Chapter 1

Fundamentals of Pressure-Driven Membrane Separation Processes

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1.1 INTRODUCTION

Membrane processes are one of the fastest growing and fascinating fields in separation technology. Even though membrane processes are a relatively new type of separation technology, several membrane processes, particularly pressure-driven membrane processes including reverse osmosis (RO), nanofiltration (NF), ultrafiltration (UF), and microfiltration (MF), are already applied on an industrial scale to food and bioproduct processing.

The concept of membrane processes is relatively simple but nevertheless often unknown. Membranes (lat.: membrana = thin skin) might be described as conventional filters (like a coffee filter) but with much finer mesh or much smaller pores to enable the separation of tiny particles, even molecules! In
general, one can divide membranes into two groups: porous and nonporous. The former group is similar to classical filtration with pressure as the driving force; the separation of a mixture is achieved by the rejection of at least one component by the membrane and passing of the other components through the membrane (see Fig. 1.1). However, it is important to note that nonporous membranes do not operate on a size exclusion mechanism. It should be pointed out that this chapter focuses on pressure-driven membrane processes using porous membranes for its close relevance to food and bioproduct processing.

Membrane separation processes can be used for a wide range of applications and can often offer significant advantages over conventional separation such as distillation and adsorption since the separation is based on a physical mechanism. Compared to conventional processes, therefore, no chemical, biological, or thermal change of the component is involved for most membrane processes. Hence membrane separation is particularly attractive to the processing of food, beverage, and bioproducts where the processed products can be sensitive to temperature (vs. distillation) and solvents (vs. extraction).

1.2 PROCESSES

1.2.1 Process Classification

There are four major pressure-driven membrane processes that can be divided by the pore sizes of membranes and the required transmembrane pressure (TMP): MF (0.1–5 μm, 1–10 bar), UF (500–100,000 Da, 1–100 nm, 1–10 bar), NF (100–500 Da, 0.5–10 nm, 10–30 bar), and RO (<0.5 nm, 35–100 bar). Figure 1.2 presents a classification on the applicability of different membrane separation processes based on particle or molecular sizes. RO process is often used for desalination and pure water production, but it is the UF and MF that are widely used in food and bioprocessing.
While MF membranes target on the microorganism removal, and hence are given the absolute rating, namely, the diameter of the largest pore on the membrane surface, UF/NF membranes are characterized by the nominal rating due to their early applications of purifying biological solutions. The nominal rating is defined as the molecular weight cut-off (MWCO) that is the smallest molecular weight of species, of which the membrane has more than 90% rejection (see later for definitions). The separation mechanism in MF/UF/NF is mainly the size exclusion, which is indicated in the nominal ratings of the membranes. The other separation mechanism includes the electrostatic interactions between solutes and membranes, which depends on the surface and physiochemical properties of solutes and membranes.

### 1.2.2 Definitions

In contrast to Figure 1.1, real membrane separations split the feed mixture stream into two streams with different compositions as shown in Figure 1.3.

The feed stream \( \dot{m}_F \) to a membrane module is split into (i) the retentate stream \( \dot{m}_R \), which is the stream that has been retained by the membrane
containing both the material that has been rejected by the membrane and a quantity of material that would not be rejected by the membrane but has yet not been given the opportunity to pass through the membrane; and (ii) the permeate stream $m_P$, the stream that has passed through the membrane, containing much less or no bigger molecules or particles than the pores.

Like any separation processes, the membrane separation processes can be evaluated by two important parameters, efficiency and productivity. The productivity is characterized by the parameter permeate flux, which indicates the rate of mass transport across the membrane. In general terms, the local mass transport of a component $i$ through a membrane element is related to its concentration on the feed side $C_{Ri}$ and the permeate side $C_{Pi}$ (see Fig. 1.3). The flow of a component $i$ through a membrane element can be referred to as its flux $J_i$. This flux is a velocity and is commonly expressed in [kg/(m$^2$ s)] or [kmol/(m$^2$ s)]. When $n$ components are permeating through the membrane a total flux $J_{\text{tot}}$ can be defined as:

$$J_{\text{tot}} = \sum_{i=1}^{n} J_i \quad (1.1)$$

The retention factor $R_i$ of a component $i$ can be defined and used as a measure of performance.

$$R_i = 1 - \frac{C_{Pi}}{C_{Ri}} \quad (1.2a)$$

where $C_P$ and $C_R$ are the concentration of component $i$ in the permeate and the retentate.

Actually pressure-driven membrane processes can be operated in two different modes: dead-end and cross-flow operations. In the dead-end mode, one stream of the feed enters the membrane module and flows vertically toward the membrane surface, and only one stream leaves the membrane module. In the cross-flow mode, one stream of the feed flows tangentially to the membrane surface, and there are two streams leaving the membrane module with one for the retentate flow and the other for the permeate flow (as shown in Fig. 1.3). The dead-end mode is employed mostly in MF for clarification and sterilization, where the feed is relatively clean. In most applications, the accumulation of the rejected particles or molecules is so severe that dead-end operation becomes impractical and cross-flow operation has to be adopted. The tangential flow in the cross-flow mode can help to shear away the accumulated rejected species at the membranes, limit the heights of cake layers, and hence maintain the permeate flux. The schematic diagrams of the dead-end mode and the cross-flow mode, and their effects on the permeate flux and the height and resistance of the cake layer, are shown in Figure 1.4.

In most applications dealing with aqueous solutions in food and bioproduct processing, the solvent permeate is largely water and permeate flux is often conveniently presented as [m$^3$/(m$^2$ s)], i.e., m/s or [L/(m$^2$ h)], LMH,
which is the volume of $m_p$ produced per unit of membrane area per unit time. Usually there is only one species, microparticle or macromolecule, to be interested, and the rejection will only be referred to the concerned species. Often the permeate flow rate is much less than the retentate flow rate in a single pass, hence the change of concentration in the retentate is not significant. The rejection can then be conveniently calculated by:

$$R = 1 - \frac{C_p}{C_F}$$  \hspace{1cm} (1.2b)

where $C_F$ is the feed concentration.

The driving force in pressure-driven membrane separation is of course the pressure, or the pressure difference between the upstream and the downstream of the membrane, or between the feed and the permeate. This is referred to as transmembrane pressure. As the pressure may vary in the membrane module due to crossflow, an averaged pressure difference over the module is used:

$$\text{TMP} = \frac{(P_F - P_p)_\text{in} + (P_F - P_p)_\text{out}}{2}$$  \hspace{1cm} (1.3)

1.3 MEMBRANES

1.3.1 Membrane Structures

Porous membranes can be divided into two types according to their structures: microporous membranes and asymmetric membranes. Microporous membranes are characterized by the membrane pores throughout the membrane
bodies. The pores are of uniform size (isotropic) or nonuniform size (anisotropic). Microporous membranes are designed to reject all the species above their ratings. However, they tend to be blocked by the species that are of similar sizes as the pores. The asymmetric membrane has a selective skin layer on the top of its membrane body. The membrane body is usually void, only giving mechanical support to the selective skin layer. Compared to the microporous membranes, the asymmetric membranes rarely get blocked. Most UF, NF, and RO membranes are of asymmetric structure, while most polymeric MF membranes are of microporous structure.

1.3.2 Membrane Materials

In terms of materials, membranes can be classified into polymeric or organic membranes and ceramic or inorganic membranes. Organic membranes are usually made up of various polymers, among which the typical ones are cellulose acetate (CA), polyamide (PA), polysulfone (PS), polyethersulfone (PES), polyvinylidene fluoride (PVDF), polypropylene (PP), etc. Polymeric membranes are relatively cheap, easy to manufacture, available in a wide range of pore sizes, and they have been widely used in various industries. Nevertheless, most of the polymeric membranes have limitations on one or more operating conditions (either pH, or temperature, or pressure, or chlorine tolerance, etc.), which hinder their wider applications. For example, CA is the classic material usually used to produce the skinned membranes. However, it has many disadvantages such as low temperature limit (30−40 °C), narrow pH range (2−8, preferably 2−6), and low chlorine tolerance (under 1 mg/L free chlorine).

Inorganic membranes have been commercialized since the early 1980s. Due to their obvious advantages of high mechanical strength, and chemical and thermal stability over the conventional polymeric membranes, they have extended the application of membrane technology into many new areas. Inorganic membranes (such as γ-alumina/α-alumina, borosilicate glass, pyrolyzed carbon, zirconia/stainless steel, or zirconia/carbon) have strong tolerance to even extreme operating conditions. For instance, they have wide limits of temperature, pH, and pressure, and have extended lifetime. However, inorganic membranes are very brittle, so the membranes can be easily damaged by dropping or unduly vibrating. Additionally the availability of such membranes is only limited to mostly UF membranes and MF membranes today. In addition, cost is the biggest disadvantage in the applications of inorganic membranes. They are far more expensive than polymeric membranes.

1.3.3 Membrane Modules

Membrane module is the way the membrane is arranged into devices and hardware to separate the feed stream into permeate and retentate streams. So far, there are four kinds of membrane modules that have been widely used in
industry. They are (1) tubular modules, (2) hollow fiber modules, (3) flat sheet modules, and (4) spiral-wound modules. These membrane modules are designed and developed by industry manufacturers in order to achieve different characteristics on the hydrodynamic conditions, filtration areas, energy consumptions, etc.

Tubular modules are composed of a number of membrane tubes assembled in a shell-and-tube arrangement. The membrane tubes are usually made up of porous fabric or plastic support with selective membranes on the inside. The internal diameters of the tubes generally range from 5 to 25 mm, and the tube lengths are in the range of 0.6 to 6 m. Tubular modules have some important characteristics: (1) due to their large internal diameters, tubular modules are capable of dealing with the feed stream containing fairly large particles. Furthermore, they can be easily cleaned by using either mechanical or chemical cleaning methods; (2) they need large pumping capacity, because they are usually operated under the turbulent flow conditions with the Reynolds numbers greater than 10,000; (3) they have the lowest surface area-to-volume ratio among all the four membrane configurations. The holdup volumes of tubular modules are also high, which need large floor space to operate.

Hollow fiber modules are actually the “thin” tubular membranes in compact modules, but in the form of self-support that enables them to withstand high backpressure. Normally, hollow fiber modules are composed of 50–3000 individual hollow fibers, bundled and sealed together on each end with epoxy in a hydraulically symmetric housing. The fiber diameters typically range from 0.2 to 3 mm (except those used in RO, which may be as thin as 0.04 mm and can withstand much higher pressure). The fiber lengths range from 18 to 120 cm. In MF and UF, hollow fiber modules are operated in the inside-out mode with selective skin layers on the inner sides of the fibers, while in RO, they are operated in the outside-in mode with selective skin layers on both sides of the fibers. Hollow fiber modules have some very different characteristics from tubular modules: (1) they are recommended to operate with the Reynolds numbers in the range of 500–3000, therefore, most of them are run in the laminar flow region. Additionally, the pressure rating of hollow fiber modules is low, normally with a maximum of 2.5 bar; (2) due to the combination of low cross flow rate and low pressure drop, hollow fiber modules are one of the more economical modules in terms of energy consumption; (3) hollow fiber modules have the highest surface area-to-volume ratio among all the four membrane configurations, and their holdup volumes are low; (4) because the fibers are self-supported, hollow fiber modules have good backwash capacity and are hence easy to clean; and (5) one distinct disadvantage of hollow fiber modules is that, their thin fibers are susceptible to get blocked by the feed with large particles, when they are operated in the inside-out mode. Therefore, the pretreatment to reduce particle size to 100 μm is usually required for hollow fiber modules.
Flat sheet modules comprise a selective flat sheet membrane on the top and a flat plate at the bottom, between which a net-like material is placed to provide space for the permeate removal, and on the other side of the flat plate, another sheet membrane and another net-like material are placed in mirror to form a sandwich-like module. Flat sheet modules have channel gaps ranging from 0.5 to 10 mm and are of lengths ranging from 10 to 60 cm. The superficial Reynolds numbers for flat sheet modules are in the laminar flow region; however, good mixing can be achieved when a screen is placed in the feed channel. The pretreatment to 150 μm is recommended for flat sheet modules. With regard to packing density, energy consumption, and cost, flat sheet modules lie in between tubular modules and spiral-wound modules.

The design of spiral-wound modules is similar to that of flat sheet modules. In the spiral-wound modules, two membrane sheets are separated by a mesh-like spacer with the active membrane sides facing away. Three edges of the two membrane sheets are glued together with the fourth edge open to a perforated center tube for the permeate removal. On the other two sides of “the envelope,” another two mesh-like spacers with thicknesses in the range of 0.56–3 mm are placed as the feed channel spacers. The whole assembly is rolled around the perforated center tube in a spiral configuration. The characteristics of spiral-wound modules are as follows: (1) spiral-wound modules are operated in the turbulent flow region because of the presence of feed spacers; (2) due to the additional drag generated by feed spacers, the pressure drop in spiral-wound modules is relatively high; (3) spiral-wound modules have fairly high surface area–volume ratio and are the lowest in terms of capital cost, among all the four kinds of membrane modules; and (4) suspended particles can easily block the mesh-like spacers and then partially block the feed channel. Therefore, spiral-wound modules require relatively clean feed that are with minimum content of suspended particles. The pretreatment to reduce suspended particles is needed for spiral-wound modules.

1.4 OPERATION

1.4.1 Concentration Polarization

Concentration polarization refers to the reversible accumulation of rejected molecules close to the membrane surface. In membrane processes all components in the feed are transported to the membrane surface by convection, and the rate increases as the permeation through the membrane increases. The selectivity of the membrane holds back the less permeable components. At steady state, these less permeable components have to be transported back into the bulk of the feed stream. As the flow next to the membrane surface is laminar, this transport can only be diffusive. The transport has to be based on the established concentration gradient, i.e., an enrichment of the less
permeable components at membrane surface, as shown in Figure 1.5. It is a natural consequence of membrane selectivity and is equivalent to the mass transfer boundary. If driving force is removed, permeation ceases, and such a concentration polarization phenomenon disappears.

Under steady-state conditions, the following relationships describe the relevant fluxes based on Figure 1.5:

Component 1

\[ J_{1,\text{con}} = J_1 \] (1.4)

Component 2

\[ J_{2,\text{con}} = J_{2,\text{diff}} + J_2 \] (1.5)

The following assumptions are made to obtain a mass balance on the feed side of the membrane:

- the process is steady state,
- the diffusion is described by Frick’s law,
- there is no chemical reaction,
- the concentration gradient parallel to membrane can be neglected,
- the density is constant, and
- the coefficient is independent from the solute concentration.

Hence in general,

\[ J_i \cdot C_i = J_{i,p} \cdot C_{i,p} - D_{ji} \frac{dC_i}{dz} \] (1.6)
Integration of Equation (1.6) taking the following boundary conditions into account,

\[
\begin{align*}
  z &= 0 \quad C_i = C_{i,M} \\
  z &= l_b \quad C_i = C_{i,b}
\end{align*}
\]

yields

\[
J_i = \left( \frac{D_{ji}}{l_b} \right) \cdot \ln \left( \frac{C_{i,M} - C_{i,P}}{C_{i,b} - C_{i,P}} \right)
\]  \hspace{1cm} (1.7)

In Equation (1.7), the term \((D_{ji}/l_b)\) can be described as a mass transfer coefficient \(k_{i,b}\).

For one interested species to be rejected and the solvent to be just water, Equation (1.7) can be rewritten as:

\[
J = k \cdot \ln \left( \frac{C_M - C_P}{C_B - C_P} \right)
\]  \hspace{1cm} (1.8)

and for a total rejection operation where \(C_P = 0\), we have

\[
J = k \cdot \ln \left( \frac{C_M}{C_B} \right)
\]  \hspace{1cm} (1.9)

According to Equation (1.9), a higher mass transfer coefficient, \(k\), and a higher membrane surface concentration lead to a higher permeate flux.

Equation (1.9) helps to explain the commonly observed UF behaviors as shown in Figure 1.6. In the pressure control region, the increase in TMP increases flux, leading to a higher \(C_M\). But if \(C_M\) reaches a certain value of...
macromolecule’s gelation concentration or the solubility of the rejected salt, gelation or salt precipitation occurs and $C_M$ reaches its maximum value. Further increase in TMP does not have any effect on $C_M$, and hence the flux, $J$, does not change — a region known as pressure-independent region.

On the other hand, increase in the mass transfer coefficient, $k$, by increasing cross-flow velocity leads to a higher permeate flux, as indicated in Equation (1.9).

The mass transfer coefficient can be estimated on the basis of heat and mass transfer analogy (so-called Colburn analogy) using the semiempirical Sherwood correlation. This correlation can be written as:

$$\text{Sh} = a \cdot \text{Re}^b \cdot \text{Sc}^c \cdot \left(\frac{d}{T}\right)^d$$  \hspace{1cm} (1.10)

The adjustment of the variables $a$, $b$, $c$, $d$ in Equation (1.10) is based on the flow regime and the module (see Tables 1.1 and 1.2).

<p>| TABLE 1.1 Variables $a$, $b$, $c$, $d$ for the Sherwood Correlation, Equation (1.10) |
|---------------------------------|----------------|----------------|----------------|----------------|</p>
<table>
<thead>
<tr>
<th>Flow regime</th>
<th>$a$</th>
<th>$b$</th>
<th>$c$</th>
<th>$d$</th>
<th>System</th>
</tr>
</thead>
<tbody>
<tr>
<td>Laminar$^1$</td>
<td>1.62</td>
<td>0.33</td>
<td>0.33</td>
<td>0.33</td>
<td>Hollow fiber</td>
</tr>
<tr>
<td>Turbulent$^2$</td>
<td>0.04</td>
<td>0.75</td>
<td>0.33</td>
<td>$-$</td>
<td>Tubular</td>
</tr>
<tr>
<td>Laminar$^3$</td>
<td>1.615</td>
<td>0.33</td>
<td>0.33</td>
<td>0.33</td>
<td>Flat sheet</td>
</tr>
<tr>
<td>Turbulent$^4$</td>
<td>0.026</td>
<td>0.8</td>
<td>0.3</td>
<td>$-$</td>
<td>Flat sheet, tubular</td>
</tr>
</tbody>
</table>

$^1$Re < 2100 hydrodynamic fully developed profile, not fully developed concentration boundary layer.  
$^2$Re < 10,000.  
$^3$Re < 2300 hydrodynamic fully developed profile, not fully developed concentration boundary layer.  
$^4$Re > 2300.

<table>
<thead>
<tr>
<th>TABLE 1.2 Dimensionless Numbers</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reynolds number:</td>
</tr>
<tr>
<td>Sherwood number:</td>
</tr>
<tr>
<td>Schmidt number:</td>
</tr>
<tr>
<td>Hydraulic diameter:</td>
</tr>
<tr>
<td></td>
</tr>
</tbody>
</table>

$d_h$, equivalent hydraulic diameter; $D$, diffusivity of the rejected species; $\rho$, density of the feed solution; $\mu$, viscosity of the feed solution; $\nu$, kinematic viscosity of the feed solution.
1.4.2 Membrane Fouling

Fouling is generally defined as a process resulting in a loss of performance of a membrane due to deposition of suspended or dissolved substances onto its external surface. Fouling cannot be removed simply by stopping the filtration process. Fouling is often the main limitation to the successful membrane application of food and biotech industries.

Fouling can be seen as a reduction in the active area of the membrane and leads therefore to a reduction in flux below theoretical capacity of the membrane. Several parameters influence the fouling rate such as:

- nature and concentration of solutes and solvents,
- membrane type,
- pore size distribution,
- surface characteristics and material of membranes, and
- hydrodynamics of membrane module.

Fouling can be related to different modes such as adsorption, chemical interactions, cake formation, and pore blocking by particles. These modes can lead to blockage or partial blockage of the active membrane area or to deposition of a layer onto the membrane surface. In Table 1.3, examples of foulants in membrane processes are given.

<table>
<thead>
<tr>
<th>Foulants</th>
<th>Fouling mode</th>
</tr>
</thead>
<tbody>
<tr>
<td>Large suspended particles</td>
<td>Particles present in the original feed or developed in the process by scaling can block module channels.</td>
</tr>
<tr>
<td>Small colloidal particles</td>
<td>Colloidal particles can rise to a fouling layer. Fouling of membranes in recovery of cells from fermentation broth.</td>
</tr>
<tr>
<td>Macromolecules</td>
<td>Gel or cake formation on membrane. Macromolecular fouling within the structure of porous membranes.</td>
</tr>
<tr>
<td>Small molecules</td>
<td>Some small organic molecules tend to have strong interactions with plastic membranes (e.g., antifoaming agents such as polypropylene glycols used during fermentation foul certain plastic ultrafiltration membranes).</td>
</tr>
<tr>
<td>Proteins</td>
<td>Interactions with surface or pores of membranes.</td>
</tr>
<tr>
<td>Chemical reactions</td>
<td>Concentration increase and pH increase can lead to precipitation of salts and hydroxides.</td>
</tr>
<tr>
<td>Biological</td>
<td>Growth of bacteria on the membrane surface and excretion of extracellular polymers.</td>
</tr>
</tbody>
</table>
Generally speaking, four fouling mechanisms for porous membranes can be observed, as shown in Figure 1.7:

(a) complete pore blocking,
(b) internal pore blocking,
(c) partial pore blocking, and
(d) cake filtration.

The following differential equation can be used to describe the influence of fouling on the flux through the membrane in the absence of any cross-flow effect

\[
J = J_0 \left[ 1 + K \cdot (2 - n) (A \cdot J_0)^{(2 - n)} \cdot (2 - n) \right]^{1/(1 - 2 - n)}
\] (1.11)

In this equation, the phenomenological coefficients \( n \) and \( K \) depend on the fouling mechanism. In Table 1.4, the different values of \( n \), their phenomenological background, their effect on the mass transport, and the relevant transport equations are given.

1.5 PREDICTION AND ENHANCEMENT OF PERMEATE FLUX

1.5.1 Flux Prediction Models

Numerous different models have been developed to predict the permeate flux as a function of operating parameters, membrane properties, and feed properties in UF. However, due to the limitations of application conditions and model assumptions, or not enough understanding of the phenomena that take place around the membrane surface, no model so far is universally applicable or fully satisfactory. Nevertheless, these models can help to understand the operation and performance links.

The pore model is applied to predict the permeate flux in the pressure control region, under the conditions of no fouling and negligible concentration polarization. In this model, it can be assumed that (1) membrane pores
<table>
<thead>
<tr>
<th>Fouling Mechanism</th>
<th>$n$</th>
<th>Phenomenological Background</th>
<th>Effect on Mass Transport</th>
<th>Transport Equation</th>
</tr>
</thead>
<tbody>
<tr>
<td>Complete pore blocking (see Fig. 1.7a)</td>
<td>2</td>
<td>Particles larger than the pore size; the active membrane area (pores) reached by particles is blocked.</td>
<td>Reduction of the active membrane area. Depending on feed velocity, permeate might be increased by increasing transmembrane driving force (pressure).</td>
<td>$J = J_0 \cdot K_b A \cdot t$ (1.12)</td>
</tr>
<tr>
<td>Internal pore blocking (see Fig. 1.7b)</td>
<td>1.5</td>
<td>Particles smaller than pore size enter the pores and get either adsorbed or deposited in the pore. Reduction in pore volume leads to blinding of pores.</td>
<td>Increase in membrane resistance due to pore size reduction. Internal pore blocking is independent from feed velocity. No limiting might be observed, $J_{\text{min}} = 0$.</td>
<td>$J = J_0 \cdot \left[1 + \frac{1}{2} \cdot K_s \cdot (A \cdot J_0)^{0.5} \cdot t \right]^{-2}$ (1.13)</td>
</tr>
<tr>
<td>Particle pore blocking (see Fig. 1.7c)</td>
<td>1</td>
<td>Any particles reaching a pore might seal it over time. Particles might bridge a pore and not block it completely.</td>
<td>Reduction of active membrane area. The effect is similar to pore blocking but not so severe.</td>
<td>$J = J_0 \cdot \left[1 + K_i \cdot (A \cdot J_0) \cdot t \right]^{-1}$ (1.14)</td>
</tr>
<tr>
<td>Cake filtration (see Fig. 1.7d)</td>
<td>0</td>
<td>Formation of a cake on the membrane surface of particles that do not enter the pores.</td>
<td>The overall resistance becomes the resistance of the membrane plus the resistance of the cake.</td>
<td>$J = J_0 \cdot \left[1 + 2 \cdot K_c \cdot (A \cdot J_0)^{2} \cdot t \right]^{-1/2}$ (1.15)</td>
</tr>
</tbody>
</table>
are ideal cylindrical channels and are uniformly distributed on the membrane surface; (2) the permeate passing through membrane pores are laminar flow in the steady state; and (3) the applied feed is of constant density (e.g., incompressible) and Newtonian (with no dependence on the shear rate). The permeate flux can be calculated using the Hagen–Poiseuille equation, which is based on the momentum balance:

\[ J_v = \frac{n_p d_p^4 \Delta P}{128 \mu l_p} \]  

(1.16)

where \( J_v \) is the permeate flux, \( n_p \) the number of cylindrical pores per unit area, \( d_p \) the pore diameter, \( \Delta P \) the TMP, \( l_p \) the pore length, and \( \mu \) the viscosity of the permeate.

The equation shows that the permeate flux is directly proportional to the TMP and inversely proportional to the viscosity. The viscosity is primarily controlled by the solvent type, feed composition, and temperature. Therefore, in the pressure control region, increasing the temperature and pressure, and decreasing the feed concentration can increase the permeate flux.

The resistance model is developed to express the entire TMP-flux behavior in MF and UF, both in the pressure control region and in the mass transfer control region. This model is based on the resistance-in-series concept, which is a common concept in heat transfer. With the ideal membrane and the ideal feed that lead to no fouling, the model can be expressed as:

\[ J_v = \frac{\Delta P}{\mu R_m} \]  

(1.17)

where \( J_v \) is the permeate flux, \( \Delta P \) the TMP, \( \mu \) the viscosity of the permeate, and \( R_m \) the hydraulic resistance of the membrane, which is a constant value for each membrane and can be determined by measuring the pure water flux.

By consideration of the effect of concentration polarization, the increase in solute concentrations near the membrane surface results in the increase in osmosis pressure, which effectively reduces the TMP. The permeate flux is then calculated by:

\[ J_v = \frac{\Delta P - \Delta \pi}{\mu R_m} \]  

(1.18)

where \( \Delta \pi \) is the osmosis pressure difference between the solution at membrane surface and the permeate. Noting \( \Delta \pi \) is dependent on \( C_m \), Equation (1.18) can be used in conjunction with Equations (1.9) and (1.10) to calculate permeate flux in UF and NF if the osmotic pressure dependence on solute concentration is given.

In the filtration of real feeds, both concentration polarization and membrane fouling occur to add additional resistances to the membrane and hence to the permeate to pass through. Therefore, the resistance of the polarized
layer $R_p$ and the fouling resistance caused by the physiochemical interactions between the solutes and the membrane $R_f$ need to be taken into account. The equation is expressed as follows:

$$J_v = \frac{\Delta P - \Delta \pi}{\mu(R_m + R_p + R_f)}$$

(1.19)

In MF, the rejected particle exerts insignificant osmotic pressure, and the concentration layer resistance and fouling layer resistance can be combined into one parameter — the cake resistance.

1.5.2 Flux Enhancement and Fouling Control

Flux enhancement can mean one of the following:
- Increase permeate flux
- Increase flux with same energy consumption
- Decrease energy consumption while maintaining same flux.

Fouling control is an important issue in order to maintain an acceptable flux level. In addition, chemical cleaning represents a significant fraction of operational cost of membrane process and downtime too. It is generally accepted that fouling is a fact of life in practical operation of membrane processes, but fouling can be controlled to an acceptable level with better understanding of the process operation and the feed characterization.

To control fouling, different approaches have been developed and used separately or in combination:

(i) hydrodynamic management,
(ii) back flushing and pulsing,
(iii) membrane surface modification,
(iv) feed pretreatment,
(v) flux control, and
(vi) effective membrane cleaning.

*Hydrodynamic management* aims at promoting local mixing close to the membrane surface and enhancing the back diffusion of the rejected molecules or particles. In doing so, the concentration at the membrane surface is reduced, also leading to reduced adsorption of the molecules (adsorption isotherm). Simply increasing cross-flow velocity leads to limited effect but much higher energy consumption. Instead much effort has been directed to increase local turbulence or mixing close to the membranes using various methods to introduce flow instability and secondary flow. These include turbulence promoters, corrugated membrane surface, various spacers, vibrating/rotational membranes, reverse or pulsatile flow, use of sponge balls, gas bubbles, etc. It has been demonstrated that these techniques generate secondary flows which improves local mixing and therefore improves mass transfer.
Back flushing or pulsing is an approach to remove cake layers on the feed side and, therefore, reduce the influence of fouling. It is carried out by reversing the flow of the permeate through the membrane and, therefore, dislodges the foulant and reestablishes the flux at a high level (Fig. 1.8). In order to maintain a high flux, back flushing is carried out periodically and requires module types with a high-pressure resistance, e.g., pressure stable capillary modules. Such back flushing can be carried out with rather high frequency in a very short period of time, which may be termed as back pulsing. Back flushing or pulsing is widely used in MF of high solid content feed or feed with high fouling tendency.

Membrane surface modification and feed pretreatment both act to alter the interactions between the filtered molecules or microparticles and the membrane surface. As such interactions are often dominated by electrostatic interactions and nonspecific interactions, membrane surface modification often focuses on introducing charge groups and increasing hydrophilicity. Feed pretreatment can adjust the pH and salt concentration to alter the charge effect, but more importantly can charge the particle size distribution by promoting or demoting aggregation and hence improve the “filterability,” i.e., reducing fouling tendency.

Flux control is based on the understanding of critical flux, a level of flux under which fouling is minimal. This flux is defined as a critical flux. The critical flux hypothesis is that if flux is controlled on start-up of a membrane operation, there exists a critical flux below which a decline in flux over time does not occur. The region of operating under which no fouling is found is termed subcritical. Experiments showed that MF can be operated at a constant flux with no increase in transmembrane pressure if the operating pressure is low. Fouling is slight or negligible. This can be understood by analyzing the forces acting onto the rejected particles (or macromolecules). The convective force due to permeation pushes the particle toward the membrane pore; diffusional force drives the particle away from the membrane due to concentration gradient and more importantly velocity.
gradient-induced diffusion. Electrostatic interaction also repels the particles. Whether the particle arrives at the membrane surface depends on the balance of the forces. At a higher flux, the particle will arrive at the membrane at the surface; and fouling may occur. Below the critical flux, it will not. The concept of critical flux helps to understand fouling and to guide the operation in theory, but difficulty appears in practical applications as (i) its value may be too low to be practically applied and (ii) it cannot be predicted largely because the feed is often a complex mixture.

Chemical cleaning is required (i) as often fouling is inevitable and (ii) as an integrated part of regulatory requirement for food and bioprocessing (cleaning in place). However, chemical cleaning, including the selection of cleaning agents and formulation, the operational procedure, is largely based on experience, and the outcome cannot be predicted with confidence. This is because of the lack of fundamental understanding of fouling, the complexity of foulant composition and fouling process, and poor characterization of the feed mixtures. This is an area that urgently needs research, as chemical cleaning represents significant operational cost (cost of chemicals, loss of productivity, etc.) in membrane applications in food and bioprocessing.

1.6 SUMMARY

This chapter presents some basic concepts related to membranes and membrane processes. Common issues to membrane application in food and bioprocessing are outlined. It provides a basis for detailed discussion on specific applications later in this book.

FURTHER READINGS