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Clarification and concentration of sugar cane juice through ultra, nano and reverse osmosis membranes

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Abstract. The performance of ultrafiltration (UF) membranes with molecular weight cut off (MWCO) of 1000 and 3500 Da in clarifying sugar cane juice was investigated, as well as the performance of a nanofiltration (NF) membrane with MWCO of 200 Da and a reverse osmosis (RO) membrane in concentrating sugar cane juice. For both cases the sugar cane juice had been limed and partially clarified. The UF membranes were found to be effective at clarifying the sugar cane juice in terms of purity rise and reduction in turbidity, colour, starch and protein. A purity rise of approximately 6 was achieved by both UF membranes at trans-membrane pressures (TMP) from 15 to 25 bar. However, Brix reduction in the permeate was between 14.5 and 41.85% and 12.11 and 26.52% for 1000 Da and 3500 Da membranes respectively. For the 200 Da and RO membranes the Brix in the concentrate was increased from 7.65 to 12.3 after 3 hours of operation for the 200 Da membrane at a TMP of 10 bar, whilst the Brix in the concentrate was increased from 15.65 to 27.6 after 3 hours of operation for the RO membrane at a TMP of 35 bar. Overall, UF membranes were found to be unsuitable for clarification of sugar cane juice since significant amount of Brix is reduced in the permeate, whilst RO membranes were found to be effective for concentration of sugar cane juice.

Keywords: concentration; clarification; sugar cane juice; ultrafiltration; nanofiltration; reverse osmosis

1. Introduction

The sugar industry worldwide continues to use conventional treatment processes for the production of sugar from sugar cane. In general the process consists of the crushing of sugar cane to produce raw juice, liming of the raw juice, clarification, concentration and crystallisation (Bhattacharya *et al*. 2001). Problems associated with this conventional process arise mainly in the clarification and concentration stages. Conventional clarification techniques result in the inefficient removal of substances which adversely affect the colour of the final product (Bhattacharya *et al*. 2001). Whilst concentration by evaporation, which is operated by steam energy, requires an enormous amount of energy and is considered the most energy intensive operation in a sugar mill (Madaeni and Zereshki 2006, 2008). In addition to this, it has been suggested that heating of the sugar juice during evaporation could change the colour and flavour, thus lowering the product quality (Madaeni and Zereshki 2008). In response to these problems experienced by the conventional treatment process, many investigations into the use of membrane technology for clarification and concentration of

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sugar juice have been conducted (Bhattacharya *et al*. 2001, Madaeni and Zereshki 2006, Ghosh and Balakrishnan 2003, Ghosh *et al*. 2000, Hamachi *et al*. 2003, Jegatheesan *et al*. 2009, Sim *et al*. 2009). From these investigations it has been concluded that microfiltration (MF), ultrafiltration (UF), nanofiltration (NF) and reverse osmosis (RO) have the potential to be incorporated into the sugar treatment process. UF in particular has been found to be effective at juice clarification, producing permeate with lower colour, lower viscosity and higher clarity compared to the clarified juice obtained by conventional means (Ghosh and Balakrishnan 2003, Ghosh *et al*. 2000, Hamachi *et al*. 2003). In one study the average purity rise achieved by UF was greater than 1.5 units when clarified juice was used as the feed and greater than 2 units when raw juice was used as the feed. This was found to be a significant improvement upon the unit purity rise obtained by conventional clarification of 0.5 – 1 (Ghosh *et al*. 2000). Furthermore, a turbidity reduction of 99.7%, colour reduction of 15%, reduction in starch concentration by 80% and increased purity by 1.4% were achieved in a previous study that used NF to treat limed and partially clarified sugar cane juice (Sim *et al*. 2009). In contrast to this, RO has been investigated as a means of concentrating sugar juice.

A two-stage RO system was used to pre-concentrate clarified sugar juice before being passed to conventional evaporators for further concentration in order to reduce the overall energy costs of the system. Calculations showed that a 33% energy saving was achieved by pre-concentrating the clarified sugar juice from 15 to 20° brix prior to final concentration in evaporators. Sugar loss caused by sugar flux across the RO membrane was also found to be negligible (Madaeni and Zereshki 2008). Whilst this study shows that the concentration of sugar juice using RO membranes has the potential to reduce energy costs in the concentration process, there is insufficient evidence to make any conclusions regarding the implementation of RO on an industrial scale. As such, the aims of this study are to (i) evaluate the clarification of sugar cane juice using different UF membranes by investigating the flux, purity rise and reduction in turbidity, colour, starch, protein and Brix concentration. (ii) evaluate the concentration of sugar cane juice using NF and RO membranes by investigating the flux, increase in Brix concentration and ionic rejection.

2. Materials and methods

Four different membranes were used to filter sugar cane juice that had been lime treated and clarified using a 0.05 μm ceramic membrane. Details of the 0.05 μm ceramic membrane set-up can be found elsewhere (Jegatheesan *et al*. 2009). The four different membranes consisted of spiral wound 1000 Da and 3500 Da molecular weight cut-off (MWCO) (GK1812) UF membranes, a spiral wound thin film 200 Da MWCO (DK1812) NF membrane and a spiral wound polyamide RO (SG-1812) membrane. All membranes were purchased from General Electric Company. The effective area of each of the membranes was 0.46 m^2. The operating conditions of each experimental run are given in Table 1, where a flow rate of 8 LPM and temperature of 35°C was used for each run. A schematic diagram of the experimental setup is shown in Fig. 1. The membrane unit consisted of membrane machine (RNF-0460), supplied by Jiangsu Jiuwu Hitech Company Ltd., which contained a 1.5 kW plunger pump with a 380 W, 50 Hz power supply and a 15 L stainless steel feed tank with a built in heat exchange jacket. A B.Braun Frigomix 1495 heater/cooler with a Thermomix 1442D temperature controller was used to maintain the temperature of the feed solution. An OHAUS electronic mass balance connected to a computer was also used in order to measure the mass of permeate at one minute intervals.

Table 1 Operating conditions of the experimental runs

Membrane	Initial feed pH	Initial feed brix (°Brix)	Feed volume (L)	TMP (bar)	No. of runs	Operating time (hr)
3,500 Da	7.5	16	10.5	1.5, 3, 5	2 per TMP	3
	7.5	16	14	15, 20, 25	1 per TMP	2
	5.7	16	14	15, 20, 25	1 per TMP	2
1,000 Da	7.5	16	10.5	1.5, 3, 5	2 per TMP	3
	7.5	16	14	15, 20, 25	1 per TMP	2 – 3
	5.7	16	14	15, 20, 25	1 per TMP	2 – 3
200 Da	7 – 7.5	7.5	10.5	7, 10	2 per TMP	3
RO	7 – 7.5	7.5	10.5	25, 35	2 per TMP	3

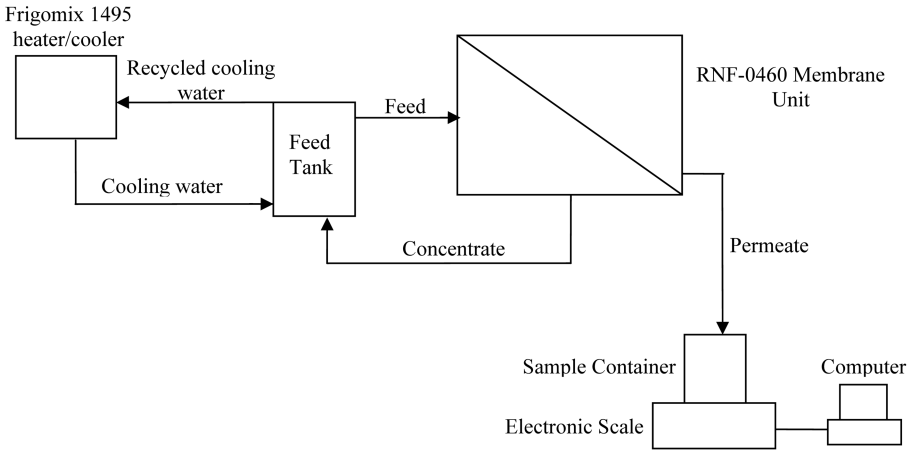

Fig. 1 Schematic diagram of the experimental set-up

Sugar cane juice obtained from Mossman Central Mill Company Ltd. (Queensland) was used for the UF experiments. For the 200 Da and RO experiments, sugar cane variety Q200 was collected from Paluma (Queensland, Australia) and stored inside a cold room at a temperature of 10°C at the Mechanical Engineering workshop at James Cook University. The raw sugar cane was crushed using a sugar cane miller and the raw juice was filtered through a 250 μm sieve to remove large fibers. Both the milled sugar cane juice and the juice obtained from Mossman Central Mill Company Ltd. were treated with $Ca(OH)_2$ and the pH was adjusted to either 5.7 or 7.5 using hydrochloric acid or sodium hydroxide, depending on the pH requirements of the subsequent experimental run. The juice was then allowed to settle for 1 hour and the supernatant juice was siphoned and filtered through a 125 μm sieve and diluted with DI water to adjust the sucrose content to around 16° Brix. The diluted solution was then used as a feed for the 0.05 μm membrane unit, which was used to produce adequate amounts of clarified juice for use in the experiments as outlined in Table 1. For the 200 Da and RO membrane runs the clarified juice was diluted further using DI water. Since higher Brix would foul 200 Da and RO membranes quicker, lower initial Brix concentration was used in those applications.

The appropriate membrane was installed into the membrane unit and was cleaned using DI water

for a period of 1 hour. The clarified juice from the 0.5 μm membrane unit was then placed in the feed tank and was fed through the membrane using the operating conditions specified for each experimental run. Samples of 400 mL of the feed and permeate solution were collected at intervals of 0, 1.5 and 3 hours for the 3 hour experimental runs and 0, 1 and 2 hours for the 2 hour experimental runs for analyses. The sample volume of the permeate was recorded and added to the total permeate volume calculated from the weight data for the flux computation. After each experimental run the membrane was rinsed with deionised water and cleaned for 40 minutes using the cleaning solution supplied by the membrane unit manufacturer, Jiangsu Jiuwu Hitech Company Ltd.

3. Analytical methods

Brix is a measurement of the refractometric dry substances (RDS) in a solution and is the mass ratio of dissolved sugar to water. The Brix of the feed, retentate and permeate were measured by a digital refractometer (Palette PR-101, Atago). The total polarized substances in the juice, Pol, are used to represent the sucrose content of the juice. Pol was measured using the dry lead method (Laboratory Manual for Queensland Sugar mills 1970). Two grams of subacetate of lead per 100 mL of solution was added to 50 mL of feed solution juice and mixed thoroughly. The solution was left for several minutes for the precipitate to settle and the supernatant was then filtered through Whatman filter paper. A polarimeter (SQF_WXG4, Vanco) calibrated in sugar degree (^{o}Z) was used to measure the Pol reading of the filtered supernatant. A 200 mm length pol tube was filled with the filtered supernatant and the polarity was measured at a temperature of 20^{o}C. Pol percent was calculated using the following equation

$$Pol\ percent\ juice = \frac{Pol\ reading}{Pol\ factor} \tag{1}$$

Where

$$Pol\ factor = \frac{100 \times apparent\ density\ at\ 20°C}{26,000} \tag{2}$$

The purity and purity rise were calculated using the following equations

$$Purity = \left(\frac{Pol}{Brix}\right) \times 100 \tag{3}$$

$$Purity\ rise = (Purity)_{permeate} - (Purity)_{feed} \tag{4}$$

The turbidity was measured according to GS7-21 method (International Commission for Uniform Methods of Sugar Analysis (ICUMSA 1994)) using a spectrophotometer (HP-8453) at a wavelength of 900 nm. The following equation was used to calculate turbidity

$$Turbidity = \left(100\frac{A_S}{b}\right) \tag{5}$$

The colour of the juice was measured according to GS1-7 method (ICUMSA 1994). The sample was adjusted to a pH of 7 using 0.1N HCl and 0.1N NaOH and was filtered through a 0.45 μm

filter in order to measure the absorbance at a wavelength of 420 nm using a spectrophotometer (HP-8453). The following equation for colour was used

$$Colour(IU) = \frac{10^8 \times A_S}{b \times RDS \times \rho} \tag{6}$$

Starch concentration was measured using the following method (Chavan and Jadhav 1992): 1.2 mL of 2N acetic acid, 0.25 mL of 10% KI and 2.5 mL of 0.001667 M KIO_3 solution were added to 1 mL of sample and mixed thoroughly. The optical density of this solution was measured at a wavelength of 600 nm using a spectrophotometer (HP-8453) against a blank containing the reagents and 1 mL of water. Finally, the pH of the solutions was measured by a digital pH meter (AQUA –pH, TPS).

4. Results and discussion

4.1 Flux

Tables 2 and 3 detail the mean hourly flux and overall mean flux for each membrane for pH 7.5 and 5.7 respectively, whilst Fig. 2 shows a graph of overall mean flux at each TMP for each membrane for ease of comparison. The flux achieved by the 3500 Da membrane for pH 7.5 increased for TMP of 1.5 bar up to 15 bar. The flux then experienced a decrease at a TMP of 20 bar and an increase again at 25 bar. This behaviour indicates the dominance of fouling effects over the effect of increasing the TMP at a TMP of 20 bar. This same trend is evident for experiments conducted at pH 5.7. In contrast to this, the 1000 Da membrane experienced an increase in flux as TMP was increased for both pH 7.5 and 5.7. The 1000 Da membrane not experiencing a decrease

Table 2 Mean hourly and overall flux for each membrane at different TMP, temperature of 35°C and pH 7.5

Type of membrane module	TMP (bar)	Flux 1st hour ($L/m^2.hr$)	Flux 2nd hour ($L/m^2.hr$)	Flux 3rd hour ($L/m^2.hr$)	Overall mean flux ($L/m^2.hr$)
3500 Dalton	1.5	3.32	3.25	3.23	3.27
	3	5.33	4.64	4.72	4.90
	5	7.85	7.79	7.42	7.68
	15	12.17	9.85	-	11.01
	20	9.33	8.10	-	8.71
	25	10.19	8.45	-	9.32
1000 Dalton	1.5	1.30	1.26	1.28	1.28
	3	2.18	2.09	2.06	2.11
	5	3.65	3.39	3.41	3.48
	15	7.61	7.19	6.05	6.95
	20	10.93	8.70	6.66	8.76
	25	12.61	10.11	-	11.36
200 Dalton	7	1.45	0.77	0.17	0.80
	10	5.94	1.99	0.46	2.79
RO	25	3.52	1.98	0.93	2.15
	35	5.22	1.91	0.60	2.57

Table 3 Mean hourly and overall flux for each membrane at different TMP, temperature of 35°C and pH 5.7

Type of membrane module	TMP (bars)	Flux 1st hour ($L/m^2.hr$)	Flux 2nd hour ($L/m^2.hr$)	Flux 3rd hour ($L/m^2.hr$)	Overall mean flux ($L/m^2.hr$)
3500 Dalton	15	15.98	10.27	-	13.13
	20	15.03	7.23	-	11.13
	25	15.21	10.08	-	12.65
1000 Dalton	15	9.58	7.66	6.44	7.89
	20	11.90	9.03	6.57	9.17
	25	14.16	9.63	-	11.89

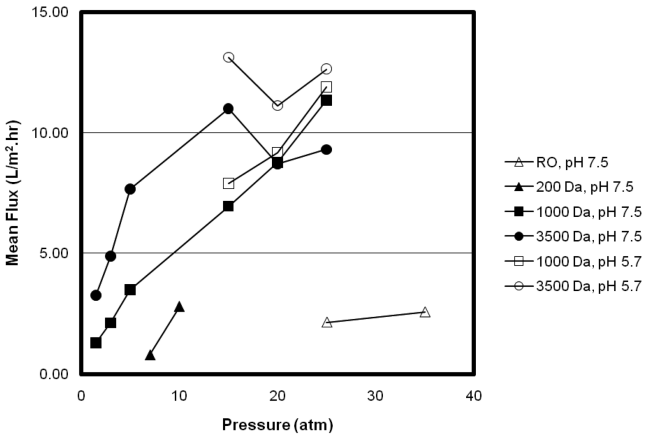

Fig. 2 Graph of overall mean flux for each membrane at different operating pressures and pH

in flux at higher pressures when compared to the 3500 Da membrane could be due to particles blocking the internal pores of the 3500 Da membrane yet being too small to block the internal pores of the 1000 Da membrane. In order to confirm this as a possible reason for the difference in flux behaviour between the 1000 Da and 3500 Da membranes, fouling models were fitted to the experimental data as explained in section 4.2. Furthermore, both the 3500 Da and 1000 Da membranes achieved higher flux for pH 5.7 compared to pH 7.5. The pH of 5.7 may have caused the membrane surface to become positively charged thus attracting the slightly negatively charged water molecules and resulting in an increase in flux compared to a pH of 7.5. This effect was more significant for the 3500 Da membrane compared to the 1000 Da membrane. The 200 Da and RO membrane both yielded an increase in flux as TMP was increased.

4.2 Fouling models

Fouling models were fitted to the experimental data for both 1000 Da and 3500 Da membranes in order to explain the flux behaviour as discussed earlier. The models consisted of the cake filtration model, pore narrowing (progressive internal fouling) model, combination of external and progressive internal fouling model and the complete pore blocking model. The equations for each of these models are shown below with term descriptions given in the nomenclature. Further details on each

of these models can be obtained in Jegatheesan *et al*. (2009).

Cake filtration model

$$\frac{t}{V_f} = \frac{1}{Q_0} + \frac{\alpha C_w V_f}{2A_0 R_{m0} Q_0} \tag{7}$$

Pore narrowing model

$$\frac{1}{\sqrt{Q_f}} = \frac{1}{\sqrt{Q_0}} + \frac{C\sqrt{Q_0}}{V_p} t \tag{8}$$

Combination of external and progressive internal fouling model

$$\frac{1}{Q_f} = \left[\frac{\mu_f}{P_{tm} A_0}\right]\left(\frac{\alpha C_w}{A_0} + \frac{2C}{V_p}\right) V_f + \left[\frac{\mu_f R_{m0}}{P_{tm} A_0}\right] \tag{9}$$

Complete pore blocking model

$$\ln[Q_f] = \ln[Q_0] - \sigma J_0 t \tag{10}$$

Figs. 3 and 4 show the fitting to the cake filtration model, whilst Figs. 5 and 6 show the fitting to the combination of external and progressive internal fouling model of the experimental data for 1000 Da and 3500 Da membranes at pH 7.5. The 1000 Da membrane fits best to the cake filtration model for TMP of 15, 20 and 25 bar, with R^2 values of 0.9505, 0.964 and 0.9504 respectively. These can be compared to the R^2 values of 0.8413, 0.87 and 0.9252 for the combination of external and progressive internal fouling model for TMP of 15, 20 and 25 bar respectively. The 3500 Da membrane fits best to both the cake filtration model and combination of external and progressive internal fouling model at a TMP of 15, with R^2 values of 0.998 and 0.9888 respectively. Similarly for a TMP of 20 bar, the 3500 Da membrane fits best to both the cake filtration model and the combination of external and progressive internal fouling model with R^2 values of 0.9415 and 0.9404 respectively. Finally, for a TMP of 25 bar, the 3500 Da membrane fits best to the cake filtration

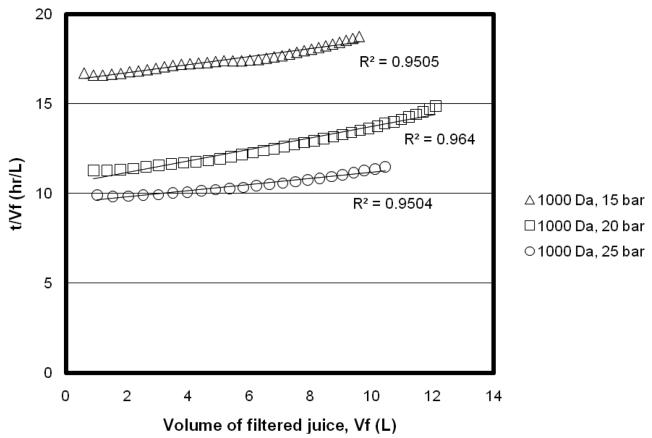

Fig. 3 Graph of cake filtration model fitting to the experimental data for 1000 Da membrane at temperature of 35°C and pH 7.5

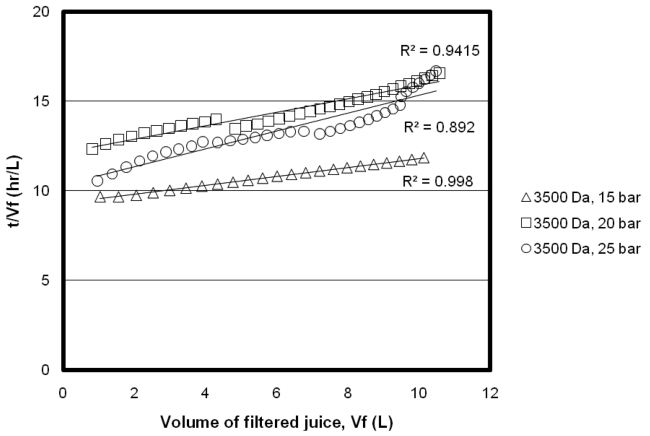

Fig. 4 Graph of cake filtration model fitting to the experimental data for 3500 Da membrane at temperature of 35°C and pH 7.5

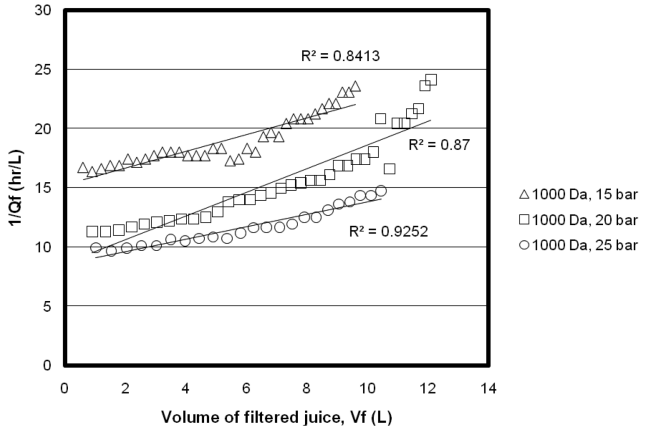

Fig. 5 Graph of combination of external and progressive internal fouling model fitting to the experimental data for 1000 Da membrane at temperature of 35°C and pH 7.5

model with an R^2 value of 0.892 which can be compared to R^2 values of 0.698, 0.828 and 0.841 when compared to the combination of external and progressive internal fouling, pore narrowing and complete pore blocking models.

Based on these models and the flux results obtained by the 3500 Da membrane, it is likely that the membrane experienced cake fouling up to a TMP of 15 bar, followed by a transition into a combination of external and progressive internal fouling at a TMP of 20 bar caused by particles blocking the internal pores of the membrane, thus resulting in a decrease in flux from a TMP of 15 bar to 20 bar. Finally, further fouling at a TMP of 25 bar was then governed solely by cake fouling allowing for a slight increase in flux from a TMP of 20 to 25 bar, yet still less than the flux obtained at a TMP of 15 bar. The above explanation has been given based on the model fitting and it is reasonable to expect that the internal pore blocking could occur when the TMP increases up to a certain value which in this case is 20 bar and further increase in TMP does not contribute to pore

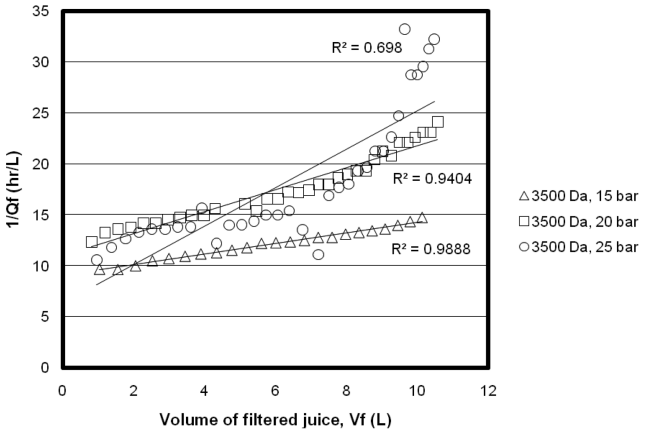

Fig. 6 Graph of combination of external and progressive internal fouling model fitting to the experimental data for 3500 Da membrane at temperature of 35°C and pH 7.5

blocking. In contrast to this, the 1000 Da membrane experienced cake fouling which can most likely be attributed to the smaller pore size of the membrane, preventing internal fouling from occurring and resulting in a continuous increase in flux as TMP was increased.

4.3 Clarified sugar cane juice quality

Tables 4 and 5 shows the mean purity rise and mean turbidity, colour, starch, protein and Brix reduction for the 1000 Da and 3500 Da membranes at pH 7.5 and 5.7 respectively. From these tables it can be seen that the 1000 Da membrane generally achieves a greater purity rise for a given

Table 4 Mean purity rise and turbidity, color, starch, protein & °Brix reduction for 1000 and 3500 Da membranes at different TMP, temperature of 35°C and pH 7.5

Type of membrane module	TMP (bars)	Mean purity rise	Mean turbidity reduction (%)	Mean color reduction (%)	Mean starch reduction (%)	Mean protein reduction (%)	Mean °Brix reduction (%)
1000 Dalton	1.5	1.38	87.90	67.60	50.53	49.50	14.48
	3	3.50	87.65	70.80	53.55	49.85	14.90
	5	3.75	92.68	69.05	58.30	52.10	16.31
	15	6.30	99.95	79.60	66.15	76.20	32.15
	20	6.80	89.30	85.25	59.20	77.85	37.97
	25	6.40	92.40	83.40	36.45	75.15	41.85
3500 Dalton	1.5	2.90	85.00	53.80	40.33	53.63	12.11
	3	4.68	86.38	69.45	52.40	53.73	16.49
	5	4.45	90.33	67.98	51.53	54.10	18.52
	15	6.35	96.90	75.25	19.15	66.85	19.87
	20	6.00	99.30	80.60	43.45	68.65	23.66
	25	5.30	87.10	66.60	39.85	63.65	26.52

Table 5 Mean purity rise and turbidity, colour, starch, protein & °Brix reduction for 1000 and 3500 Da membranes at different TMP, temperature of 35°C and pH 5.7

Type of membrane module	TMP (bars)	Mean purity rise	Mean turbidity reduction (%)	Mean color reduction (%)	Mean starch reduction (%)	Mean protein reduction (%)	Mean °Brix reduction (%)
1000 Dalton	15	5.40	100.00	85.05	26.95	78.50	30.76
	20	5.15	91.00	79.35	37.80	77.30	31.54
	25	6.80	100.00	86.05	40.70	79.45	28.77
3500 Dalton	15	5.10	100.00	81.45	51.05	71.70	17.98
	20	4.95	99.95	86.80	63.15	74.25	20.02
	25	3.45	96.40	85.80	37.55	97.05	24.82

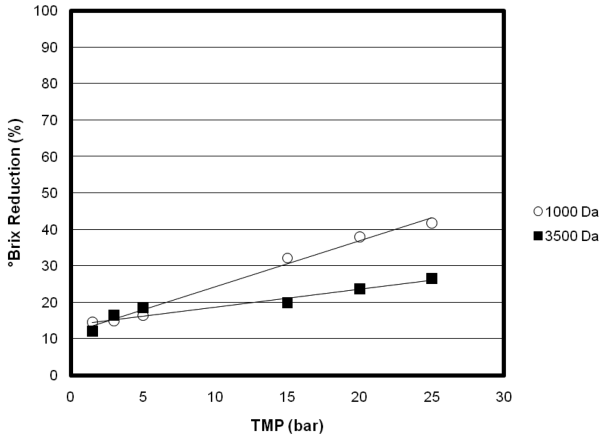

Fig. 7 Graph of mean °Brix reduction at different TMP for 1000 and 3500 Da membranes at pH 7.5

pressure when compared to the 3500 Da membrane. Similarly, a greater reduction of turbidity, colour, starch, protein and Brix was generally achieved by the 1000 Da membrane compared to the 3500 Da membrane. The increase in purity rise and greater reduction in the other parameters can be attributed to the smaller pore size of the 1000 Da membrane. Furthermore, as the TMP was increased, greater values for all of the parameters were generally achieved for both the 1000 Da and 3500 Da membrane. The Brix reduction for each membrane at the different TMP's is shown in Fig. 7 and an increasing linear trend can be seen between TMP and Brix reduction. A Brix reduction of between 14.5 – 41.85% and 12.11 – 26.52% was achieved by the 1000 Da and 3500 Da membrane respectively at a pH of 7.5.

4.4 Concentrated sugar cane juice quality

Fig. 8 shows the mean Brix in the concentrate for the 200 Da and RO membranes at different operating times. The graph shows that the RO membrane achieved higher Brix in the concentrate compared to the 200 Da membrane. The Brix in the concentrate for both membranes increased as the TMP was increased. The 200 Da membrane achieved a maximum value for Brix in the concentrate

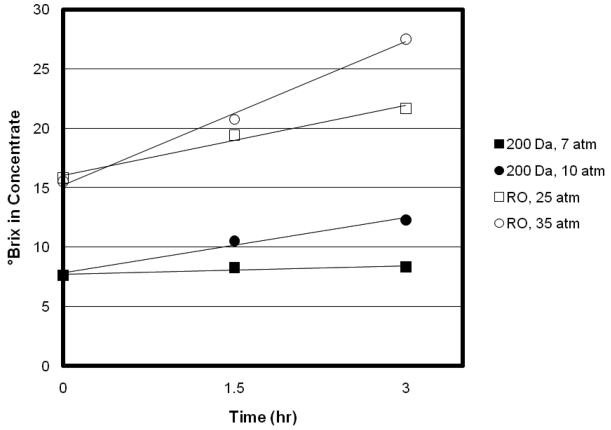

Fig. 8 Graph of mean °Brix in concentrate for 200 Da and RO membranes at various operating times and pressures

at a pressure of 10 bar with Brix values of 7.65, 10.5 and 12.3 at 0, 1.5 and 3 hours respectively. The RO membrane achieved a maximum value for Brix in the concentrate at a pressure of 35 bar with Brix values of 15.65, 20.7 and 27.6 at 0, 1.5 and 3 hours respectively. Whilst the 200 Da membrane achieved a similar percentage increase in Brix to the RO membrane, its application in the sugar industry is limited as it can only operate effectively at lower initial Brix concentrations. The results for the RO membrane are similar to those obtained by another study (Madaeni and Zereshki 2008) which used a two-stage RO system to pre-concentrate sugar-cane juice, before conventional concentration, from 15 Brix to 20 Brix, resulting in an estimated 33% energy saving compared to a conventional concentration process. Table 6 details the mean ion concentrations and rejection achieved by the 200 Da and RO membranes. The 200 Da membrane achieved rejections

Table 6 Mean ion concentrations and rejection achieved by 200 Da and RO membranes at a temperature of 35°C and pH 7.5

Sample	Sample time (hours)	Mean ion concentration (mg/L)							
		Na^{+}	K^{+}	Ca^{2+}	Mg^{2+}	$Fe^{(3,2+)}$	Al^{3+}	Si^{4+}	P
$200Da_{feed}$ (Adjusted to 7.65 °Brix initially)	**0**	5.48	584	118	47.2	0.1	0.417	9.55	7.14
	3	6.96	742	189	81.7	0.741	1.11	12.4	11.4
$200Da_{permeate}$	**0**	3.09	305	11.7	2.23	0.1	0.2	6.96	0.1
	1.5	4.09	417	20.1	3.82	0.1	0.2	8.43	0.1
	3	4.88	489	33.3	7.96	0.1	0.2	7.49	0.1
Rejection at 3Hrs (%)		**29.89**	**34.10**	**82.38**	**90.26**	**86.50**	**90.99**	**39.60**	**99.12**
RO_{feed} (Adjusted to 15.65 °Brix initially)	**0**	14.3	1210	267	118	13	1.4	26.6	19.5
	3	22.3	1800	392	153	18.3	2.8	44.7	32.1
$RO_{permeate}$	**0**	4.16	84.6	1.22	0.532	0.1	0.2	0.352	0.1
	1.5	1.75	134	2.4	1.06	0.1	0.2	0.531	0.1
	3	2.05	164	4.42	1.71	0.1	0.2	1.02	0.1
Rejection at 3Hrs (%)		**90.81**	**90.89**	**98.87**	**98.88**	**99.45**	**96.43**	**97.72**	**99.69**

greater than of 82 – 99% for divalent ions whilst only achieving rejections of 29.9 – 34% for monovalent ions. In contrast to this, RO achieved rejections of approximately 90% for monovalent ions and 96 – 99.7% for divalent ions.

5. Conclusions

The performance of 3500 Da and 1000 Da UF membranes for the clarification of sugar cane juice was evaluated by investigating the flux, purity rise and reduction in turbidity, colour, starch, protein and Brix concentration of the permeate under different operating conditions. The performance of 200 Da NF and RO membranes for the concentration of sugar cane juice was evaluated by investigating the flux, increase in Brix in the concentrate and ionic rejections under different operating conditions.

The 200 Da and RO membranes were found to be effective at concentrating the sugar cane juice with respect to the measured parameters. Whilst the flux was only approximately 2.6 and 2.8 L/m^2.hr for the RO and 200 Da membrane at operating pressures of 35 bar and 10 bar respectively, the overall flux could be increased on an industrial scale by operating multiple membrane elements in series and parallel, as is the case in seawater desalination. The Brix in the concentrate was increased from 7.65 to 10.5 and 12.3 after 1.5 and 3 hours of operation respectively for the 200 Da membrane at a TMP of 10 bar, whilst the Brix in the concentrate was increased from 15.65 to 20.5 and 27.6 after 1.5 and 3 hours of operation respectively for the RO membrane at a TMP of 35 bar.

The rejection of ions was also found to be high for both the 200 Da and RO membrane, which indicates that sucrose would be retained in the concentrate, resulting in concentrated sugar cane juice with negligible sucrose loss and a high purity water permeate. Finally, RO membranes have the potential to reduce energy costs in the concentration process and further studies using RO membranes in continuous flow operation rather than batch flow operation are recommended in order to determine the Brix concentration that can be obtained using this flow regime, as continuous flow would be required for industrial application.

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CC

Nomenclature

α	Specific cake resistance per unit mass
ρ	Density of solution
σ	Plugging potential of the feed suspension
A_0	Total membrane area
A_S	Spectrophotometer Absorbance at a given wavelength
b	Spectrophotometer cell length
C	Fraction of solute adsorbed by membrane
C_w	Rejected particle concentration near the membrane surface
J_0	Initial permeate flux
Q_0	Initial flow rate of the permeate
Q_f	Filtration flow rate through membrane
RDS	Refractometric dry substances
R_{m0}	Intrinsic membrane resistance
t	Operation time of membrane system
TMP	Trans-membrane pressure
V_f	Volume of filtered juice
V_p	Total initial pore volume