, at moderate concentration TWAF reduction is around 75% on an average, that attributes to the possible internal flow mechanism and the fluid behaviour that leads to such reduction in the TWAF in RFMM. According to the operational procedure for both the modules, TMP exerts on the membrane surface with the help of a flow restrictor (shown as backpressure regulator valve (BPRV) in Fig. 2) attached to the retentate line. Hence, it is obvious in

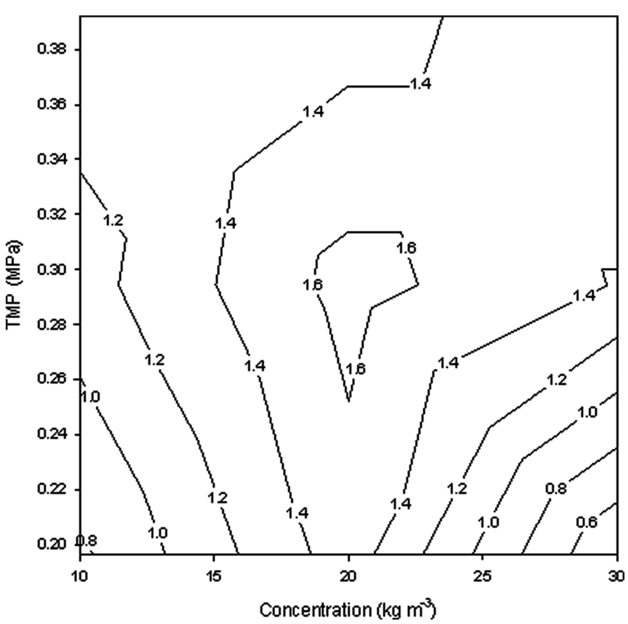

Fig. 5 Variation of TWAF ($\langle J\rangle \times 10^6$ m^3 m^{-2} s^{-1}) from RFMM equipped with PSf membrane with varying inlet feed concentration and TMP.

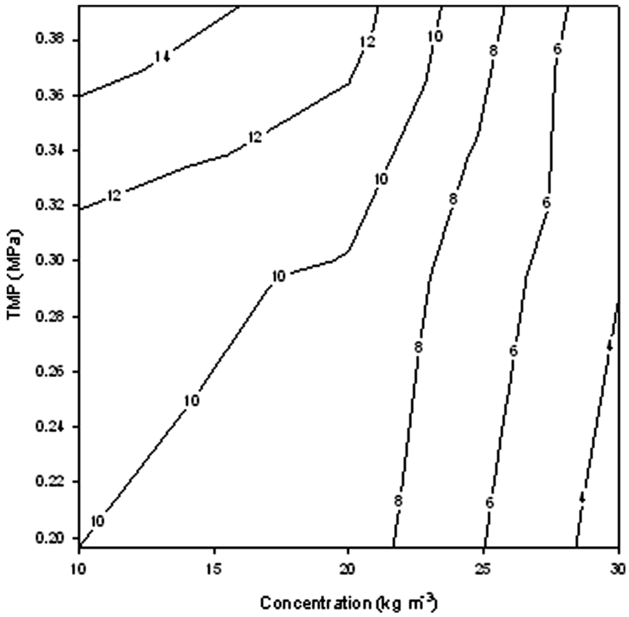

Fig. 6 Variation of TWAF ($\langle J\rangle \times 10^6$ m^3 m^{-2} s^{-1}) from TFMM equipped with PES membrane with varying initial feed concentration and TMP.

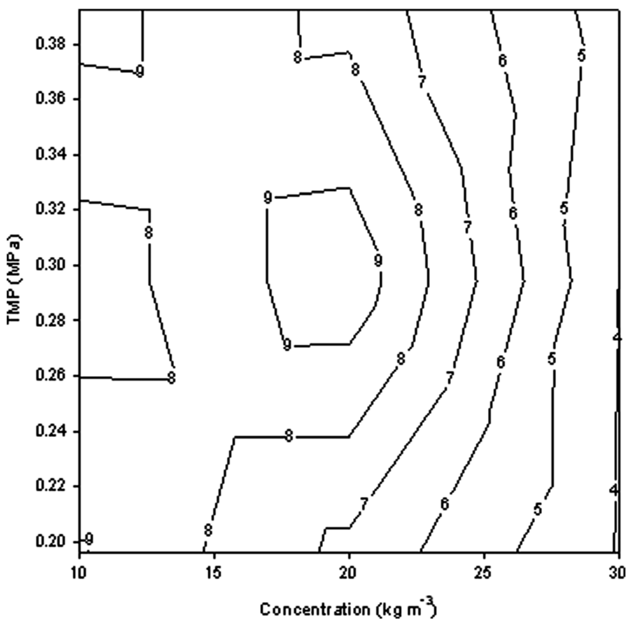

Fig. 7 Variation of TWAF ($\langle J\rangle \times 10^6$ m^3 m^{-2} s^{-1}) from TFMM equipped with PSf membrane with varying inlet feed concentration and TMP.

both the modules that less TMP results a high flow over the membrane, which eventually increases the shear rate over the membrane surface. In RFMM, the principle towards the creation of high shear development in order to alleviate the gel layer is the mass transport away from the membrane surface because of likely eddies formed over the membrane surface. While, in TFMM such mass transport occurs because of the swept gel layer by tangential displacement of the fluid element over the membrane surface during the rotation of the blades. Fig. 4 and

6 can enlighten the effect of such different hydrodynamics within the modules on TWAF, as the effect of comparatively more hydrophilic PES membrane on TWAF is almost trivial in these cases.

A quick glimpse to Fig. 4 and 6 manifests two almost likely profiles for TWAF with the variation in initial feed concentration and TMP, although with some minor unlikeness in the profiles if it will be scrutinised carefully. According to Etemad and Gholamhosseini,[18] the viscosity of PEG solution increases with the increase in shear rate. Consequently, such increase in viscosity will lead to more accumulation of the solute over the membrane surface because of ceased flow of the solution. Now, accumulation might attribute to a built-up gel layer or because of the increased shear rate between the fluid and the membrane surface. In case of RFMM, as discussed before, the module efficacy attributes to the formation of eddies. With high TMP, cr-eated using the flow restrictor, such formation of eddies either will be ceased or the formation will be followed by the rupture of eddies. This reduces the possibility of reduction in the gel layer due to the less transport of accumulated solute away from the membrane. Moreover, with time the permeation of water through the membrane increases the concentration of PEG solution over the membrane that consequently increases the viscosity of the solution. On the contrary, in case with TFMM, high TMP restricts the rotation of the turbine blade manifesting a low shear rate compared to that of with low TMP. Therefore, the viscosity of the PEG solution was less affected by such low shear rate along with the alleviation of gel layer by mild shear. Hence, not significant, but flux was on slight higher side at high TMP (0.4 MPa) for all the initial feed concentrations.

At 0.2 MPa and 0.3 MPa, for 10 kg m^{-3} initial feed concentration, the flux from RFMM is higher than that of from TFMM because of the enhanced shear rate that increases the viscosity of the solution in addition to the water permeation with time that result enhanced gel layer formation. Moreover, as the concentration increases with these TMPs, TFMM plays an important role in an effective removal of gel layer compared to RFMM by its cross-flow type flow arrangement until a threshold concentration level, which in this case was 20 kg m^{-3}. After this point, it was seen that at 30 kg m^{-3}, the TFMM performance decreases compared to that one observed with 20 kg m^{-3} because of the increased initial feed concentration and the increased shear rate at low TMPs. At 0.2 MPa and with 30 kg m^{-3}, there was practically no difference in TWAF from both TFMM and RFMM.

Fig. 8(a) shows the ratios between the TWAF from TFMM and RFMM with varying feed concentration and TMP in proof of the above discussion. However, as said before, with PSf membrane the scenario is quite different from that was seen with PES membrane. From Fig. 5 and 7, at high TMP of 0.4 MPa, unlike PES membrane, PSf membrane gives low ratios between the TWAF from TFMM and that from RFMM with increased concentration (Fig. 8(b)). One of the facets here could be PEG attachment on the PSf membrane surface due to the formation of hydrogen bond, which can be re-conferred from the wave number at around 667 cm^{-1} in the IR spectra taken with post-run PSf membrane as shown in Fig. 9. An accrued electron cloud over the sulfonic group due to the presence of neighbouring electron repelling –CH_3 group might lead to the formation of such hydrogen bond in case with a PSf membrane. Ensuing peaks at wave number 1150 cm^{-1} because of C–SO_2–C symmetric stretching[19] (Fig. 10) confirming such electron sharing events. In contrast with the PES equipped membrane, at 30 kg m^{-3} initial feed concentration, performance for TFMM increases compared to RFMM with the reduction in TMP. In case with the PSf membrane, the water permeation rate is slow.[17] Therefore, the rate of increase in the concentration of PEG over the membrane becomes lower that limits the enhanced anti-thixotropic behaviour of the PEG solution with concentration as described by Etemad and Gholamhosseini.[18] Therefore, in this case the comparative behaviour of both the modules depends on the enhanced shear field to alleviate the gel layer. With lowering the TMP, TFMM acts much promptly compared to RFMM because of its cross-flow arrangement

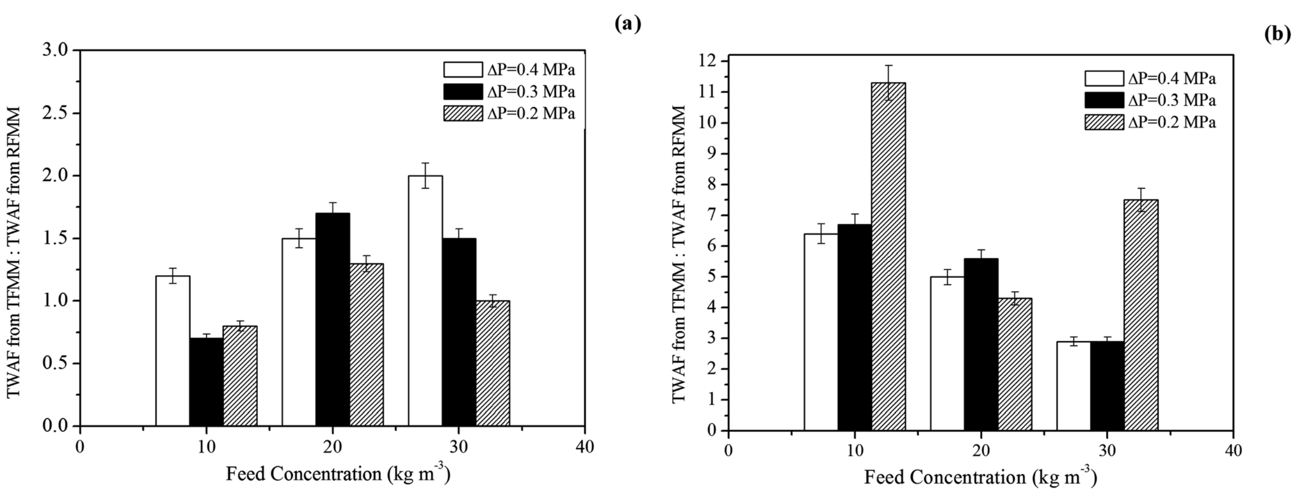

Fig. 8 (a) Variation of TWAF from PES membrane equipped TFMM over PES membrane equipped RFMM with varied inlet feed concentration and TMP. (b) Variation of TWAF from PSf membrane equipped TFMM over PSf membrane equipped RFMM with varied inlet feed concentration and TMP.

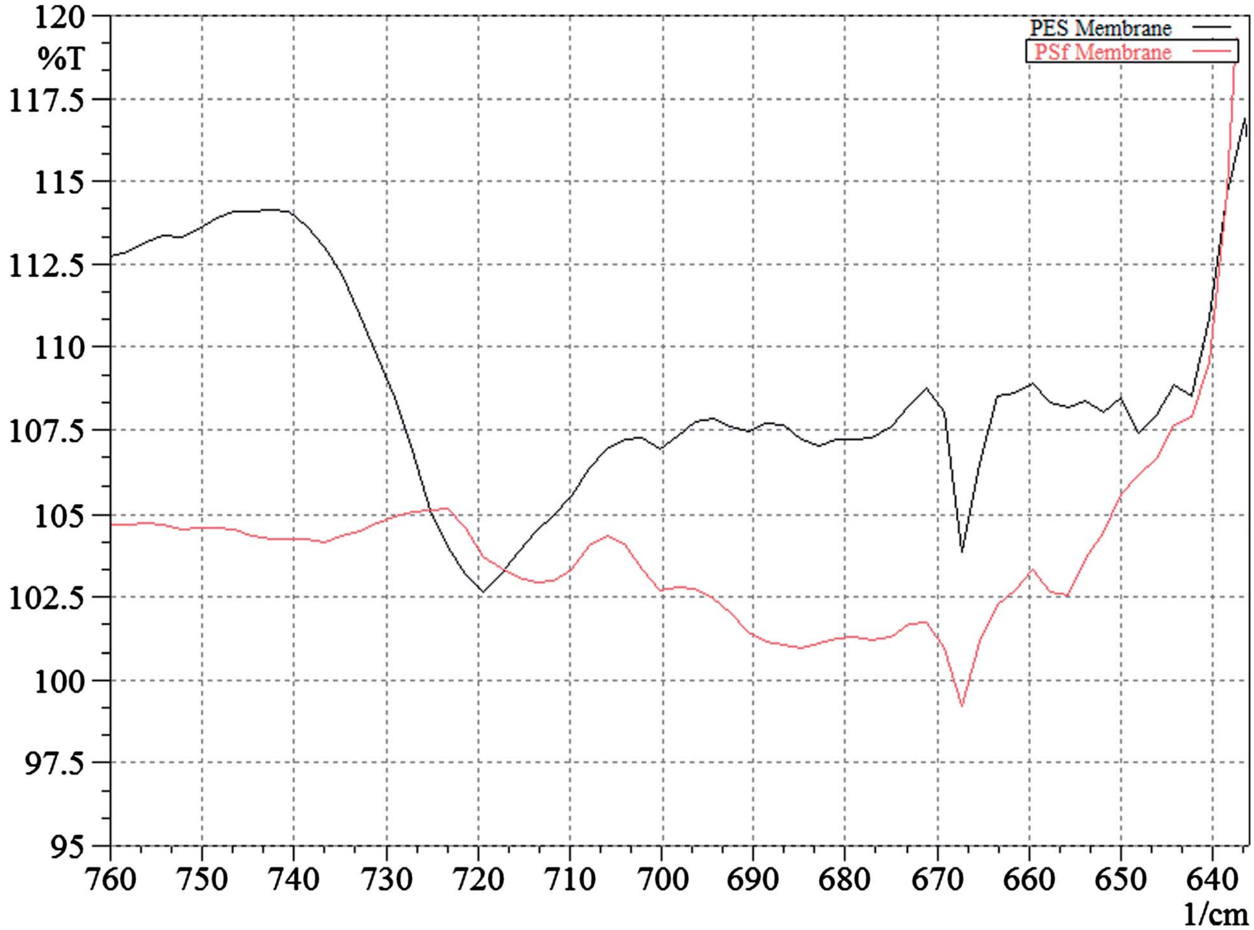

Fig. 9 FTIR spectra post-run PSf membrane displaying the range 637–750 cm^{-1}.

within the module. Hence, in compliance to the present context, one of the critically inquired facts is the power requirement by the modules that truly portrays the features of high sheared membrane modules, such as TFMM here, alleviating the resistance over the membrane against permeation. Such power consideration has been investigated in the following sections.

3.2 Comparative understanding of the power consumption by the feed pump between the two modules

As discussed, in TFMM, blades are acting as a shear enhancer over the membrane as seen in case with a stirrer equipped rotating disc membrane module (RDMM).[20] However, TFMM will be unveiled as a more advantageous module over the stirrer equipped RDMM in terms of energy consumption in order to alleviate the dispersion of gel layer and henceforth, to obtain a moderate permeate flux. In a stirrer equipped RDMM the prime energy involvement includes the motor power to rotate the stirrer and thus to create a proper mixing over the membrane in order to distort the gel layer. On the contrary, in TFMM, blades are rotating at angular speed 'ω_{blade}' by the virtue of the energy, generated because of dynamic head of the introduced fluid, spent against the backpressure developed by the flow restrictor (eqn (3)). Hence, the rest rotational kinetic energy developed by the blades becomes responsible for the flow in retentate line ($Q_{retentate}$) and is given by eqn (3). However, the formation of eddies and their structure prediction because of the varied backpressure by doing eddy simulation will be a complicated procedure and has been avoided in the present scope of the paper.

$$\frac{1}{2} I_{blade} {\omega_{blade}}^2 = \frac{1}{2} (V_{hold} \rho_{retentate}) \left[\frac{Q_{retentate}}{\pi({d_i}^2/4)} \right] \tag{3}$$

where, the moment of inertia (I_{blade}) is given by,

$$I_{blade} = M_{solid} \frac{L^2}{3}$$

where, M_{solid} is mass of the propeller; L is the length of the blade from the axis; d_i is the internal diameter of the retentate pipe; V_{hold} is the holdup volume within the module and $\rho_{retentate}$ is the specific gravity of the retentate w.r.t water at 25 °C.

However, one of the key concerns in the present module is adjusting the pump delivery flow rate (Q) and the backpressure in such a way so that it will exert a moderate TMP as well as creating enough shear to reduce the built-up resistances on the membrane surface. According to the operational principle of the module, TMP over the membrane can be increased by

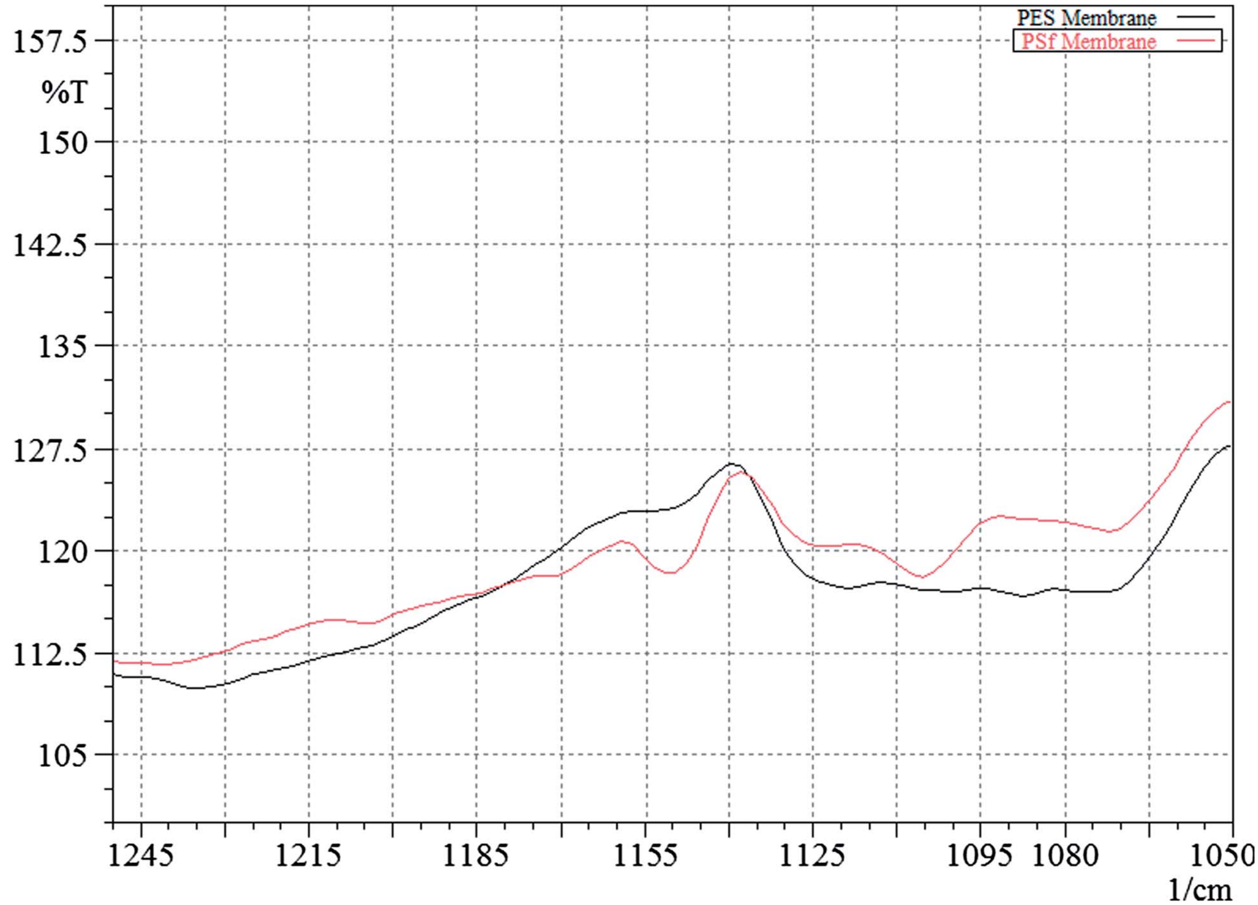

Fig. 10 FTIR spectra post-run PSf membrane displaying the range 1095–1200 cm^{-1}.

properly adjusting the pump bypass valve and BPRV. Eqn (4)[21] provides a convenient formulation to calculate the pump energy required per ml of permeate across unit area (m^2) (P) of the membrane at a certain TMP. However, in case of TFMM, the reduction in the resistant layer over the membrane is much depending on the generated shear rate by the rotating vanes over the membrane. The rotating blades can be hypothesized a disk (consisting of infinite number of blades) rotating over the membrane as the blades' revolution per second is very high, in the order of 10^2. Hence, subsequently mean shear rate enumeration applying eqn (3) in eqn (5)[7] in case with the flow between the blades and the membrane will provide a better understanding of the implicit relation between the backpressure and the shear rate. Fig. 11 shows the variation of shear rate with TMP and feed concentration. With increase in the TMP, the shear rate is continuously decreasing and hence, the chances of shear thickening effect have been reduced leading to a moderate flux from the module. Figure shows that at 0.4 MPa, the shear rate (γ) is same for all the initial concentration of the feed manifesting the restricted rotational speed of the blades.

$$P = \frac{P_{\text{m,pump}}}{J_{\text{s}}} = \frac{P_{\text{h,pump}}}{\eta_{\text{m}} J_{\text{s}}} = \frac{(\Delta PQ)}{\eta_{\text{m}} J_{\text{s}}} \tag{4}$$

$$\gamma = 0.0164(0.65\omega_{\text{blade}})^{1.8} R_{\text{membrane}}^{1.6} \left(\frac{\mu_{\text{retentate}}}{\rho_{\text{retentate}}}\right)^{-0.8} \tag{5}$$

where, $P_{\text{m,pump}}$ is the mechanical power supplied by the pump; $P_{\text{h,pump}}$ is the Hydraulic power supplied by the pump; J_{s} is the steady state permeate flux; η_{m} is the mechanical efficiency of the

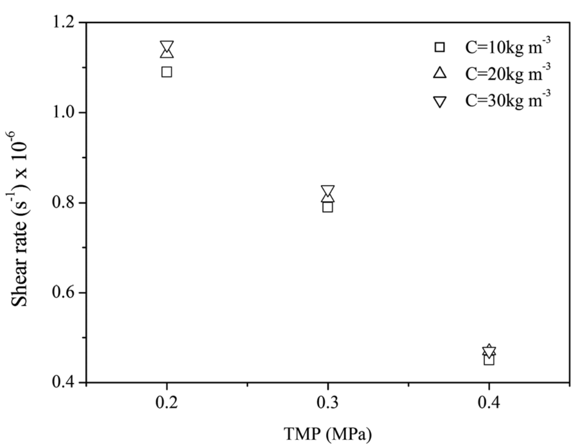

Fig. 11 Variation of shear rate of PEG 6000 solution on the membrane surface imparted by the rotating turbine blades attached within TFMM with varying TMP at different initial feed concentration.

pump; $\mu_{retentate}$ is the viscosity of the retentate and $R_{membrane}$ is the effective radius of the membrane.

Fig. 12 shows a comparative figure on the fold increase in feed pump's energy (eqn (6)) consumed per ml of permeate across unit area (m^2) of the membrane of RFMM over TFMM with varying feed concentration. The figure shows that with RFMM, the energy requirement is high almost every time compared to TFMM to generate 1 ml of permeate flux across unit area (m^2) of the membrane. The possible reason might indicate the more energy requirement in case of RFMM because of the mass transport away from the membrane surface by instable eddy transport. Especially, with increased concentration the deposition will be more and hence requires more eddy transport in order to alleviate the resistance. Furthermore, the inclusion of PSf membrane makes the process so energy intensive reasoned to possible PEG bind with the membrane (Fig. 9 and 10) or the comparatively more hydrophobic nature of the PSf membrane showing to PEG solution.[17] On the contrary, at low TMP and initial feed concentration, the shear action is prominent with TFMM in order to alleviate the polarisation and manifest moderate permeate collection. Therefore, with PES, the fold increase is less than '1' at 10 kg m^{-3} feed concentration and at low to moderate TMP.

Fig. 13 shows the power number (N_P) variation with the Reynold's number ($Re_{rotational}$) for different TMP. Hence, lowering the rotational Reynold's number attributes to high TMP according to the operational principle of the module. Therefore, at high TMP, the drag force has minimum impact on the rotating blade manifested by a flattened slope in the power number. At 0.2 MPa and 0.3 MPa, there is a sharp decrease in power number with increasing $Re_{rotational}$ indicates more power consumption against an increasing drag force because of shear thickening rheopexy of PEG 6000 attributing to high shear rate (Fig. 11).

3.3 Conferring the reusability of the membrane based on the regaining of post-washing water flux

One of the bottom line facts in any membrane separation process is the reusability of membrane or might be the life of the membrane unit. Such reusability has been adjudicated based on the post-wash percentage water flux regain after each experiment that implicitly manifests the extent of permanent fouling of the membrane. Permanent fouling or more specifically the clogging of the membrane pores might be incurred due to prolonged exposure of the membrane surface to the deposited macrosolutes over the membrane impinges within the pores because of the applied pressure. Hence, reduction in the residence time of such solutes on the membrane surface reduces the possibility of such permanent clogging of the pores. This reduction in time will be modulated by the enhanced shear action over the membrane surface or less membrane-solute interaction that potentially reduces the energy barrier to remove the deposited layer over the surface. Therefore, it is quite apprehensible that the membrane and the membrane module, both are having substantial effect on conferring the reusability of the membrane. Fig. 14 shows a comparative graph on the percentage regain of the water flux for each combination of module and membrane in two

$$\text{Pump energy fold increase} = \frac{\text{Pump energy required per ml of permeate across unit membrane area from RFMM}}{\text{Pump energy required per ml of permeate across unit membrane area from TFMM}} \quad (6)$$

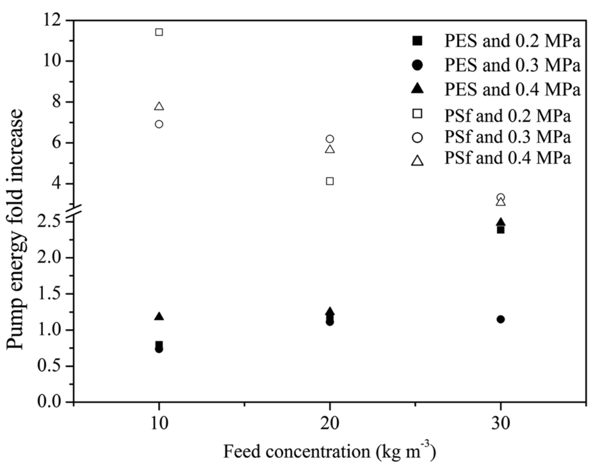

Fig. 12 Variations of fold increase of feed pump energy per ml of permeate collection across unit area (m^2) of the membrane from RFMM over TFMM, equipped with both PES and PSf membrane subsequently, with initial feed concentration at varying TMP.

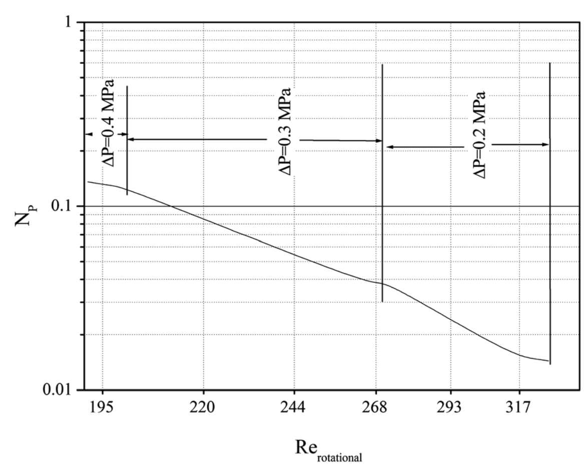

Fig. 13 Power number variation with rotational Reynold's number enumerated for the fluid flow within the TFMM (irrespective of the membrane) with TMP.

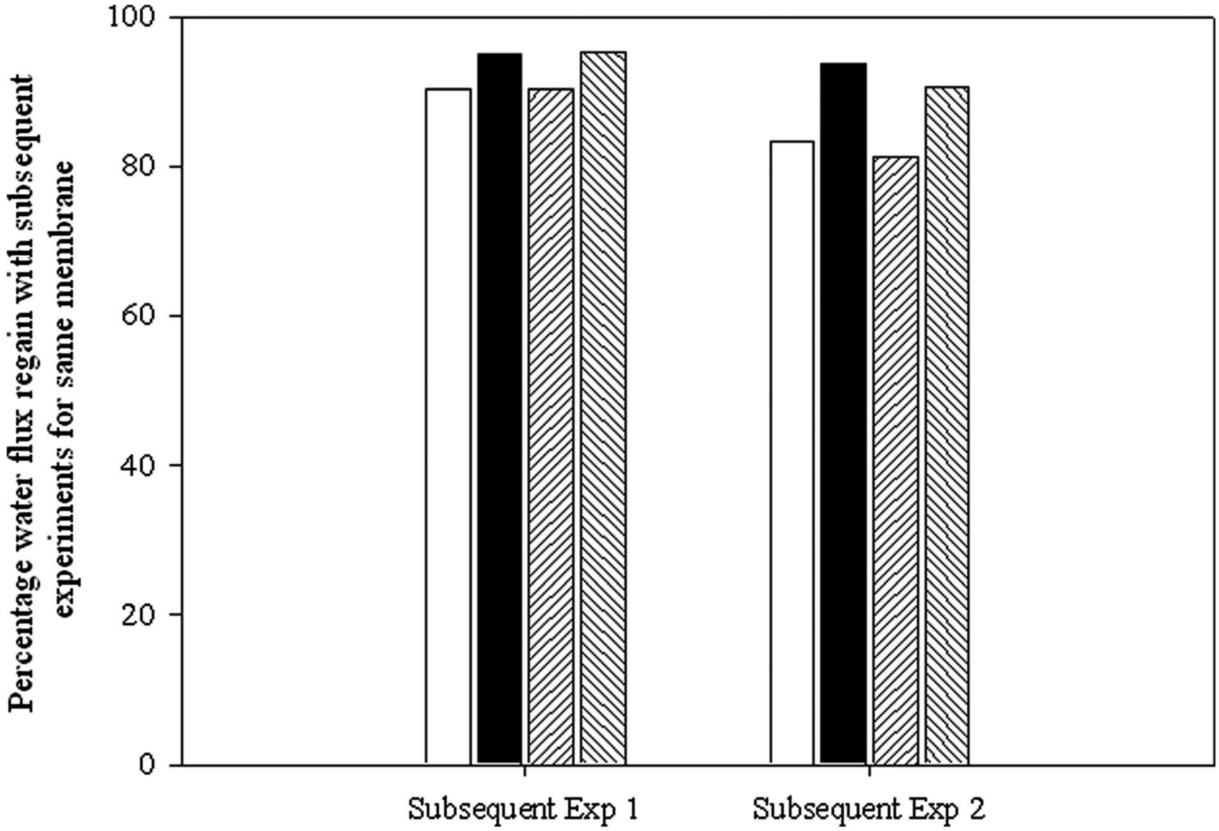

Fig. 14 Comparative water flux percentage regain for subsequent experiments with the same membrane in each module (□: PES membrane fitted in RFMM; ■: PSf membrane fitted in RFMM; ▨: PES membrane fitted in TFMM; ▧: PSf membrane fitted in TFMM).

consecutive runs with the same membrane and chemical wash in between. Decline in the water flux during subsequent experiments, even after adequate chemical wash of the membrane, exhibits the extent of permanent fouling. From the figure, the continuous decrease in the percentage water flux regain showing PES membrane is more vulnerable to fouling compared to stable PSf membrane[17] under weakly protonated condition of the PEG solution.[19]

3.4 Understanding the influence of membrane selection and module selection on permeate flux through analysis of variance (ANOVA) test

A two-way ANOVA test was performed on TWAF with varying TMPs and feed concentration to understand the influence of membranes and the membrane modules. Minitab 15.1.1 software was used to execute two-way ANOVA test and consequently, the results are given in Table 1. Smaller '*P*' value from ANOVA suggests less chances of making type I error, an error results because of wrongly rejecting null hypothesis among the groups even if it is true. Thus, smaller 'P' value implies that the corresponding parameter or the condition has significant impact on controlling the process performance. From Table 1, it is obvious that the membrane has more controlling effect on obtaining TWAF compared to the membrane module and thus, the result manifests the fact of membrane interaction with the solutes depending on the nature of the membrane used. Hence, membrane choice can be considered more typical than the module in concentrating PEG 6000 solution. While, in case with the selection of membrane module the only differences that can alter the separation efficiency is the in-house hydrodynamics, discussed earlier, to reduce the gel layer on the membrane surface. The results can be well judged from the variation of the permeate flux with time (Fig. 15), for different sets of membrane modules, equipped with both PSf and PES membranes. Figure shows that with PES membrane the average flux (eqn (7)) is only 11% more in TFMM than RFMM. While, in the case of PSf membrane equipped in TFMM, average flux was reduced by almost 19% compared to PES equipped TFMM.

Table 1 ANOVA test for the effect on TWAF ($\langle J \rangle \times 10^6$ m^3 m^{-2} s^{-1}) due to the variations in membrane and membrane modules[a]

Source	DF	SS	MS	F	P($\alpha = 0.05$)
Membrane module	1	88.827	88.827	7.46	0.010
Membrane	1	179.841	179.841	15.11	0.000
Interaction	1	59.192	59.192	4.97	0.033
Error	32	380.851			
Total	35	708.711			

[a] DF: Degrees of freedom; SS: sum squared error; MS: mean squared error; *F*: *F*-value; P: probability of making type-I error.

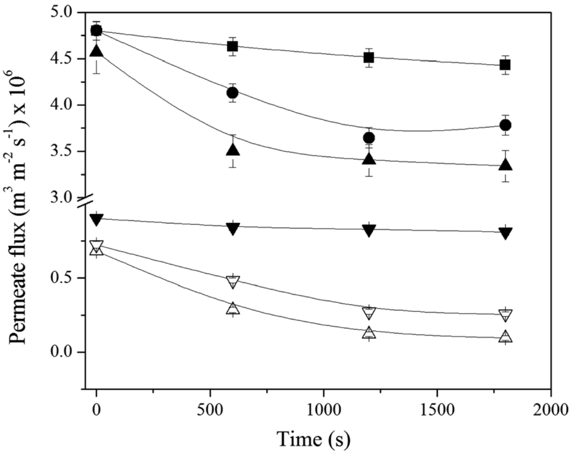

Fig. 15 Variation of permeate flux with time at 0.2 MPa and with 20 kg m^{-3} initial feed concentration for the membrane modules equipped with PES and PSf individually (-■-: TFMM equipped with PES membrane; -●-: RFMM equipped with PES membrane; -▲-: TFMM equipped with PSf membrane; -▼- RFMM equipped with PSf membrane; -△-: Amicon stirred cell equipped with PES membrane; -▽-: Amicon stirred cell equipped with PSf membrane).

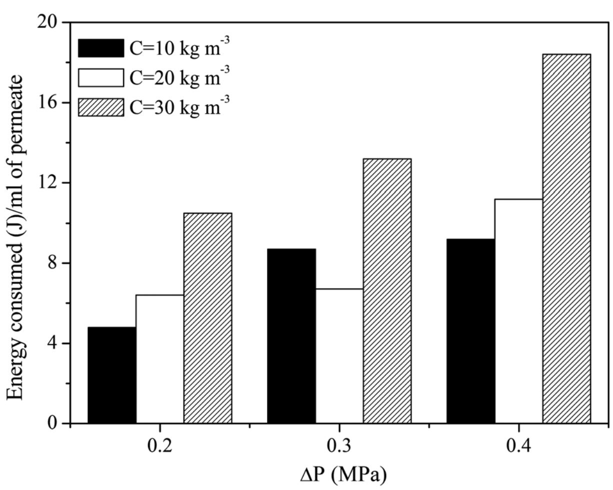

Fig. 16 Variations of energy requirement per ml of permeate collection across unit area (m^2) of the membrane for TFMM equipped with PSf membrane at different TMP and initial feed concentration.

Table 2 Consumption of energy per ml of permeate collection across unit area (m^2) of the membrane for previously studied different membrane modules along with TFMM at operating conditions 0.2 MPa TMP and 10 kg m^{-3} PEG 6000 solution

Membrane module	Membrane attached	Energy required (J) per ml of permeate collection across unit area (m^2) of the membrane
Cross-flow membrane module	CA[15]	4.7×10^6
Spinning Basket Membrane (SBM) module	PES[22]	7.2×10^6
TFMM (Proposed here)	PSf	4.8×10^6

The performance of TFMM had also been understood after comparing it with the well known Amicon stirred cell equipped with both PES and PSf membrane. With PES, Amicon stirred cell shows 60% decrease in the permeate flux w.r.t. initial flux compared to TFMM. While with PSf membrane it manifests almost 53% flux drop w.r.t. initial flux compared to TFMM.

$$\bar{J} = \frac{\sum_{t=0}^{t_f} J(t)}{n} \tag{7}$$

4 Conclusions

The present investigation illustrates a comparative survey on the application of two indigenously fabricated membrane modules RFMM and TFMM for concentrating PEG 6000 solution. In conjunction, the performances of both the modules are being evaluated after equipping the modules with PES and PSf membrane. Moreover, as a primary concern, two native modules were fabricated in order to obtain moderate flux with low energy consumption and in addition, ensuring the prolonged reusability of the membrane. Hence, the key finding from the present observations with the aqueous solution of anti-thixotropic PEG 6000 confers:

With PES, practically there are no differences in TWAF from both the modules. However, with PSf membrane, flux from TFMM increases around 7–12 fold compared to RFMM at low/high feed concentration but with low TMP. However, PSf fitted TFMM, gives a moderate flux as compared to the PES membrane, a mere 19% reduction towards obtaining the steady state flux.

Both the modules equipped with PSf membrane exhibits almost 97% water flux regain after every subsequent wash and this attributes to the enhanced possibility of membrane reusability.

However, the energy requirement by the feed pump per ml of permeate collection across unit area (m^2) of the membrane was high with the RFMM compared to TFMM, when both the modules are equipped with PSf membrane. Especially, with RFMM the energy goes upto around 12 fold increase compared to TFMM at 0.4 MPa TMP and initial feed concentration of 10 kg m^{-3}. On the contrary, with PES membrane, energy requirement for RFMM is almost same as that of TFMM to collect 1 ml of permeate flux across unit area (m^2) of the membrane at all TMPs and initial feed concentration. According to the experimental result, consumption of energy is minimum at 10 kg m^{-3} of initial feed concentration and 0.2 MPa (Fig. 16), while the shear rate is maximum at this condition. Table 2 shows a comparative figure on the energy consumption per ml of permeate collection across unit area (m^2) of the membrane for different modules studied earlier, which helps to provide the economy of the proposed TFMM equipped irrespective of the membranes.

Hence, after inspecting the pros and cons of the two membrane modules with the existing one, along with the membrane fitted, it is realised that TFMM equipped with PSf membrane provides an excellent opportunity to separate out PEG 6000 at the expense of low energy with utmost shear rate to alleviate the polarisation phenomenon over the membrane.

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