

# MEMBRANE IN WATER AND WASTEWATER TREATMENT

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## Abstract

*The use of ultrafiltration technology for water applications is a relatively recent concept, although in the beginning, it is already commonly used in many industrial applications such as food or pharmaceutical industries. Ultrafiltration is proven to be a competitive treatment compared with conventional ones. In some cases, combination of ultrafiltration with conventional process is also feasible particularly for high fouling tendency feed water or for removal of specific contaminants. Recently, ultrafiltration has been recognized as competitive pre-treatment for reverse osmosis system. A system designed with an ultrafiltration as pre-treatment prior to reverse osmosis system has been referred to as an Integrated Membrane System (IMS). The application of IMS is a must for sites require very extensive conventional pre-treatment or where wide fluctuation of raw water quality is expected. However, the UF design was generally dismissed as commercial alternative to conventional filtration due to its high membrane cost. Nevertheless, today, the UF membrane price has gone far down, even below conventional treatment system with the new coming Asian membrane industries. Therefore, there is no doubt, UF is now becoming a competitive pretreatment system for RO in a wide range of raw water quality. Meanwhile, the application of membrane to replace secondary clarifier of conventional activated sludge, known as membrane bioreactor (MBR), has also led to a small footprint size of treatment with excellent effluent quality. The use of MBR eliminates almost all disadvantages encountered in conventional wastewater treatment plant such as low biomass concentration and washout of fine suspended solids. Today, there are more than 1000 installations of MBR all over the world. However, fouling still become a main drawback. To minimize membrane fouling, a new configuration of submerged membrane bioreactor for aerobic industrial wastewater treatment has been developed. In this configuration, a bed of porous particle is applied to cover the submerged ends-free mounted ultrafiltration membrane into which a new configuration is made. Membrane performance was assessed based on flux productivity and selectivity. A reasonably high and stable flux around 11 l/m<sup>2</sup>.h was achieved with COD removal efficiency of more than 99% from wastewater containing high organic matter. The fouling analysis also show that this newly configured ends-free membrane bioreactor exhibit lower irreversible resistance compared with the submerged one. The performance of pilot scale system, with 10 m<sup>2</sup> membrane area in a 120 L tank volume, was also studied. The resulting flux from the pilot scale system is around 8 l/m<sup>2</sup>.h with COD removal of more than 99%. Hence, this study has demonstrated the feasibility of the newly configured submerged ends-free MBR at larger scale.*

## I. INTRODUCTION

Membrane can be described as a thin layer of material that is capable of separating materials as a function of their physical and chemical properties when a driving force is applied across the membranes. Physically membrane could be solid or liquid. In membrane separation processes, the feed is separated into a stream that goes through the membrane, i.e., the permeate and a fraction of feed that does not go through the membrane, i.e., the retentate or the concentrate. A membrane process then allows selective and controlled transfer of one species from one bulk phase to another bulk phase separated by the membrane. The major breakthrough in the development of membrane technology was recorded in the late of 1950s. However, industrial application was just started ten years later, by the application of thin layer asymmetric cellulose acetate reverse osmosis membrane for seawater desalination.

Membrane process can be classified in many ways, i.e., based on its nature, structure, or driving force. Hydrostatic pressure differences are used in microfiltration (MF), and nanofiltration (NF), as well as reverse osmosis (RO) and gas separation (GS) as driving force for the mass transport through the membrane. Ultrafiltration (UF) as the main topic in this paper is also one of the membrane process based on pressure difference as its driving force. Ultrafiltration in its ideal definition as mentioned by Cheryan [1] is a fractionation technique that can simultaneously concentrate macromolecules or colloidal substances in process stream. Ultrafiltration can be considered as a method for simultaneously purifying, concentrating, and fractionating macromolecules or fine colloidal suspensions.

In the beginning, most application of ultrafiltration is in medical sector, i.e., kidney dialysis operations. Nowadays, ultrafiltration is applied in wide variety of fields, from food and beverage industries to chemical industries. Water and wastewater treatment are also the potential field of ultrafiltration application. Today, UF technology is being used worldwide for treating various water sources. The use of UF technology for municipal drinking water applications is a relatively recent concept, although as mentioned before, it is commonly used in many industrial applications such as food or pharmaceutical industries [2]. The recent global increase in the use of membranes in water application is attributed to several factors, i.e., increased regulatory pressure to provide better treatment for water, increased demand for water requiring exploitation of water resources of lower quality than those relied upon previously, and market forces surrounding the development and commercialization of the membrane technologies as

well as the water industries themselves [3]. In this paper, the application of ultrafiltration in water treatment, the system design, and its performance as pre-treatment for reverse osmosis system are presented.

## II. ULTRAFILTRATION MEMBRANE

Ultrafiltration membranes can be made from both organic (polymer) and inorganic materials. There are several polymers and other materials used for the manufacture of UF membrane. The choice of a given polymer as a membrane material is based on very specific properties such as molecular weight, chain flexibility, chain interaction, *etc.* Some of these materials are polysulfone, polyethersulfone, sulfonated polysulfone, polyvinylidene fluoride, polyacrylonitrile, cellulose, polyimide, polyetherimide, aliphatic polyamides, and polyetherketone. Inorganic materials have also been used such as alumina and zirconia [4].

The structure of UF membrane can be symmetric or asymmetric. The thickness of symmetric membrane (porous or nonporous) is range from 10 to 200  $\mu\text{m}$ . The resistance to mass transfer is determined by the total membrane thickness. A decrease in membrane thickness results in an increased permeation rate. Ultrafiltration membranes have an asymmetric structure, which consist of very dense top layer or skin with thickness of 0.1 to 0.5  $\mu\text{m}$  supported by a porous sublayer with a thickness of about 50 to 150  $\mu\text{m}$ . These membranes combine the high selectivity of a dense membrane with the high permeation rate of a very thin membrane. The resistance to mass transfer is determined largely or completely by thin top layer.

In porous membranes, the dimension of the pore mainly determines the separation characteristics. The type of membrane material is important for chemical, thermal, and mechanical stability but, in most cases, not for flux and rejection. Therefore, the aim of membrane preparation is to modify the material by means of an appropriate technique to obtain a membrane structure with morphology suitable for a specific separation. The most important techniques are sintering, stretching, track-etching, phase-inversion, sol-gel process, vapour deposition, and solution coating. However, the technique usually use for the preparation of UF membrane is mainly phase-inversion and sol-gel process.

Characterisation method of porous membranes can be performed based on structure-related parameters (determination of pore size, pore size distribution, top layer thickness, surface porosity) and permeation-related parameters (cut-off measurements) [4]. The molecular weight cut-off (MWCO) is a specification used by membrane suppliers to describe the retention capabilities of UF membrane, and it refers to the molecular mass of a macrosolute (typically polyethylene glycol, dextran, protein) for which the membrane has a retention capability greater than 90%. The MWCO can therefore be regarded as a measure of membrane pore dimensions [5]. UF covers particles and molecules that range from about 1000 in molecular weight to about 500,000 Daltons [6]. Other techniques beside cut-off measurements for characterising UF membranes are thermoporometry, liquid displacement, and permoporometry.

## III. TRANSPORT MECHANISM

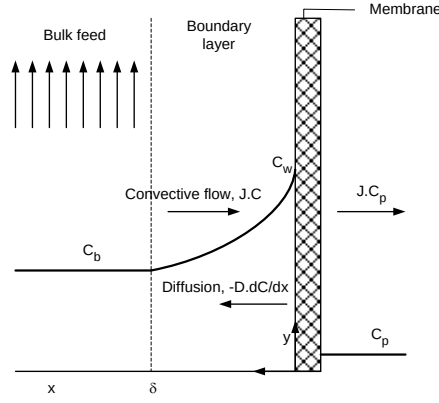
One of the critical factors determining the overall performance of an ultrafiltration system is the rate of solute or particle transport in the feed side from the bulk solution toward the membrane. As shown in Fig. 1, the pressure-driven flow across the membrane convectively transports solutes toward the upstream surface of the membrane. If the membrane is partially, or completely, retentive to a given solute, the initial rate of the solute transport toward the membrane,  $J.C$ , will be greater than the solute flux through the membrane,  $J.C_p$ . This causes the retained solute to accumulate at the upstream surface of the membrane. This phenomenon is generally referred to as concentration polarization, *i.e.*, a reversible mechanism that disappears as soon as the operating pressure has been released [7]. The solute concentration of the feed solution adjacent to the membrane varies from the value at the membrane surface,  $C_w$ , to that in bulk solution,  $C_b$ , over a distance equal to the concentration boundary layer thickness,  $\delta$ . The accumulation of solute at the membrane surface leads to a diffusive back flow toward the bulk of the feed,  $-D.dC/dx$ . Steady state conditions are reached when the convective transport of solute to the membrane is equal to the sum of the permeate flow plus the diffusive back transport of the solute, *i.e.*:

$$J.C - D \frac{dC}{dx} = J.C_p \quad (1)$$

where  $J$  is the permeate flux,  $C$  is the solute concentration profile in  $x$  direction,  $D$  is the diffusion coefficient, and  $C_p$  is the solute concentration in the permeate.

The boundary conditions are:

$$\begin{aligned} x = 0 &\Rightarrow C = C_w \\ x = \delta &\Rightarrow C = C_b \end{aligned}$$



**Fig. 1. Concentration polarization under steady-state conditions**

Integration of eq. (1) results in

$$\ln \frac{C_w - C_p}{C_b - C_p} = \frac{J\delta}{D} \quad (2)$$

If we introduce the ratio between the diffusion coefficient  $D$  and the thickness of the boundary layer  $\delta$  called the mass transfer coefficient  $k$ , *i.e.*

$$k = \frac{D}{\delta} \quad (3)$$

then eq. (3) becomes

$$J = k \ln \left( \frac{C_w - C_p}{C_b - C_p} \right) \quad (4)$$

The flux-limiting value for a totally retained solute ( $C_p = 0$ ) at gel layer conditions is given by eq. (4) as

$$J = k \ln \left( \frac{C_w}{C_b} \right) \quad (5)$$

The surface concentration ( $C_w$ ) may be obtained by extrapolation of a plot of  $J$  versus  $\ln C_b$ . It has, however, been shown that the information obtained on the surface concentrations is frequently not reliable. For identical solutions different authors have found widely varying values at  $C_w$ . In addition, it has been shown that feed solutions of various macrosolutes with concentration  $C_b = C_w$  did not give zero flux [8]. Assumption of  $k$  constant with concentration also remains questionable.

The accumulation of solutes/particles at the membrane surface can affect the permeate flux in two distinct ways. First, the accumulated solute can generate an osmotically driven fluid flow back across the membrane from the permeate side toward the feed side, thereby reducing the net rate of solvent transport. This effect generally will be most pronounced for small solutes, which tend to have large osmotic pressures (*e.g.*, retained salts in reverse osmosis). However, very high concentrations of dextran and whey protein solutions at the membrane surface have a substantial osmotic pressure [9]. Second, the solutes/particles can irreversibly foul the membrane due to specific physical and/or chemical interactions between the membrane and various components present in the process stream, thereby providing an additional hydraulic resistance to the solvent flow in series with that provided by the membrane. These interactions can be attributed to one or more of the following mechanisms: (a) adsorption, (b) gel layer formation, and (c) plugging of the membrane pores. Its severity depends on the membrane material, the nature of solutes, and other variables such as pH, ionic strength, solution temperature and operating pressure [10].

Membranes fouling typically manifests itself as a decline in permeate flux with time of operation, and consequently, this is often accompanied by an alteration in membrane selectivity. These changes often continue throughout the process and eventually require extensive cleaning or replacement of the membrane. It should be noted that the effect of membrane fouling on the flux can often be very similar to those associated with concentration polarization. For this reason, it is first necessary to distinguish between membrane fouling and concentration polarization, although both are not completely independent of each other since fouling can be resulted from polarization phenomena. In addition, flux decline can also be caused by changes in membrane properties as a result of

physical deterioration of the membrane and/or change in feed properties. So far, a number of different mathematical formulations have been proposed to predict permeate flux.

When the osmotic pressure difference  $\Delta\Pi$  across the membrane can then become substantial, the driving force of the fluid transport across the membrane is given by  $\Delta P - \sigma\Delta\Pi$  [11]. The reflection coefficient  $\sigma$  indicates the degree of perm-selectivity of the membrane. When  $\sigma = 1$  the solute is totally retained and when  $\sigma = 0$  it is totally permeable. The resistance of the accumulated solute at the membrane surface is sometimes represented as a hydraulic resistance  $R_s$ . If we introduce hydraulic resistance  $R_m$  instead of permeability in Darcy's equation and take the osmotic pressure of the solute into consideration, the flux may be described by the generalized equation:

$$J = \frac{\Delta P - \sigma\Delta\Pi}{\mu(R_m + R_s)} \quad (6)$$

The theoretical models that often be related to eq. (6) are the osmotic pressure model, the gel layer model and the resistance in series model. In the osmotic pressure model, the solute hydraulic resistance  $R_s$  is substituted by a continuous, steep, concentration gradient at the membrane, resulting in a substantial osmotic pressure:

$$J = \frac{\Delta P - \sigma\Delta\Pi}{\mu R_m} \quad (7)$$

Taking the osmotic pressure at the membrane wall into account, Wijmans *et al.* [12] have derived a relation between pressures and permeate flux. They also used the following relationship between the osmotic pressure and the concentration at the membrane wall:

$$\Pi_w = aC_w^n \quad (8)$$

where  $a$  and  $n$  are solution-dependent constants.

When the solute is completely retained ( $\sigma = 1$  and  $C_p = 0$ ), and hydraulic resistance of the solute,  $R_s$ , is neglected, combination of eq. (7) and (8) gives the following expression:

$$J = \frac{\Delta P - aC_b^n \exp(nJ/k)}{\mu R_m} \quad (9)$$

Eq. (9) shows that flux declines faster for the high permeability membrane than for the low permeability membrane. In addition, the derivative  $\partial J/\partial \Delta P$  shows how the permeate flux changes with pressure:

$$\frac{\partial J}{\partial \Delta P} = \left[ \mu R_m + aC_b^n \frac{n}{k} \exp\left(\frac{nJ}{k}\right) \right]^{-1} \quad (10)$$

Combining eq. (8) and (9) and substituting the result into eq. (10) leads to

$$\frac{\partial J}{\partial \Delta P} = \frac{1}{\mu R_m} \left( 1 + \frac{\Delta\Pi n}{\mu R_m k} \right)^{-1} \quad (11)$$

Using eq. (11), the extent of the permeate flux deviation from the pure water flux can be easily demonstrated, that is given by the second term,  $\Delta\Pi n/\mu R_m k$ . It is clear that the effect of a pressure increase depends on membrane permeability (the effect of  $R_m$ ), solution temperature (which effects  $\mu$ ), osmotic pressure ( $\Delta\Pi$  and  $n$ ), and cross-flow velocity (which effects  $k$ ).

On the contrary, in the gel layer model the osmotic pressure is assumed to be zero. The fluid flow is then described by:

$$J = \frac{\Delta P}{\mu(R_m + R_g)} \quad (12)$$

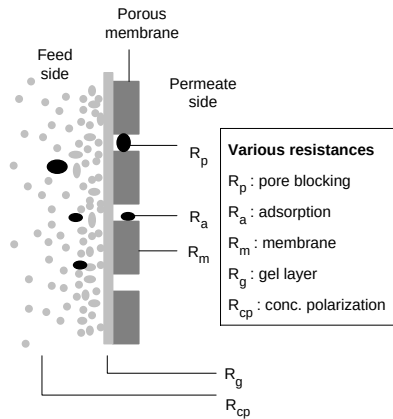
The gel layer model predicts the flux to be independent of operating pressure. An increased pressure merely results in a thicker gel layer (larger  $R_g$ ), which retards the flux to its original value. The gel layer model has been used to correlate experimental limiting fluxes [13-15]. The limiting flux for a totally retained solute ( $C_p = 0$ ) at gel layer conditions is given by eq. (5) as

$$J = k \ln\left(\frac{C_g}{C_b}\right) \quad (13)$$

Lastly, resistance to flow may be accounted for by a number of resistances: the resistance of the membrane ( $R_m$ ), the boundary layer resistance ( $R_{cp}$ ), the gel layer resistance ( $R_g$ ), the pore blocking resistance, and the adsorbed layer resistance ( $R_a$ ) as shown schematically in Fig.2.

Equation (6) may then be written as:

$$J = \frac{\Delta P}{R_m + R_{cp} + R_g + R_p + R_a} \quad (14)$$

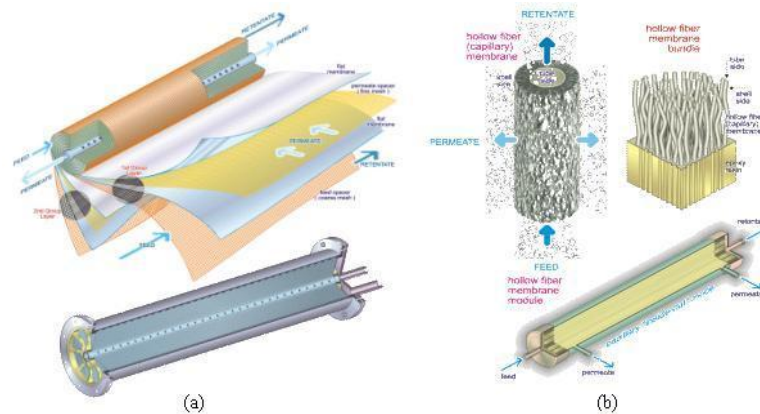


**Fig. 2. Various resistances hindering mass transfer through a UF membrane based on the resistance in series model**

#### IV. ULTRAFILTRATION SYSTEM DESIGN

Ultrafiltration (UF) is a low-pressure operation at transmembrane pressures of, typically, 0.5 to 5 bars. This is not only allows nonpositive displacement pumps to be used, but also the membrane installation can be constructed from synthetic components, which has cost advantage.

UF membranes can be fabricated essentially in one of two forms: tubular or flat sheet. Membranes of these designs are normally produced on a porous substrate material. The single operational unit into which membranes are engineered for use is referred to as a module. This operational unit consists of the membranes, pressure support structures, feed inlet, concentrate outlet ports, and permeate draw-off points. Two major types of UF modules can be found in the market, *i.e.*, hollow fibers (capillary), and spiral wound (Fig. 3). Other modules are plate and frame, tubular, rotary modules, vibrating modules, and Dean vortices.



**Fig. 3. Major types of UF modules: (a) spiral wound and (b) hollow fiber**

Each type of modules have its particular characteristics based on its packing density, ease of cleaning, cost of module, pressure drop, hold up volume and quality of pre-treatment required. Hollow fiber module has the highest packing density compare with other types of modules, including the easiest to clean and relatively cost competitive as well as spiral wound module. Based on pressure drop, the tubular module and rotating disc/cylinder have the lowest pressure drop compare with others. Hold up volume of hollow fiber module is the highest, followed by plate and frame, spiral wound, tubular, and rotating disc/cylinder module. Requirement of pre-treatment is lowest in tubular and rotating disc/cylinder modules [16].

Current membrane systems are typically modular with high packing density. Most are suitable for scale-up to larger dimensions. A broad range of membrane devices, useful for small-scale separation in the laboratory or large industrial-scale operation, is available [5]. Full-scale membrane facilities comprise series/parallel modules and operate according to various modes, range from intermittent single-stage system to the continuous multistage system [16].

Operation of UF membrane can be performed in two different service modes, *i.e.*, dead-end flow and cross-flow. The dead-end flow mode of operation is similar to that of a cartridge filter where there is only a feed flow and filtrate flow. The dead-end flow approach typically allows optimal recovery of feed water on the 95 to 98% range, but is typically limited to feed streams of low suspended solids (<1 NTU). The cross-flow mode different with dead-end mode in which there is an additional flow aside from feed flow and filtrate flow (permeate), *i.e.*, the concentrate. The cross-flow mode of operation typically results in lower recovery of feed water, *i.e.*, 90 to 95% range [17].

Nowadays, full-scale membrane elements are designed in a number of ways to optimise membrane area to element size. The design of facilities has also been optimised with the increasing plant capacities. Individual units (skids mounted units) are usually used for small plant capacities whereas for larger plant capacities (10,000 m<sup>3</sup>/d and above) racks with ancillary equipment designed. Today, racks comprised of up to 48 membrane modules are being constructed and additional scale-up savings are therefore observed [2]. Typical large scale UF plant is shown in Fig. 4.

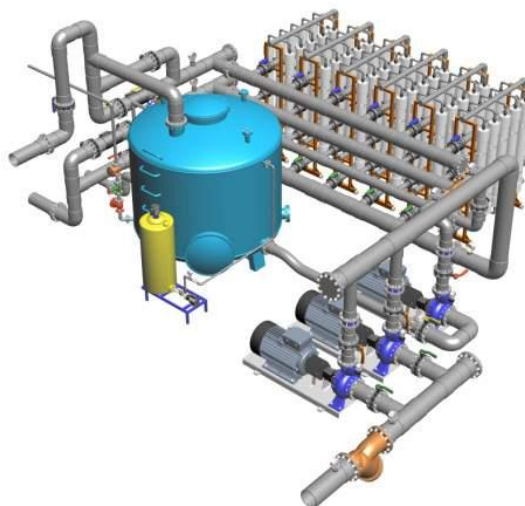
Flux decline has a negative influence on the economics of a given membrane operation. Flux decline usually attributed to fouling phenomenon. Fouling control strategies can be categorized as: tailor or membrane treatment, modify or pre-treat the feed water, adjustment of operating condition, and cleaning [18, 19]. Membrane cleaning is the removal of foreign material from the surface and body of the membrane and associated equipment to reduce fouling to some extent. The frequency of cleaning is a critical economic factor, since it has a profound effect on the operating life of a membrane. Cleaning and sanitizing membranes is desirable for several reasons, that is, laws and regulations may demand it in certain applications (*e.g.*, the food and biotechnological industries), reduction of microorganisms to prevent contamination of the product stream, and process optimisation. A clean membrane can be defined in three terms according to Cheryan [6], *i.e.*, physically clean membrane, chemically clean membrane, and biologically clean membrane. Flux recovery to initial flux of a new membrane after cleaning can be used as indication of clean membrane.

Four cleaning methods can be distinguished, *i.e.*, hydraulic cleaning, mechanical cleaning, chemical cleaning, and electrical cleaning. The choice of cleaning method mainly depends on the module configuration, the type of membranes, the chemical resistance of the membrane and the type of foulant encountered. Hydraulic cleaning methods include back flushing, alternate pressurising and depressurising and by changing the flow direction at a given frequency. In backflush technique, the direction of the permeate flow through the membrane is periodically reversed. However, backflushing also reduces the effective operation time, and gives a loss of permeate to the feed solution. The impact of backflushing in industrial application is very limited, because of its fundamental limitation, *i.e.* loss of permeate and operation time, therefore the backflush process needs adequate optimisation. The backflush process is optimized both for the duration of the backflush and for the backflush interval. The improvement of the product rate upon backflushing is mainly a function of the backflush pressure and the interval between two backflushes. Recently, the time interval of back flushing has been reduced to seconds which implies that the cake resistance remains low since it has no time to built up a layer.

A novel backflush technique with high frequency and extremely short duration have been introduced. It was found that excellent results could be obtained using very short backflush time (0.06 second) with maximum interval time of 5 seconds (preferable 1-3 seconds) [20, 21]. Since the effective backflush time is very short and the backflush pressure is relatively high (typically 1 bar over the feed pressure) then it was called “backshock” technique. Backshock technique combined with the use of reversed asymmetric structures allows filtration at extremely low cross-flow velocities with very stable permeate fluxes [21]. By employing this method, UF fouling during the filtration of solution containing high solid concentration could be controlled [22-25].

Mechanical cleaning using oversized sponge balls can only be applied in tubular systems. Several researchers are developing other mechanical cleaning using ultrasonic wave. Chemical cleaning is the most important method for reducing fouling, with a number of chemicals being used separately or in combination. The concentration of the chemical and the cleaning time are also very important relative to the chemical resistance of the membrane. Electrical cleaning is a very special method of cleaning. By applying an electric field across a membrane, charged particles or molecules will migrate in the direction of the electric field. Electrical cleaning can be applied without interrupting the process and the electric field is applied at certain time intervals [4].

Fouling may also be controlled by operating UF under its critical flux [26]. When UF is operated under its critical flux, foulant deposition on membrane surface can be avoided. Thus membrane can be operated with stable flux.



**Fig. 4. Typical large scale UF plant**

## V. ULTRAFILTRATION IN WATER TREATMENT

Water has the ability of dissolving and containing various substances. Fresh water from surface water or groundwater is utilized for industrial or domestic purpose, either for potable or non-potable use. Due to the intended purposes, a water treatment plant is needed to fulfil the requirements of treated water. In general, conventional water treatment plant usually consists of physical treatment (screening, sedimentation, flotation, filtration) and chemical treatment (pH adjustment, coagulation-flocculation process, oxidation-reduction process, adsorption process) [27]. The degree of the complexity of the treatment plant also depends on the quality of raw water and treated water requirement. In industrial processing, water is used in numerous applications requiring likewise different qualities of water. Examples of different use are cooling water, water for rinsing and chemical production, boiler feed water, purified water, water for injection, *etc.* The growth in population, the increasing costs of treatment and distribution, contamination of fresh water source, and the sophistication of end user, somehow forces the development for better improvement of water treatment technology [5].

Ultrafiltration (UF) is proven to be a competitive treatment compare with conventional ones. The production of clear and sparkling water that is safe as far as disease is concerned usually require chemical precipitation, adsorption, sedimentation, and filtration [5]. Each step of this process has to be controlled to get an optimal performance of the overall process, which results in a complex control system [28]. Nowadays, UF is used to replace clarification step in conventional water treatment plant, *i.e.*, coagulation, sedimentation, and filtration and can be defined as a clarification and disinfections membrane operation. UF membranes are porous, however, all particulate contaminants such as viruses and bacteria, including macromolecules are rejected. The main advantages of low-pressure UF membrane processes compare with conventional clarification (direct filtration, settling/rapid sand filtration, or coagulation/sedimentation/filtration) and disinfection (post chlorination) processes are no need for chemicals, size-exclusion filtration as opposed to media depth filtration, good and constant quality of treated water in terms of particle and microbial removal regardless of raw feed water quality, process and plant compactness, and simple automation [5].

Source water quality directly impacts UF membrane performance. Therefore, in practice, depending on the quality of raw water, UF can be operated as single operation or combination with other process (coagulation, adsorption, *etc.*) or hybrid membrane system (UF/MF). In water application, UF can be the main process or as pre-treatment for example in RO system.

Today, more than 2 millions m<sup>3</sup>/d (750 mgd) of drinking water is produced worldwide using low-pressure membranes, including microfiltration (MF) and ultrafiltration. More than 50 UF plants for producing drinking water from surface water are in operation in the world [29]. Out of the low-pressure membrane full-scale plants identified worldwide, UF applications represent about 74% of the total installed capacity. A six years operation of UF membrane in Amoncourt, France showed no loss performances in terms of production capacity and water quality produced. In addition, mechanical properties of the membrane material over time did not show any important losses [2]. Existing UF membrane plants worldwide treated various source of raw water, *e.g.* groundwater, surface water, clarified surface water, to produce drinking water with the capacity 0.01-14.53 gpd. Some are located in France, UK, US, Tahiti, and Japan [5].

As mentioned before, application of UF for drinking water supply can be in form of single operation, *i.e.*, without any pre-treatment except a common screen filter [28]. UF can be used on its own for treating drinking water where the feed water is not too high in terms of organic content [2]. Membrane filtration has become the preferred alternative to conventional technology to remove water-borne pathogens in the preparation of drinking water [30]. Therefore, it is possible to reduce the necessity of water disinfection after the treatment process [31]. UF technology has been found to exceed current water regulation for turbidity, *Giardia*, and also virus removal [2]. Removal of viruses and bacteria using UF could achieve a percentage removal 90-100% [5, 32].

Apart of the increasing number of UF plant, fouling and membrane costs are still the main limitations to UF development and widespread use [33, 34]. However the cost of UF technology has significantly decreased within past five years. Capital costs were found to depend not only on the raw water quality (flux) and the plant capacity but also on the year on construction [2]. The term fouling includes the totality of phenomena responsible for decreases of permeate flux over a period of time, except those linked to membrane compaction and mechanical characteristics modification [5]. Numerous research studies have been conducted to study the mechanisms and factor affecting flux decline as related to the fouling phenomenon including main source of fouling during membrane processes.

Several researchers have studied effects of fouling materials, that is, clay-organic substances, humic acid, microbial decomposition products, on the fouling of membrane [35-39]. Teixeira, *et al.* [40] found the important role of the pH on the UF performance controlling the interactions of membrane with fouling matter. Natural organic matter (NOM) rejection and NOM transport across the membrane also had been studied [41]. Natural organic matter present in raw water not only impart colour to water, but can also cause health risks associated with disinfections by-products (DBP). The most common DBPs found in drinking water are trihalomethanes (THMs) and haloacetic acids (HAAs), which are formed when NOM reacts with chlorine or chlorine based disinfectants [42]. Membrane processes allow the reduction or elimination of NOM (*e.g.* humic acid, fulvic acid) as THM precursors and prevent the formation of substances posing hazards to human health [31].

In cases where the feed water contains high turbidity levels or high fouling tendencies, combination with conventional pre-treatment (adsorption, coagulation, oxidation) is required to allow the membranes to operate

efficiently [43]. UF alone also is not very effective for removing DBPs and dissolved substances in general, and have limited capability in removing organic matter. The use of powdered activated carbon (AC) in combination with UF membrane is attracting increasing interest for the removal of organic compounds in drinking water treatment [44]. This hybrid process utilizes the capabilities of activated carbon to adsorption of impurities and the microorganisms and particles removal ability of the membranes [31]. Coupled with PAC, UF can be used to treat groundwater contaminated by micropollutants such as pesticides or surface water with high organic matter load [2]. Effect of filtration time, membrane reactor volume, carbon dosing procedure, carbon dose and carbon particle size on the adsorption removal of selected micro pollutants and dissolved organic matter has also been studied [44-46]. Yuasa [47] found that combination of UF with PAC/GAC could improve the removal of organics and other micropollutants such as agrochemicals. Currently there are already several installations of water treatment plant that has been applied using hybrid of AC/UF with the capacity range from 200-65,000 m<sup>3</sup>/day [2].

Combination of coagulation/UF can also be considered for surface waters containing fairly high level of organics and also to minimize membrane fouling potential [2]. Coagulation pre-treatment may enhance permeate flux by reducing foulant penetration into membrane pores, conditioning the layer of materials deposited on the membrane, and improving particle transport characteristics [48]. Guigui, *et.al.* [49] reported that the addition of coagulant before UF unit with or without settling may increase NOM removal for a better reduction of DBP. Determination of optimum coagulation condition, removal efficiency, effect of configuration and module design of combination UF/coagulation has also been studied [34, 50, 51].

Other potential application of UF is the production of ultrapure water. Usually, UF is act as pre-treatment of RO unit to produce ultrapure water. However, Oosterom, *et. al.* [52] proposed an innovative alternative process to use rainwater followed by low-pressure MF/UF to produce demineralised water.

Recently, membrane technology has been considered as an alternative water treatment in aquaculture [53]. A sufficient supply of good quality water is essential to any aquaculture operation. Water quality affects reproduction, growth, and survival of aquatic organism. The criteria for good quality water established by safe level of physical, chemical and biological properties of water, which have significant adverse effects on aquatic organism growth and survival. To increase the quality of water input, the use of UF will surely retain the pathogen and generate highly free pathogenic water. As the UF pore size still allowed ions to pass the membrane pore, the use of UF to treat seawater for example in shrimp culture is perfect. A study showed that growth rate, survival rate and production of Black Tiger Shrimp *Penaeus monodon* postlarve is directly influenced by water quality and hygienic condition in culture system [54, 55]. Therefore, in aquaculture system, UF is needed to reject suspended solid and pathogenic microorganism from the culture water [53].

## VI. THE REVERSE OSMOSIS SYSTEM

One of the first membrane applications for the utilization of membrane technology was the conversion of seawater into drinking water by reverse osmosis (RO). RO system separate dissolved solutes (includes single charged ions, such as Na<sup>+</sup>, Cl<sup>-</sup>) from water via a semipermeable membrane that passes water in preference to the solute. RO can be described as diffusion-controlled process in which mass transfer of ions through RO membranes is controlled by diffusion. Physical holes may not exist in an RO membrane, which distinguishes RO membrane with other filtration system. RO membrane is very hydrophilic; therefore water will be able to readily diffuse into and out of the polymer structure of the membrane.

Four types of modules are used for RO membrane, *i.e.*, plate and frame, tubular, hollow fiber, and spiral wound. However, the spiral-wound element is the most common by far for the production of drinking water. RO configurations include single stage, two stages, and two-pass systems. The selection among these configurations depends on the desired quality of the product water. The pass system gives the highest purity product and it is suitable for preparation of make-up boiler water. The single stage system is the simplest layout and quite common for use on various desalination applications. Meanwhile, the two-stage system is common for brackish water use where it is necessary to increase the overall recovery ratio [56].

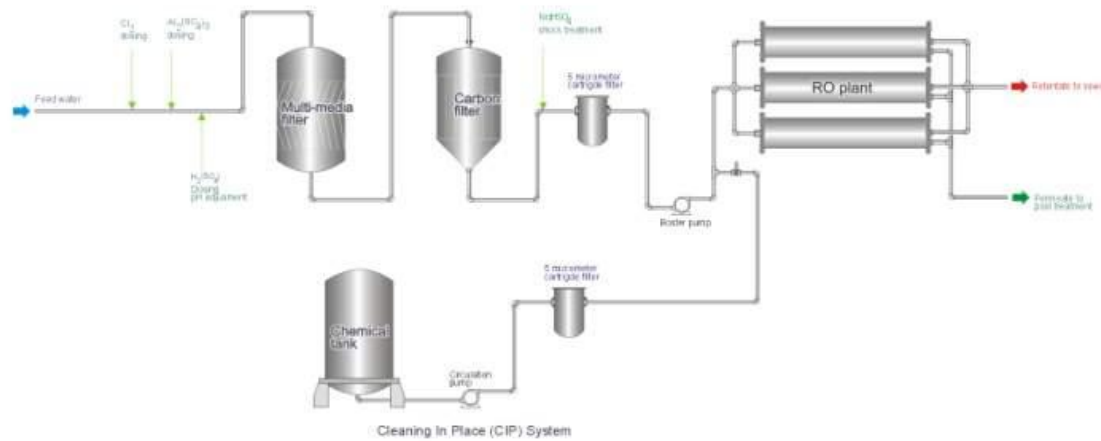
Nowadays, RO system has become a popular water treatment technology in industry requiring separation of dissolved solute from its solvent (water) including desalination and also, residentially, to improve the taste of water as well as to remove potentially unhealthy contaminants. RO has increased the water supply by making possible the use of brackish waters for potable water supply. Desalination using RO has become a major source to produce fresh water in many arid regions including remote area where the fresh water is hardly found. Recent advances particularly in improvements of the membrane materials and pre-treatment have meant that RO desalination has now become economically attractive even at seawater concentrations [57]. The scale of membrane applications is now very large, plants with capacity in excess of 19,000 m<sup>3</sup>/d are common [57].

The success of RO technology has been due mostly to the economics of its operation and to its simplicity. Rapid developments in RO membrane are addressed to new membrane working at lower pressure and increasing salt rejection from the original cellulose acetate membrane requiring 28 bar to modern polyamide thin-film membranes requiring only 7 bar net driving pressure. The increase of salt rejection of RO membrane from 97 to 99.5% with some special membrane types exhibiting even higher separation efficiency [58]. Byrne [59] also noted that newer membranes, because of its ability to reject more salts and pass more water at a particular pressure, is having greater



energy efficiency. The simplicity of RO desalination process layout compare with the large-scale thermal desalination process is also one of the main attractive feature of RO system. Its modular design allows for simple expansion and increase of the production capacity. Specific power consumption of RO is low, around 5 kWh/m<sup>3</sup>. This amount is almost equivalent to the pumping power for the major thermal desalination process, which include MSF and ME [56].

Yet, available RO membranes are generally not robust enough to operate directly on surface feed seawater [60]. RO membranes are more sensitive than thermal desalination processes to scaling, fouling, chemical and biological attack. The susceptibility to fouling is one major shortcoming of RO membrane. Hence, RO has developed into an energy efficient alternative to thermal processes but it still continues to face competition due to the requirements of pre-treatment. Schematic of RO with extensive pretreatment system can be seen in Fig. 5.



**Fig. 5. RO system with extensive pre-treatment**

## VII. IMS, INTEGRATED MEMBRANE SYSTEM

The successful long-term performance of RO seawater desalination plant is highly depend on proper pre-treatment. Pre-treatment of RO system is designed to prevent fouling of the membrane, maintain performance of the system, and extend the lifetime of the membranes [60]. The selection of pre-treatment RO system is based on the raw water quality, the reliability, the investment cost, and the RO membrane type [61].

Where potential fouling waters are the only available source for processing into high purity water as in marginal waters, the conventional pre-treatment process methods may not be adequate. Marginal waters are difficult to treat due to the fouling problems that can occur with insufficient pre-treatment in a membrane plant [62]. High fouling surface water and low fouling beach well water sources need different complexity of pre-treatment. As in the case of direct seawater intake or municipal wastewater reuse, extensive pre-treatment is required up-stream the RO process compare with beach well water.

As mentioned before, pre-treatment of process water before RO is very important for membrane life and the economical operation of the RO plant [5]. Pre-treatment by conventional means (*i.e.*, coagulation, flocculation, and media-filtration) is known to be complex, labour intensive and space consuming. Many SWRO (sea water RO) plants operate successfully for many years with conventional pre-treatment [63]. However, if conventional pre-treatment is not designed and operated carefully, RO plants can have problems with membrane fouling [64].

Primarily, the development of UF technology in water application is focused in producing filtrate for drinking water [17]. Recently, UF has become an efficient pre-treatment for reverse osmosis (RO) system [28]. It is important to re-evaluate the cost and operating benefits of UF as pre-treatment particularly for high fouling feed water source such as surface water, a wastewater, or an open-intake seawater [17]. A system designed with a MF/UF membrane system as pre-treatment prior to RO system has been referred to as an Integrated Membrane System (IMS) [17, 65]. IMS combines the advantages of UF for particle removal with the selectivity of RO [61]. IMS to achieve the water quality objectives is considered very seriously, and several studies are currently on going to evaluate the feasibility of such dual membrane system [2]. A major reason for the re-emergence of UF technology has been improvements in the control of fouling during the service operation by the use of short-duration periodic backwashing. Periodic backwashing is designed to minimize the need for chemical cleaning to once every month to six months [17]. The IMS design approach to water treatment systems has some significant advantages over RO systems designed with conventional pre-treatment. The important features of UF pre-treatment are continuous and easily automated operation, no breakthrough as occurs in granular media filtration, good downstream protection of RO membranes, no addition of chemicals, simple chemical shock disinfections treatment and compact design of pre-treatment equipment [66].

The pre-treatment of feed water prior to RO is intended to lower the silt density index (SDI), remove excessive turbidity or suspended solids, and adjust and control the pH [60]. The SDI is the most widely used fouling index. The SDI of feedwater of an RO plant should be less than 2 to minimize the rate of colloidal fouling [67]. The significantly lower SDI filtrate produced by UF membrane as RO pre-treatment have also been proven by several researchers. The quality of feed water produced by the UF system, operating parallel with the conventional system, was very little affected by the fluctuation of the seawater quality [68]. The surface seawater SDI of 13-25 was reduced to below 1 whereas the conventional pre-treatment failed to reduce it below 2.5 [69]. A pilot plant conducted by Van Hoof (2001) showed that UF membranes used for RO pre-treatment produced water with SDI<sub>15</sub> values as low as 0.4 and showed stable operation. Glucina, *et.al.* (2000) also found that UF could produce filtrate with the average SDI of 1.2, a reasonably low value when compared to the maximum advised by the RO membrane manufacturer. Good quality of RO feed make it possible to reduce RO cleaning frequencies due to colloidal fouling.

The dead-end UF mode coupled with operation at low pressure allowing very low power consumption, approximately 0.1 kWh per m<sup>3</sup> of permeate [70]. UF system also require less time and is easier to operate than some conventional filtration processes, particularly those prone to system upsets. UF concentrated waste streams are easier to dispose of relative to chemically enhanced conventional pre-treatment processes [17]. Drioli, *et al.*, [71] also mentioned that an interesting way for further reducing fouling phenomena and extending the life time of RO membranes is related to the use of UF for pre-treatment. The field test results of UF membrane pre-treatment, tested at two different sites confirm that the membrane pre-treatment is reliable technology capable of providing consistently good quality feed water for RO seawater system independently of the raw water quality fluctuation [68]. Meanwhile, the specific flux of the UF membrane also remained stable as found by Teuler, *et.al.* [70].

The IMS system choice depends on the fouling properties of the feed water, which may necessitate additional (pre) treatment and the local circumstances. Additional pre-treatment inevitably leads to extra investment costs. However operating and maintenance costs may be lower due to a more stable system performance with lower cleaning frequencies and longer life time of the membranes [65]. Typically the only pre-treatment requirement to UF system is coarse filtration by the use of strainers rated at 100 to 150 micron. Occasionally the use of a coagulant aid like a ferrous salt is considered [17]. The combination of UF with a pre-coagulation at low dose helped in controlling the UF membrane fouling and providing filtered water in steady state condition [69]. The seawater system operating with UF membrane pre-treatment can be designed to operate at the higher limit of the permeate flux range due to the very low concentration of suspended solids in the UF filtrate [68]. Ability to operate RO seawater unit at higher flux and recovery rate enables optimisation of the RO process and reduction of water cost [68].

The reason why the trend of pre-treatment RO system goes for integrated membrane system are mainly feasibility, process reliability, plant availability, modularity, relative insensitivity in cases of raw water changes and lower operating costs. UF allows the membrane inventory of an RO plant to be reduced by some 20% and have simplified the RO pre-treatment process resulting in lowering the operating costs of the plants [62]. Bates [17] reported that operating costs and chemical costs are competitive and in some scheme less. The demand of UF system as pre-treatment for RO will be accentuated by the increasing scarcity of low-fouling feed water sources (well water) and the need to treat more difficult feed water sources (surface waters, industrial wastewater, and municipal sewer waters). In future the coagulation-sedimentation-filtration (CSF)-UF-RO scheme will compete with the CSF-SSF (slow sand filtration) scheme as estimated by Nederlof, *et.al.* [65].

Although UF provides high quality feed water for RO, the UF design was generally dismissed as commercial alternative to conventional pre-treatment for a number of reasons, *i.e.*, capital costs were too high for treatment of surface waters. Glueckstern, *et.al.* [68] reported that the cost of membrane as pre-treatment is more expensive than the conventional pre-treatment. As cited by Redondo [62] from several authors, the application of IMS is currently not frequently used to lower costs although this may change. However, since the energy requirement is very low, consequently the cost is mainly directed to the membrane price. Nowadays, the UF membrane price has gone far down, even below conventional treatment system with the new coming Asian membrane industries. Therefore, there is no doubt, UF is now becoming a competitive pre-treatment system for RO in a wide range of raw water quality, from excellent to poor quality of raw water.

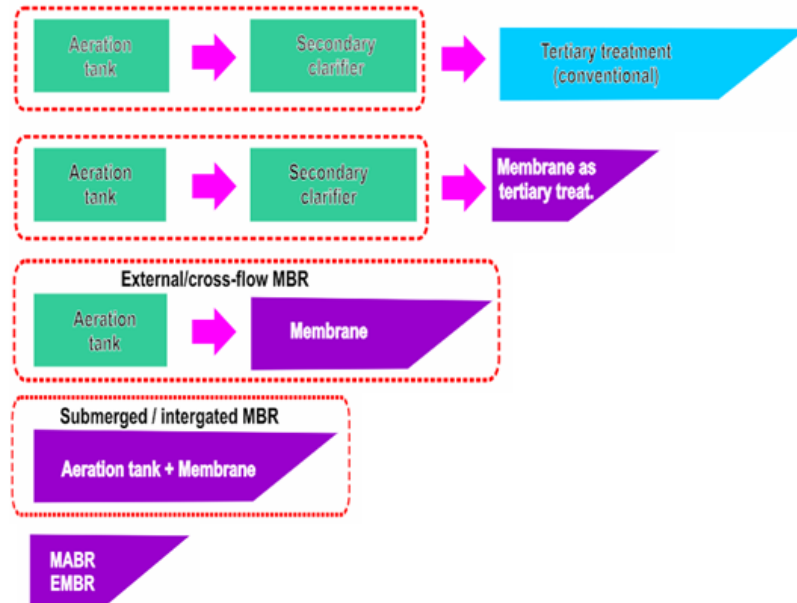
## VIII. MEMBRANE BIOREACTOR

The discharge of the improperly treated wastewater directly to water body has led to a serious impact to the environment. Membrane technology as a separation process appears as the best alternative to improve the performance of conventional wastewater treatment. The application of membrane unit to replace secondary clarifier of conventional activated sludge has led to a small footprint size of treatment plant with excellent effluent quality. This combination of membrane technology with biological process is known as membrane bioreactor (MBR). Today there are more than 1000 installations of MBR all over the world [72].

The use of membrane bioreactor eliminates almost all disadvantages encountered in conventional wastewater treatment plant, such as low biomass concentration (3000-4000 mg/l), washout of fine suspended solids in the effluent (>20 mg/l), and no elimination of germs [73]. The sub-micron size of membrane pore allows membrane bioreactor to completely retain the biomass, including bacteria and viruses, therefore it also act as disinfections device. The ability of MBR to completely retain biomass also allowed the operation of MBR on very high biomass concentration, higher

than 8,000 mg/l [74, 75]. In some cases, the biomass concentration can reach 50,000 mg/l [76], far beyond the biomass concentration commonly found in conventional biological wastewater treatment, *i.e.*, 3,000-4,000 mg/l.

MBR can be divided into two main configurations, *i.e.*, external-MBR and submerged-MBR. In external MBR the membrane module is placed outside the bioreactor meanwhile in submerged MBR, the membrane module is placed directly inside the bioreactor. Nowadays, there is a strong trend to employ submerged configuration compared with external one due to the lower cost of fabrication, maintenance, and energy consumption. The schematic development of MBR configuration can be seen in Fig. 6.



**Fig. 6. Development of MBR configuration**

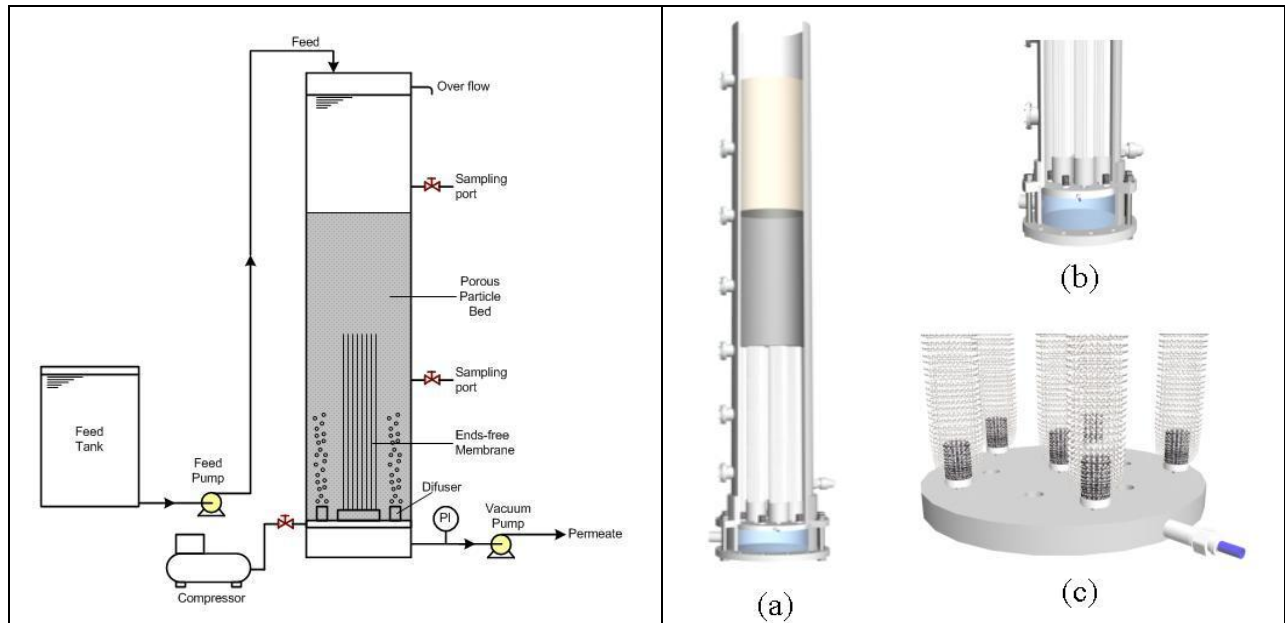
In MBR operations, fouling still become a main drawback. The presence of fouling is usually characterized by flux decline during filtration process. Fouling in membrane process can be defined as irreversible, or not easily reversible, deposition of retained species onto or into the membrane [77]. Fouling makes frequent membrane cleaning and consequently replacement, which then increases maintenance and operating costs. Numerous fouling reduction techniques have been studied by many researches to minimize membrane fouling particularly in submerged configuration as can be seen in Table 1. By far, aeration is the most common method that used extensively for fouling reduction considering its ability to create shear stress on the surface of the membrane. However, several drawbacks are reported from the use of aeration such as the trapping of bubbles between the membrane fibers, and not evenly distributed bubbles flow to the fibers. Bouhabila, *et al* [78] observed that an increase in the air flow rate partly stimulated the cake removing efficiency, but there was a critical value beyond which the air-flow rate increase have virtually no effect on the cake-removing efficiency.

**Table 1. Fouling reduction techniques**

| Techniques                                                                                        | References        |
|---------------------------------------------------------------------------------------------------|-------------------|
| Aeration                                                                                          | [75, 78-82]       |
| Hybrid with activated carbon completely mixed in the bioreactor                                   | [83-91]           |
| Zeolite addition                                                                                  | [92]              |
| ELDE-MBR configuration (external loop dead-end MBR)                                               | [80]              |
| Utilization of riser and downcomer                                                                | [93]              |
| Modification of reactor chamber                                                                   | [94, 95]          |
| Intermittent filtration                                                                           | [92, 93, 96-103]  |
| Backflush                                                                                         | [30, 79, 80, 104] |
| Operating at critical flux                                                                        | [78, 105, 106]    |
| Optimization of the distance between the membrane module and the wall of bioreactor               | [107]             |
| Utilization of inclined-plate                                                                     | [108]             |
| Utilization of moving carrier (polyurethane cubes coated with activated carbon, plastic granules) | [109, 110]        |
| Polymer addition                                                                                  | [111]             |
| Membrane surface modification                                                                     | [112-115]         |
| The use of aerobic biogranules activated sludge                                                   | [86, 116]         |

All of the above mentioned method to reduce fouling still enable the direct contact of mixed-liquor of activated sludge with membrane surface, hence, promote membrane fouling. In this work, a newly configured submerged MBR has been developed. The submerged membrane is covered by porous particle bed in which the

porous particle bed will act as a protection means to minimize direct contact of mixed-liquor with membrane surface and also give a scouring effect as *in situ* pre-treatment filter to minimize membrane fouling and to extend the lifetime of the membrane. Furthermore, the growth of microbes on the porous structure of the porous particles apart from microbe grown in suspended form is also expected to give additional biodegradation effect on small organics responsible for membrane fouling. The configuration of this newly configured submerged ends-free MBR (eMBR) is schematically described in Fig. 7. The difference of this configuration compared with conventional submerged MBR is the existence of porous particle bed on the surface of the membrane. The ends-free hollow fiber membrane unit was immersed in bioreactor tank and covered by porous particle, i.e., granular activated carbon or zeolite. The activated sludge was mixed and aerated by airflow supplied at the bottom of bioreactor, which act to fluidize the porous particle bed. The membrane unit was operated in outside-in mode.



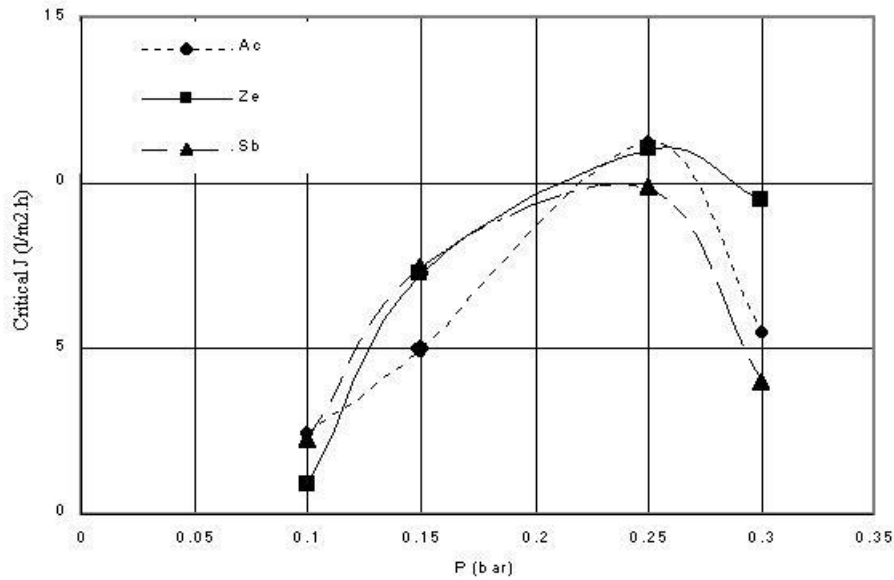
**Fig. 7. Schematic of newly configured submerged ends-free MBR. (a: full-system; b: inner side of the MBR; c: air-flow pattern from diffuser located in the bottom of MBR)**

The rejection characteristics of newly configured submerged ends-free MBR (eMBR) compared with conventional MBR and activated sludge is shown in Table 2. Activated sludge process contribute to 90% of COD removal. Meanwhile, by incorporating membrane, as in submerged MBR, an increase of COD removal in average of 97.65% is observed. In the meantime, both Ze-eMBR and Ac-eMBR show better COD removal compare with submerged one, i.e 97.68% and 99.36%, respectively. The residual COD found on the permeate represents the soluble non-biodegradable COD of the treated water or soluble biodegradable COD that have yet been biodegraded by microorganisms. The difference between the rejection characteristics of Ze-eMBR and Ac-eMBR also implies the importance of porous particle selection according to its effectiveness in adsorbing the soluble contaminants.

**Table 2. COD removal comparison**

| Run     | Activated<br>sludge | Rejection (%)    |         |             |
|---------|---------------------|------------------|---------|-------------|
|         |                     | Submerged<br>MBR | e-MBR   |             |
|         |                     |                  | Zeolite | Act. carbon |
| I       | 85.94               | 97.97            | 97.34   | 99.09       |
| II      | 87.67               | 97.19            | 96.92   | 99.67       |
| III     | 91.67               | 97.99            | 99.72   | 99.00       |
| IV      | 90.71               | 96.85            | 99.11   | 99.79       |
| V       | 90.15               | 97.65            | 97.68   | 99.36       |
| Average | 90.15               | 97.65            | 97.68   | 99.36       |

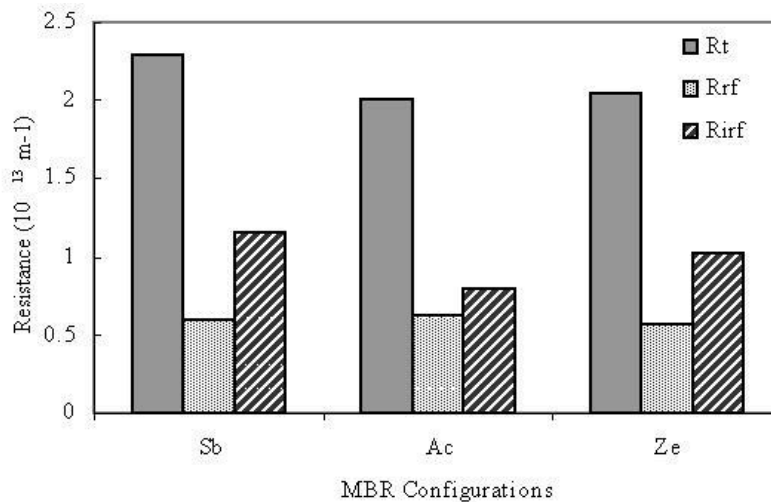
During trials under different TMPs tested for each configuration, the hydraulic resistance is rapidly increased at the beginning of the filtration period. However, the initial flux is reasonably high. This result indicate that fouling still occur in the e-MBR configuration but with different magnitude as can be seen from critical flux shown in Figure 9 below. Critical flux determination was done according to Le Clech, *et al.* [117].



**Fig. 8. Effect of pressure on the critical flux for each MBR configuration**

Figure 8 present critical flux of all configurations at TMP between 0.10 – 0.30 bar. It can be seen that by increasing TMP up to 0.25 bar, higher critical flux is obtained. However, the increase of TMP to 0.30 bar give a contrary result. According to Cheryan [6] flux will be affected by the TMP under conditions where concentration polarization effects are minimal. These also emphasize that at TMP range of 0.10-0.25 bar, the filtration process is in the pressure-controlled region. Therefore, the optimum TMP for all configurations is achieved at TMP of 0.25 bar. Further, both Ze-eMBR and Ac-eMBR show higher flux compared with normal submerged configuration, in which the highest is achieved by Ac-eMBR. It is shown that fouling still occur on eMBR configuration but the effect is less.

Fouling on membrane bioreactor is very complex phenomena because it involves a combination of physical, chemicals, and biological aspects. Fouling can occur through three mechanisms, *i.e.*, particles deposition, adsorption (specific interaction between membrane materials and foulant), and pore blocking mechanism. The effects of fouling on membrane performance can be expressed in term of hydrodynamic resistances. At the beginning of filtration, initial flux will be dominantly influenced merely by the membrane resistance. However with prolonged filtration time, the value of the total resistance will be changed because of the pore blocking and cake resistance formed on the membrane surface. In this experiment, the resistance-in-series model was applied to quantitatively evaluate the fouling resistance in each MBR configurations as shown in Fig. 9.



**Fig. 9. Series of resistance for each MBR configuration. (Rt: total resistance; Rrf: reversible fouling; Rirf: irreversible fouling)**

Figure 10 shows series of resistances for each MBR configuration at TMP of 0.25 bar. Rt, Rrf, and Rirf respectively denote the value of total fouling resistance, reversible fouling resistance, and irreversible fouling resistance. The results clearly show that the total fouling resistance of submerged MBR is highest compared with the Ze-eMBR and the Ac-eMBR. This can be the reason for the lowest final flux obtained from submerged configuration

in previous discussion. Meanwhile, for the eMBR configuration, the use of activated carbon give lower total fouling resistance compared with zeolite.

Figure 10 also demonstrate that in all configurations, the major resistance is contributed by irreversible fouling. Irreversible fouling is usually resulted from adsorption of foulant into the membrane surface and the entrapment of foulant into the inner structure of the membrane. The highest irreversible fouling resistance is shown by submerged configuration. This implies that submerging the membrane directly into the mixed-liquor cause stronger tendency for the foulant to be adsorbed on the membrane and more chance of the solutes to be entrapped on the inner structure of the membrane. In the meantime, the Ac-eMBR shows lowest irreversible resistance. This may be due to the role of activated carbon bed which hinder direct contact between foulant and membrane surface by the existence of scouring effect and ability to adsorb solutes which responsible for the irreversible fouling.

Nevertheless, the reversible fouling resistance is also highest for Ac-eMBR. Reversible fouling is initiated by formation of cake layer on the membrane surface. From the above data, it can be seen that the e-MBR configuration still allow the formation of cake layer on the membrane surface, though this can be easily removed by surface cleaning. Fig. 10 shows photograph of membrane after trial for submerged MBR (a) and Ac-eMBR (b). The eMBR one shows cleaner surface compared with the submerged one in which thick sludge deposit is observed. However, from the analysis on membrane fouling, the cake resistance of eMBR configuration was higher than the submerged one. It is assumed that in eMBR configuration, the cake layer resistances is partially attributed to the carbon active bed because, as seen on the picture, there are some activated carbon particles unevenly distributed on the membrane inter fiber. A significant scouring effect is also shown as there is no thick deposit cake layer observed on the membrane surface. Therefore, subsequent experiments were focused on Ac-eMBR configuration.

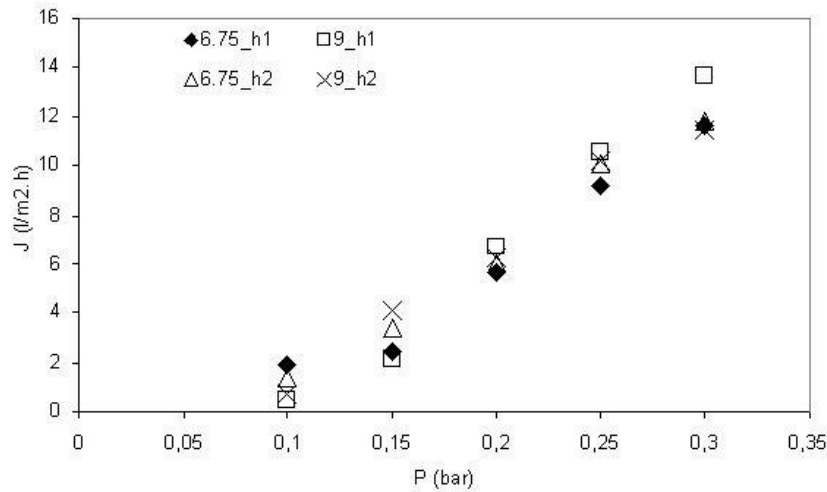


**(a) submerged MBR**

**(b) Ac-eMBR**

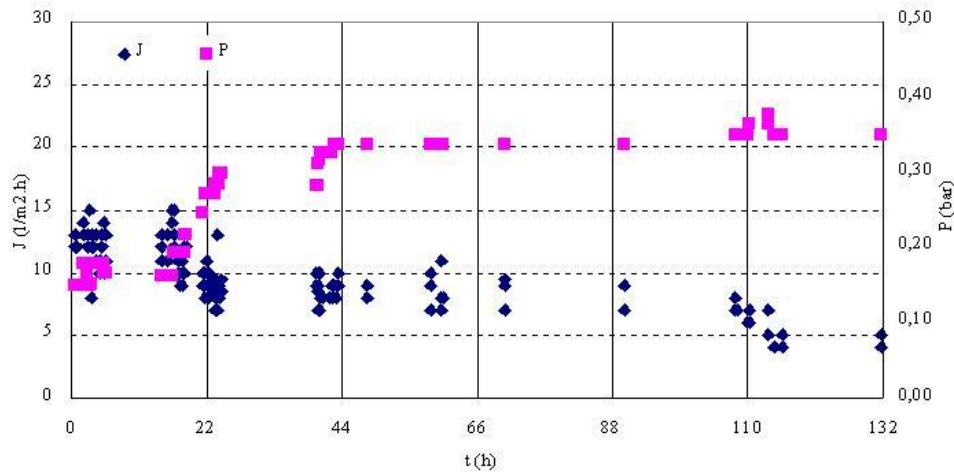
**Fig. 10. Picture of membrane after trial**

As can be concluded from the previous discussion, it is necessary to observe the effect of aeration and activated carbon bed thickness to further study fouling phenomena that occur in Ac-eMBR configuration. Figure 11 show relationship between flux of pure water at various TMP and aeration rate at different activated carbon bed thickness. Increased aeration rate is assumed to increase flux meanwhile increasing bed thickness increases resistance for mass transfer passing the membrane. In general, bed thickness of 8 cm and aeration rate of 9 l/min give highest flux compared with other variations at TMP between 0.2-0.3 bar. Meanwhile, for TMP less than 0.2 bar, highest flux is achieved by bed thickness/aeration rate of 8 cm/6.75 l/min and 15 cm/9 l/min, correspondingly. It can be seen that there is no clear pattern observed and the increase of flux is merely attributed to the increase of TMP. Hence, it is assumed that at a certain TMP, fouling resistance contributed by the difference thickness of activated carbon is not significant.



**Fig. 11. Profiles of stable flux of pure water at various bed thickness and aeration rate on TMP**

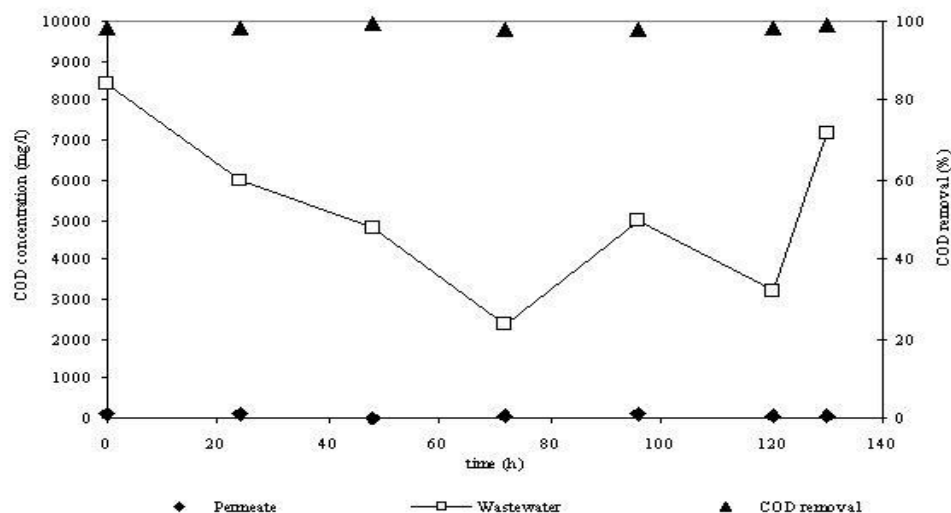
In this experiment, the lab-scale Ac-eMBR was operated in continuous mode. The feed was continuously added to the Ac-eMBR and the permeate was continuously sucked and placed in permeate tank. In this experiment, the initial biomass concentration was  $\pm 8.000$  mg/l. Figure 12 shows the relationship between flux and pressure in continuous mode. The flux and TMP are in the range of 5-15 l/m<sup>2</sup>.h and 0.1-0.35 bar, respectively. A reasonably high and relatively stable flux is obtained for more than 90 hours operation time. Vigneswaran, *et al.* [118] also has mentioned that the performance of membrane combined with adsorption process is influenced by reactor configuration, mode of operation, carbon dosage, adsorption, and influent characteristics.



**Fig. 12. Relationship between flux and pressure for Ac-eMBR in continuous mode**

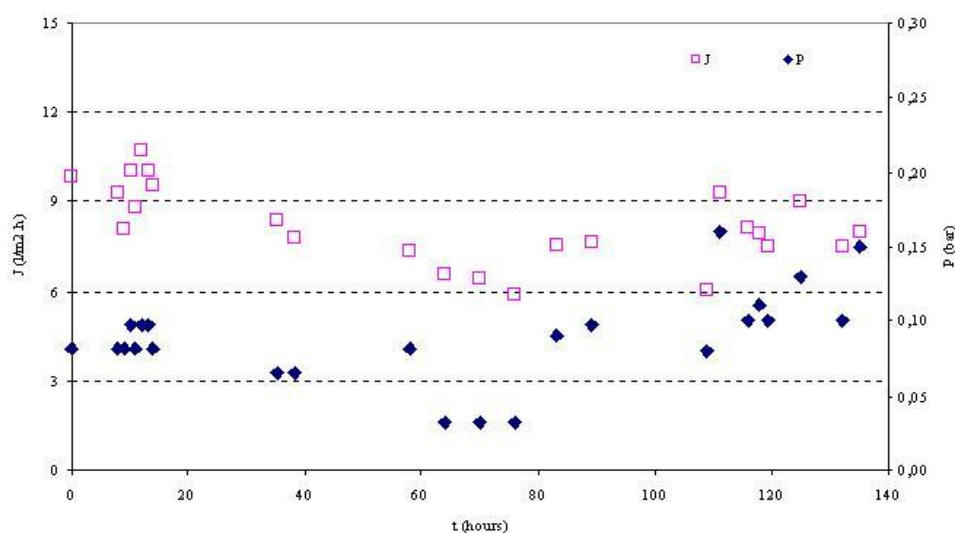
Meanwhile, even though the resulted flux is dropped to 5 l/m<sup>2</sup>.h, the rejection characteristic was excellent. The COD concentration in permeate are always below 100 mg/l at fluctuated feed concentration as shown in Figure 14. The average COD removal were approximately 98% with hydraulic retention time less than 24 hours.





**Fig. 13. Rejection characteristics for Ac-eMBR in continuous mode**

Pilot-plant test was also performed for more than 100 hours operating time. During the pilot-plant test, the flux is relatively high ranging from 6-11 l/m<sup>2</sup>.h, with operating pressure ranging from 0.03-0.17 bar as shown in Fig. 14. It is likely that the ends-free module type and the presence of activated carbon can significantly reduce membrane fouling. The fluxes are stable and the operating pressures are only slightly increased for almost 140 hours operation time. According to Fang *et al.* [91], the addition of activated carbon may reduce the film resistance because of its capacity to absorb EPS as one of the main source of fouling in MBR. Guo *et al.* (2006), also has mentioned the role of activated carbon as pre-adsorption of dissolved organic substance which reduce the membrane fouling and maintain consistent permeate flux. Meanwhile, Li *et al.* (2005), stated that in long-term operation, the membrane fouling could be reduced effectively by adding PAC, and operating intervals could be extended about 1.8 times compared to the normal activated sludge system. The total resistance was also 40% lower than that of the activated sludge system. In the meantime, the COD is successfully removed in which the COD permeate were below 50 mg/l. These results demonstrated the feasibility of the Ac-eMBR at larger scale. The removal of small molecular weight compound is due to the biodegradation by bacteria grown on PAC particles and the slow formation of biofilm on the membrane surface. Vigneswaran *et al.* [118], has shown that the addition of PAC could keep the organic removal efficiency constant without the need for chemically cleaning the membrane for a long time.



**Fig. 14. Pilot-scale performance of Ac-eMBR**



## IX. CONCLUSION

Ultrafiltration has been recognized as competitive pre-treatment for reverse osmosis system. A system designed with an ultrafiltration as pre-treatment prior to reverse osmosis system has been referred to as an Integrated Membrane System (IMS). The application of IMS is a must for sites require very extensive conventional pre-treatment or where wide fluctuation of raw water quality is expected. However, the UF design was generally dismissed as commercial alternative to conventional filtration due to its high membrane cost. Nevertheless, today, the UF membrane price has gone far down, even below conventional treatment system with the new coming Asian membrane industries. Therefore, there is no doubt, UF is now becoming a competitive pretreatment system for RO in a wide range of raw water quality. Meanwhile, the application of membrane to replace secondary clarifier of conventional activated sludge, known as membrane bioreactor (MBR), has also led to a small footprint size of treatment with excellent effluent quality. The use of MBR eliminates almost all disadvantages encountered in conventional wastewater treatment plant such as low biomass concentration and washout of fine suspended solids. Today, there are more than 1000 installations of MBR all over the world. However, fouling still become a main drawback. To minimize membrane fouling, a new configuration of submerged membrane bioreactor for aerobic industrial wastewater treatment has been developed. In this configuration, a bed of porous particle is applied to cover the submerged ends-free mounted ultrafiltration membrane into which a new configuration is made. Membrane performance was assessed based on flux productivity and selectivity. A reasonably high and stable flux around  $11 \text{ l/m}^2\cdot\text{h}$  was achieved with COD removal efficiency of more than 99% from wastewater containing high organic matter. The fouling analysis also show that this newly configured ends-free membrane bioreactor exhibit lower irreversible resistance compared with the submerged one. The performance of pilot scale system, with  $10 \text{ m}^2$  of membrane area in a 120 L tank volume, was also studied. The resulting flux from the pilot scale system around  $8 \text{ l/m}^2\cdot\text{h}$  with COD removal of more than 99%. Hence, this study has demonstrated the feasibility of the newly configured submerged ends-free MBR at larger scale.

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