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Groundwater treatment by ceramic membranes: pilot tests at the Commonwealth Games Village in Delhi, India

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Abstract

Ceramic membranes are increasingly gaining popularity in the field of water treatment. High quality, better operability, robustness, ability to work in harsher environments and efficient removal of organic contaminants are just some of the characteristics which are making it an interesting alternative for water treatment facilities worldwide. This paper aims to evaluate the performance of an ultrafiltration (UF) ceramic membrane pilot plant installed in an 1 MGD (million gallons per day) plant in Delhi, to adapt the process and to optimize the operational features, so that it can be suggested as an economical and technical alternative to existing UF polymeric membrane installations in Indian municipal facilities. The investigation is conceptualized with the production of potable water after filtration at a volumetric flow rate of 1.2 m^3/h through an UF ceramic membrane having a pore size of 50 nm. The process parameters such as the duration of filtration, flow rate, flux, flocculant concentration, and chemicals were varied to check its adaptability and effectiveness for better performance. Laboratory tests, evaluation of the results and troubleshooting were used to regulate the quality aspects of the study.

Key words: ceramic membrane, drinking water treatment, flocculation, pressurized backwash, ultrafiltration

INTRODUCTION

Membrane filtration is rapidly becoming one of the most popular techniques of water treatment in India and is gaining reputation as a more competent technology than its predecessors. This treatment is based on selective filtration through pores of different sizes by classifying it into four main categories: microfiltration, ultrafiltration, nanofiltration and reverse osmosis. Their utilization is based on process parameters like operating pressure, permeability and flux rates.

Ceramic membranes are a state-of-the-art technology to enter the water treatment field. Originally used in niche applications, it faces competition with polymeric membranes because of its high initial capital investment. They are made from inorganic materials like alumina, zirconia oxides, titanium, recrystallized silicon carbide and other materials. Hashim *et al.* (2017) and Baker (2012) found ceramic membranes to work in harsher environments as compared to polymeric membranes and that they can be treated with efficient cleaning procedures having good flux stability. Its use of a pressurized backflush and other advantages made it imperative to be used in drinking water treatment. Recent developments have also reduced the capital involved in the construction of ceramics.

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Further studies by Kabsch-Korbutowicz & Urbanowska (2010), Hofs *et al.* (2011), Meyn *et al.* (2011) and Kenari & Barbeau (2016) helped in understanding the performance of ceramic membranes vs. their polymeric counterparts, and gave a broader understanding to its filtration capabilities.

Ceramic membranes perform well within a range of varying raw water quality and are effective as a barrier against microbes as described by Lerch *et al.* (2005), Kabsch-Korbutowicz & Urbanowska (2010) and Anderson *et al.* (2012). This was verified on the basis of pathogen removal with pretreatment steps of coagulation/flocculation steps prior to membrane filtration. The said process involves dosing of iron or aluminum salts to coagulate the formed flocs at certain conditions along with ensuring that the remnants of the flocs are not adsorbed on the membrane surface with the optimized backwashing procedure. Unlike conventional treatment processes with coagulation/sedimentation followed by rapid sand filtration, it is not necessary to form larger flocs because the microfiltration/ultrafiltration (MF/UF) will also retain flocs of smaller sizes as found by Anderson *et al.* (2012).

OBJECTIVES OF THE STUDY

Ceramic membrane filtration is a relatively new technology in the field of drinking water treatment in India, and hence, it was necessary to conduct trials before full-scale expansion. The pilot tests were started at the Commonwealth Games Village (CWGV) water treatment plant (WTP) with a capacity of 4,545 m^3/day (4.5 MLD, million liters per day) in New Delhi to gain more knowledge in the intricacies of the membrane.

Polymeric membranes are widely used for drinking water filtration and are the final treatment step with membrane filtration in the full-scale WTP in Delhi before disinfection. But polymeric membranes have high sensitivity towards cleaning chemicals, narrow limitations on operating pressure (transmembrane pressure, TMP), a delicate preserving procedure and low life time due to fiber breakage. Hence, one of the main objectives of piloting the ceramic membrane plant was to test it as an alternative to UF polymeric membranes.

The objectives of the study are to:

- test the ceramic membrane from different suppliers and integrate them in the process;
- determine and optimize operational parameters like flux rates, filtration times, chemical dosing rates, chemically enhanced backwash (CEB) sequences and cleaning-in-place (CIP) regimes;
- study the feasibility and the reduction of process steps;
- calculate capital expenditures (CAPEX) and operating expenditures (OPEX);
- techno-commercially compare with polymeric UF membranes.

Full-scale CWGV WTP

The raw water at the WTP, located in India's capital city – New Delhi, is treated in order to achieve the specified water quality conforming to WHO standards for potable water. The raw water is drawn from a Ranney well and three bore wells located in the flood plains of the river Yamuna. The conceptual approach of treatment includes bicarbonate hardness removal, suspended particles, colloids and harmful microorganisms as described by Lahnsteiner *et al.* (2012). The salient unit processes that are involved within the scope of treatment are aeration, lime softening, coagulation, flocculation, clarification, neutralization and pH adjustment of soft water in a recarbonization tank followed by UF polymeric membrane filtration with disinfection by ultraviolet light or chlorination as the final step. Sludge thickening and dewatering of thickened sludge is carried out for treating the generated sludge. A brief representation of the process scheme is shown in Figure 1.

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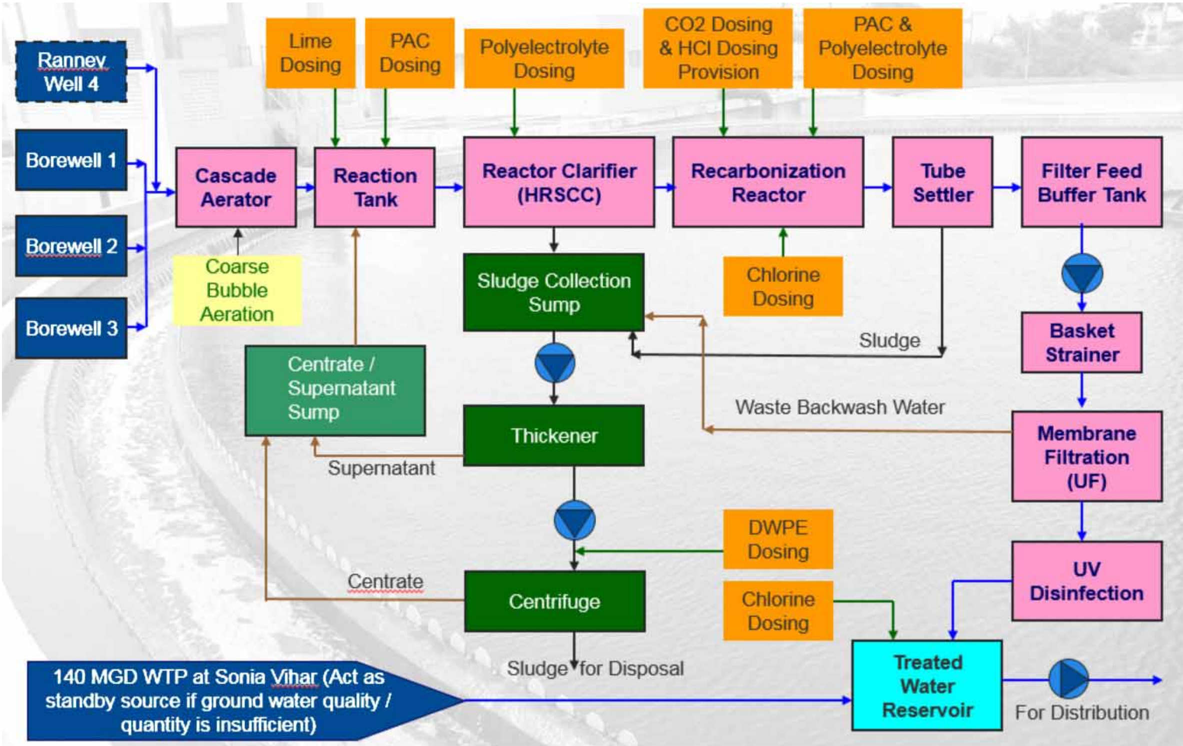

Figure 1 | Process scheme of the 4.5 MLD WTP.

MATERIALS AND METHODS

Water is softened in a high rate solid contact clarifier (HRSCC) in the WTP and then there is a need to stabilize the softened water and reduce its pH. For this reason, water is received from the HRSCC in a recarbonization reactor where the pH is reduced by dosing CO_2 gas and later flocculated to enable the settling process for drawing sludge in the WTP. In this study as shown in Figure 2, the feed pump

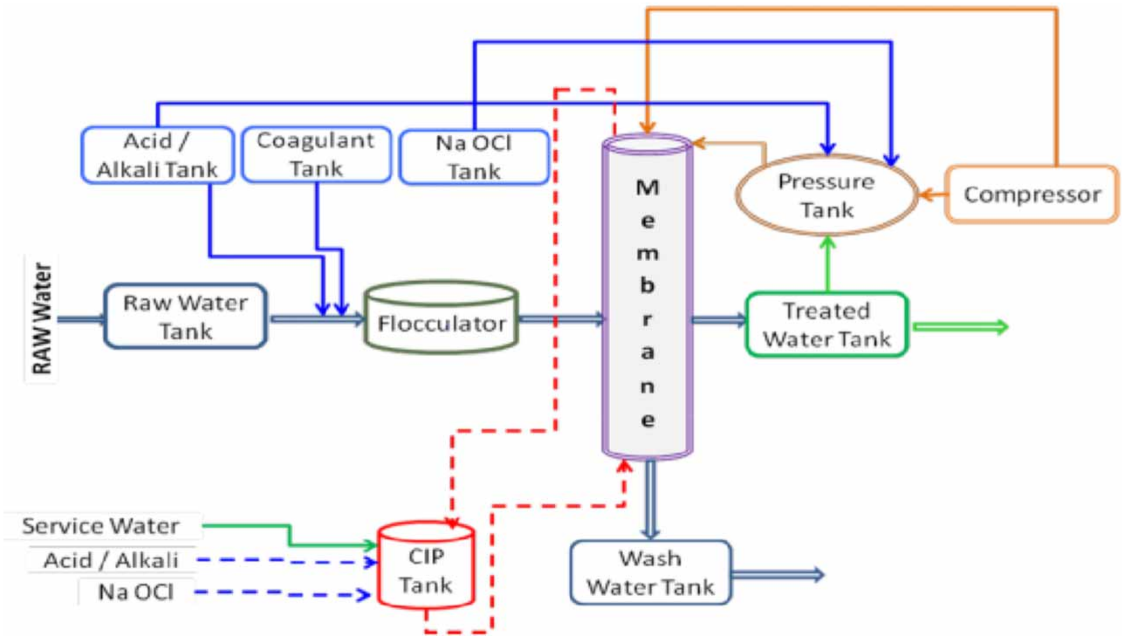

Figure 2 | Schematic flow chart of ceramic membrane process.

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draws raw water from the recarbonization reactor in the WTP and pushes it through the ceramic membrane in the pilot plant. For better separation of solids, flocculant is added to the raw water, which is mixed in-line in a static mixer. The pH adjustment can be carried out by the addition of HCl or NaOH solution to obtain optimal flocculation, if needed. In the dead-end process, flocculated water is filtered through a ceramic membrane (Figure 3) where permeate is directed to a storage tank. The blockages and depositions are removed during backwashing by the pulse-like application of permeate to the membrane surface via a pressure vessel and rinsed in several steps with compressed air with action of the permeate water being normally in the downwards direction.

The specifications of the installed membrane and operational details of the pilot plant are described in Table 1. The control system for the process is based on the supervisory control and data acquisition (SCADA) and programmable logic controllers (PLC) system. The flocculant and the cleaning chemicals are dosed according to control philosophy at pre-decided rates of dosing and concentration. Provision for compressed air is made for the air flushing and the initial pressure in the pressure tank. A sludge conditioning option has been provided to increase the yield; otherwise, the backwash sludge water is collected in a sludge tank and discharged. In pre-decided intervals, CEB and CIP were executed to remove blockages which cannot be cleaned with a normal backwash.

Figure 3 | Tubular-type ceramic membrane.

Table 1 | Ceramic membrane specifications and plant operation

Installed membrane surface area	8 m^2
Type of membrane	Ceramic (tubular type)
Membrane material	Al_2O_3
UF membrane	50 nm pore size
Flow rate	1.2 m^3/h
Current flux	150 LMH
Flocculation time	6 min (approx.)
Present TMP	Around 0.6–0.8 bar
$FeCl_3$ dosing	5 ppm (approx.)
Type of filtration	Dead-end filtration, inside out
Type of backflush	With 5 bar pressure (bottom to top mode)

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Total organic carbon analysis

Total organic carbon (TOC) is often used as an indicator of drinking water quality and membrane fouling. This test can often be used to estimate the amount of natural organic matter (NOM) and other synthetic compounds in drinking water as mentioned by Hendricks (2007). Hence, the permeate was analyzed using Hach Lange test kits on a regular basis. Furthermore, a dissolved organic carbon (DOC) test (liquid chromatography/organic carbon detection (LC/OCD) analysis) was conducted at a laboratory in Germany for DOC fractionation into humics, building blocks, biopolymers and inorganic colloids.

Overview of the trials

The trials were started in January 2016 where the plant was monitored on a continuous basis, and the critical parameters like TMP, permeability and flux were recorded regularly. Laboratory tests were conducted to monitor the turbidity and pH value at raw and permeate side. TOC tests were conducted for one grab sample per day. Daily observations and results were noted from the SCADA as well for comparison with the laboratory tests.

Jar tests

Calculations were done to check adequate dosage of $FeCl_3$ as well as for the chemicals used for the CEB procedure. For determining adequate dosage of the flocculant, volumetric tests were conducted along with jar tests. Primarily, the calculation of retention time was an important constituent to find out the adequate dosage to facilitate flocculation.

With the retention time being found to be around 6 min, jar tests were conducted to calculate the nominal dosing of the flocculant. The tests were done with three jars of 1 L capacity filled with the raw water dosed with different predefined concentrations of $FeCl_3$. The jars were stirred with a mechanical stirrer rapidly for 5 min and then mixed slowly for around 10 min. The dosage giving optimum flocculation was then used as an indicator to adjust $FeCl_3$ in the pilot plant.

RESULTS AND DISCUSSION

Transmembrane pressure

The trials were conducted initially at a flux rate of around 100 L/(m^2/h) where the process conditions were stable within a TMP of around 0.6 bar. To test the efficiency of the membrane, the plant was tested at higher flux rates to mimic full-scale conditions. Hence, the flux rate was increased to 150 L/(m^2/h) and filtration duration of 30 min with provision for one CEB in 1 day. When the filtration duration was decreased to 20 min and the number of backwashes increased, the TMP remained more or less constant. Increasing the number of CEBs with HCl gave inconsistent performance. It was observed that around 5 ppm of dosing gave optimum flocculation and this value was decided to be adopted for the pilot plant.

As found in Figure 4, from mid February, higher TMP values were found – from 0.95 to 1.9 bar. CIP treatment was performed to remediate this problem. CEB was normally performed during the start of the trials after adjusting backflush ratio to give two CEBs per day. The CEBs were initially being done with dosing of HCl at that point. Due to inconsistent TMP values, CEB regimes were done incorporating both NaOCl and HCl on alternate days from March 2016 onwards. Since NaOCl alone could not help achieve high pH values necessary to reduce organic fouling, a solution by mixing NaOCl and NaOH in the ratio of 35:15 was made from the end of March 2016 and used for CEB and CIP procedures. This gave satisfactory results and led to an increase of 'free residual chlorine' (FRC)

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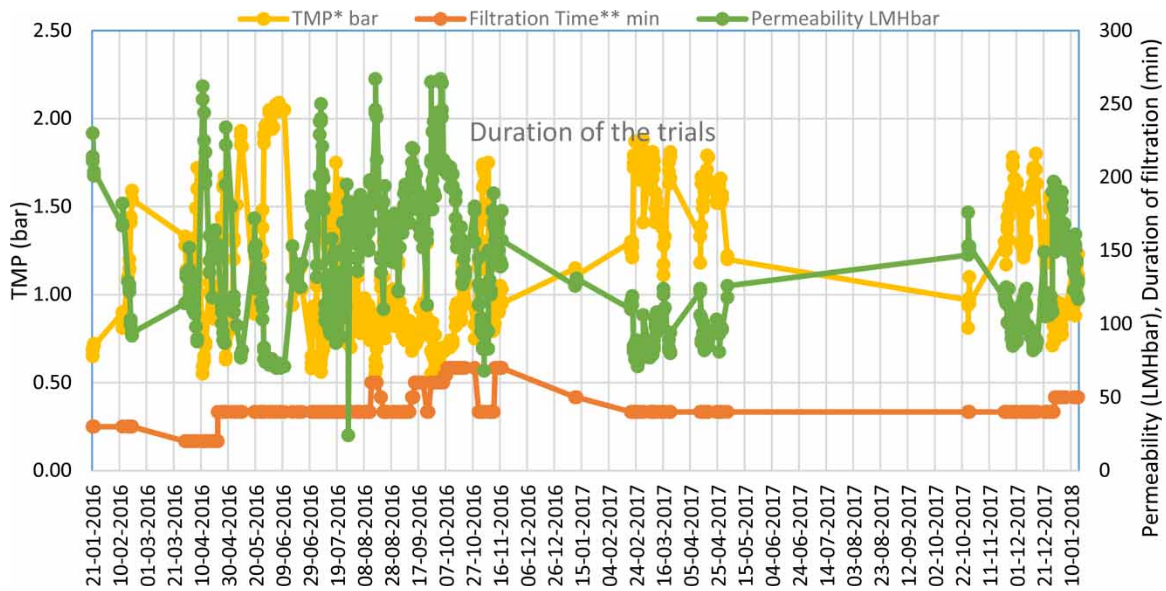

Figure 4 | TMP, permeability and filtration duration during pilot plant operation between 18.01.2016 and 15.01.2018.

levels to around 50 mg/l after dosing point of the pump for NaOCl was raised. Dilutions were carried out by mixing stock solution of ferric chloride and water in the ratio of 1:3 to reduce membrane choking by excessive deposition of flocs on the membrane surface.

CIP was carried out with only HCl at the initial duration of the trials and later with both HCl and NaOCl to clean out possible problems with organic fouling. The alkali–acid CIP regimes were done initially with sodium hypochlorite and caustic solution to remove organic foulants, and after an interval of at least five backwashes a CIP with HCl was carried out to treat inorganic scalants. Normally, a temperature of around 40 °C was maintained along with a pH of 0.5 for the acidic CIP and a temperature of 45 °C with pH of around 11.5 for the alkaline CIP.

The variation in TMP as seen in Figure 4 can be attributed to the stoppage times caused due to errors during the operation of the pilot plant. The operational value of TMP during stable operation was observed in the range of 0.6–1 bar. Since the flux rate was kept constant at 150 LMH, major deviations did not occur in the same and the values were more or less near 150 LMH. Further operation from July until October 2016 saw uninterrupted operation possible which provided a much deeper understanding of the performance of the membrane with respect to quality of the permeate, membrane permeability and transmembrane pressure. This led to testing of increased filtration times to obtain better yield rates for the plant along with improvement in the TMP.

Permeability

Lower permeability values indicate an incidence of membrane fouling. Stable operation led to improving permeability values as seen in Figure 2. Lower TMP, better CEB and backflush regimes gave values closer to the respective pure water flux for a 0.05 μm ceramic membrane module. Better permeability rates are a prime indicator of good membrane performance, and these are something to aim for when continuous and stable operation of the pilot plant is achieved.

Turbidity and TOC

The turbidity of water was relatively unaffected by the increase in TMP and issues with permeability of the membrane at the start of operation in Delhi as shown in Figure 5. This, however, changed with continued and improved operation of the plant. Introduction of new backwash regimes and better

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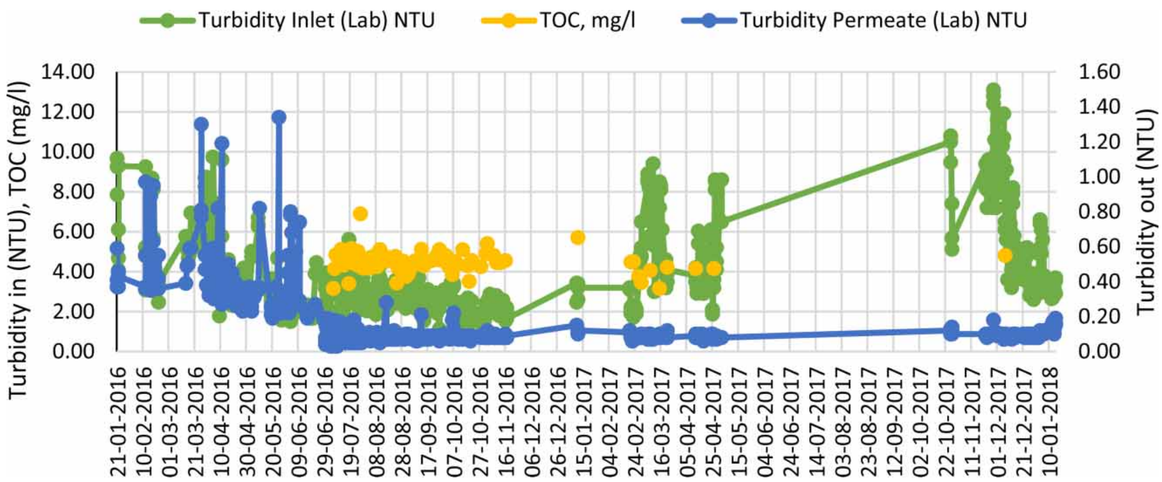

Figure 5 | Plot of turbidity in the raw water, turbidity in the permeate side and TOC over the operational time.

CEB with increased FRC rates led to the turbidity values of the permeate to remain in the range of 0.1 NTU as expected. TOC of the permeate samples was found to be in the range of 3–4 mg/l with maximum values obtained in the range of 5 mg/l.

Changing the mode of backwash led the turbidity to go below 0.1 NTU – the target value in the case of ceramic membranes as shown in Figure 5. The turbidity of the raw water from the main plant kept varying from 2.5 to 9.5 NTU. This had relatively less effect on the turbidity of the water obtained on the permeate side which shows that the ceramic membrane handled removal of flocs and other dissolved solids from the feed sample.

COMPARISON OF POLYMERIC AND CERAMIC MEMBRANES

One of the main objectives of the pilot plant trials was to compare the ceramic membrane pilot plant and the full-scale unit and showcase the advantages that ceramics have over UF polymeric membranes. Some points are as follows:

1. Recovery rate/yield. The polymeric membrane plant was designed for a recovery of 92%. At current operating conditions, the pilot plant showed a yield of 97% giving it an edge as far as yield is concerned. With the ‘sludge recovery’ feature in the pilot plant, a recovery rate of 99% can also be achieved.
2. Backwash and CEB procedures. CEB conducted in the main plant with normal backflow is incomparable to the pressurized backwash facility, automatic CEB procedures and CIP provision in the pilot plant.
3. Flux rates (inter-related to flux). The ceramic membrane was operated at flux of around 150 L/(m^2/h), while the 1 MGD plant was operated at less than one-third of these flux rates. Increasing of flux in the polymeric plant can lead to complete stoppage of the filtration process since it is not designed for higher flux rates.
4. Turbidity at the permeate side. The pilot plant was successful in obtaining a turbidity value lower than 0.1 NTU mostly independent of the TMP or permeability conditions.
5. Chemical consumption during CEB. Consumption of chemicals during CEB is more or less comparable for both. Further calculations have showed that ceramic membranes utilized less chemicals because of their better regeneration capabilities contributing to lower operational costs and lesser maintenance charges.
6. Fouling characteristics. Comparison studies by Hofs *et al.* (2011) in terms of organic fouling also showed that ceramic membranes have a better capacity of handling organic fouling and better regeneration of the membrane than polymeric membranes.

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Table 2 | Techno-commercial comparison for 4.5 MLD plant

Description	Polymeric membranes	Ceramic membranes
UF feed flow rate (L/h)	205,200	
Flux (approx.) in L/(m^2/h)	53.4	150
Net area of filtration (calculated from gross area and recovery rate for ceramics) in m^2	3,840	1,410
CAPEX		
Estimated cost/m^2 (INR)	1,100	9,000
Initial CAPEX of membranes (INR)	4,224,000	12,692,784
Additional CAPEX for tube settler (INR)	1,931,079	0
OPEX		
Membrane replacing cost in INR (considering polymeric replacement every 8th year)	12,672,000	0
Maintenance and manpower cost in INR (assuming 10% membrane replacement cost)	1,267,200	0
25-year cost for CEB chemicals (INR)	12,775,000	12,372,500
25-year estimated power cost (INR)	33,443,000	35,115,150
Calculated recovery rate (%)	92.4	97.0
Indicative water savings for 25 years in INR (Delhi water costs, i.e. approx. INR 20/m^3 * 10^3 * water saved in MLD due to higher recovery in ceramic membranes)	0.00	−41,343,696
Total 25-year OPEX cost (INR)	*60,157,200*	*6,143,954*
Net present value of OPEX calculated for 25 years at 10% interest rate (INR)	*17,019,847*	*2,230,757*
CAPEX + NPV of OPEX (INR)	*23,174,926*	*14,923,540*
Ceramic plant lower by approx. (INR)		*8,251,386*

Indicative CAPEX vs. OPEX

On average, ceramic membranes are at least five times more expensive than polymeric membranes in terms of CAPEX. On the other hand, it is assumed to have a lifetime of more than 25 years. This would necessitate a replacement of at least three to four polymeric membranes in the same 25-year period. An estimated comparison for a capacity of 4.5 MLD is shown in Table 2.

As shown above, ceramic membranes have an edge essentially because of its ability to recover from fouling as also observed by Hofs *et al.* (2011), minimal maintenance costs and less backwash water utilization with observed flux of around 550 l/m^2/h. This aspect has been captured in the indicative calculations above where water recovered means more water savings for the municipal water supplier. Pretreatment steps like the tube settler in the CWGV polymeric membrane plant are not needed in the ceramic membrane process which can lead to a lower footprint. Although OPEX is more or less comparable as per the trials, higher yield and elimination of some pretreatment steps required by the polymeric membranes brings about savings in power and chemicals as well.

CONCLUSION

The trials conducted were operated at a flux rate of 150 L/(m^2/h), with an overall yield of 97% during stable operation. After changing the backwash procedure and introduction of stronger CEB regimes with both HCl and NaOCl, it was observed that the hypo-caustic solution helped in bringing down the TMP levels and efficiently recovering the membrane. A possible alternative is conducting a CIP using citric acid in place of HCl for efficient cleaning of inorganic scalants if present on the membrane surface. Better FRC obtained in the backflush would be an indication to effectively treat organic foulants. The change executed by reversing the backwash mode and intensive CEB procedures led to an improvement in permeate quality.

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An optimum distribution of the flocculant gave better opportunities for floc formation as verified by the jar tests. The comparison between ceramic and polymeric membrane showed ceramic as a good option for potable water production due to its high flux, long life, better recovery rate and higher flux rates over the operation period. The quality of potable water obtained from the pilot plant established the ceramic membrane as a premier technology for water treatment.

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