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Performance and cost modeling of low-pressure membrane filtration processes

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Rice University, 1994
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PERFORMANCE AND COST MODELING OF LOW-PRESSURE
MEMBRANE FILTRATION PROCESSES

by

SANDEEP SETHI

A THESIS SUBMITTED
IN PARTIAL FULFILLMENT OF THE
REQUIREMENTS FOR THE DEGREE
MASTER OF SCIENCE

APPROVED, THESIS COMMITTEE

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January, 1994
Abstract

Performance and Cost Modeling of Low-Pressure Membrane Filtration Processes

Sandeep Sethi

An integrated numerical model for examining the technical performance and estimating associated costs of low-pressure membrane filtration processes is presented. A model for removal of contaminants predicts permeate quality with respect to colloidal and organic material. A shear-induced diffusion model for estimating permeate flux is modified to also include Brownian diffusion as a particle transport mechanism. These two diffusion mechanisms lead to a minimum in permeate flux through a typical ultrafiltration membrane for particles in the size range of 0.1 μm in diameter. Lower shear rates are predicted to move this unfavorable size towards larger particle diameters. The effect of raw water quality and operating parameters on the performance and cost of typical ultrafiltration and microfiltration facilities is investigated. Particle size appears to be an important variable in determining both performance and cost. A product of this research is a software application package, MeMsys, developed for the PC-DOS environment.
Acknowledgments

Dr. Mark R. Wiesner for letting me pursue this research, for his never failing guidance, encouragement and support. Dr. Philip B. Bedient for the time and consideration to serve on my thesis committee. Dr. John E. Dennis for making me feel comfortable with not so comfortable nonlinear math, teaching me a few good numerical techniques and for being on my thesis committee. Jeff Adams and Dick Eilers at EPA for making this project possible. In particular (and in no particular order) Mae, Vasu, Faten, Shankar, Duce, Greg, Erin, Draha, Reji, John and Susan for being grand friends and colleagues and for constructive philosophical discussions at Valhalla. Bob, Vicky and Kathy for all their help and smiles. My mom for being my mom and my dad for teaching me the worth of good work. For all their love and affection. Mona for being a wonderful sister and friend. Akshay for making beer taste best in his company. Their new daughter and my niece Nikita for being the cutest thing in life. Kumud and Anurag for sharing "100 mph" rides and similar fun. Sam, Sunit and Mohit for musik, all-nighters and more. Bobby for jokes and times. Ravi for jogs and "nightwalks". Bonny for "ghosts". Pink Floyd and Bach for nourishing the right half of my brain. Asterix and Obelix for laughs. Plato for philosophy and I. Asimov for science. Bits and bytes for their binary help. And, none the least, Curiosity for walking along with me...

This material is based upon work supported in part by the Risk Reduction Engineering Laboratory, United States Environmental Protection Agency, Grant R10800
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Chapter 1

Introduction

1.1 Significance

Standards which define the quality of drinking water have seen an ever increasing regulatory impetus as human knowledge on various contaminants has progressed. The principal statute pertaining to drinking water quality set by the Safe Drinking Water Act (SDWA) of 1974 has long been revised by the amendments in 1986, which required setting standards and monitoring requirements for eighty-three contaminants, and more recently by demanding regulations such as the Disinfectants-Disinfection By-product Rule and the Surface Water Treatment Rule.

There is a need for new, more efficient separation processes, which can be used to meet these more stringent regulatory requirements. Relatively new technologies such as nanofiltration (NF), ultrafiltration (UF) and microfiltration (MF) are capable of removing a wide variety of materials and appear to offer promising new possibilities for producing water of very high quality. The efficiency of membrane filtration in terms of removal has been well established. Ultrafiltration membranes as well as nanofiltration and reverse osmosis membranes may potentially serve as absolute barriers to the passage of microbial contaminants.

Microfiltration membranes have also been shown to effectively
remove bacteria, protozoans and viruses from water. The previous decade has seen improvements in the technical efficiency of membrane systems and a growth in membrane markets. These factors have combined to potentially make membranes an economically cost competitive alternative to conventional filtration.

1.2 Nature and Scope of Problem

Membrane processes may use various types of driving forces, such as gradients in pressure, concentration, electrical potential and temperature as the basis for transporting water or contaminants. This study focuses on low-pressure membrane processes, such as ultrafiltration and microfiltration. Other pressure driven membrane processes, such as reverse osmosis (RO) have been in successful practice for desalination since many years.

Membrane separation processes employ semi-permeable membranes which restrict the passage of various chemical species in a rather specific manner. When employed in the treatment of drinking water the performance and economics of a membrane process depends on a large number of raw water quality as well as hydrodynamic parameters. The object of this research is to present an integrated mathematical model for estimating the performance and cost of low-pressure membrane filtration processes taking into account selected raw water quality parameters and the hydrodynamics of the membrane system.
The rate of permeation, a critical measure of membrane performance, plays a primary role in determining the cost of membrane filtration. As such it is required that mathematical models provide a very good estimate of the permeation rate. Models to estimate the performance of membrane filters seek to determine the accumulation of rejected materials near the membrane surface (concentration polarization layer) and on the membrane (cake or gel layer) since the resistance offered to permeation by these layers principally determines the permeate flux. Previous models have focused on a single mechanism of particle back-transport to predict the permeate flux. In this study a previous model (Romero and Davis, 1988) for shear-induced diffusion, which predicts permeate flux as a function of axial distance along the filter, is extended to account for Brownian diffusivity as a mechanism of particle transport. This model predicts that the conflicting quantitative variance of these two diffusive phenomena results in an unfavorable range of particle diameters, around 0.1 \( \mu \text{m} \), where back transport of colloids approaches a minimum. This in turn is seen to correspond to a decline in the permeate flux and consequently high filtration costs.

To determine the quality of the permeate from membrane filters, a contaminant removal model, founded on the work of Ferry (1936), has been incorporated in the performance model. The removal model is based on theoretical aspects of hindered solute transport in membrane pores. Experimental data on solute removal have been fit with the theoretical model to derive a correlation between the molecular weight of natural organic material and its effective "size". This is also considered in the light
of the morphology of the organic material.

The performance model describing the permeate flux and quality is integrated with a cost model which determines both the capital as well as the operating costs of membrane filtration. The cost model is based on the work of Pickering (1991).

1.3 Objectives

This study focuses on formulating an integrated mathematical model which predicts the physical and economic performance of low-pressure membrane filtration systems as a function of various operating and cost related factors. More specific objectives of this work were:

(1) To extend a model for contaminant transport and permeate flux based on the concept of shear-induced diffusivity to include contaminant transport by Brownian diffusion.

(2) To include a model for permeate quality. Besides predicting removal of particulate material a means to predict the removal of organic matter was also investigated.

(3) To accommodate in an approximate fashion a means of extending models for monodisperse suspensions to describe heterodisperse suspensions typically encountered in practice.

(4) To review and integrate a cost model for capital and operation costs of membrane filtration plants with the mathematical model describing the
performance of the membrane filter.

(5) To evaluate the effects of raw water quality as well as several important hydrodynamic variables on performance and cost of membrane filtration.

(6) To develop a software application package including the performance and cost models in a menu-driven user friendly environment. This software was coded in FORTRAN and runs under the PC-DOS environment.
Chapter 2

Literature Review

2.1 Membranes in Water Treatment

Membranes have been used widely in the chemical, food and beverage, medical and biotechnology industries as well as in water and wastewater treatment (Turbak, 1981). Reverse osmosis now provides almost a third of the world’s desalting capacity (Mason and Lonsdale, 1990). This process is capable of removing ionic species and its efficiency in removing various organic materials has also been well documented (Reinhard et al., 1986; Clair et al., 1991). UF and MF have been only recently introduced to water and wastewater treatment. In contrast with RO, UF and MF membranes are permeable to most dissolved species, operate at relatively lower pressures and produce higher rates of permeate flux.

The Surface Water Treatment Rule (SWTR) for potable water treatment requires

- 99.9% (3 log) or more removal-inactivation of Giardia cysts and
- 99.99% (4 log) or more removal-inactivation of enteric viruses.

In addition, for direct and conventional filtration plants a turbidity of less than 0.5 NTU must be accomplished in at least 95% of the samples collected each month and should never exceed 5 NTU in any individual
sample. Meeting these goals could necessitate improved chemical disinfection and, in some cases, addition of filtration facilities. Common and economical techniques of chemical disinfection by themselves may not be feasible from a regulatory perspective due to constraints imposed by anticipated disinfection by-product (DBP) regulations on suspected carcinogens formed from the reaction of chemical disinfectants with natural organic matter. For example, increasing chlorine dose and/or contact time is likely to result in greater production of trihalomethanes (THMs) and other DBPs. Lead and copper removal from drinking water is hoped to be achieved through optimum corrosion control. Again, a simple means to achieve this, such as by increasing the pH, may favor formation of certain DBPs. These dilemmas due to DBPs could perhaps be resolved using membrane processes which would remove as much organic matter as possible and hence minimize DBP formation. Thus, membrane technologies may be useful in resolving apparent conflicts between existing and proposed regulations.

There have been several pilot investigations (Jacangelo et al., 1989; Taylor et al., 1989; Wiesner et al., 1991; Heneghan et al., 1991; Laine et al., 1991; Jacangelo et al., 1991; Olivieri et al., 1991) representing and warranting the ability of low-pressure membrane filtration in efficient removal of particulate, organic and microbial material from source waters with moderate to high concentrations of these contaminants. Together, these studies conclude that membrane filtration processes are proficient in consistently delivering water of high quality. Consistency is inferred due to the apparent independence of the removal capability of membrane filtration
to changes in source water quality as well as operating parameters such as transmembrane flux.

Jacangelo et al. (1989), Heneghan et al. (1991) and Laine et al. (1991) examined typical UF membranes, with a MWCO of 100,000 Daltons, for removal of particulate material. In all cases effluent water turbidities observed were less than 0.1 NTU while influent water turbidities ranged up to as high as 60 NTU. In analyzing the efficacy of UF in removing micro-organisms (viruses, Giardia and coliform bacteria) of concern from untreated source waters Jacangelo et al. (1991) reported more than 6.5 log removals of virus and more than 4.0 log removals of Giardia. Based on removal data they concluded that UF was capable of meeting SWTR requirements without the use of chemical disinfection. Taylor et al. (1989) explored the promise membranes offer in trihalomethane (THM) control. From their data on THM formation potential (THMFP) in source and treated waters, an average reduction of about 95% in THMFP can be deduced. The pilot study led Taylor et al. to conclude that membrane processes can be successfully employed to control the formation of THMs even in case of source waters carrying high concentrations of organic matter.

Wiesner et al. (1991) and Olivieri et al. (1991) presented pilot plant evaluation of MF in combination with chemical addition. Observed product water turbidities ranged from 0.1-1 NTU for MF membranes with a nominal pore size of 0.2 μm. Average THMFP reduction reported in both cases was about 60%.
These investigations have thus demonstrated the effectiveness of membrane processes in the removal of contaminants for achieving drinking water standards as required under the SWTR. Indeed, membranes appear to be sufficiently effective in removing microbes that additional concerns over the formation of DBPs may encourage the use of membranes for disinfection.

2.2 Modeling

There have been various approaches to describing the dynamics of mass transfer in membrane processes. The important categories of mathematical models pertaining to fluid and solute dynamics in cross-flow pressure driven filtration processes are reviewed and summarized in the following sections.

2.2.1 Models for Predicting Permeate Flux

*Flux Decline*

The permeate flux across a membrane typically declines and/or the pressure drop increases over operating time due in part to the accumulation of rejected material near the membrane surface (concentration polarization) or on the membrane (cake or gel formation). For low-pressure membrane systems accumulated materials may act as a dynamic secondary membrane on the primary membrane structure, resulting in increased hydrodynamic resistance to solvent flow through the membrane (Porter, 1972). In most practical cases resistance of the cake is significantly greater than the
resistance of the concentration polarization layer. Permeate flux decline due to accumulated material is often reversible by means such as backflushing and/or pulsing (fastflushing) of the membrane with water and/or air. These flushing procedures along with chemical cleaning of the membranes are collectively referred to as flux enhancement. Chemical cleaning may be required to reverse the effects of adsorption or precipitation of material on membranes as well as removing bacterial films and their by-products. Reductions in permeate flux which are either hydrodynamically or chemically reversible are termed "colmatage". Irreversible decreases in permeate flux are referred to as membrane fouling.

Models for predicting permeate flux

Perhaps the simplest model for permeate flux considers the passage of a fluid through a membrane with uniformly distributed, evenly sized pores in the membrane, and negligible effects due to solute and suspended materials. Under these assumptions a good estimate of flux through a microporous membrane is given by the Hagen-Poiseuille law for streamline flow thorough channels (Cheryan, 1986). The equivalent form for flow through porous membranes is expressed as:

\[ V_w = \frac{\varepsilon r_o^2 \Delta P}{8 \mu \Delta l} \quad (2.2.1) \]

where \( V_w \) is the permeate flux, \( r_o \) is the pore radius, \( \Delta P \) is the applied transmembrane pressure, \( \mu \) is the fluid viscosity, \( \Delta l \) is the membrane skin thickness and \( \varepsilon \) is the surface porosity of the membrane. The Carmen-
Kozeny equation developed for flow through porous bed of particles and taking the tortuosity into account, can also be used to estimate the permeate flux through clean membranes (Gutman, 1987):

\[ V_w = \frac{\varepsilon^3 r_o^2 \Delta P}{150 \mu \Delta l (1 - \varepsilon)^2} \quad (2.2.2) \]

Both these models predict a linear relationship of flux with applied pressure and an inverse relationship with the viscosity. In fact, permeate flux is often found to approach a pressure independent regime at high pressures. Such "mass transport limited" flux is often described in terms of the effects of concentration polarization and cake (or gel) formation. Since concentration polarization effects have not been considered in Equations 2.2.1 and 2.2.2, the experimentally observed asymptotic relationship of permeate flux with pressure is not predicted by these models. In addition to high operating pressures, concentration polarization may be favored under conditions of high feed concentration or low cross-flow velocity. Hence, these models would hold only under the pressure controlled region of Figure 2.2.1.

Models for predicting limiting permeate flux

In practice, the permeation rate through membranes is limited by colmatage and fouling produced by materials convected to the membrane. The convective-diffusion equation serves as the governing expression for describing mass transfer in this case. Under the simplifications of an incompressible fluid, steady-state, and no axial diffusion (as it is comparatively very small in all boundary layer flows), the required
expression in two-dimensions is:

$$\frac{\partial uc}{\partial x} + \frac{\partial vc}{\partial y} = D \frac{\partial^2 c}{\partial y^2} \quad (2.2.3)$$

where $D$ is the diffusion coefficient. The solution of Equation 2.2.3 usually entails the following boundary conditions (Kleinstreuer and Belfort, 1984):

$$[c]_{y=0} = c_0 \quad (2.2.4)$$

$$[\frac{\partial c}{\partial y}]_{y=0} = 0 \quad (2.2.5)$$

$$V_w c_w = [D \frac{\partial c}{\partial y}]_{y=0} \quad (2.2.6)$$

Figure 2.2.1 Pressure controlled and mass transfer controlled behavior of permeate flux.
where \( c_o \) is the concentration at the channel inlet, \( c_w \) is the concentration at the deposited layer on the membrane wall and the permeation rate, \( V_w \), is assumed to be constant. Equation 2.2.4 implies a uniform feed concentration at the channel entrance. The boundary condition of no solute flux at the wall is represented by Equation 2.2.5. Finally, Equation 2.2.6 states that at steady-state, the convective flux of solute towards the membrane is balanced by diffusive back transport away from the membrane and towards the bulk suspension.

Equation 2.2.6 also forms the basis of one of the simplest and widely used models for permeate flux, commonly known as the "film theory model". The premise of this model is that the resistance to mass transport is due to a thin stagnant film at the membrane surface. Based on a lumped parameter approach (Brian, 1965), the mass balance over the polarized layer is expressed as:

\[
(V_w c_b) = V_s = D \frac{dc}{dx}
\]  

(2.2.7)

where \( c_b \) represents the bulk suspension concentration and \( V_s \) is the solute flux. Integration over the boundary layer leads to the following expression for permeate flux:

\[
V_w = \frac{D}{\delta} \ln \frac{c_w}{c_b} = K \ln \frac{c_w}{c_b}
\]  

(2.2.8)

where \( K \) is the mass transfer coefficient and \( \delta \) is the thickness of the concentration boundary layer. It should be noted that this model is valid
only in the pressure-independent or mass transfer controlled region (Figure 2.2.1). The mass transfer coefficient is usually calculated using the classical Lévêque correlation for laminar flow. This correlation is commonly expressed in terms of the Sherwood number $Sh$, the Reynolds number $Re$ and the Schmidt number $Sc$:

$$Sh = 1.86 Re^{0.33} Sc^{0.33} \left(\frac{d_h}{L_c}\right)^{0.33} \quad (2.2.9)$$

where $d_h$ is the hydraulic diameter of the channel, and $L_c$ is the channel length.

When Stokes-Einstein diffusivities are assumed and applied to the film theory model, it predicts permeate fluxes in colloidal ultrafiltration with sufficient accuracy, as shown by Porter (1972). However, when larger particle sizes are considered the predictions of permeate flux are highly inaccurate in that the values are as much as two orders of magnitude lower than observed. This phenomenon is commonly known as the "flux paradox for colloidal suspensions". In trying to describe solute polarization in UF it lead Blatt et al. (1970) to conclude that either:

1. "the 'back-diffusion flux' of particles from the polarized layer is, in some fashion, substantially augmented over that expected to occur by Brownian motion" or
2. "that the removal of particles" from the polarized layer "takes place by a completely different mechanism".

These two conclusions summarize the motivation for several categories of
models as discussed later in this chapter.

One model that overcomes the limitation of the previously discussed models and correctly describes the pressure-dependent as well as the pressure-independent behavior of the permeate flux is the "resistance model". Based on a resistance in series concept this model considers the resistance due to the membrane, $R_m$, the "fouling layer", $R_f$, and the polarization layers, $R_p$, as a lump sum resistance to permeation and describes the permeate flux as:

$$ V_w = \frac{\Delta P}{\mu (R_m + R_f + R_p)} \quad (2.2.10) $$

Both $R_m$ as well as $R_f$ are generally considered unaffected by operating conditions. The polarization resistance, $R_p$, however varies with the operating conditions and is dependent on the applied pressure. The resistance model is therefore seen to conceptually describe the observed behavior of the permeate flux over all magnitudes of the driving force.

The main focus of all categories of mathematical models which endeavor to predict permeate flux as a function of operating conditions and physical properties is to describe the complex phenomena arising from particle-fluid, particle-particle and/or particle-membrane interactions. Consequently, these models, which hold in the mass transfer controlled region of Figure 2.2.1, differ primarily in that they consider the different transport processes for foulant species within the membrane elements. The three most frequently considered transport mechanisms in laminar flow
are:

(1) convection of materials toward and across the membrane,
(2) inertial lift on particles and,
(3) shear-induced diffusion

Although, these mechanisms are not mutually exclusive, they are considered separately in the following three sections.

Convective models

The concept of convection as a removal mechanism was first hypothesized by Blatt et al. in 1970 and later supported by others (Schneider and Klein, 1982; Leonard and Vassilieff, 1984; Davis and Birdsell, 1987). These authors assert that for particulate suspensions, unlike for micro-molecular and macro-molecular suspensions, the predominant mechanism of removal is convective. Diffusive phenomena are assumed to play no role and consequently back-transport of particles is not considered. Instead, it is proposed that under sufficient hydrodynamic shear the particle layer begins to flow tangentially along the membrane surface at a rate which balances the convective deposition of particles, thus maintaining the steady-state thickness. The convective models of Leonard and Vassilieff (1984) and Davis and Birdsell (1987) attempt to quantify these hypotheses. In their analysis Leonard and Vassilieff assume the fluid being divided into two distinct regions; one with a fixed bulk concentration and the other with a fixed concentration corresponding to the close packing of particles in the concentrated layer. In both these regions concentration is
invariant with respect to space and time. The authors then solve the solute mass balance to obtain the boundary, $\delta(x, t)$, dividing the two regions. Their analysis leads to the following coupled equations for particle layer thickness, $\delta(x, t)$ and permeate flux, $V_w$:

$$\frac{\dot{\gamma}}{2} \frac{\delta^2}{c_w - c_b} = V_w \frac{c_b}{c_w - c_b} x$$  \hspace{1cm} (2.2.11)

$$\delta = V_w \frac{c_b}{c_w - c_b} t$$  \hspace{1cm} (2.2.12)

where $\dot{\gamma}$ is the shear rate. The model of Davis and Birdsell (1987) relaxes some of the simplifications imposed by Leonard and Vassilieff; the most important ones being the determination of parabolic instead of linear velocity profiles and the consideration of the axial variance of the permeate flux. These authors use a numerical technique involving determination of the nonlinear velocity profiles and then solving a system of equations for permeate flux, transmembrane pressure, and particle layer thickness as they vary axially along the membrane length.

The convective models described above treat the flowing layer as a Newtonian fluid. The effective viscosity of such Newtonian layers depends on the local particle concentration. However, these models do not predict that and instead assign a fixed value to the effective viscosity.

*Inertial Lift Models*

Several investigators have tried to explain the 'flux paradox' by
invoking the inertial lift on particles as a mechanism which augments Brownian back transport of colloidal particles from the membrane surface (Porter 1972).

Radially directed particle lift forces arise due to the fluid velocity difference on either side of a particle. Serge and Silberberg (1961) first reported the radial displacements of particles in a Poiseuille flow of suspensions in straight tubes. Since then there have been many studies related to estimating the lift velocity of particles due to inertial forces; Saffman (1965), Cox and Brenner (1968), Ho and Leal (1974), Vasseur and Cox (1976), Madsen (1977).

Altena and Belford (1984) adapted the lift theory to porous channels in formulating a mathematical model to describe the performance of ultrafiltration membranes. Their two-dimensional analysis, for Poiseuille flow in a slit with one porous wall, was later extended to three dimensional analysis of flow in a porous tube by Weigand et al. (1985). Both these analyses of porous channels were based on an earlier assertion by Green and Belford (1980) that the net lateral force on a particle can be expressed as the vector sum of the inertial force, calculated assuming a non-porous channel, and the permeation drag force. Thus, inertial lift effectively reduces the convective flux of particles towards the membrane.

Mathematical analyses of inertial lift assume a dilute suspension of non interacting particles, and are inapplicable near filter walls (Romero and Davis, 1988).
Shear Induced Diffusion Models

A consideration of orthokinetic phenomena, arising from particle motions under shear flow, has been undertaken by a number of investigators. Some of the earliest pioneers in the field include; A. Einstein, Von Smoluchowski, G. B. Jeffery, F.R. Eirich and G.I Taylor (Mason, 1977). Particle-Particle interactions in a flowing suspension result in lateral migrations of particles from their instantaneous trajectories. Two types of motions contribute towards this phenomenon; a rotary motion as well as a translational motion. Particle rotation leads to the creation of a velocity field that exerts a drag force on neighboring particles. Particles translating on faster streamlines overtake particles on slower streamlines and the consequent interaction of velocity fields again causes a net lateral migration to occur.

Eckstein and co-workers (1974) were able to experimentally determine an empirical expression for the self-diffusion coefficient of particles in a concentrated suspension under the conditions of a simple shear flow. The experimental technique involved observing the lateral position of radioactively labeled particles in a Couette apparatus. To exclude inertial and gravitational effects, making self-diffusion the dominant phenomenon, neutrally buoyant particles were suspended in a very viscous oil in a cylindrical Couette device with a small ratio of gap width to mean radius. By means of dimensional analysis and application of one-dimensional random walk theory to their experimental data Eckstein et al. arrived at the following expressions for the diffusion coefficient:
\[ D_{sh} = 0.02 \, a_p^2 \, \dot{\gamma} \, \phi \quad \text{0 < \phi < 0.2} \]  
\[ D_{sh} = 0.025 \, a_p^2 \, \dot{\gamma} \quad \text{0.2 < \phi < 0.5} \]  

where \( \phi \) represents the particle volume fraction. The diffusion coefficient was seen to vary linearly with volume fraction up to \( \phi = 0.2 \). At larger volume fractions, the behavior was not clear due to experimental scatter. It was found from dimensional analysis that the diffusion coefficient followed a linear relationship with shear rate and with the square of particle diameter. Eckstein et al. also concluded that within the experimental accuracy, spherical and disk-like particles had the same value for shear enhanced diffusivity.

In their work on plasmapheresis Zydney and Colton (1986) incorporated shear-enhanced diffusivity into the film theory to develop a model for predicting permeate flux during microfiltration. Using Lévéque's correlation for determining the mass transport coefficient and the value of shear-enhanced particle diffusivity from Eckstein's work, they were able to arrive at equations predicting the axial variance of particle layer thickness and flux along the membrane:

\[ \delta(x) = 0.578 \left( a_p^4 \, x \right)^{1/3} \]  
\[ V_w(x) = 0.052 \left( \frac{a_p^4}{x} \right)^{1/3} \gamma_w \ln \left( \frac{c_w}{c_b} \right) \]
The independence of $\delta(x)$ with respect to wall shear rate results from an assumed linear shear rate dependence for Eckstein's diffusivity. Assuming that the shear rate at the wall remains constant with axial position in the filter, integration of Equation 2.2.16 over the length of the membrane to express the length averaged flux yields:

$$V_w = 0.078 \left( \frac{d_p}{L_e} \right)^{1/3} \gamma_w \ln \left( \frac{c_w}{c_b} \right)$$  \hspace{1cm} (2.2.17)

Zydney and Colton found this model to be in good agreement with plasmapheresis data.

In 1987 Leighton and Acrivos (1987a) enlarged on Eckstein et al.'s (1974) work with additional work aimed at determining the coefficient of self-diffusion in concentrated suspensions. However, this time they employed a different measurement technique, since they proved that the earlier method of Eckstein violated conditions of unbounded shear field and was thus constrained by wall limitations. Their procedure involved measuring variations in the transit time taken by a single suspended particle to complete successive circuits in a Couette device. The results indicated that the diffusion coefficient was an increasing function of the volume fraction, $\phi$, and that it was approximately proportional to $\phi^2$ in the dilute limit. The linear dependence of the diffusion coefficient on concentration reported by Eckstein et al. (1974) was thus incorrect and would lead to a gross underestimation of diffusivity, particularly at higher concentrations.
Based on detailed experimental and theoretical work on shear-induced diffusion of particles in concentrated suspensions, Leighton and Acrivos (1987b) presented an explicit expression for the diffusion coefficient:

\[ D_{sh} = \dot{\gamma} a_P^2 \hat{D}_{sh}(\phi) \]  

(2.2.18)

where \( \hat{D}_{sh} \) is a dimensionless function of \( \phi \) and was determined empirically for suspensions of rigid spheres in a Couette viscometer as:

\[ \hat{D}_{sh}(\phi) = 0.33 \phi^2 (1 + 0.5 e^{8.8\phi}) \]  

(2.2.19)

This expression was obtained for particle volume fractions up to 0.5. The authors also noted that this expression should be considered approximate due to a 25-50 percent scatter in the base data, and that its usage would probably result in underestimation of the diffusivity at low concentrations. In a simultaneous study (Leighton and Acrivos, 1986), the same authors incorporated this correlation for shear-induced diffusivity to describe the resuspension of an initially settled layer of negatively buoyant particles in a shear flow. They showed that, even under laminar conditions, on application of shear the settled layer could be expanded, entrained into the bulk fluid, and convected out of the filter. Explaining this phenomenon in terms of a shear-induced diffusion process which they balanced with the downward flux of particles due to gravity, they were able to predict the equilibrium height, \( h \), achieved by the expanded layer as a function of the dimensionless shear diffusivity, \( \hat{D}_{sh} \), and the relative viscosity \( \eta(\phi) \):
\[ h = h_0 + \frac{9 \tau}{2 g} \int_0^{\phi_s} \frac{\phi_s - \phi}{\phi_s} \frac{\hat{D}_{th}}{\eta(\phi)} d\phi \]  

(2.2.20)

where \( \phi_s \) is the initial, settled volume fraction, \( g \) is the acceleration due to gravity, \( \tau \) is the applied shear stress and \( f \) is the particle settling hindrance factor.

The correlation for relative viscosity had been determined by the same authors (Leighton and Acrivos, 1987b), using concentrated suspensions of rigid spheres as:

\[ \eta(\phi) = \left[ 1 + 1.5 \frac{\phi}{1 - \phi/0.58} \right]^2 \]  

(2.2.21)

Equation 2.2.21 was found to be accurate for volume fractions up to 0.5 (Leighton and Acrivos, 1987b). Figure 2.2.2 illustrates the comparison of Equation 2.2.21 with Einstein's formula for relative viscosity (Equation 2.2.22). The agreement is seen to be good (Figure 2.2.2) in the range of dilute concentrations even with the original first order equation due to Einstein (Equation 2.2.22a). As higher order terms are considered (Equation 2.2.22b) the agreement gets better over higher concentrations.

\[ \eta(\phi) = 1 + 2.5 \phi + O(\phi^2) \quad \text{(Einstein, 1906)} \]  

(2.2.22a)

\[ \eta(\phi) = 1 + 2.5 \phi + 6.2 \phi^2 + O(\phi^3) \quad \text{(Batchelor, 1977)} \]  

(2.2.22b)
Figure 2.2.2 Comparison of Einstein's formula for relative suspension viscosity with correlation derived by Leighton and Acrivos (1987b).

A model for crossflow filtration with shear-induced particle diffusion as the dominant phenomenon governing the steady-state boundary layer thickness, was first proposed by Davis and Leighton (1987). Their model predicted nonlinear velocity and concentration profiles within a flowing layer of concentrated particles at the membrane surface which forms due to the convection of particles to the membrane with the permeate flux. Convective flow to the layer is balanced by two mechanisms of transport out of the layer (and eventually the filter); shear-induced back diffusion which transports particles away from the membrane surface and expands the concentrated layer, and axial convection which transports the particles along the membrane surface and out of the element. Their model also predicted the formation of a stagnant cake beneath the flowing particle layer when a maximum packing density is reached at the wall. The particle
layer thickness, $\delta(x)$, at any "local" axial position was then determined assuming a known value for the permeate flux, $V_w(x)$. This model was later extended into a "global" model by Romero and Davis (1988) where the assumption on $V_w(x)$ being known a priori was removed and standard filtration theory was used to describe permeate flux as a function of cake and membrane resistance. Extensions to include time dependent analysis were also later incorporated into the model for viscous resuspension (Chapman and Leighton Jr., 1991) and the global model (Romero and Davis, 1990). The transient global model predicted that the time required for buildup of a flowing boundary layer is usually very small, particularly for all conditions of practical interest, and is of the order of $\delta_f \Phi / V_w \phi_b$. It was also concluded (Romero and Davis, 1990) that the stagnant cake layer, which usually governs flux decline due to its dominant hydrodynamic resistance to permeation, should take several minutes to hours to form, depending on the operating conditions. Therefore, it can be inferred that inclusion of transient effects is largely of theoretical interest. However, at high bulk concentrations ($\phi_b > 0.1$) transient effects might become important in practical considerations (Romero and Davis, 1990).

In recent experimental work, Romero and Davis (1991) verified the steady-state (Romero and Davis, 1988) and transient (Romero and Davis, 1990) models of crossflow microfiltration using suspensions of spherical polystyrene and acrylic particles, made neutrally buoyant using a glycerol solution. They conducted two series of experiments, one using a rectangular glass walled cross-flow microfilter and the other using a hollow-fiber microfiltration module. Overall results indicated an average
absolute deviation between theoretical and experimental flux values of 11% for the first series and of 27% for the second series of experiments. By comparison, alternate models under predict flux by an order or two of magnitude, giving rise to the phrase "flux paradox".

For potable water treatment, which typically involves removal of particles that vary in size over a wide range in the feed suspension, there has been interest in considering various mechanisms of particle transport together in an integrated approach. Wiesner et al. (1989) considered the importance of Brownian diffusion, shear induced diffusion and inertial lift transport mechanisms as a function of particle size and predicted that if all the three mechanisms are considered, particles approximately 0.3 μm in diameter should be preferentially accumulated near the surface of a hollow fiber ultrafiltration membrane under typical operating conditions. According to Chellam and Wiesner (1991) inertial lift forces are important in reducing the transport of larger particles to the membrane; particularly at smaller permeation rates. For typical hydrodynamics existing in UF and MF modules, fluid drag has been predicted to control transport of bacteria-sized particles and smaller particles, leading to colmatage of UF and MF membranes (Chellam and Wiesner, 1992). As these smaller particles accumulate near the membrane they are subject to the effects of Brownian and shear-induced diffusion.

2.2.2 Models for Predicting Permeate Quality

The efficacy of any filtration process is primarily and necessarily
determined by the quality of filtrate achieved through the process. The following sections summarize the work done on modeling the removal of solutes by ultra- and microfiltration membranes.

Membranes employed for low-pressure UF and MF processes have a distinct porous structure. Consequently, the phenomenon of rejection by UF and MF membranes is primarily a "sieving" phenomenon. On the other hand, rejection by RO membranes, which even under the highest resolution of microscopes exhibit no porous structure, requires a consideration of solute membrane interactions within the membrane. The following discussion is confined to a consideration of rejection by UF and MF membranes.

Ferry (1936) considered the theoretical aspects of hindered solute transport in membrane pores. Based on particle hydrodynamics he derived a relationship between the sieve constant, defined as the ratio of permeate to bulk concentrations, and the dimensions of the solute and the pore. This analysis included the following simplifications:

(1) Hard spherical particles and cylindrical pores
(2) Laminar flow within the pore, Newtonian fluid
(3) Negligible wall effects (solute velocity = solvent velocity)
(4) Negligible diffusive transport (high Peclet numbers)
(5) Uniform concentrations in the pores (equal to the bulk concentration)
(6) No concentration polarization or membrane-solute interactions.
Consider a spherical particle of radius $a_p$ in a cylindrical membrane pore of radius $r_o$ as depicted in Figure 2.2.3. Let $\beta$ denote the relative radial position of the particle in the pore and $\lambda$ the particle to pore size ratio. The parabolic velocity profile for the solvent within the membrane pore is given by:

$$v(y) = 2\langle v_y \rangle \left[1 - \left(\frac{y}{r_o}\right)^2\right] = 2\langle v_y \rangle [1 - \beta^2] \quad (2.2.23)$$

The probability of a particle being able to penetrate a pore is governed by the geometrically constrained area available for entry. For particles with radii larger than or equal to the pore radius ($\lambda \geq 1$) this probability is zero. Particles that enter a pore in a region within a radius equal to the difference between the pore and particle radii (the shaded area A in Figure 2.2.4), are assumed to have a pore penetration probability of unity. Particles entering outside of this critical region are assumed to be excluded from passage through the pore.

![Figure 2.2.3. Schematic of a spherical solute particle in a cylindrical membrane pore.](image)

Ferry (1936), integrated Equation 2.2.23 over the geometrically
constrained area available for particle penetration to arrive at an expression for the "sieving coefficient" representing the ratio of solute concentrations

Figure 2.2.4. Schematic of geometrical constraints to pore penetration by a particle.

on either side of the pore:

\[
\kappa = \begin{cases} 
1 - \lambda & \lambda < 1 \\
\int_0^1 4\beta (1 - \beta^2) \, d\beta & \lambda \geq 1
\end{cases}
\]

(2.2.24)

Ferry’s (1936) comparison of theoretical predictions with experimental data for two proteins; serum albumin and hemocyanin, and a virus showed good qualitative agreement of filterability curves. Quantitative agreement was also considered to be good in light of the experimental error and the simplifications inherent in the model.

Ferry’s analysis assumes that the concentration in the pore is not a function of the radial position and is equal to the bulk concentration. Amongst other simplifications Ferry’s analysis also assumes that the solute moves with the same velocity as the surrounding fluid. However, due to
the presence of the pore walls, the solute experiences a retarding effect and moves at a velocity lower than the velocity of the surrounding fluid. The consideration of wall effects which result in retarding the solute velocity, is applied by modifying Ferry's equation with a "lag factor". This "lag factor", or retardation factor, can be estimated by considering the drag force on a spherical particle moving with velocity $v_p$ within a cylindrical pore:

$$F_{\text{drag}} = 6\pi \mu a_p [v_p \kappa_p(\lambda, \beta) - v \kappa(\lambda, \beta)]$$  \hspace{1cm} (2.2.25)

where $\kappa_p$ and $\kappa$ are weighting solute and fluid component drag coefficients, respectively. Under steady state conditions, the drag forces sum to zero and the lag factor can be estimated as:

$$G(\lambda, \beta) = \frac{v_p}{v} = \frac{\kappa(\lambda, \beta)}{\kappa_p(\lambda, \beta)}$$  \hspace{1cm} (2.2.26)

Using the same simplified solute and pore geometry, it was possible to estimate the lag factor for solute to pore size ratios up to 0.9 (Paine and Scherr, 1975; Anderson and Quinn, 1974; Zeman and Wales, 1981). Paine and Scherr (1975) theoretically calculated $\kappa(\lambda, 0)$ and $\kappa_p(\lambda, 0)$ for rigid spheres up to $\lambda = 0.9$, neglecting the dependence on relative radial positions of these coefficients. They noted that the error introduced by this "center-line approximation" ($\beta = 0$) has been known to be small and cited several works to the effect. Zeman and Wales (1981) fitted Paine and Scherr's data to arrive at the following correlation for the lag coefficient:
\[ G(\lambda, 0) = \exp(-0.7472\lambda^2) \]  

(2.2.27)

The sieve coefficient can now be corrected for solute lag by introducing the lag coefficient in Equation 2.2.24. The integrated form of \( \kappa \), after accounting for the lag correction, can be expressed as:

\[ \kappa = (1 - \lambda)^2 \left[ 2 - (1 - \lambda)^2 \right] G(\lambda, 0) \]  

(2.2.28)

It is noted that Equation 2.2.28 is a function of only the solute and pore sizes. Hence a knowledge of the particle size and the rated membrane pore size will serve to estimate the sieving coefficient.

Zeman and Wales (1981) consider the effect of most of the simplifications inherent in Ferry's (1936) work. Their conclusions, supported by experimental and theoretical comparison, categorize solute polydispersivity, pore size distribution and concentration polarization as importantly effecting the rejection by membranes. Wall effects and van der Waals attractions were concluded to be factors of relatively minor importance.

The extension to consideration of non-spherical solutes is another important aspect which has been under research. Giddings et al. (1968) defined a "mean external diameter" as the important parameter uniquely determining removal for rigid molecules of various shapes. They concluded that this parameter which is a geometrical function of the equatorial diameter and polar length for any rigid solute, characterized
removals in their studies better than several other parameters proposed earlier, such as the radius of gyration, equivalent hydrodynamic radius and equivalent molecular weight.

Chao and Tojo (1987), investigated removal of non-spherical organic solutes from fishery processing wastewaters and found that the original Ferry model failed to fit their experimental data satisfactorily. They attributed this to the assumption of spherical particles in contrast with the presence of non-spherical organics in their system. Using the concept of equivalent molecular weight to modify Ferry's model, they were able to better correlate theoretical predictions with experimental data. Anderson and Quinn (1974) incorporated the theory developed for non-spherical particles by Giddings et al. (1968), and added a diffusive component of transport to their model.

2.2.3 Cost Models

The usual trend in deriving equations for the cost of water treatment has been to correlate data available from previously established plants with cost information. Unfortunately, such data are very limited for pressure driven membrane filtration facilities; particularly low pressure processes like UF and MF. Pickering and Wiesner (1991) review previous studies related to the economic analysis of membrane filtration plants. In comparison to cost data for low pressure filtration processes for drinking water treatment, cost data for RO and NF are more available.
A cost analysis of RO and NF membranes, for use in water softening and trihalomethane removal, was presented by Taylor et al. (1989). Based on data collected from a pilot plant in Florida, they found that membrane systems were effective and cost competitive with alternate processes in the treatment of ground waters with high concentrations of organic compounds. They also concluded that membrane plants would be less labor intensive and easier to operate than conventional softening plants.

Capital and operating costs for a facility include: the one time costs of building the facility and initially procuring the membrane modules, and the operating costs arising from energy requirements, membrane replacement, chemicals and concentrate disposal. The cost equations developed by Pickering and Wiesner (1991) for the previously mentioned cost components were used directly in this study and are reviewed in the following paragraphs.

CAPITAL COSTS

Pickering (1991) developed cost functions for plant and membrane cost components. Capital costs are divided into two components, a non-membrane cost representing all the equipment (besides the membranes themselves) required to house and operate the membrane modules, and the initial cost of the membranes. Both of these costs are dependent on the number of modules required and hence the area of the membranes required to generate the design flow. The required membrane area for a design flow of $Q_{des}$ is calculated from the ratio of the required design capacity to the
net permeate flux:

\[ A_{\text{mem}} = \frac{Q_{\text{des}} t_{\text{tot}}}{V_w t_o - V_{\text{bf}} t_{\text{bf}}} \]  
(2.2.29)

where \( t_{\text{tot}} \) represents the total time for one complete operating and flux enhancement cycle, \( t_o \) is the operating time between two flux enhancement cycles, \( V_{\text{bf}} \) is the backflush flux and \( t_{\text{bf}} \) is the backflush duration.

The number of modules corresponding to this membrane area is determined from this estimate of the requisite area and the membrane area per module, \( A_{\text{mod}} \), as:

\[ N_{\text{mod}} = \text{Nearest Integer} \left( \frac{A_{\text{mem}}}{A_{\text{mod}}} + 0.5 \right) \]  
(2.2.30)

The module number is rounded to the next higher integer, since fractional values for this parameter have no significance. A correlation for the estimate of the non-membrane annualized capital costs expressed as cost per unit volume of water treated was developed from data on hollow fiber UF systems:

\[ C_{\text{plant}} = \frac{150037.56 \ N_{\text{mod}}^{0.74} (A/P)}{Q_{\text{des}}} \]  
(2.2.31)

Equation 2.2.31 incorporates the requirement for annualizing the cost by including the amortization factor, \((A/P)\). This factor is calculated as a function of the discount rate for capital investments, \(i_c\), and the design life
of the plant, $DL$, :

$$(A/P) = \frac{i_c(i_c + 1)^{DL}}{(i_c + 1)^{DL} - 1} \quad (2.2.32)$$

The initial membrane costs were similarly expressed in an annualized form:

$$C_{membrane} = \frac{c_{mod} N_{mod} (A/P)}{Q_{des}} \quad (2.2.33)$$

where $c_{mod}$ is the cost of one membrane module. The sum of $C_{plant}$ and $C_{membrane}$ gives the total capital costs for a plant in terms of cost per unit of treated water.

**OPERATING COSTS**

Operating costs considered in the Pickering-Wiesner model include those associated with energy, chemicals, membrane replacement and concentrate disposal.

*Energy Cost*

The pumping requirements for the feed water, recycling, and backflushing are considered in calculating energy costs for a module in a "feed and bleed" configuration (Figure 3.1.1). The ratio of the permeate flow, $Q_p$, to the feed flow denotes the "recovery", $R_{rec}$, of the system and consequently the feed flow can be expressed as:
\[ Q_f = \frac{Q_p}{R_{rec}} = \frac{V_w A_{mod} N_{mod}}{R_{rec}} \]  

(2.2.34)

Energy utilized by the feed pump is expressed as:

\[ E_f = \frac{P_f Q_f}{\eta_f} \]  

(2.2.35)

where \( P_f \) is the feed pressure and \( \eta_f \) the efficiency (expressed as a fraction) of the feed pump.

The headloss across the length of the membrane element and the recycle rate determines the consumption of energy by the recycle pump. For an element with a circular cross section the axial pressure drop is derived from a mechanical energy balance across the module length:

\[ P_r = \frac{4 L_e}{D_e} \left[ f_f - 3 a_{ke} \frac{V_w}{U_m} + \frac{V^3_w}{U^3_m} \right] \left[ \frac{1}{2} \rho \overline{U}^2_m \right] \]  

(2.2.36)

where \( D_e \) is the element diameter, \( a_{ke} \) is a kinetic energy coefficient, \( f_f \) is the fanning friction factor, \( \rho \) is the fluid density and \( \overline{U}_m \) is the average crossflow velocity at the midpoint of the membrane element \( (\overline{U}_m = U_o - 2 V_w L_e / D_e \) where \( U_o \) represents the crossflow velocity at the channel inlet). The values of \( a_{ke} \) and \( f_f \) depend on the Reynolds number and can be approximated from correlations for flow through smooth pipes:

For laminar flow \( (Re \leq 4000) \)

\[ f_f = 16 / Re \quad a_{ke} = 2 \]  

(2.2.37)
For turbulent flow \((Re > 4000)\)

\[ f_f = 0.0791 / Re^{0.25} \quad a_{ke} = 1 \quad \text{(2.2.38)} \]

The recycle rate, \(Q_r\), is estimated from a flow balance on the system (see Section 3.1) and the energy consumed by the recycle pump is then calculated as:

\[ E_r = \frac{P_r Q_r}{\eta_r} \quad \text{(2.2.39)} \]

where \(\eta_r\) is the efficiency of the recycle pump.

The flow rates for fastflushing and backflushing are calculated as:

\[ Q_{ff} = A_x U_{ff} N_f N_{mod} \quad \text{(2.2.40)} \]

\[ Q_{bf} = V_{bf} A_{mod} N_{mod} \quad \text{(2.2.41)} \]

where \(A_x\) is the cross sectional area of the membrane element, \(N_f\) is the number of elements per module and \(U_{ff}\) is the velocity used for fastflushing.

The energies associated with flux enhancement are then calculated as:

\[ E_{bf} = \left( \frac{P_{bf} Q_{bf}}{\eta_{bf}} \right) \frac{t_{bf}}{t_{tot}} \quad \text{(2.2.42)} \]
\[ E_{ff} = \left( \frac{P_f Q_{ff}}{\eta_f} \right) \frac{t_{ff}}{t_{tot}} \]  \hspace{1cm} (2.2.43)

where \( \eta_{bf} \) is the efficiency of the backflush pump and \( t_{ff} \) the fastflush duration.

Finally, if \( c_{kw} \) is the cost for a unit of energy then the total energy costs per unit treated water for operating the system can be expressed as:

\[ C_{energy} = \frac{c_{kw} (E_f + E_r + E_{ff} + E_{bf})}{Q_{des}} \]  \hspace{1cm} (2.2.44)

**Chemical Cost**

The cost of treating feed water with chemicals at the head of the plant are calculated from the chemical dosage, \( CH_d \), and the bulk cost of the chemical, \( CH_c \):

\[ C_{chemical} = \frac{Q_f CH_d CH_c}{Q_{des}} \]  \hspace{1cm} (2.2.45)

In cases where several chemicals are needed, the chemical dose and chemical cost represent concentration and mass-weighted averages.

**Membrane Replacement Cost**

Assuming that membranes are replaced at fixed intervals based on life expectancy of membranes, the cost associated with replacement of membranes is annualized and expressed as:
\[ C_{mr} = \frac{c_{mod} N_{mod} (A / F)}{Q_{des}} \]  

(2.2.46)

where \((A / F)\) is the amortization factor, calculated as a function of the discount rate for membrane replacement, \(i_f\), and the design life of the membrane, \(ML\):

\[ (A / F) = \frac{i_f (i_f + 1)^{ML}}{(i_f + 1)^{ML} - 1} \]  

(2.2.47)

**Concentrate Disposal Cost**

The minimum cost related to the disposal of waste or concentrate is equal the cost of the energy and chemicals invested in the water. The waste water flow rate can be calculated from the following relation:

\[ Q_w = (1 - R_{rec}) Q_f \]  

(2.2.48)

Disposal costs per unit volume of water treated are then expressed as:

\[ C_{disposal} = \left( \frac{c_{kw} P_f Q_w}{\eta_f} \right) + \left( CH_d CH_c Q_w \right) \]  

(2.2.49)

This expression for concentrate disposal considers only the resources "lost" with the concentrate and does not account for any of the costs associated with sewerage, deep well injection, or other disposal options.
Chapter 3

Model Development

In this chapter, an integrated model for estimating the performance and cost of UF and MF systems is presented. The efficiency of any filtration process is determined by two essential performance parameters. One is the ability to remove the influent contaminants to the required degree, and the second is the rate at which the filtrate is obtained. When considered together, these two parameters determine both the necessary and sufficient conditions required in establishing the technical efficiency of a filtration process in a particular situation. However, to evaluate the economic feasibility of a process, it becomes essential to relate these parameters to the associated costs for treatment. The following sections describe the mathematical models which were integrated to estimate the permeate flux, contaminant removal, and the associated unit cost for low-pressure membrane filtration.

3.1 Module Configuration

Figure 3.1.1 details the standard system configuration modeled in this study. The system comprises cylindrical membrane modules (or pressure vessels) each of which consists of a shell enclosing a number of tubular or hollow fiber membrane elements. The feed, with a suspension volume fraction of $\phi_o$, flows at the rate $Q_f$. This rate is determined by the permeate flow rate, $Q_p$, and the specified recovery (Equation 2.2.34). $Q_p$
is itself calculated as a function of the permeate flux, which must be predicted from a model for permeate flux, and the required capacity of the plant (Equation 2.2.34). \( Q_w \) represents the waste flow rate and is determined from a simple flow balance on the system (control system A) as:

\[
Q_w = Q_f - Q_p
\]  \hspace{1cm} (3.1.1)

Recycling is considered to take place at a rate \( Q_r \) and is determined by another flow balance on the system (control system B):

\[
Q_r = Q_t - Q_p - Q_w
\]  \hspace{1cm} (3.1.2)

\( Q_r \) is the total flow into the membrane modules and is determined as:

\[
Q_r = A_x N_f \overline{U}_o N_{mod}
\]  \hspace{1cm} (3.1.3)
Since the total flow rate should be equal to or greater than the feed flow rate, the following equation should always be satisfied:

\[ A_x \ N_f \ \bar{U}_o \ N_{mod} \geq \frac{V_w}{R_{rec}} \frac{A_{mod} \ N_{mod}}{R_{rec}} \]  

(3.1.4)

This expression, when simplified, leads to an upper limit on the permeate flux possible through a membrane element of a particular configuration. For a tubular (or hollow fiber) configuration:

\[ (V_w)_{max} = \frac{D_e \bar{U}_o R_{rec}}{4 L_e} \]  

(3.1.5)

The permeate flux model requires knowledge of the bulk suspension volume fraction, \( \phi_b \), which at steady state, is determined from the rejection of particles on the membrane surface and the recovery rate \( R_{rec} \). \( \phi_p \) represents the permeate suspension volume fraction determined as a function of solute rejection on the membrane and the bulk suspension concentration.

### 3.2 Mathematical Modeling of Permeate Flux

Predictions of the permeate flux in the integrated performance and cost model are based in large part on the work of Romero and Davis (1988). However, their model for shear-induced diffusion is modified to account for Brownian diffusivity as a mechanism of particle transport. In an additional departure from the global model presented by Romero and Davis, the modified model for permeate flux approximately accounts for a
"leaky" membrane in which some mass in the boundary layer may pass through the membrane. This effect manifests itself as a decrease in the bulk concentration as calculated from the model for contaminant rejection (Section 3.3). In the following sections, a modification of the Romero and Davis model is presented that accounts for the effects of Brownian as well as shear-induced diffusion.

In crossflow filtration, the feed suspension to be filtered flows through the membrane channel in a direction tangential to the membrane surface (Figure 3.2.1). The driving force for separation is the pressure difference across the membrane, $\Delta P$. Higher pressure within the membrane channel drives relatively clean solvent across the membrane, while colloidal materials that are larger than the membrane pore size are rejected at the membrane surface. Some of the solute molecules smaller than the pore size, that are not rejected by the membrane, exit with the permeate. The tangential flow of the bulk suspension, under sufficient shear stress, may transport particles retained near the membrane out of the membrane element.

Crossflow filtration limits the buildup of rejected particles on the membrane and, at steady-state, the particle layer formed on the walls is predicted to attain a thickness that is invariant with time but increases with axial distance from the filter entrance. This constant thickness is determined by a balance between the convective and diffusive mechanisms of transport which add to or remove particles from the membrane surface.
Figure 3.2.1 Schematic of crossflow membrane filtration.

This layer of particles provides an additional resistance that may reduce the permeate flux or increase the pressure drop across the membrane. In practice, long term fouling and/or compaction of the membrane may be additional causes of permeate flux decline.

3.2.1 Mechanisms of Particle Transport

Consider a tubular membrane filter in longitudinal section as shown in Figure 3.2.2. There are three essential mechanisms by which the particles in the bulk suspension may be transported:

(1) *Convective particle transport towards the membrane.* The driving force of differential pressure across the membrane brings the suspension towards the membrane. The smaller solvent and solute molecules pass through the membrane across to the low pressure permeate side while larger particles are retained on the surface of the membrane.
(2) **Diffusive particle transport away from the membrane.** The accumulation of materials near the membrane causes concentration gradients at the membrane surface which induce Brownian diffusion of materials away from the membrane into the bulk suspension. The crossflow also induces a shear-induced diffusive mechanism in the concentrated particle layer which augments the Brownian diffusion.

(3) **Bulk convective flow of the particle layer along the membrane surface and out of the filter.** Under sufficient shear stress, the whole particle layer experiences a sweeping tangential flow in the axial direction.

Under steady-state, a balance between these three mechanisms determines the thickness of the particle layer or cake formed on the filter. These mechanisms will be described further in the following section.

**Figure 3.2.2** Mechanisms of particle motion and convective transport of the particle layer.
3.2.2 Reformulation of Transport Equations to Include Brownian Diffusion

Consider the flow of a Newtonian fluid in a channel of circular cross-section and having porous walls (Figure 3.2.3). The channel geometry is specified by an element length $L_e$ and a radius $R_o$. The axial coordinate is measured from the channel entrance and is denoted by $x$. The transverse coordinate $y$ is orthogonal to $x$ and represents the distance from the wall towards the center line of the channel. The fluid is assumed to contain rigid neutrally buoyant spherical particles at a bulk volume fraction of $\phi_b$. The average velocity at which the bulk suspension flows in the axial direction is denoted by $\bar{U}$.

Figure 3.2.3 Schematic of steady-state stagnant cake layer and flowing particle layer formation in a cross-flow membrane filter.
Under the driving force of differential pressure, $\Delta P$, the solvent permeates across the membrane at a rate $V_w$, while the particles brought to the walls due to this convective motion either accumulate near the membrane or pass through the membrane depending on their size. In the general case, the dynamic dependence of the permeation velocity, $V_w$, is on the particle layer thickness, $\delta$, and the pressure drop, $\Delta P$. Both of these may be dependent on the axial position in the channel. The model of Romero and Davis (1988) solves for the functions $\delta(x)$ and $V_w(x)$ in terms of known parameters such as the pressure drop, the bulk particle volume fraction, the average down-channel suspension velocity, the channel radius, and the hydraulic resistance of the particle layer. In their analysis the model is formulated assuming absolute rejection of solutes at the membrane walls and back diffusion due to shear-induced phenomena. This study employs the same model for the estimation of the permeate flux, but with the modification that Brownian diffusivity be considered as another mechanism of back diffusion. Absolute rejection on the membrane surface is also not assumed and the permeate quality model (Section 3.3) is used to predict the expected bulk suspension volume fraction at steady state.

The model of Romero and Davis considers the formation of a flowing as well as a stagnant cake layer on the membrane surface (Figure 3.2.3). Near the filter entrance, the particles depositing near the membrane surface form a flowing layer, which increases in thickness in the axial direction. At a "critical" axial point, the volume fraction of particles at the membrane surface reaches a limiting value, due to maximum packing, and
the particle layer is unable to accept any more particles. This situation is alleviated by the formation of a stagnant layer of particles beneath the flowing layer, which causes constriction of the filter and consequently an increase in shear stress. The shear-induced diffusion coefficient varies linearly with shear stress, and this causes an increase in the thickness of the flowing layer which is now able to accommodate more particles. The flowing and stagnant layers are hence expected to increase in thickness with distance from the membrane inlet. A thicker particle layer offers a greater resistance to permeate flow and consequently the permeation rate decreases over the length of the filter as the thickness of the cake increases.

If $R_m$ is the membrane resistance, $\delta_{fl}$ and $\delta_{st}$ are the thicknesses of the flowing and stagnant particle layers, and $R_{fl}$ and $R_{st}$ the respective specific resistances of the particle layers, then the permeate flux according to a "resistance in series" model is given by:

$$V_w = \frac{\Delta P}{\mu_o \left( R_m + \delta_{fl}R_{fl} + \delta_{st}R_{st} \right)} \quad (3.2.1)$$

The resistance due to the flowing layer is usually much less than the membrane resistance and hence negligible in all cases of practical interest (Romero and Davis, 1988). The specific resistance of the cake layer, is estimated using the Blake-Kozeny correlation:

$$R_{st} = \frac{37.5 \phi_{max}^2}{a_p^2 (1 - \phi_{max})^3} \quad (3.2.2)$$
where $\phi_{\text{max}}$ is the volume fraction of the maximally packed particles in the cake layer. For non-deformable particles a value of 0.58 can be assumed for $\phi_{\text{max}}$. Polydispersivity effects are considered by calculating a diffusivity-weighted mean particle size (as detailed in Section 3.4).

The estimation of $\delta(x)$ and $V_w(x)$ involves formulating a microscopic mass balance on an element within the particle layer, as well as a macroscopic mass balance on the particle layer as a whole. The microscopic analysis is employed to include the convective and diffusive effects normal to the membrane surface while the macroscopic balance incorporates consideration of convective transport of the particle layer along the membrane surface.

*Microscopic mass balance*

A microscopic mass balance on an elemental volume is given by the convective-diffusion equation. In two-dimensional flow, at steady state, neglecting diffusion in the axial direction, the convective diffusion equation reduces to:

$$\frac{\partial u\phi}{\partial x} + \frac{\partial v\phi}{\partial y} = \frac{\partial}{\partial y}(D \frac{\partial \phi}{\partial y})$$  \hspace{1cm} (3.2.3)

where $D$ is the effective diffusivity, calculated as the sum of Brownian and shear-induced diffusion components. In the boundary layer the axial convective term is considered negligible compared to the other two terms (Romero and Davis, 1988). Equation 3.2.3 is integrated considering the remaining two terms to give the following integrated form:
\[ D \frac{d\phi}{dy} + V_w \phi = 0 \]  \hspace{1cm} (3.2.4)

The Brownian component of diffusivity is calculated using the Stokes-Einstein equation:

\[ D_{brn} = \frac{kT}{6 \pi \mu_o a_p} \]  \hspace{1cm} (3.2.5)

The shear-induced hydrodynamic diffusion component is calculated as a function of the local shear rate in the particle layer, \( \dot{\gamma} \), and the particle size:

\[ D_{sh} = \dot{\gamma} a_p^2 \hat{D}_{sh}(\phi) \]  \hspace{1cm} (3.2.6)

\( \hat{D}_{sh} \) is a dimensionless function of \( \phi \), which was experimentally determined by Leighton and Acrivos (1986) as:

\[ \hat{D}_{sh}(\phi) = 0.33 \phi^2 \left(1 + 0.5e^{8\phi} \right) \]  \hspace{1cm} (3.2.7)

The local shear rate represents the velocity gradient within the particle layer:

\[ \dot{\gamma} = \left| \frac{du}{dy} \right| = \frac{\tau_w}{\mu_o \eta(\phi)} \]  \hspace{1cm} (3.2.8)

where \( \eta(\phi) \) is given by Equation 2.2.21. In presenting Equation 3.2.8 Romero and Davis assumed that the particle layer is thin compared to the
channel radius, so that the shear stress within the particle layer can be approximated as a constant, and equal to the stress at the membrane wall, \( \tau_w \). For elements with circular cross-sections, \( \tau_w \) is calculated as:

\[
\tau_w = 4 \bar{U} \mu_o \eta(\phi_b) / R
\]  

(3.2.9)

where \( \bar{U} \) is the average crossflow velocity and \( R \) is the effective tube radius, calculated by accounting for the constriction of the channel due to cake formation.

**Macroscopic mass balance**

At steady state, the flux of particles per unit filter width flowing across the plane of constant \( x \) in the concentrated layer is equal to the total volume of particles which have been deposited onto the layer everywhere upstream of \( x \) or

\[
\int_0^{\delta_b} u(\phi - \phi_b) dy = \int_0^x V_w \phi_b \, dx = Q
\]  

(3.2.10)

where \( Q \) is defined as the "excess particle flux". Equation 3.2.4, Equation 3.2.8 and the first integral of Equation 3.2.10 are solved, after changing the variable of integration from \( y \) to \( \phi \), to give the following expression for the excess particle flux:

\[
Q = \left( \frac{\tau_w}{\mu_o^3 V_w^2} \right) \left[ \int_{\phi_b}^{\hat{\phi}_w} \left( \int_{\phi}^{\hat{\phi}_w} \frac{\tau_w a_p^2 \hat{D}_{sh}(\phi') / \eta(\phi') + kT / 6\pi\mu_o a_p}{\phi' \eta(\phi')} d\phi' \right) \right]

(\phi - \phi_b) \left( \frac{\tau_w a_p^2 \hat{D}_{sh}(\phi) / \eta(\phi) + kT / 6\pi\mu_o a_p}{\phi} \right) d\phi
\]  

(3.2.11)
The excess particle flux increases with axial distance from the channel inlet due to more particles accumulating in the particle layer. Before the formation of a stagnant cake the wall shear stress has the constant value $\tau_{wo}$, calculated using Equation 3.2.9 with the clean channel radius $R_0$ as the effective radius. Under this condition the double integral in Equation 3.2.11 is only a function of $\phi_b$ and $\phi_w$. At the onset of the stagnant cake layer the particle volume fraction at the wall has reached its maximum value, $\phi_{\text{max}}$, which for non-deformable particles is approximated as 0.58. The permeate flux up to this point also has a constant value $V_{wo}$ (assuming the flowing particle layer contributes negligible resistance to permeation). Representing the value of the excess particle flux, $Q$ (Equation 3.2.11), at the point the cake starts forming or the "critical point", by $Q_{cr}$, and then equating Equation 3.2.11 and the second integral in Equation 3.2.10, the "critical length" can be expressed as:

$$x_{cr} = \frac{\tau_{wo} q_{cr}}{\mu_0^3 V_{wo}^2 \phi_b} \quad (3.2.12)$$

where $q_{cr}$ represents the double integral in Equation 3.2.11 and is defined as:

$$q_{cr} = \frac{\mu_0^3 V_w^2 Q_{cr}}{\tau_w} \quad (3.2.13)$$

Thus, the solution to the problem prior to the formation of the stagnant cake, i.e. for $x < x_{cr}$ can be summarized as:
\[ \delta_{st} = 0, \quad \phi = \phi_b, \quad \tau_w = \tau_{wo}, \quad V_w = V_{wo} \quad (3.2.14) \]

Beyond the critical point the numerical solution to the permeate flux and the cake thickness, as they vary with axial distance, is based on the fact that the upper limit of both the integrals in Equation 3.2.11 is the maximum particle volume fraction at the wall, \( \phi_{\text{max}} \). Solution entails first discretizing the length of the membrane represented by the axial coordinate \( x \), and then iteratively solving the system of equations involving the variable parameters, \( \delta_{st}(x) \), \( V_w(x) \), \( Q(x) \) and \( \tau_w(x) \). The Equations involved are Equation 3.2.1, Equation 3.2.11, the second integral in Equation 3.2.10 and the following Equation for the shear stress:

\[ \tau_w = \tau_{wo} \frac{R_0^3}{(R_0 - \delta_{st})^3} \quad (3.2.15) \]

Equation 3.2.15 is derived from a mass balance on the suspension flow through the restricted channel \( (\bar{U}_o \ R_o^2 = \bar{U} \ R^2) \). It is noted that such a mass balance neglects the small fraction of fluid lost as permeate with the finite difference segment. The algorithm coded to solve this system of complex equations was based on a hybrid quasi-Newton method and is presented in Chapter 4.

### 3.3 Solute Rejection and Permeate Quality

The model for estimating the quality of permeate achieved through membrane filtration is based on the concept of the "sieve coefficient" introduced by Ferry (1936). The original relation for this coefficient due to
Ferry has been modified by other workers (Paine and Scherr, 1975; Zeman and Wales, 1981) to account for solute lag in pores. Details of the relevant theory have already been presented in Section 2.2.2 and are not repeated here.

The sieve coefficient used in this study is calculated using Equation 2.2.28. Figure 3.3.1 depicts the sieve coefficient plotted for a range of particle to pore size ratios. It is noted from the figure that for a particle half the size of the membrane pore ($\lambda = 0.5$), the sieve constant is calculated to be 0.36, or a rejection $(1 - \lambda)$ of 64% is predicted. At a particle diameter equal to a tenth of the pore diameter ($\lambda = 0.1$), the rejection is about 4%. For $\lambda = 0.9$, a rejection of about 99% is predicted, that is, almost all the particles are estimated to be unable to penetrate the membrane pore.

![Graph](image)

*Figure 3.3.1. Sieve coefficient, representing the ratio of permeate to bulk suspension concentrations, for a range of solute to pore size ratios.*
Estimation of Bulk and Permeate Suspension Concentration

At steady state within a membrane module, the bulk suspension volume fraction can be estimated from the feed suspension volume fraction, the recovery rate, and a term representing the rejection of solute on the membrane; \((1 - \kappa)\):

\[
\phi_b = \frac{\phi_o}{1 - [(1 - \kappa) R_{rec}]} \tag{3.3.1}
\]

The volume fraction of solute in the permeate stream is then calculated from the definition of the sieve coefficient:

\[
\phi_p = \kappa \phi_b \tag{3.3.2}
\]

Equation 3.3.2 thus serves to estimate the quality of the permeate produced from filtration. Calculation of the bulk suspension volume fraction by Equation 3.3.1 is further employed in estimating the average permeate flux using the permeate flux model. For fixed filtration rate and mass transfer conditions near the membrane, a decrease in the bulk concentration corresponds to a decrease in the concentration near the membrane. Thus, modification of the bulk concentration by Equation 3.3.1 simultaneously accounts for changes in rejection and mass accumulation near the membrane.

3.4 Polydispersivity Considerations

Polydispersivity considerations become important for any realistic analysis
of solute species in water. The method adopted to incorporate particle size
distributions in the performance analysis of membrane systems was to
arrive at an "average" sieve coefficient and "effective" particle size as
parameters representing polydispersivity.

**Mean Sieve Coefficient**

For a polydisperse suspension, an average sieve coefficient can be
defined as:

\[
\tilde{K} = 1 - \left[ 1 - \phi_o / \sum_i \frac{\phi_i}{1 - (1 - \kappa_i) R_{rec}} \right] / R_{rec} \tag{3.4.1}
\]

where \( \phi_{o,i} \) is the feed suspension volume fraction corresponding to size
class \( i \), and \( \kappa_i \) is the associated sieve coefficient.

**Mean Particle Size**

A concentration and diffusion weighted estimate for an "effective" or
average particle size is defined as:

\[
\overline{a_p} = \frac{\sum_i a_{p,i} \phi_{b,i}}{\sum_i \frac{\phi_{b,i}}{D_i}} \tag{3.4.2}
\]

where \( i \) represents a size class in the distribution and \( D_i \) the corresponding
total diffusivity for particles of this size. Equation 3.4.2 uses a summation
in which particles with lower effective diffusivity contribute more towards
the calculation of the mean particle size.

The total diffusivity, $D_i$, is the sum of the Brownian and shear-induced diffusion components for particles in size class $i$. It should be noted that shear-induced diffusivity for a particle in a suspension flowing through a membrane is a function of the shear rate in the particle layer and the volume fraction at the wall. Both these parameters actually vary axially along the length of the membrane. For the purposes of calculating the mean particle diameter, Equation 3.4.2 is solved using the shear rate at the channel entrance and assuming a particle volume fraction of 0.29 representing an estimate of the mean volume fraction in the flowing layer.

### 3.5 Cost Estimation

To assess the economic utility of a membrane filtration process, the total cost associated with the facility needs to be estimated as a function of various capital as well as operating components. Such components would include: the one time costs of building the plant and initially procuring the membrane modules as well as the operating costs of energy requirement, membrane replacement, chemicals and concentrate disposal. The cost equations developed by Pickering and Wiesner (1991) for these components were used directly in integrating the performance analysis with a cost estimation model. These equations have been reviewed in Section 2.2.3.
3.6 Model Options

Some options are provided in the model with respect to specifying operating time between flux enhancement cycles and the cross-flow velocity.

3.6.1 Determining Operating Time Between Hydrodynamic Cleaning

If the system is to be operated in a "feed and bleed configuration" (Figure 3.1.1) in which a waste stream is continually been generated then the operating time between flux enhancement cycles must be specified. However, another possible means of operating a membrane system is to waste only during fastflushing and backflushing. In this case, the recovery of the system is defined by the frequency of backflushing. Conversely, the operating time between two flux enhancement cycles can be calculated as a function of the desired recovery. This can be done using Equation 2.2.29:

\[
A_{mem} = \frac{Q_{des} \cdot t_{tot}}{V_w \cdot t_o - V_{bf} \cdot t_{bf}} \tag{2.2.29}
\]

The first term in the denominator of Equation 2.2.29 represents the permeate flow. If it is assumed that the waste flow is represented by the amount of water utilized in backflushing then the operating time can be determined from the following relation:

\[
R_{rec} = \frac{Q_p}{Q_p + Q_w} = \frac{V_w \cdot t_o}{V_w \cdot t_o + V_{bf} \cdot t_{bf}} \tag{3.6.1}
\]
Equation 3.6.1 can be solved to give the following relation for the operating time:

\[ t_o = \left( \frac{V_{bf}}{V_w} \right) \frac{R_{rec}}{(1 - R_{rec})} \]  \hspace{1cm} (3.6.2)

Equation 3.6.2 was employed to calculate the operating time and hence the backflushing and fastflushing frequency in the simulations presented in this work (Chapter 5).

3.6.2 Specifying Crossflow Velocity or Axial Pressure Drop

Practical operation of membrane filtration units is usually limited by the allowable pressure drop across membrane modules. This constraint is taken into consideration by either working with a specified crossflow velocity or calculating the corresponding crossflow velocity for a specified pressure drop. For the latter case, the nonlinear equation for pressure drop (Equation 2.2.36) is solved for crossflow velocity assuming a clean water permeate flux.

3.7 Model Limitations and Simplifications

(1) The permeate flux model assumes the shear stress within the flowing particle layer to be approximated as a constant. The model is therefore limited to situations where the flowing particle layer is thin relative to the channel radius. Also, the model holds for cake thicknesses which are small compared to the channel radius since the bulk suspension concentration is
assumed to remain constant. The solution would thus be an approximate one in situations where the channel gets "plugged". However, in practice, the plugging scenario does not occur or is avoided.

(2) The model is valid only under laminar flow. Turbulent flow has not been investigated.

(3) The permeate flux model is limited to a steady-state condition. Behavior with time has not been analyzed.

(4) The removal model is based on a simple pore and particle geometry and other simplifications as detailed in Section 2.2.2.

(5) Polydispersivity considerations are included by calculating an average parameter or mean particle size.

(6) The model is limited by the assumption that effects of gravity on particle transport can be neglected. In practice UF and MF modules are often mounted vertically with respect to crossflow direction and gravity effects on particle transport are therefore negligible.

(7) The resistance to permeation due to long term effects such as pore blocking and adsorption on pore walls has not been considered. The model is limited to situations where the stagnant cake and/or the membrane are the only limiting resistances to permeation.
Chapter 4

Development of Computer Program

The models for performance and costing of membrane filtration were transformed into computer code and incorporated into a numerical software package; "MeMsys". All the coding was done in FORTRAN language and the program compiled under Microsoft Fortran Compiler Version 5.1.

4.1 Overall Framework

As depicted in Figure 4.1.1, the program consists of a main driver module which controls the basic flow and menu interface. The driver module manages several sub-modules comprising the performance model, the cost model, a graphics module and other modules containing various subroutines for user-interface, data control features such as editing, saving and loading, as well as error handling.

A flowchart for the program is shown in Figure 4.1.2. It should be noted that the interface allows the user to access various menu options in any order, and as such, Figure 4.1.2 represents just one logical flow pattern. In actuality, once the user starts the program, the sub-modules can be chosen to run from the main menu in any order. On start of execution, the program loads several predefined files containing default data. The user can then edit any of these data through the data entry module or load other
files containing previously stored data. The performance and cost models can then be run using these data. The program also provides the option of viewing graphs of cake thickness, permeate flux and the wall shear stress as a function of axial distance along the membrane element. Output is produced on the screen and can also be saved to disk files. Details of program capabilities, data requirements, output generated and other features are covered in the user’s guide (Appendix A).

Figure 4.1.1 Overall framework of the computer model
Figure 4.1.2 Sample program flowchart. Selections from the main menu determine program flow.
4.2 Performance Module - Numerical Estimation of Permeate Flux

The permeate flux model requires numerical integration along the filter length. The finite difference procedure employed within the model to perform these calculations is detailed in this section. Figure 4.2.1 depicts the grid scheme used.

![Grid scheme used in the finite difference algorithm employed to solve for the permeate flux.](image)

Figure 4.2.1 Grid scheme used in the finite difference algorithm employed to solve for the permeate flux.

Initial permeate flux at channel inlet, i.e. $V_{wo}$ at point $x=0$, is calculated from Equation 3.2.1 with $\delta_{st}(x=0)$ equal to zero (the flowing cake resistance, $\delta_{fl}R_{fl}$, is always neglected by the model). The critical length is then evaluated using Equation 3.2.12. If the critical length is greater than the length of the filter element, then the length-averaged permeate flux equals $V_{wo}$ and the average cake thickness is zero; i.e. no cake is formed on the filter. When the critical point falls within the element length, a finite difference solution is begun using the grid scheme depicted in Figure 4.2.1. The element length over which the cake forms, $L_e - x_{cr}$,
is divided into a grid of \( N \) points. \( \Delta x \) denotes the stepsize of the axial grid:

\[
\Delta x = \frac{L_x - x_{cr}}{N}
\]  

(4.2.1)

The permeate flux at a distance \( x_i \) can be estimated as a function of the cake thickness at location \( i \) using the following relation:

\[
V_{wi} = \frac{V_{wo}}{\left(1 + \frac{R_{st} \delta_{st}}{R_m}\right)}
\]  

(4.2.2)

From Equation 3.2.15 the wall shear stress at grid point \( i \) is represented as a function of the cake thickness as:

\[
\tau_{wi} = \tau_{wo} \left(\frac{R_o}{R_o - \delta_{st}}\right)^3
\]  

(4.2.3)

The second integral in Equation 3.2.10 which represents the excess particle flux, is discretized over the grid in the following manner:

\[
\int_{x_o}^{x_i} V_{wi} \phi_b \, dx = V_{wo} \phi_{b,cr} + \sum_i V_{wi} \phi_b \Delta x
\]  

(4.2.4)

Using Equations 4.2.2 and 4.2.3, Equation 4.2.4 is then equated with Equation 3.2.11 for the excess particle flux to define the nonlinear function which needs to be solved over the grid for the cake thickness. The algorithm to solve this function is described later in this section. The converged solution for the cake thickness returned from the algorithm, at
each point, is substituted back into Equation 4.2.2 to calculate the permeate flux at that point. The length averaged flux from the filter is then calculated in the following manner:

$$V_w = \frac{V_{wo}x_{cr} + \sum_{i}^{N} V_{wi} \Delta x}{L_e}$$  \hspace{1cm} (4.2.5)

The flowchart of the performance subroutine formulated for estimation of the permeate flux is presented in Figure 4.2.2. The performance subroutine calls a nonlinear equation solver subroutine to estimate the cake thickness at each grid point. The first grid point is likely to have a very small cake thickness, since it is closest to the critical point where the cake growth is predicted to start. Hence a value of zero is passed to the equation solver subroutine as an initial guess. The value of the guess for the next iteration is updated, at each iteration, to the solution at the current iteration.

The algorithm formulated to solve the nonlinear function with the cake thickness as the unknown parameter is based on Newton's method. Global convergence is assured by hybridizing Newton's method with the bisection method. Employing such a strategy attempts to combine global convergence with the fast local convergence of Newton's method (Dennis and Schnabel, 1983). The flowchart depicting the algorithm contained in the nonlinear solver subroutine is presented in Figure 4.2.3. The algorithm starts with a current estimate of the solution, \( s_c \), as equal to the guess passed by the performance subroutine. The value of the nonlinear function at a point \( s \) is represented by \( f(s) \). The stepsize, \( h_c \), for estimating the
Figure 4.2.2 Permeate flux estimation module: Simplified flowchart
Figure 4.2.3 Hybrid quasi-Newton algorithm to solve a nonlinear function
finite difference gradient, \( a_c \), of the function, at \( s_c \), is calculated as a function of the machine precision, macheps, and \( s_c \). A Newton iteration is then employed to get a better estimate of the solution, \( s_+ \). If this iteration results in an increase in the absolute value of the function at \( s_+ \) compared to the value at \( s_c \), then a backtracking strategy is employed to assure global convergence. A stopping criterion is examined by checking for the relative error between \( s_+ \) and \( s_c \). If convergence is acceptable then the value of \( s \) at the last iteration, \( s_+ \), is returned as the solution. Otherwise the algorithm is repeated until convergence is achieved. Failure to converge within a specified maximum number of iterations results in an error condition and the algorithm is forced to stop.

The average permeate flux and other parameters are displayed on the screen as they are calculated. The user is then given the option of either saving the results to a disk file or returning to the main menu.

4.3 Cost Module

The flowchart depicting the cost module is represented in Figure 4.3.1. The cost module can be run following the performance module or independently if a user defined permeate flux value is specified. The cost components and other system parameters such as flow rates and energy requirements of the different pumps are calculated and displayed on the screen. An option to store the results to a disk file is also available.
4.4 Other Features

Data handling in the model is done through a data entry module. Using this module it is possible to view the values of all program variables
and edit them as required. Variables are stored in different files depending on their relation to the cost module or performance module. Loading different data files and saving modified files can also be accomplished within this module.

A graphics module is included in the program which permits viewing graphic display of the cake thickness, permeate flux and the wall shear stress as calculated by the performance module.
Chapter 5

Results and Discussion

5.1 Model Behavior - Permeate Flux and Cost

The computer code described in the previous chapters was used to examine the effect of raw water quality, on selected performance and cost parameters for membrane filtration. A sensitivity analysis was performed to evaluate the impact on capital and operating costs of design and operating variables such as plant capacity, recovery, feed pressure, element diameter and length for a fixed raw water quality. The effect of particle size and concentration in the raw water was evaluated using a fixed set of values for these variables. In the course of this analysis the effect of raw water quality (as described by particle size and concentration) on several performance parameters such as the permeate flux, cake thickness, and the critical point was also studied.

For the purpose of simulating the model behavior, two typical low-pressure configurations, one for UF, and one for MF were selected. The values selected for the various parameters in the two configurations are given in Table 5.1.1. The UF system employs values typical for a hollow fiber membrane with a MWCO of 100,000 Dalton. The MF configuration employs a tubular membrane with a pore size of 0.2 µm. It should be noted that the model for permeate flux considered in this study is valid only for laminar flow. The crossflow velocity for the MF simulations is
Table 5.1.1 Baseline values of parameters for the UF and MF simulations.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Microfiltration</th>
<th>Ultrafiltration</th>
</tr>
</thead>
<tbody>
<tr>
<td>Membrane Pore Size (1/cm)</td>
<td>0.2 μm</td>
<td>100,000 Dalton</td>
</tr>
<tr>
<td>Membrane Resistance (mm)</td>
<td>3.33 X 10⁹</td>
<td>1.46 X 10¹⁰</td>
</tr>
<tr>
<td>Element Diameter (cm)</td>
<td>85</td>
<td>120</td>
</tr>
<tr>
<td>Feed Pressure (bars)</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td>Cross-flow Velocity (cm/s)</td>
<td>125</td>
<td>90</td>
</tr>
<tr>
<td>Recovery (%)</td>
<td>95</td>
<td>95</td>
</tr>
<tr>
<td>Temperature (°C)</td>
<td>20</td>
<td>20</td>
</tr>
<tr>
<td>Design Capacity of Plant (m³/hr)</td>
<td>200</td>
<td>200</td>
</tr>
<tr>
<td>Module Packing Density (%)</td>
<td>10.69</td>
<td>11.54</td>
</tr>
<tr>
<td>Module Diameter (cm)</td>
<td>32</td>
<td>32.5</td>
</tr>
<tr>
<td>Cost of Capital (%)</td>
<td>10</td>
<td>10</td>
</tr>
<tr>
<td>Earning Interest Rate (%)</td>
<td>8</td>
<td>8</td>
</tr>
<tr>
<td>Design Life of Plant (yrs)</td>
<td>20</td>
<td>20</td>
</tr>
<tr>
<td>Design Life of Membrane (yrs)</td>
<td>5</td>
<td>5</td>
</tr>
<tr>
<td>Coagulant Dose (mg/L)</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Coagulant Cost ($/1000kg)</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Energy Cost ($/kWhr)</td>
<td>0.07</td>
<td>0.07</td>
</tr>
<tr>
<td>Module Cost ($)</td>
<td>8000</td>
<td>8000</td>
</tr>
<tr>
<td>Efficiency of Feed Pump (%)</td>
<td>70</td>
<td>70</td>
</tr>
<tr>
<td>Efficiency of Recycle Pump (%)</td>
<td>70</td>
<td>70</td>
</tr>
<tr>
<td>Efficiency of Backflush Pump (%)</td>
<td>70</td>
<td>70</td>
</tr>
<tr>
<td>Fastflush Duration (sec)</td>
<td>5</td>
<td>15</td>
</tr>
<tr>
<td>Fastflush Velocity (cm/s)</td>
<td>210</td>
<td>111</td>
</tr>
<tr>
<td>Backflush duration (sec)</td>
<td>1</td>
<td>45</td>
</tr>
<tr>
<td>Backflush Flux (L/m²/hr)</td>
<td>4428</td>
<td>387</td>
</tr>
<tr>
<td>Backflush Pressure (bars)</td>
<td>6</td>
<td>2</td>
</tr>
<tr>
<td>Operating Time (min)</td>
<td>(see Section 3.6.1; Equation 3.6.2)</td>
<td></td>
</tr>
</tbody>
</table>
therefore kept artificially low so that the flow may be interpreted as laminar. In practice tubular membranes are operated at higher crossflow velocities, well into the turbulent regime. The values for membrane resistance were calculated using available experimental data on permeate flux of clean water using UF (Wiesner, 1993) and MF (Nazzal, 1993) membranes of the specified pore size/MWCO.

5.1.1 Particle Size Effects

A particle size range of 0.001 µm to 10 µm was considered in generating the graphs presented in this section. Since the model considers polydispersivity by estimating a "mean" particle size; a crude approximation, this size range can be assumed to either represent an average particle size for a polydisperse suspension or the absolute particle size for a monodisperse suspension. Values for a single parameter were varied while holding the remaining parameters listed in Table 5.1.1 constant. The operating time was calculated as a function of recovery as detailed in Section 3.6.1, for the simulations presented in this section. A set of five values representative of the range of solute concentration in source raw waters for water treatment was assumed; viz., 10, 20, 50, 100, and 200 mg/L. The focus in this section, however was on particle size effects. A similar analysis, where the focus is on concentration effects, is covered in Section 5.1.2.

5.1.1.1 Diffusivity

The effect of particle size on total diffusivity is illustrated in Figure
5.1.1 which was generated using data for the UF and MF membranes. Shear induced diffusivity was calculated using the shear rate at the channel inlet and assuming a mean value of 0.29 for the particle volume fraction.

![Figure 5.1.1 Effect of particle size on diffusivity. Shear induced diffusion component calculated using wall shear stress at channel inlet and an assumed mean solute volume fraction of 0.29 in the flowing boundary layer. Reynolds number calculated at channel inlet.](image)

The purpose of Figure 5.1.1 is to illustrate the interplay between Brownian and shear induced diffusion coefficients and the presence of a minimum in diffusivity at a critical particle size. The curve in Figure 5.1.1 results due to an inverse relationship of Brownian diffusivity with particle radius (Equation 3.2.5) and a direct relationship of shear-induced diffusivity with the square of the particle radius (Equation 3.2.6). Brownian diffusion dominates for particles about a tenth of a micron or smaller in diameter while shear induced diffusion is seen to dominate for particles larger than a
tenth of a micron (It is noted that the shear-induced component of diffusivity is calculated very approximately and for a specific case in Figure 5.1.1. The location of the minimum in the diffusivity curve is hence only a rough estimate. Different concentrations in feed water and different shear rates of operation are other factors which, in addition to particle size, determine shear-induced diffusivity and hence the location of the minimum in the diffusivity curve).

The "tenth of a micron" range thus represents a band of unfavorable particle sizes. The minimum is seen to move slightly towards larger particles at higher Reynolds numbers. This can be explained by the fact that the shear rate is lower in the case of the MF scenario, resulting in a lower value of the shear induced diffusion coefficient. At the channel inlet, the shear rate for the MF scenario is calculated to be lower by a factor of about 2.3 than for the UF element. A smaller value of the cross-flow velocity to element diameter ratio is the cause for the lower shear rate for the MF element as compared to the UF element (Equation 3.2.9). However, as mentioned earlier in Section 5.1, the cross-flow velocity in the MF simulations has been kept artificially low so that the flow may be interpreted as laminar.

5.1.1.2 Critical Point

The average cake thickness and consequently the permeate flux are both determined by the occurrence of the "critical point" which denotes the onset of stagnant particle layer growth. For UF the theoretical occurrence
of the critical point as predicted by the model is illustrated in Figure 5.1.2.

![Graph showing the effect of particle size on critical point](image)

**Figure 5.1.2** Effect of particle size on critical point as predicted from UF membrane simulations.

The cake layer is seen to form very close to the channel inlet for particles which have been predicted to have lowest diffusivity. For relatively smaller and larger particles effective diffusivity is seen to move the critical point further away from the inlet almost exponentially. For points in the figure above the dotted line, which represents an element length of 120 cm, no cake formation is predicted to occur in the membrane due to the high diffusivity associated with the particles in this size range. For the case of MF (Figure 5.1.3.), a stagnant cake formation is predicted over almost all of the particle size range; all of the curves fall fully under the dotted line representing the element length of 85 cm. The critical point forms relatively closer to the membrane inlet, as compared with the UF
membranes, due to lower shear rate as well as higher initial permeate flux (Equation 3.3.12). Concentration changes have a significant effect on the critical length; an increase in concentration from 10 mg/L to 200 mg/L moves the critical point closer to the channel inlet by approximately one order of magnitude for both the UF and MF membranes.

![Graph showing critical point vs. particle diameter for different concentrations](image)

**Figure 5.1.3** Effect of particle size on critical point as predicted from MF membrane simulations.

### 5.1.1.3 Cake Growth

The average cake thicknesses as simulated for the UF and MF facilities are depicted in Figure 5.1.4 for UF and Figure 5.1.5 for MF. In both cases the importance of particle size in determining the cake thickness is well illustrated. Maximum cake growth occurs in the particle size range where particle diffusion is at a minimum. For UF, a cake thickness less
than one micron is predicted for materials up to 0.01 μm in size, while for MF this number moves up slightly to 0.02 μm. This behavior is attributed to a lower shear stress for the MF element, as described in Section 5.1.1.1. The UF membrane is predicted to form no cake at all even at high concentrations of 200 mg/L, for particles either smaller than about 0.002 μm or larger than about 4 μm; due to the critical length being predicted to be greater than the element length (Figure 5.1.2). However, at 4 μm particle size cake thickness is very significant for MF (Figure 5.1.5). This can again be attributed to the fact that the shear induced diffusivity component is comparatively lower for the MF element at this particle size range. It can be inferred from Figures 5.1.6 and 5.1.7 that shear effects "start working" for UF at a particle size of about 0.3 μm while for the MF scenario considered here, the corresponding particle size is about 2 μm.

Figure 5.1.4  Effect of particle size on average cake thickness as predicted from UF membrane simulations.
Figure 5.1.5  Effect of particle size on average cake thickness as predicted from MF membrane simulations.

Figure 5.1.6  Effect of particle size on average cake thickness as predicted from UF membrane simulations. Cake thickness represented on a linear scale.
Figures 5.1.6 and 5.1.7 represent the cake thicknesses on a linear scale instead on a logarithmic scale to highlight the increase in cake thickness in the unfavorable particle size range. Cake thickness is seen to increase with increasing feed suspension concentrations, as expected.

![Graph showing effect of particle size on average cake thickness](image)

**Figure 5.1.7** Effect of particle size on average cake thickness as predicted from MF membrane simulations. Cake thickness represented on a linear scale.

### 5.1.1.4 Permeate Flux

The permeate flux at any specified pressure is a function of the total resistance offered to permeation; which constitutes the intrinsic resistance of the membrane and the hydrodynamic resistance due to the cake (Equation 3.2.1). The resistance offered to permeation by the stagnant particle layer, or cake, is a function of both the thickness of the cake as well as its porosity. Smaller particles form highly compact cakes due to
porosity being inversely proportional to the square of the particle radius (Equation 3.2.2). The variance of the average permeate flux with particle size, as predicted by the model, is depicted in Figures 5.1.8 and 5.1.9 for UF and MF respectively.

![Figure 5.1.8](image.png)

**Figure 5.1.8** Effect of particle size on average permeate flux as predicted from UF membrane simulations.

The "clean water" flux for UF, when only the membrane restricts permeate flow, was calculated to be 245 L/m²/hr. This value is approached at the extremes of the particle size range, since no cake formation is predicted at these sizes. Very low flux rates are observed for both the simulation sets in the range of unfavorable particle sizes due to very resistive cakes. In the case of the UF membrane (Figure 5.1.8) for particles in the size range of 0.1 μm to 0.01 μm flux increases with decreasing particle size even though the cake porosity decreases, due to an
increase in Brownian diffusion which results in a rapid decline in the cake thickness. From 0.01 \( \mu m \) to 0.001 \( \mu m \) this increase in flux is all the more rapid due to a very steep decline in cake thickness to the point where no cake is formed at all (Figure 5.1.4) and flux is limited only by the membrane resistance. From 0.1 \( \mu m \) to 1 \( \mu m \), the size range where very thick cakes have been predicted, the decrease in porosity of the cakes with increasing particle size results in a rise in permeate flux. Shear effects help to increase the flux very rapidly. The resistance of the cake still exceeds the membrane resistance and limits the permeation flux. Finally, from 1\( \mu m \) to 10 \( \mu m \) the flux is again predicted to increase to a maximum limiting value due to shear effects reducing cake thickness drastically. Towards the upper limit of this particle size range the cake resistance is very low and the membrane resistance limits the permeate flux.

![Graph showing the effect of particle size on average permeate flux](image)

Figure 5.1.9 Effect of particle size on average permeate flux as predicted from MF membrane simulations.
The "clean water" flux value for the MF membrane was calculated to be 1075 L/m²/hr. Figure 5.1.9 shows that this value is approached only for particles larger than about 5 μm where significantly thick (Figure 5.1.7) but very porous cakes do not offer much resistance to permeate flow and the membrane itself limits permeation. The general trend of flux change with particle size is similar in the case of MF as for UF. The more "spread out" nature of the permeate flux curves for MF, as compared to the UF curves, is due to the fact that the MF membrane has a lower membrane resistance (Table 5.1.1). Hence, the curves in Figure 5.1.9 represent permeate flux values which are primarily limited by the cake resistance over (0.001 μm to 5 μm) almost the entire particle size range.

5.1.1.5 Costs

The significant effect of particle size on the calculated permeate flux is reflected on the cost of facility design and treatment. Total cost, capital cost and operating cost for UF are presented in Figures 5.1.10, 5.1.11 and 5.1.12 respectively. Total cost, capital cost and operating cost for MF are presented in Figures 5.1.13, 5.1.14 and 5.1.15 respectively.

The minimum in particle diffusivity (Figure 5.1.1) produces a maximum in the total cost function. A maximum in UF costs is predicted at a particle diameter of about 0.07 μm (Figure 5.1.10). MF costs are predicted to be at a maximum for particle diameters near 0.15 μm (Figure 5.1.13). In all cases, costs increase with increasing feed suspension concentration. The cost of treatment by conventional flocculation-
Figure 5.1.10 Effect of particle size on total cost as predicted from UF membrane simulations.
Figure 5.1.11 Effect of particle size on capital cost as predicted from UF membrane simulations.

Figure 5.1.12 Effect of particle size on operating cost as predicted from UF membrane simulations.
Figure 5.1.13 Effect of particle size on total cost as predicted from MF membrane simulations.
Figure 5.1.14 Effect of particle size on capital cost as predicted from MF membrane simulations.

Figure 5.1.15 Effect of particle size on operating cost as predicted from MF membrane simulations.
sedimentation-filtration is estimated to be about 0.5 $/m^3$ for a 200 m$^3$/hr capacity plant (James M. Montgomery, 1992). In a worst case analysis, where all particles in the raw water are in the range of 0.01 $\mu$m to 0.1 $\mu$m, it appears that UF may be cost effective as a treatment alternative for feed waters containing about 20 mg/L or less of solutes (Figure 5.1.10).

At a plant capacity of 200 m$^3$/hr, which was the constant baseline value in the simulations presented here, capital costs for both UF and MF (Figure 5.1.16) contributed 75% to 85% of the total treatment cost. The capital and operating components of total costs depend on the raw water quality. The percent operating cost is seen to increase in the particle size region of unfavorable transport. However, the cost of energy is not

![Figure 5.1.16 Capital and operating costs as a percent of total costs for UF and MF simulations. Cost data corresponds to that generated for the 20 mg/L feed suspension concentration of raw water.](image)
responsible for this behavior. Rather, the membrane replacement costs increase more rapidly than the capital cost with an increased membrane area requirement. It is also noted from Figure 5.1.16 that the percent increase in operating cost is more for the MF facility than for the UF facility. This can be attributed to the above discussion coupled with the fact that the MF membrane cost per unit area is more than the corresponding cost for the UF membrane (Table 5.1.1).

5.1.2 Concentration Effects

Effects of the concentration of the feed suspension on cost and other parameters such as permeate flux are evident in the sensitivity analysis presented in the previous section. This section focuses on the role of mass concentration as a raw water quality variable affecting capital and operating costs. A set of five particle sizes is chosen to study the effect of different particle sizes over a range of concentration; 0.001 µm, 0.01 µm, 0.1 µm, 1 µm and 10 µm. The range of feed suspension concentration considered is from 10 mg/L to 200 mg/L, representing typical values expected for raw water sources in water treatment.

5.1.2.1 Permeate Flux

Figures 5.1.17 and 5.1.18 show the predicted effect of concentration on permeate flux for various particle sizes, for UF and MF respectively. In all cases permeate flux is predicted to decrease with increasing mass concentration. For the MF membrane (Figure 5.1.18) this is seen to be true for all the particle sizes considered, but the effect is seen to be much
more pronounced for the smaller particles (0.001 μm) as well as for the particles in the "unfavorable size range" (0.01 μm and 0.1 μm). This behavior can be attributed to the fact that for smaller particles the shear induced diffusion component is negligible and concentration effects are not appreciable. For larger particles, as the concentration increases, so does the shear induced diffusivity, and hence the flux is seen to "recover" from concentration effects in the sense that the flux decline is comparatively very less. However, for the UF simulations (Figure 5.1.17) both very small particles (0.001 μm) and very large particles (10 μm) seem to be independent of concentration effects. This paradox is resolved by the fact that for the UF membrane the model had predicted no cake formation for particles as small 0.001 μm (Figure 5.1.6) at all concentrations up to 200 mg/L. For UF, the critical point for the 0.001 μm and 10 μm particles is predicted to fail "outside" the length of the element (Figure 5.1.2).

![Figure 5.1.17](image.png)

Figure 5.1.17 Effect of feed suspension concentration on average flux as predicted from UF membrane simulations (curves for 0.001 μm and 10 μm particles overlap).
5.1.2.2 Costs

Figures 5.1.19 and 5.1.20 present the cost results with respect to concentration effects. The effect of concentration on costs is most dynamic for particles in the unfavorable size range (0.01 μm and 0.1 μm) for both UF and MF, and observable for smaller particles (0.001 μm) in the case of MF. This can be inferred to follow directly from the discussion in the previous section on the larger particles (1 μm and 10 μm) being able to "recover" from concentration effects. Costs are small and almost constant, over the whole concentration range, for the 1 μm and the 10 μm particles. The curve for the 0.001 μm particle size is also associated with very low costs, but rises with increasing concentration in case of MF. For the 0.01 μm and the 0.1 μm particles this rise in cost with concentration is more
Figure 5.1.19 Effect of feed suspension concentration on total cost as predicted from UF membrane simulations (curves for 0.001 μm and 10 μm particles overlap).

Figure 5.1.20 Effect of feed suspension concentration on total cost as predicted from MF membrane simulations.
evident. Table 5.1.2 presents the percent increase in cost as the concentration increases from 10 mg/L to 200 mg/L. Within the concentration range considered, MF is seem to be more sensitive to concentration effects than UF.

Table 5.1.2 Percent increase in total treatment cost with increase in feed suspension concentration from 10 mg/L to 200 mg/L, for various particle diameters.

<table>
<thead>
<tr>
<th>Process</th>
<th>Percent increase in cost</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.001 $\mu$m</td>
</tr>
<tr>
<td>UF</td>
<td>0 %</td>
</tr>
<tr>
<td>MF</td>
<td>92 %</td>
</tr>
</tbody>
</table>

5.1.3 Cost Analysis

Module geometry and operating conditions are important parameters in determining treatment costs and, unlike the raw water quality, can be controlied in an effort to achieve optimai configurations and operating conditions. The effects of feed pressure, recovery and element diameter and length on treatment costs are investigated in this section. Cost behavior as a function of plant capacity is also studied. All the simulations presented are based on a feed suspension concentration of 20 mg/L. Sensitivity analyses are performed on the independent variable in each graph. The operating time varies as a function of this sensitivity parameter to produce the specified recovery. Unless, mentioned otherwise, all other parameters presented in Table 5.1.1 are fixed at the constant baseline values.
5.1.3.1 Effect of Plant Capacity

The effect of plant capacity is illustrated in Figure 5.1.21 for UF and Figure 5.1.22 for MF. Three different capacities are considered; 200 m³/hr, 2000 m³/hr and 20000 m³/hr. The total cost per unit of water produced decreases significantly over the entire range of particle size with increasing capacity. At higher capacities the marginal cost of building the plant decreases very rapidly (Equation 2.2.31). Figures 5.1.21 and 5.1.22 depict that the most dramatic effect of change in cost with a change in capacity is at the particle size where cost peaks occur. This is true in an absolute sense, but as illustrated in Table 5.1.3, the percent decrease in cost at the peaks is actually less than at the extremes of the x-axis representing the smallest and largest particle sizes. This is because of the decreasing effect of economies of scale with increasing number of membrane modules (Equation 2.2.31). Moreover, Table 5.1.3 illustrates that the effect of capacity is slightly greater for UF costs than for MF costs, since UF numbers are higher than corresponding MF numbers.

Table 5.1.3 Percent decrease in total treatment cost with increase in plant capacity for various particle diameters.

<table>
<thead>
<tr>
<th>Increase in capacity (m³/hr)</th>
<th>UF</th>
<th></th>
<th></th>
<th>MF</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.001 μm</td>
<td>0.08 μm (Peak)</td>
<td>10 μm</td>
<td>0.001 μm</td>
<td>0.16 μm (Peak)</td>
</tr>
<tr>
<td>200 - 2000</td>
<td>33.7%</td>
<td>29.3%</td>
<td>33.7%</td>
<td>34.7%</td>
<td>25.7%</td>
</tr>
<tr>
<td>2000 - 20000</td>
<td>27.9%</td>
<td>22.7%</td>
<td>27.9%</td>
<td>27.3%</td>
<td>19.1%</td>
</tr>
</tbody>
</table>
Figure 5.1.21  Effect of plant capacity on total cost as predicted from UF membrane simulations.

Figure 5.1.22  Effect of plant capacity on total cost as predicted from MF membrane simulations.
5.1.3.2 Effect of Feed Pressure

The effect of feed pressure on cost is illustrated in Figures 5.1.23 (UF) and 5.1.24 (MF). Changes in feed pressure affect the permeate flux produced through the membrane and hence directly affect process costs. The overall effect is a decrease in cost with increasing feed pressure; this can be explained by the fact that though the costs of energy are expected to increase with an increased feed pressure, the flux increase and hence decrease in membrane area required to produce the design flow is so much more in this case, that the lowered capital costs more than offset the increase in energy costs. Moreover, as described in Section 5.1.1.5 operating costs account for only about 20% of the total cost (Figure 5.1.16) for a 200 m³/hr capacity plant. All the curves are seen to merge

![Graph showing the effect of feed pressure on total cost as predicted from UF membrane simulations.](image)

Figure 5.1.23 Effect of feed pressure on total cost as predicted from UF membrane simulations.
Figure 5.1.24 Effect of feed pressure on total cost as predicted from MF membrane simulations.

together in the range of 0.01 - 0.03 μm particle size due to similar values of limiting permeate flux in this region for all the three values of the feed pressure. That is, for this region, permeate flux was seen to asymptotically approach similar minimum values. This happens due to the mass transport limited behavior of permeate flux for particles in this size range as illustrated in Figure 5.1.25. The curves separate again at very small particle sizes (Figures 5.1.23 and 5.1.24) due to the permeate flux being limited only by the membrane resistance. Figure 5.1.25 thus signifies that the effect of feed pressures on treatment cost would be minimal or significant depending on the raw water quality.
Figure 5.1.25 Illustration of mass transport limited behavior of permeate flux for certain sized particles. Curves correspond to UF membrane simulations.

5.1.3.3 Effect of Element Diameter

The size of the element diameter controls the membrane area available in a module for a specified packing density. Hence, it is expected that smaller elements would result in an decreased cost of treatment due to increased membrane area available per module. However, there is a potential for more operational problems associated with smaller elements since element plugging and large pressure drops across modules could result. Figures 5.1.26 and 5.1.27 present the effect of element diameter on cost. To study the effect of element diameter on the costs of treatment in a reasonable fashion, the "clean water" pressure drops across modules were constrained to the value obtained for the baseline element diameter. The model option of calculating the cross-flow velocity for a specified pressure
Figure 5.1.26 Effect of element diameter on total cost as predicted from UF membrane simulations.

Figure 5.1.27 Effect of element diameter on total cost as predicted from MF membrane simulations. Analysis was restricted to element diameters 3 mm or smaller to avoid turbulent flow.
drop was then employed to estimate cross-flow velocity for the different element diameters. The cost analysis in Table 5.1.4 illustrates that for both UF and MF the treatment cost decreases very significantly with decreasing element size. Contrary to the effect of capacity on cost, the effect is absolutely as well as relatively most dramatic for the particles that are poorly transported by diffusion processes.

Table 5.1.4 Percent decrease in total treatment cost with decrease in element diameter for various particle diameters. Cross-flow velocity for different element sizes calculated assuming similar "clean water" pressure drops as for the baseline element size.

<table>
<thead>
<tr>
<th>Decrease in diameter (mm)</th>
<th>0.001 µm</th>
<th>Peak</th>
<th>10 µm</th>
<th>Decrease in diameter (mm)</th>
<th>0.001 µm</th>
<th>Peak</th>
<th>10 µm</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.84-0.92</td>
<td>40.9%</td>
<td>50.4%</td>
<td>40.9%</td>
<td>3-2</td>
<td>23.3%</td>
<td>32.6%</td>
<td>22.8%</td>
</tr>
<tr>
<td>0.92-0.46</td>
<td>17.1%</td>
<td>47.2%</td>
<td>17.1%</td>
<td>2-1</td>
<td>17.3%</td>
<td>53.2%</td>
<td>1.0%</td>
</tr>
</tbody>
</table>

5.1.3.4 Effect of Element Length

The effect of element length on treatment costs was examined (Figures 5.1.28 and 5.1.29) using cross-flow velocities calculated by assuming a constant "clean water" pressure drop (that corresponding to the baseline element length), as described in the previous section (Section 5.1.3.3). The relative effect is more at the particle size extremes and less at the peaks (Table 5.1.5). However, the percent change in cost is significant throughout the particle size range. Longer elements, according to the model, are expected to be more cost effective. This happens because at the specified capacity capital costs account for about 80% of the total
Figure 5.1.28 Effect of element length on total cost as predicted from UF membrane simulations.

Figure 5.1.29 Effect of element length on total cost as predicted from MF membrane simulations. Analysis was restricted to element length 85 cm or larger to avoid turbulent flow.
Table 5.1.5 Percent decrease in total treatment cost with increase in element length for various particle diameters. Cross-flow velocity for different element sizes calculated assuming similar "clean water" pressure drops as for the baseline element size.

<table>
<thead>
<tr>
<th></th>
<th>UF</th>
<th>MF</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Increase in length</td>
<td>Increase in length</td>
</tr>
<tr>
<td></td>
<td>(cm)</td>
<td>(cm)</td>
</tr>
<tr>
<td></td>
<td>0.001 μm</td>
<td>0.001 μm</td>
</tr>
<tr>
<td>60-120</td>
<td>35.8% 13.6% 35.9%</td>
<td>85-127.5 8.5% 4.0% 18.3%</td>
</tr>
<tr>
<td>120-240</td>
<td>25.4% 11.1% 25.4%</td>
<td>127.5-170 2.6% 3.4% 14.5%</td>
</tr>
</tbody>
</table>

costs and even though longer elements result in higher energy costs, this increase is far outweighed by the decrease in capital costs.

5.1.3.5 Effect of Recovery

The effect of changing recovery on the total cost of treatment is depicted in Figure 5.1.30 (for UF) and Figure 5.1.31 (for MF). It is seen that the effect is variable depending on the particle size and hence the raw water quality. It should be noted that besides recovery the other variable in these simulations is the backflush/fastflush frequency. The effect of recovery on permeate flux was consistent over the particle size range considered in this work. Both UF and MF permeate fluxes decreased with increasing recovery. The number of modules required, however, is calculated as a function of the permeate flux as well as the backflush frequency. The backflush frequency in these simulations is itself calculated as a function of the permeate flux and the recovery (see Section 3.6.2). For UF (Figure 5.1.30), it is seen that flux controls costs at the "unfavorable" particle sizes; flux decreases with increasing recovery and consequently the
Figure 5.1.30  Effect of recovery on total cost as predicted from UF membrane simulations.

Figure 5.1.31  Effect of recovery on total cost as predicted from MF membrane simulations.
cost increases. Over the favorable particle sizes, where flux is quite high, the frequency of backflushing is seen to control costs; though the flux decreases with increasing recovery, the cost too is seen to decrease since the backflushing frequency decreases. For MF (Figure 5.1.31) this interplay between flux and frequency of hydrodynamic cleaning is less prominent, and the latter is seen to almost always control costs.

5.2 Model Behavior - Permeate Quality and Removal of Natural Organic Matter (NOM)

Model predictions on technical and economic performance of UF and MF membranes as a function of colloidal particle size and concentration have been addressed in the previous section (5.1). In this section the capability of the model to predict removal of organic matter, which usually constitutes complex molecular structure, is investigated. The removal model is "tuned" to handle organic matter by calibrating it with an average geometrical descriptor of organic molecules.

There are numerous factors determining the overall efficiency of a filtration process. However, selection of a particular process in achieving drinking water quality, as specified by the regulatory agencies, should necessarily consider the goals to be met on the permeate quality. Hence, permeate quality is an important limiting criterion in choosing membranes. UF and MF membranes have been shown to effectively remove particles as well as microbial contaminants; Jacangelo et al., 1989; Heneghan et al., 1991; Lainé et al., 1991; Jacangelo et al., 1991; Wiesner et al., 1991;
Olivieri et al., 1991. These studies report that typical UF membranes can achieve effectively absolute removal of colloidal matter and microorganisms such as bacteria, protozoans and viruses.

There is growing concern on the role of NOM as a precursor to DBP formation. This necessitates investigations into removal of NOM through membranes. The removal model studied in this thesis can be utilized to predict the rejection of organic matter by the method detailed in the next section. Results on a few test cases are also presented in Section 5.3 where raw waters having varying quantities of colloidal as well organic matter are considered for UF treatment. It should however be noted that UF membranes are not typically capable of removing large fractions of NOM.

**Calibration of the Permeate Quality Model for Dissolved Organic Carbon (DOC) Removal**

The model for estimating solute rejection by membranes can be used for predicting the removal of organics, given a size distribution of the organic material and the pore size of the membrane. However, when either or both of solute size and membrane pore size are given in terms of molecular weight or MWCO then a correlation is required to relate molecular weight to dimensional size. For organic matter, this essentially requires finding an "average descriptor" for the geometry/shape of organic molecules in the raw water. This average descriptor or equivalent diameter can then be used to determine the removal of the organic matter on a specific membrane.
The relation between effective diameter, $d_p$, and molecular weight, $MW$, is usually of the form of a power law:

$$d_p = k'(MW)^\alpha \quad (5.2.1)$$

where $k'$ and $\alpha$ are fitting constants. The exponent $\alpha$ assumes a range of values from 1/3 to 1. A value of 1/3 corresponds to a spherical geometry and values increasing to unity signify a more and more linear geometry. If the molecules are fractal, then Equation 5.2.1 can also be considered in the light of the fractal dimension ($D_f$) of a molecule. In terms of fractal dimensions Equation 5.2.1 can be expressed as:

$$d_p = k' (MW)^{V/D_f} \quad (5.2.2)$$

The dependence expressed in Equation 5.2.1 is thus seen to correspond to a range of fractal dimensions from 3 to 1. In this case a change in fractal dimension from 3 to 1 represents a change from spherical to linear morphology of molecules.

In this analysis it is assumed that the MWCO of membranes is estimated using spherical solutes ($D_f=3$) as is typically done using globular proteins. Under this assumption the corresponding correlation for the membrane pore size as determined by Equation 5.2.2 and reported in literature is:

$$d_o(\mu m) = 0.147 \times 10^{-4} (MWCO)^{0.33} \quad (5.2.3)$$
Using membranes as a series of "measuring sticks", data on removal by the membranes can be interpreted in terms of an average descriptor of molecule morphology; here considered as a fractal dimension. A calibration was carried out to determine the likely fractal dimension for DOC by using experimental data available in literature on DOC removal by membranes. The data of Taylor et al. (1987) on removal of organic material from raw water obtained from Florida groundwaters were employed for this purpose. In their work Taylor et al. reported DOC removals from Florida groundwater sources using membranes of several MWCO's. Data for one such site are summarized in Table 5.2.1. Based on these data the molecular weight distribution of DOC in the raw water can be calculated as in Table 5.2.2.

Figure 5.2.1 shows the calibration of the removal model for DOC removal using the data presented in Tables 5.2.1 and 5.2.2. The experimental removal values reported by Taylor et al. are represented by the solid circles. The removals, as calculated from the removal model for the molecular weight distribution determined from the experimental data, were calculated for a range of MWCO's and are represented by the solid line. Calibration was achieved using the following correlation for solute size:

\[ d_p (\mu m) = 0.08 \times 10^{-3} (MW)^{0.36} \]

(5.2.4)

It is interesting to note that the value of the exponent in this case (0.36) corresponds to a fractal dimension of 2.78; this appears to indicate a fairly
compact geometry for these natural organic materials. Natural organic materials are likely to differ significantly from one raw water to another; the value of the exponent in Equation 5.2.4 should therefore vary as well. Nonetheless, for the purposes of subsequent calculations, it is assumed that the organic materials in hypothetical waters considered contain organic materials similar to those in the raw waters considered by Taylor and co-workers.

Table 5.2.1 Experimental data from Taylor et al. (1987) on removal of DOC from raw water. DOC concentration in the raw water was stated to be 15.0 mg/L.

<table>
<thead>
<tr>
<th>MWCO of membrane (Dalton)</th>
<th>DOC Removal (mg/L)</th>
</tr>
</thead>
<tbody>
<tr>
<td>100</td>
<td>0.6</td>
</tr>
<tr>
<td>400</td>
<td>1.4</td>
</tr>
<tr>
<td>2000</td>
<td>7.4</td>
</tr>
<tr>
<td>10000</td>
<td>12.6</td>
</tr>
<tr>
<td>20000</td>
<td>14.2</td>
</tr>
</tbody>
</table>

Table 5.2.2 MW distribution of DOC in the raw water as determined from the removal data presented in Table 5.2.1

<table>
<thead>
<tr>
<th>Molecular Weight (Dalton)</th>
<th>DOC Concentration (mg/L)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0-100</td>
<td>0.6</td>
</tr>
<tr>
<td>100-400</td>
<td>0.8</td>
</tr>
<tr>
<td>400-2000</td>
<td>6.0</td>
</tr>
<tr>
<td>2000-10000</td>
<td>5.2</td>
</tr>
<tr>
<td>10000-20000</td>
<td>1.6</td>
</tr>
<tr>
<td>&gt;20000</td>
<td>0.8</td>
</tr>
</tbody>
</table>
Figure 5.2.1 Calibration of the removal model for DOC removal using data from Taylor et al. (1987)

5.3 Performance and Cost Calculations for Example Waters

Sections 5.1 and 5.2 have been utilized to explore the capabilities of the model formulated in this study to predict the permeate flux and the permeate quality with respect to both colloidal as well as organic matter when using low pressure membrane processes. In this section, calculations for a set of four hypothetical raw waters are reported. The approach was to consider the following combinations of particle concentration and TOC concentration:

1. Low Particle Concentration - Low TOC
2. Low Particle Concentration - High TOC
(3) High Particle Concentration - Low TOC
(4) High Particle Concentration - High TOC

The test cases were based on experimentally available raw water quality data. The data for low and high particle concentrations correspond to the Trinity and San Jacinto rivers (Texas), respectively (Wiesner, 1993). Low TOC data are from analyses on samples obtained from the river Oise in France and the high TOC data are from the river Sam Houston (Texas) (Ducellier and Devitt, 1993). The particle size distributions and the TOC distributions are presented in Figures 5.3.1, 5.3.2, 5.3.3 and 5.3.4. Total TOC in the low concentration case is 3 mg/L while for the high concentration case it is 10 mg/L. The case for low particle concentration has 28.86 mg/L of suspended particles while the high concentration case has 51.76 mg/L of particles.

In calculating the removal of TOC it is assumed that the correlation (Equation 5.2.4) developed for the data from Taylor et al. (1987) can be used to describe the geometry of the TOC in the test waters considered in this section. In other words, it is assumed that the organics in general all share the same morphology independent of their molecular weight, concentration and origin. In fact, it would appear likely that the morphology of natural organic matter would vary as well as its MW distribution, composition and other characteristics.

Results

The equivalent pore size of the 100,000 Dalton MWCO UF
Figure 5.3.1  Particle size distribution for low concentration test case. (Total concentration = 28.86 mg/L)

Figure 5.3.2  Particle size distribution for high concentration test case. (Total concentration = 51.76 mg/L)
Figure 5.3.3 TOC molecular weight distribution for low concentration test case. (Total concentration = 3 mg/L)

Figure 5.3.4 TOC molecular weight distribution for high concentration test case. (Total concentration = 10 mg/L)
membrane employed in these simulations is about 0.007 μm (Equation 5.2.3). Since all the particle sizes (Figures 5.3.1 and 5.3.2) are larger than the membrane pore size, total rejection of particles was obtained in all of the four simulations. The removal of TOC is presented in Figure 5.3.5, for both the high TOC and low TOC cases. Removals are identical for both of

![Graph showing molecular weight vs. removal percentage](image)

Figure 5.3.5 Percent removals of the various molecular weight fractions for representative TOC distributions.

these cases because of similar size classes (however, note that the low TOC distribution does not have data for 100-300 kD size class; Figure 5.3.3). The difference in concentration has no effect on removal, according to the removal model used. The removal values plotted in Figure 5.3.5 are based on the bulk and permeate concentrations. Total removal for the high TOC case is predicted to be 86.66% while it is only 28.27% for the low TOC case. This is as expected since the high TOC case has more
molecules distributed over higher molecular weights compared to the case of low TOC (Figures 5.3.3 and 5.3.4). Removals are more appropriately calculated based on the feed concentration rather than the bulk concentration. In this case the corresponding total removals are 24.6% for high TOC case and only 2% for low TOC case (Table 5.3.1). These removals are low due to the relatively large MWCO of the UF membrane (100,000 kDalton).

Table 5.3.1 Feed, bulk and permeate concentrations for TOC representative cases.

<table>
<thead>
<tr>
<th>Concentrations/Removal</th>
<th>Low TOC</th>
<th>High TOC</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed Concentration (mg/L)</td>
<td>3.0</td>
<td>10.0</td>
</tr>
<tr>
<td>Bulk Concentration (mg/L)</td>
<td>4.1</td>
<td>56.6</td>
</tr>
<tr>
<td>Permeate Concentration (mg/L)</td>
<td>2.94</td>
<td>7.54</td>
</tr>
<tr>
<td>Removal (%)</td>
<td>2</td>
<td>24.6</td>
</tr>
</tbody>
</table>

The average permeate flux predicted by the model for the test cases is presented in Table 5.3.2. The flux is primarily determined by the mean solute size calculated using Equation 3.4.2 for the polydisperse feed suspension. Flux is highest for the low particle - low TOC case. The low particle - high TOC test case is predicted to have the lowest permeate flux. This is due to the high concentration of TOC molecules in the "unfavorable size range" of about 0.1 μm (Figure 5.3.4). The high particle - low TOC scenario has a lower predicted permeate flux compared to the low particle - low TOC case due to a higher particle concentration. The low TOC for these two cases does not have many molecules falling in the unfavorable
Table 5.3.2  Average permeate flux and costs for the representative cases. Simulations employed the UF membrane data listed in Table 5.1.1

<table>
<thead>
<tr>
<th>Test Case</th>
<th>Permeate Flux (L/m²/hr)</th>
<th>Total Cost ($/m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Low Particle Concentration - Low TOC</td>
<td>234</td>
<td>0.12</td>
</tr>
<tr>
<td>Low Particle Concentration - High TOC</td>
<td>51</td>
<td>0.40</td>
</tr>
<tr>
<td>High Particle Concentration - Low TOC</td>
<td>179</td>
<td>0.15</td>
</tr>
<tr>
<td>High Particle Concentration - High TOC</td>
<td>91</td>
<td>0.25</td>
</tr>
</tbody>
</table>

size range for permeate flux (Figure 5.3.3). Hence from low particle - low TOC case to high particle - low TOC case the permeate flux reduces due to concentration effects. Finally the high particle - high TOC case is predicted to have greater flux than low particle - high TOC probably due to the fact that in the former case large particles (≈ 1 μm or larger) are present in a higher concentration, driving the mean particle size away form the unfavorable size range. In this instance particle size effects are important (unlike in the instance of low particle - low TOC and high particle - low TOC comparison) since the TOC molecules in the unfavorable size range are not negligible, but rather prominent. Hence, in this instance particle size effects are seen to override the effect of increased concentration.

The costs associated with the treatment of the four test cases are presented in Table 5.3.2. Figure 5.3.6. depicts the capital and operating cost components. Since cost is primarily determined by the permeate flux, the cost results presented in Table 5.3.2 follow directly from the behavior of the permeate flux. The trend is seen to be similar for both the capital as
well as the operating cost components. Highest treatment costs are therefore predicted to be associated with the case of low particle - high TOC. In this case the TOC is seen to be the controlling parameter. The low particle - low TOC case is the least expensive to treat, as expected. Total costs in all instances are predicted to be quite competitive to conventional treatment costs, which are estimated to be $0.5/m³ for a 200 m³/hr facility.

Figure 5.3.6 Capital and Operating costs for the representative cases. Simulations employed the UF membrane data listed in Table 5.1.1.
Chapter 6

Conclusions

6.1 Conclusions from Computer Simulations

The size distribution of colloidal materials in the raw water appears to be an important variable in determining the performance and cost of UF systems. The quantitative variance of two diffusive mechanisms, Brownian diffusion and shear-induced diffusion, conflicting with respect to particle size, results in an unfavorable range of particle diameters, around 0.1 \( \mu m \), where the back transport of colloids approaches a minimum. Particles in this size range have maximum tendency to form deposits on the membrane surface. These deposits offer hydrodynamic resistance to permeation and result in lowered rate of permeate flux. Since permeate flux is the dominant factor in determining the cost of membrane filtration, the latter increases significantly for raw water containing particles in the size range of about 0.1 \( \mu m \). A model for particle transport that includes the effects of both Brownian and shear-induced diffusion predicts mass-transfer limited (pressure independent) permeate rates when particles in the range of about 0.005 \( \mu m \) to 0.03 \( \mu m \) accumulate near the membrane for conditions typical of hollow fiber ultrafiltration membranes.

The cost for membrane filtration increases with increasing particle concentrations. Based on the transport considerations included in the model described in this study, membrane filtration appears to be cost effective
when compared with conventional treatment processes for low particle concentrations in the raw water. For a typical UF facility, as simulated in this work, treatment costs are predicted to be comparable to conventional treatment costs for feed water concentrations of 20 mg/L or less under a worst case scenario of particle size. A more favorable size of material in the raw or feed water may improve membrane filtration performance to the extent that mass concentration as high as 200 mg/L could be treated at a cost competitive with conventional treatment. The model associates high treatment costs for a suspension containing particles of a mean diameter about 0.08 μm. However, it should be noted that pretreatment aimed at aggregating particles can shift an unfavorable particle size out of the unfavorable size range and treatment can be achieved at lower costs.

Treatment costs for membrane filtration facilities are predicted to be dominated by capital costs. Cost estimates obtained from simulations in this study indicate that, when operating conditions are kept within practical limits (constrained pressure drop across module), element geometry favorable to decreasing capital costs (smaller and longer elements) should result in optimal configuration. A similar conclusion is drawn with respect to operating parameters such as the feed pressure. In this case, even though higher feed pressure is expected to increase the cost of energy, this increase is far outweighed by the decrease in capital costs.

Some interesting conclusions can be drawn from the analysis of the results obtained from application of the model formulated in this study to several representative cases (section 5.3). Typical UF membranes (100,000
Dalton) are seen to be able to remove a very small percentage of organic matter. Most of the low molecular weight fractions are able to penetrate the membrane. Removal of natural organic matter is thus predicted to be poor using UF or MF membranes. The high molecular weight fractions, typically in the range of 100 kD - 0.2 μm, affect the permeate flux critically if present in sufficient quantity. The presence of larger colloidal material is predicted to improve the permeate flux. Raw waters with high concentrations of natural organic matter and low concentrations of particles are predicted to be the most costly to treat. This conclusion arises from the distribution assumed for the molecular weight of NOM in the high TOC scenario. It should be noted that only reversible impacts of NOM on permeate flux are considered. Raw waters with high concentrations of NOM can also be expected to adsorptively foul membranes, thereby increasing treatment costs.

6.2 Scope for Further Research

Extensive work has been done on mathematically modeling the performance of membrane filters. However, there are still many considerations which need to be studied and incorporated into an integrated mathematical theory. More work needs to be done primarily in the following areas:

(1) The effect of polydispersivity on cake formation is of great importance due to its practical significance. Mathematical models should handle more than a single particle size parameter in a more rigorous fashion.
(2) Membrane solute interactions in terms of pore blocking and adsorption affect the long term behavior of permeate flux. There is a need to incorporate mathematical theories of membrane fouling into the present analysis.

(3) High crossflow velocities are often employed in membrane filtration to minimize the effects of concentration polarization. Models need to be formulated for operation in turbulent flow regime.

(4) Current models, such as the one studied in this thesis, are limited to the consideration of thin boundary layers. The assumption of a constant wall shear stress becomes limiting for larger boundary layers. Future work should aim to develop theories which overcome this limitation.

(5) This analysis does not consider the dynamic performance of membrane filtration units. Time based behavior may become important in practice under conditions of high feed suspension concentrations.

(6) The costs of constructing and operating a membrane filtration facility are seen to be determined by a number of design parameters. An interesting area of work would be to utilize the performance and cost models in formulating an optimization problem of the membrane filtration process.
Appendix A

MEMSYS User's Guide

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A-1 Introduction

MEMSYS is an interactive and user friendly computer program for simulating the performance and cost of pressure driven membrane filtration processes. MEMSYS is a research product of an integrated mathematical model which can be used as a tool to analyze the important operational and economic parameters related to the running of a drinking water treatment facility utilizing low-pressure membrane filtration processes such as ultrafiltration or microfiltration as treatment techniques.

A-1.1 Objectives

The primary objectives of this software are:

- to estimate the average, steady state permeate flux through a membrane filter as a function of filter geometry, operating conditions and feed water quality.
- to estimate the permeate quality as a function of membrane and feed water characteristics.
- to estimate the capital and operating costs of membrane facilities.

A-1.2 Limitations and Simplifications

The primary limitations and simplifications can be summarized as under.

- the model holds in situations where the solute accumulation near the membrane is thin relative to the radius of the membrane element. The assumption of constant bulk suspension concentration limits the model to situations where the thicknesses of the cake formed on the membrane are small compared to the element radius. The solution would thus be an approximate one in situations where the channel gets "plugged".
- the performance model holds only under laminar flow of the suspension through the membrane modules.
- polydisperse suspensions are considered by calculating an average parameter or mean particle size.
- transient effects are not considered in evaluating the performance of the membrane filter. In other words, long term phenomena, such as pore fouling, are neglected.
- gravitational effects are not considered. Hence, all solute particles are assumed to be neutrally buoyant.
- the model to calculate permeate quality assumes membrane pores to be cylindrical and the solute particles to be spherical. Solute-Membrane interactions and concentration polarization are not considered.
A-1.3 Documentation and Conventions

This user's guide describes the features of the MEMSYS computer program and explains the procedures for using them. The scientific basis or the theory behind the computer program does not comprise a part of this documentation. The primary reference is the research thesis of which MEMSYS is a product.

There are a few important notational conventions followed in this documentation which are to be noted.

- CAPITAL type refers to items associated with the program.
- Italic type indicates important terms and is also used to highlight important concepts.
- Bold type is used to indicate the keys on the keyboard. (Titles of chapters and sections also appear in bold type).
- 'This kind' of type is used for computer output.

MEMSYS can be run under both the SI as well as the English system of units. However, for the purposes of illustration in this documentation, all input/output is shown in the SI system.

A-1.4 Technical Computer System Requirements

The program MEMSYS runs under the following computer environment:

- An IBM PC-AT / PC-XT or compatible. (PC-AT 386 or above highly recommended).
- A color graphics adapter (EGA, VGA or CGA) for graphics display.
- A math co-processor (optional but highly recommended).
- MS-DOS version 3.1 or later.
- At least 250 kilobytes (kB) of random-access memory (RAM).
- A parallel printer (should be connected to LPT1 or the PRN device).

MEMSYS can be run from a floppy disk too, but the performance will be slow, and storage space for output files might become limiting, hence it is recommended that the program be installed and run on a hard disk. Since it is a stand-alone application package, no other software is required in the execution of MEMSYS; the only essential requirement is the DOS operating system.

A-1.5 Installation and Getting Started

The basic requirements for getting MEMSYS up and ready for execution along with the general methodology of choosing commands from menus and status bars are discussed in the following sections.
A-1.5.1 Description of MEMSYS Files

There are 8 files on the MEMSYS disk as listed below.

<table>
<thead>
<tr>
<th>File Name</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>MEMSYS.EXE</td>
<td>executable file</td>
</tr>
<tr>
<td>HELVB.FON</td>
<td>font file used by MEMSYS.EXE</td>
</tr>
<tr>
<td>ROMAN.FON</td>
<td>font file used by MEMSYS.EXE</td>
</tr>
<tr>
<td>TMSRB.FON</td>
<td>font file used by MEMSYS.EXE</td>
</tr>
<tr>
<td>MS-GLOB.DAT</td>
<td>default data file</td>
</tr>
<tr>
<td>MS-PART.DAT</td>
<td>default data file</td>
</tr>
<tr>
<td>MS-TOC.DAT</td>
<td>default data file</td>
</tr>
<tr>
<td>MS-COST.DAT</td>
<td>default data file</td>
</tr>
</tbody>
</table>

The executable program is contained in the file 'MEMSYS.EXE'. The HELVB.FON, ROMAN.FON and TMSRB.FON files are used by the executable file to display graphical text for the graphics procedures in the program. The remaining four files with extension '.DAT' contain default data which the program tries to load automatically on startup. It is recommended to run the program with these default files present since running different simulations might require just simple edition of some of the default data. If these files are not present then the program initializes all input variables to zero.

A-1.5.2 Running MEMSYS from a Floppy Disk

This can be done by simply using the disk containing the MEMSYS files, or if required making a copy of the disk and then utilizing that to run MEMSYS. All MEMSYS files should be stored in a single directory and the program should be run from this directory. This is required for proper input/output operations. The program can be run by entering 'MEMSYS' at the DOS prompt. The following is an example when the program is stored in the directory 'MS' and run from the A drive:

```
A:MS> memsys
```

It is recommended that the program be stored and run from the hard disk. This will increase the speed of execution to a great extent.

A-1.5.3 Running MEMSYS from the Hard Disk

This requires copying all MEMSYS files to the hard disk. Follow these sequence of steps to install MEMSYS on a hard disk:

**STEP 1:** Create a directory for storing MEMSYS by using the DOS 'mkdir' command. For example to create a directory MS under the root directory enter 'mkdir MS':

```
C:> mkdir MS
```
STEP 2: Change the current directory to MS:

C:> cd MS

STEP 3: Copy all files from the MEMSYS floppy disk to the directory MS:

C:MS> copy a:*.*

This completes the installation procedure. Now the program can be run by entering 'MEMSYS' at the DOS prompt:

C:MS> memsys

A-1.5.4 Printing Requirements

MEMSYS prints files by exporting them to the PRN device. Hence printing requires that the printer be connected to the first parallel port; LPT1. To utilize the printing facility set the printer on-line and position the printer once to the beginning of a page. MEMSYS sends a form feed signal to the printer after printing a file to reset the top of page, so the printer does not have to be positioned again.

A-1.5.5 Menu Operation in MEMSYS

MEMSYS is a highly menu driven software. All options provided by the program have been organized into a hierarchy of menus. The menus share common features of operation for selecting an item or option (Fig. A-1.5.1):

```
Menu Title
Item # 1...
Item # 2...
Item # 3
Item # 4...
Item # 5
Item # 6

Enter  Esc
```

Fig. A-1.5.1 Format of a generic MEMSYS menu

- a menu item or option can be selected by using the UP arrow (↑) and DOWN arrow (↓) keys on the keyboard to highlight the required item and then pressing Enter. The Enter command is always listed at the bottom of the menu.
- another quick manner of making a selection without having to highlight an item and then pressing Enter is by pressing the highlighted character in the required item name.
• a menu (except the MAIN MENU) can be deactivated or closed by pressing the Esc key. This is used when the user decides to close a menu without making any selection or to return to a parent menu. This option, when available is always listed at the bottom of the menu.

• a menu item followed by an ellipsis (...) indicates that MEMSYS expects further information. This is usually achieved by presenting a sub-menu to the user.

A-1.5.6 Executing Commands Listed in the Status Bar

Any special commands available from a screen are always listed at the bottom of the screen in a highlighted status bar. These commands along with the keys that execute them are paired in angle (< >) brackets. For example, the following two commands are available from the screen which contains the MAIN MENU (only when the MAIN MENU is the active menu):

<F1=Help> <F3=Exit>
A-2 Input Requirements

The architecture of the program allows the performance model (or PERFORMANCE ANALYZER) and the cost model (or COST ESTIMATOR) to be run independently if required. To give some categorization to the numerous variables required as input parameters by the program, the program deals with four sets or files of data. These files organize the input data required by the PERFORMANCE ANALYZER (PA) and the COST ESTIMATOR (CE).

<table>
<thead>
<tr>
<th>DATA SET/FILE</th>
<th>REQUIRED BY</th>
</tr>
</thead>
<tbody>
<tr>
<td>Global data</td>
<td>PA &amp; CE</td>
</tr>
<tr>
<td>Particle size distribution</td>
<td>PA</td>
</tr>
<tr>
<td>TOC size distribution</td>
<td>PA</td>
</tr>
<tr>
<td>Cost data</td>
<td>CE</td>
</tr>
</tbody>
</table>

A-2.1 Naming Input Files

The files containing input data should always have an extension of ".DAT" for MEMSYS to recognize them as data files. When data files are saved using MEMSYS, the user is expected to enter the filename (up to a maximum of 8 characters) without the extension. The extension ".DAT" is automatically appended to the filename. Note that MEMSYS will not accept a filename with an extension and will repeatedly prompt the user to enter a filename till a valid filename is entered. The data file contains only numeric data corresponding to the variables belonging to the file. For the user's facility MEMSYS also creates a corresponding ".TXT" file whenever a ".DAT" file is created. The ".TXT" file is a formatted text file which contains the description of each variable along with its numeric value.

A-2.2 Contents of Input Files

The variables belonging to the four type of input files detailed at the beginning of this chapter are described in Tables A-2.1, A-2.2, A-2.3 and A-2.4. All the variables listed are of double precision type. The 'global data' file contains 10 variables. The last column in Table A-2.1 indicates whether the variable is required as an input by both the performance and cost models or just the performance model.

The particle and TOC distribution data sets or files contain data required by the PERFORMANCE ANALYZER. The particle-size and TOC molecular weight distributions are described in terms of records. The distributions can contain up to a maximum of 50 records each. Each record consists of two fields (or variables); the first field is the size class and the second field is the concentration pertaining to that size class. Note that the fields in the two distributions have different units, corresponding to standard method of reporting size analysis. Also, these are the only options of units available for describing the size distributions.
Table A-2.1. Input variables for the 'global data' input file

<table>
<thead>
<tr>
<th>Field #</th>
<th>Variable</th>
<th>SI Units</th>
<th>English Units</th>
<th>Definition</th>
<th>Required by</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>$D_e$</td>
<td>mm</td>
<td>in</td>
<td>Diameter of the membrane element</td>
<td>PA &amp; CE</td>
</tr>
<tr>
<td>2</td>
<td>$L_e$</td>
<td>cm</td>
<td>in</td>
<td>Length of the membrane element</td>
<td>PA &amp; CE</td>
</tr>
<tr>
<td>3</td>
<td>$P_f$</td>
<td>bar</td>
<td>psi</td>
<td>Feed pressure applied to the membrane</td>
<td>PA &amp; CE</td>
</tr>
<tr>
<td>4</td>
<td>$\bar{U}_o$</td>
<td>cm/s</td>
<td>ft/s</td>
<td>Average crossflow velocity at inlet of element</td>
<td>PA &amp; CE</td>
</tr>
<tr>
<td>5</td>
<td>$R_{rec}$</td>
<td>%</td>
<td>%</td>
<td>Percent recovery</td>
<td>PA &amp; CE</td>
</tr>
<tr>
<td>6</td>
<td>$T$</td>
<td>°C</td>
<td>°F</td>
<td>Temperature of the suspension</td>
<td>PA &amp; CE</td>
</tr>
<tr>
<td>7</td>
<td>$d_o$</td>
<td>µm</td>
<td>µm</td>
<td>Pore diameter of the membrane</td>
<td>PA</td>
</tr>
<tr>
<td>8</td>
<td>$R_m$</td>
<td>1/cm</td>
<td>1/in</td>
<td>Intrinsic resistance of the membrane</td>
<td>PA</td>
</tr>
<tr>
<td>9</td>
<td>$\rho_{oc}$</td>
<td>gm/cm³</td>
<td>lb/ft³</td>
<td>Density of the TOC</td>
<td>PA</td>
</tr>
<tr>
<td>10</td>
<td>$\phi_{max}$</td>
<td></td>
<td></td>
<td>Max. solute volume fraction at membranes wall</td>
<td>PA</td>
</tr>
</tbody>
</table>

Table A-2.2. Input variables for the particle size distribution input file

<table>
<thead>
<tr>
<th>Record Number</th>
<th>Field 1</th>
<th>Field 2</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>1-50</td>
<td>Size $\text{µm}$</td>
<td>Number Concentration $\text{#/mL}$</td>
<td>Field1 = Particle size Field2 = Particle number concentration</td>
</tr>
</tbody>
</table>
Table A-2.3. Input variables for the TOC molecular weight distribution input file

<table>
<thead>
<tr>
<th>Record Number</th>
<th>Field 1</th>
<th>Field 2</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>1-50</td>
<td>Size</td>
<td>Mass Concentration</td>
<td>Field1 = TOC molecular weight</td>
</tr>
<tr>
<td></td>
<td>(kD)</td>
<td>(mg/L)</td>
<td>Field2 = TOC Mass Concentration</td>
</tr>
</tbody>
</table>

Table A-2.4. Input variables for the cost data input file

<table>
<thead>
<tr>
<th>Field #</th>
<th>Variable</th>
<th>SI Units</th>
<th>English Units</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>( Q_{des} )</td>
<td>m(^3)/hr</td>
<td>mgd</td>
<td>Design capacity of the plant</td>
</tr>
<tr>
<td>2</td>
<td>( P_d )</td>
<td>%</td>
<td>%</td>
<td>Volumetric packing density of the module</td>
</tr>
<tr>
<td>3</td>
<td>( D_m )</td>
<td>cm</td>
<td>in</td>
<td>Diameter of the module</td>
</tr>
<tr>
<td>4</td>
<td>( c_{\text{mod}} )</td>
<td>$/\text{module}</td>
<td>$/\text{module}</td>
<td>Cost of one membrane module</td>
</tr>
<tr>
<td>5</td>
<td>( t_o )</td>
<td>min</td>
<td>min</td>
<td>Operating period between two flux enhancement cycles</td>
</tr>
<tr>
<td>6</td>
<td>( t_{ff} )</td>
<td>sec</td>
<td>sec</td>
<td>Duration of fastflush</td>
</tr>
<tr>
<td>7</td>
<td>( U_{ff} )</td>
<td>cm/s</td>
<td>ft/s</td>
<td>Fastflush Velocity</td>
</tr>
<tr>
<td>8</td>
<td>( t_{bf} )</td>
<td>sec</td>
<td>sec</td>
<td>Duration of backflush</td>
</tr>
<tr>
<td>9</td>
<td>( V_{bf} )</td>
<td>L/m(^2)/hr</td>
<td>gfd</td>
<td>Backflush Flux</td>
</tr>
<tr>
<td>10</td>
<td>( P_{bf} )</td>
<td>bars</td>
<td>psi</td>
<td>Backflush Pressure</td>
</tr>
<tr>
<td>11</td>
<td>( CH_d )</td>
<td>mg/L</td>
<td>mg/L</td>
<td>Coagulant Dose</td>
</tr>
<tr>
<td>12</td>
<td>( CH_c )</td>
<td>$/1000$kg</td>
<td>$/1000$lb</td>
<td>Coagulant Cost</td>
</tr>
<tr>
<td>13</td>
<td>( c_{kw} )</td>
<td>$$/\text{kw-hr}</td>
<td>$$/\text{kw-hr}</td>
<td>Energy Cost</td>
</tr>
<tr>
<td>14</td>
<td>( DL )</td>
<td>yrs</td>
<td>yrs</td>
<td>Design life of the plant</td>
</tr>
<tr>
<td>15</td>
<td>( ML )</td>
<td>yrs</td>
<td>yrs</td>
<td>Life of the membrane</td>
</tr>
<tr>
<td>16</td>
<td>( i_c )</td>
<td>%</td>
<td>%</td>
<td>Borrowing Interest Rate</td>
</tr>
<tr>
<td>17</td>
<td>( i_f )</td>
<td>%</td>
<td>%</td>
<td>Earning Interest Rate</td>
</tr>
<tr>
<td>18</td>
<td>( \eta_f )</td>
<td>%</td>
<td>%</td>
<td>Efficiency of feed pump</td>
</tr>
<tr>
<td>19</td>
<td>( \eta_r )</td>
<td>%</td>
<td>%</td>
<td>Efficiency of recycle pump</td>
</tr>
<tr>
<td>20</td>
<td>( \eta_{bf} )</td>
<td>%</td>
<td>%</td>
<td>Efficiency of backflush pump</td>
</tr>
</tbody>
</table>
The 'cost data' set or file contains data required by the COST ESTIMATOR. There are 20 variables in this file (Table A-2.4).

A-2.3 Editing Input Variables

MEMSYS completely overcomes the need to enter all input data each time it is executed. As described in the section dealing with the model architecture (Section A-4.1.2) on startup the program automatically loads four files (provided along with the program) containing standard default data.

<table>
<thead>
<tr>
<th>DATA TYPE</th>
<th>FILE CONTAINING DEFAULT DATA</th>
</tr>
</thead>
<tbody>
<tr>
<td>Global data</td>
<td>:MS-GLOB.DAT</td>
</tr>
<tr>
<td>Particle size</td>
<td>:MS-PART.DAT</td>
</tr>
<tr>
<td>TOC size distribution</td>
<td>:MS-TOC.DAT</td>
</tr>
<tr>
<td>Cost data</td>
<td>:MS-COST.DAT</td>
</tr>
</tbody>
</table>

A data management module (DATA MANAGER) is also provided which permits editing data amongst other functions (Section A-4.2.2). Thus, the user just needs to edit the value(s) of the variable(s) which change over different simulations to run various cases or simulations. Furthermore, the DATA MANAGER allows saving and loading of disk files, so data sets pertaining to different simulations can be stored and later loaded directly from the disk, without the need to re-enter all or any of the data. For particle and TOC distributions additional functions such as CREATE, ADD and DELETE are also provided to define new size distributions and manipulate file records.
A-3 Output Generated by the Program

Output is generated by running the performance model (PERFORMANCE ANALYZER) and the cost model (COST ESTIMATOR). The output variables pertaining to the permeate quality with respect to particles and TOC are listed in Tables A-3.1 and A-3.2 respectively. Tables A-3.3 and A-3.4 list the output variables for the permeate flux model and the cost model respectively.

Table A-3.1. Output variables for the permeate quality analysis with respect to TOC

<table>
<thead>
<tr>
<th>Field #</th>
<th>Unit</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>kD</td>
<td>Molecular Weight of the TOC</td>
</tr>
<tr>
<td>2</td>
<td>mg/L</td>
<td>Concentration in the feed or influent stream</td>
</tr>
<tr>
<td>3</td>
<td>mg/L</td>
<td>Concentration in the bulk stream</td>
</tr>
<tr>
<td>4</td>
<td>mg/L</td>
<td>Concentration in the permeate stream</td>
</tr>
<tr>
<td>5</td>
<td>%</td>
<td>Removal of solute by the membrane (based on solute volume fractions in the bulk and permeate suspensions)</td>
</tr>
</tbody>
</table>

Table A-3.2. Output variables for the permeate quality analysis with respect to particles

<table>
<thead>
<tr>
<th>Field #</th>
<th>Unit</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>μm</td>
<td>Size of the particles</td>
</tr>
<tr>
<td>2</td>
<td>#</td>
<td>Volume fraction in the feed or influent stream</td>
</tr>
<tr>
<td>3</td>
<td>#</td>
<td>Volume fraction in the bulk stream</td>
</tr>
<tr>
<td>4</td>
<td>#</td>
<td>Volume fraction in the permeate stream</td>
</tr>
<tr>
<td>5</td>
<td>%</td>
<td>Removal of solute by the membrane (based on solute volume fractions in the bulk and permeate suspensions)</td>
</tr>
</tbody>
</table>

By default the output is displayed on the screen after running the performance or cost models. Output pertaining to the latest simulation can also be displayed using the RESULTS DISPLAY (Section A-4.2.5) option from the MAIN MENU. Besides screen display the output can be stored on to disk files or redirected to the printer (for details see sections A-4.2.2.2.3, A-4.2.2.2.4, A-4.2.2.2.8 and A-4.2.7).

MEMSYS also has the capability to display dimensionless graphical profiles of the cake thickness, permeate flux and wall shear stress estimated by the performance model as
they vary along the axial length of the membrane element. This option is available from the MAIN MENU (Section A-4.2.6).

### Naming Output Files

The files containing output data are always saved by MEMSYS with an extension of '.OUT' if the user saves them using the SAVE option (Section A-4.2.2.2.3). If the user selects the SAVE AS option (Section A-4.2.2.2.4) then he is expected to enter the file-

<table>
<thead>
<tr>
<th>Field #</th>
<th>Variable</th>
<th>SI Units</th>
<th>English Units</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>$\phi_0$</td>
<td>#</td>
<td>#</td>
<td>Volume fraction of solute in the feed or influent stream</td>
</tr>
<tr>
<td>2</td>
<td>$d_p$</td>
<td>$\mu$m</td>
<td>$\mu$m</td>
<td>Mean particle diameter calculated for the particle and TOC distributions</td>
</tr>
<tr>
<td>3</td>
<td>$d_o$</td>
<td>$\mu$m</td>
<td>$\mu$m</td>
<td>Pore diameter of the membrane</td>
</tr>
<tr>
<td>4</td>
<td>$\phi_b$</td>
<td>#</td>
<td>#</td>
<td>Volume fraction of solute in the bulk stream</td>
</tr>
<tr>
<td>5</td>
<td>$\phi_p$</td>
<td>#</td>
<td>#</td>
<td>Volume fraction of solute in the permeate stream</td>
</tr>
<tr>
<td>6</td>
<td>$R_{cm}$</td>
<td>%</td>
<td>%</td>
<td>Total removal of solute by membrane</td>
</tr>
<tr>
<td>7</td>
<td>$R_m$</td>
<td>1/cm</td>
<td>1/in</td>
<td>Intrinsic resistance of the membrane</td>
</tr>
<tr>
<td>8</td>
<td>$R_{st}$</td>
<td>1/cm²</td>
<td>1/in²</td>
<td>Specific resistance of the cake formed on the membrane</td>
</tr>
<tr>
<td>9</td>
<td>$X_{cr}$</td>
<td>cm</td>
<td>in</td>
<td>The critical point at which cake formation is predicted to begin</td>
</tr>
<tr>
<td>10</td>
<td>$Q_{cr}$</td>
<td>cm²/s</td>
<td>in²/s</td>
<td>The solute flux at the critical point</td>
</tr>
<tr>
<td>11</td>
<td>$\delta$</td>
<td>mm</td>
<td>in</td>
<td>Length averaged thickness of the cake formed on the membrane</td>
</tr>
<tr>
<td>12</td>
<td>$V_{wo}$</td>
<td>L/m²/hr</td>
<td>gfd</td>
<td>Initial permeate flux</td>
</tr>
<tr>
<td>13</td>
<td>$V_w$</td>
<td>L/m²/hr</td>
<td>gfd</td>
<td>Length averaged permeate flux through the membrane</td>
</tr>
<tr>
<td>14</td>
<td>$P_r$</td>
<td>kPa</td>
<td>psi</td>
<td>Axial pressure drop across the membrane element</td>
</tr>
</tbody>
</table>
Table A-3.4. Output variables for the cost analysis

<table>
<thead>
<tr>
<th>Field #</th>
<th>Variable</th>
<th>SI Units</th>
<th>English Units</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>$Q_{des}$</td>
<td>m$^3$/day</td>
<td>mgd</td>
<td>Design capacity of the plant</td>
</tr>
<tr>
<td>2</td>
<td>$V_w$</td>
<td>L/m$^2$/hr</td>
<td>gfd</td>
<td>The permeate flux through the membrane used in the cost model</td>
</tr>
<tr>
<td>3</td>
<td>$A_{mod}$</td>
<td>m$^2$</td>
<td>ft$^2$</td>
<td>Surface area of membrane per module</td>
</tr>
<tr>
<td>4</td>
<td>$E_f$</td>
<td>kN.m/s</td>
<td>kft.lbf/s</td>
<td>Energy consumption of the feed pump</td>
</tr>
<tr>
<td>5</td>
<td>$E_r$</td>
<td>kN.m/s</td>
<td>kft.lbf/s</td>
<td>Energy consumption of the recycle pump</td>
</tr>
<tr>
<td>6</td>
<td>$E_{ff}$</td>
<td>kN.m/s</td>
<td>kft.lbf/s</td>
<td>Energy consumption of the fastflush pump</td>
</tr>
<tr>
<td>7</td>
<td>$E_{bf}$</td>
<td>kN.m/s</td>
<td>kft.lbf/s</td>
<td>Energy consumption of the backflush pump</td>
</tr>
<tr>
<td>8</td>
<td>$Q_f$</td>
<td>m$^3$/day</td>
<td>mgd</td>
<td>Total flow of the feed stream in the plant</td>
</tr>
<tr>
<td>9</td>
<td>$Q_p$</td>
<td>m$^3$/day</td>
<td>mgd</td>
<td>Total flow of the permeate stream in the plant</td>
</tr>
<tr>
<td>10</td>
<td>$Q_w$</td>
<td>m$^3$/day</td>
<td>mgd</td>
<td>Total flow of the waste stream in the plant</td>
</tr>
<tr>
<td>11</td>
<td>$Q_r$</td>
<td>m$^3$/day</td>
<td>mgd</td>
<td>Total flow of the recycle stream in the plant</td>
</tr>
<tr>
<td>12</td>
<td>$Q_{bf}$</td>
<td>m$^3$/day</td>
<td>mgd</td>
<td>Total flow of the backflush stream in the plant</td>
</tr>
<tr>
<td>13</td>
<td>$Q_{ff}$</td>
<td>m$^3$/day</td>
<td>mgd</td>
<td>Total flow of the fastflush stream in the plant</td>
</tr>
<tr>
<td>14</td>
<td>$P_r$</td>
<td>kPa</td>
<td>psi</td>
<td>Axial pressure drop across the membrane module</td>
</tr>
<tr>
<td>15</td>
<td>$C_{cap}$</td>
<td>$/m^3$</td>
<td>$/1000gal$</td>
<td>Capital cost</td>
</tr>
<tr>
<td>16</td>
<td>$C_{plant}$</td>
<td>$/m^3$</td>
<td>$/1000gal$</td>
<td>Plant cost</td>
</tr>
<tr>
<td>17</td>
<td>$C_{membrane}$</td>
<td>$/m^3$</td>
<td>$/1000gal$</td>
<td>Membrane cost</td>
</tr>
<tr>
<td>18</td>
<td>$C_{oper}$</td>
<td>$/m^3$</td>
<td>$/1000gal$</td>
<td>Operating cost</td>
</tr>
<tr>
<td>19</td>
<td>$C_{energy}$</td>
<td>$/m^3$</td>
<td>$/1000gal$</td>
<td>Energy cost</td>
</tr>
<tr>
<td>20</td>
<td>$C_{chemical}$</td>
<td>$/m^3$</td>
<td>$/1000gal$</td>
<td>Chemicals cost</td>
</tr>
<tr>
<td>21</td>
<td>$C_{mr}$</td>
<td>$/m^3$</td>
<td>$/1000gal$</td>
<td>Membrane replacement cost</td>
</tr>
<tr>
<td>22</td>
<td>$C_{disposal}$</td>
<td>$/m^3$</td>
<td>$/1000gal$</td>
<td>Concentrate disposal cost</td>
</tr>
<tr>
<td>23</td>
<td>$C_{total}$</td>
<td>$/m^3$</td>
<td>$/1000gal$</td>
<td>Total cost</td>
</tr>
<tr>
<td>24</td>
<td>$N_{mod}$</td>
<td>#</td>
<td>#</td>
<td>Number of modules required</td>
</tr>
</tbody>
</table>
name (up to a maximum of 8 characters) *without* the extension. The extension ".OUT" is automatically appended to the filename. Note that MEMSYS will *not* accept a filename with an extension and will repeatedly prompt the user to enter a filename till a valid filename is entered. If the SAVE option is selected MEMSYS saves the files under the following default names:

<table>
<thead>
<tr>
<th>OUTPUT DATA TYPE</th>
<th>DEFAULT FILE</th>
</tr>
</thead>
<tbody>
<tr>
<td>Permeate Quality w.r.t particles</td>
<td>:MS-PART.OUT</td>
</tr>
<tr>
<td>Permeate Quality w.r.t TOC</td>
<td>:MS-TOC.OUT</td>
</tr>
<tr>
<td>Permeate Flux</td>
<td>:MS-FLUX.OUT</td>
</tr>
<tr>
<td>Cost Analysis</td>
<td>:MS-COST.OUT</td>
</tr>
</tbody>
</table>
A-4 Model Architecture

A-4.1 Startup and Autoload

MEMSYS expects the default data files to be present under the same directory as the executable program 'MEMSYS.EXE'. Startup essentially constitutes asking the user for the choice of the units to be used for interactive input/output followed by automatically loading the default data files.

A-4.1.1 Specifying Working Units

The program can be run under both the SI system or English system of units. However, units are specified only once at the beginning of execution (Fig. A-4.1.1) and cannot be changed in between a simulation.

![Units Specification](Image)

Fig. A-4.1.1 Units Specification

Once the units are specified all input/output operations which occur interactively with the user takes place in the specified units. The program, however, always stores and reads all input data from disk files consistently in SI units. (So if the user anytime decides to edit an input data file using some text editor instead of from within
MEMSYS, he should do the edition employing values in SI units only). If English units are specified all data is automatically converted from SI to English units after loading. Output on screen, disk or printer always takes place in whichever units the user has selected.

A-4.1.2 Default Data Autoload

The default data files are automatically loaded by the program each time it is executed. The user can have his own default data loaded automatically by storing the four required files under the following names.

<table>
<thead>
<tr>
<th>DATA TYPE</th>
<th>FILE CONTAINING DEFAULT DATA</th>
</tr>
</thead>
<tbody>
<tr>
<td>Global data</td>
<td>:MS-GLOB.DAT</td>
</tr>
<tr>
<td>Particle size distribution</td>
<td>:MS-PART.DAT</td>
</tr>
<tr>
<td>TOC size distribution</td>
<td>:MS-TOC.DAT</td>
</tr>
<tr>
<td>Cost data</td>
<td>:MS-COST.DAT</td>
</tr>
</tbody>
</table>

While autoloading, any of the following conditions can lead to an error message being invoked by MEMSYS:

- a missing default data file
- a default data file not having the appropriate number of fields
- non-numeric field type in a default data file

The error message indicates the type of error encountered and the name of the file with which the error is associated is displayed at the bottom of the screen. Similarly if a file exists but does not contain an appropriate number of numeric fields then the program would again display an error-message. Errors associated with reading files are non-fatal due to the error-trapping nature of MEMSYS and serve to only caution the user about partial or complete autoloading failure. All variables associated with the default file which could not be read successfully are initialized to zero by the program. The recourse taken to an autoloading error depends entirely on the user. The user might decide to exit the program and investigate what is wrong (or do so through the DOS SHELL option of the program; section A-4.2.8), or he might continue and load other disk files utilizing the DATA MANAGER (Section A-4.2.2.1).

A-4.2 The MAIN MENU

The menu hierarchy is headed by the main driver module, or the MAIN MENU. After completion of the startup procedure (Section A-4.1) control is transferred to the main driver which displays the MAIN MENU and expects the user to select an option (Fig. A-4.2.1). Options of the main menu correspond to sub-modules of the program. After executing a sub-module the user can return to the main-menu by pressing the function key F10, i.e. this function key always has the specific function of returning the user to the MAIN MENU.

The main-menu screen serves some other functions too. As listed in the status bar at the
bottom of the screen (Fig. A-4.2.1) the user has the option of viewing some general help information on menu operation. The program can be exited anytime from this screen by pressing the F3 key and then confirming the intention. When appearing for the first time the main-menu screen confirms the success or failure of the default data autoload procedure (Section A-4.1.2) in the message window (Fig. A-4.2.1) which is located just above the status bar.

The options of the MAIN MENU will be described in order of their appearance in the menu in the following sections.

![Main Menu](image)

**A-4.2.1 GLOBAL OPTIONS for Some Inputs**

MEMSYS includes flexibility by either estimating parameters such as the membrane resistance, the cross-flow velocity and the permeate flux or letting the user enter known values for these parameters. An option of calculating the membrane pore size from the molecular weight cut-off (MWCO) of the membrane is also provided. MEMSYS assumes a cylindrical geometry of the membrane element by default. However, it is also capable of handling spiral wound modules by calculating the "hydraulic diameter" of a spiral wound membrane if the user specifies the spacer thickness. These options are shown in Fig. A-4.2.2.
A-4.2.1.1 PERMEATE FLUX Option

The COST ESTIMATOR requires the permeate flux through the membrane element as an input. By default this is calculated and supplied by the PERFORMANCE ANALYZER. However, in instances when the user wishes to run just the COST ESTIMATOR he can supply the value of the steady-state permeate flux and proceed with cost analysis. The procedure to do so is detailed below as well as illustrated in Fig. A-4.2.3:

---
STEP 1: Choose GLOBAL OPTIONS from the MAIN MENU
This will bring up the GLOBAL OPTIONS MENU.

STEP 2: Choose PERMEATE FLUX from the GLOBAL OPTIONS MENU
This will bring up the PERMEATE FLUX MENU.

STEP 3: Choose USER DEFINED FLUX from the PERMEATE FLUX MENU
This will bring up a data-entry box.

STEP 4: Enter the permeate-flux in the specified units.
---

The COST ESTIMATOR will use the user defined value for the permeate flux after it has been specified. To switch back to using permeate flux as calculated by the PERFORMANCE MODULE the user has to repeat Steps 1-2 listed above and then select the option PERFORMANCE ANALYZER FLUX from the PERMEATE FLUX MENU. Note that just running the PERFORMANCE ANALYZER does not tell the program to switch back to using permeate flux as calculated by the PERFORMANCE ANALYZER and it continues to use the user defined value.
A-4.2.1.2 MEMBRANE RESISTANCE Option

The PERFORMANCE ANALYZER needs the intrinsic resistance of the membrane to estimate the permeate flux through the membrane. By default MEMSYS expects this to be directly specified in the "global data" file (Table A-2.1). However, MEMSYS also provides two other options for calculating the membrane resistance.

Membrane resistance can be calculated using the feed pressure and viscosity if the user provides the "clean water" permeate flux:

\[ R_m = \frac{\Delta P}{\mu V_{wp}} \]

where \( \Delta P \) is the transmembrane pressure (MEMSYS considers this equal to the feed pressure \( P_f \)), \( V_{wp} \) is the clean water flux and \( \mu \) is the viscosity of the suspension. The sequence of steps involved is detailed below as well as illustrated in Fig. A-4.2.4:
STEP 1: Choose GLOBAL OPTIONS from the MAIN MENU
This will bring up the GLOBAL OPTIONS MENU
STEP 2: Choose MEMBRANE RESISTANCE from the GLOBAL OPTIONS MENU
This will bring up the RESISTANCE MENU
STEP 3: Choose EVAL. FROM CLEAN WATER FLUX from the RESISTANCE MENU
This will bring up a data-entry box
STEP 4: Enter the membrane resistance in the specified units

![Diagram of the Membrane Systems interface](image)

Fig. A-4.2.4 Specifying CLEAN WATER FLUX to estimate the membrane resistance

MEMSYS can also calculate the membrane resistance from the specified membrane pore diameter using the following correlation:

\[ R_m = \frac{3.514 \times 10^7}{d_o^{1.7}} \]

where \( R_m \) is the membrane resistance in cm\(^{-1}\) (or in\(^{-1}\)) and \( d_o \) is the membrane pore diameter in microns. This correlation was derived by collecting a few data from a few membrane manufacture catalogs. It should be noted that membrane resistance is a function of a number of parameters and varies widely with membrane types. As such this correlation is of a very limited value in this program and should only be used under circumstances where no information on membrane resistance can be obtained. It is thus recommended that the user supply the value of the membrane resistance whenever possible. The sequence of steps involved in choosing this option is same as Steps 1-2
above followed by choosing the EVAL. FROM CORRELATION option from the RESISTANCE MENU.

A-4.2.1.3 PORE SIZE Option

The PERFORMANCE MODULE requires the average size of the membrane pores as an input to estimate the permeate flux through the membrane. By default MEMSYS expects this to be directly specified in microns in the "global data" file (Table A-2.1). However, MEMSYS provides the option of estimating the pore size from user specified molecular weight cut off (MWCO) of the membrane by using the following correlation:

\[ d_o = 0.147 \times 10^{-4} (MWCO)^{1/3} \]

where \( d_o \) is the pore diameter in microns and \( MWCO \) is the molecular weight cut off of the membrane in Daltons. This correlation has been reported in literature for estimating the size of spherical solute molecules from their MWCO. The procedure to use this option is detailed below as well as illustrated in Fig. A-4.2.5:

**STEP 1:** Choose GLOBAL OPTIONS from the MAIN MENU
This will bring up the GLOBAL OPTIONS MENU.

**STEP 2:** Choose PORE SIZE from the GLOBAL OPTIONS MENU
This will bring up the PORE SIZE MENU

**STEP 3:** Choose the MWCO option
This will bring up a data-entry box

**STEP 4:** Enter the MWCO in the specified units

---

*Fig. A-4.2.5 Specifying MWCO to estimate the pore size of the membrane*
A-4.2.1.4 MODULE TYPE Option

By default MEMSYS assumes a cylindrical hollow fiber geometry of the membrane element. However, it can also handle spiral wound modules by estimating an equivalent hydraulic diameter for the "element" of a spiral wound membrane. The user needs to specify the spacer thickness for the spiral wound membrane for MEMSYS to do this estimation. The procedure to use this option is detailed below as well as illustrated in Fig. A-4.2.6:

**STEP 1:** Choose GLOBAL OPTIONS from the MAIN MENU  
This will bring up the GLOBAL OPTIONS MENU.

**STEP 2:** Choose MODULE TYPE from the GLOBAL OPTIONS MENU  
This will bring up the MODULE TYPE MENU.

**STEP 3:** Choose the SPIRAL WOUND option  
This will bring up a data-entry box.

**STEP 4:** Enter the spacer thickness in the specified units

---

Fig. A-4.2.6 Specifying spacer thickness to estimate the "element diameter" of a spiral wound membrane

---

A-4.2.1.5 CROSS-FLOW VELOCITY Option

The cross-flow velocity through the membrane element is required as an input by both the PERFORMANCE ANALYZER and the COST ESTIMATOR. By default MEMSYS expects this to be directly specified in the "global data" file (Table A-2.1). However, MEMSYS can also estimate the cross-flow velocity from user specified
"clean water" axial pressure drop across the membrane module. The procedure to use this option is detailed below as well as illustrated in Fig. A-4.2.7:

STEP 1: Choose GLOBAL OPTIONS from the MAIN MENU
This will bring up the GLOBAL OPTIONS MENU.
STEP 2: Choose CROSS-FLOW VELOCITY from the GLOBAL OPTIONS MENU
This will bring up the CROSS-FLOW VELOCITY MENU
STEP 3: Choose the EVAL. FROM PRESSURE DROP option
This will bring up a data-entry box
STEP 4: Enter the clean water pressure drop in the specified units

![Diagram](Fig. A-4.2.7 Specifying clean water pressure drop to estimate the cross-flow velocity)

A-4.2.2 Data Management using the DATA MANAGER

MEMSYS utilizes an integrated data handling module to manage all input data. However, it should be noted that optional inputs such as user defined values of the permeate flux are not handled by this module (see Section A-4.2.1 on how to enter these values). The structure of the data manager screen is shown in Fig. A-4.2.8 and the elements are described below:
DISPLAY WINDOW: This is the region where input files are displayed one at a time. The file to be displayed can be selected using the FILE MENU (see Section A-4.2.2.1).

FILENAME: The name of the data file currently displayed in the DISPLAY WINDOW is displayed at the upper left hand corner of the screen.

STATUS BAR: The status of some special function keys is stated here. For the DATA MANAGER the following keys are active when no menu is activated:
- PgUp - used to scroll a file up by a page
- PgDn - used to scroll a file down by a page
- F5 - used to bring up the FILE MENU
- F6 - used to bring up the OPTIONS MENU
- F10 - used to return to the MAIN MENU

MENU DISPLAY: This is the region where the FILE MENU and the OPTIONS MENU are displayed.

DATA ENTRY WINDOW: This is the region where the user is prompted to enter data.

MESSAGE WINDOW: This region is utilized to display helpful messages for the user while prompting for data entry and for reporting the successful completion of some actions such as loading or saving a file.

---

Fig. A-4.2.8 Structure of the DATA MANAGER screen
Details on the FILE MENU and the OPTIONS MENU are discussed in the following sections.

A-4.2.2.1 Displaying Input Files Using the FILE MENU

The FILE MENU is activated by pressing the F5 key when the STATUS BAR indicates that this key is active. As explained in the chapter on input requirements (Chapter A-2), the input data is categorized into four files. The FILE MENU allows the user to select the required file to be opened. The DATA MANAGER keeps track of the file last displayed and when selected from the MAIN MENU always opens this file. Figure A-4.2.9 illustrates the procedure to display different files. In the figure the global data file is the currently active file. To make another file active follow these sequence of steps:

STEP 1: Activate the FILE MENU by pressing the function key F5
STEP 2: Select the file to be displayed from the FILE MENU. The FILE MENU will deactivate as soon as a selection is made and the required file will be displayed

Fig. A-4.2.9 The FILE MENU of DATA MANAGER

Figures A-4.2.10 through A-4.2.13 illustrate the format of the other three input data files as displayed by the DATA MANAGER. Note that these three files are allotted two "pages" each for display and can be scrolled up/down using the PgUp/PgDn keys.
### Particle Size Distribution

<table>
<thead>
<tr>
<th>Size Class (μm)</th>
<th>Number/mL</th>
</tr>
</thead>
<tbody>
<tr>
<td>[ 1]</td>
<td>.380</td>
</tr>
<tr>
<td>[ 2]</td>
<td>.590</td>
</tr>
<tr>
<td>[ 3]</td>
<td>1.170</td>
</tr>
<tr>
<td>[ 4]</td>
<td>1.760</td>
</tr>
<tr>
<td>[ 5]</td>
<td>2.350</td>
</tr>
<tr>
<td>[ 6]</td>
<td>2.940</td>
</tr>
<tr>
<td>[ 7]</td>
<td>3.520</td>
</tr>
<tr>
<td>[ 8]</td>
<td>4.000</td>
</tr>
</tbody>
</table>

Fig. A-4.2.10 Format of a sample particle size distribution file

### TOC Molecular Weight Distribution

<table>
<thead>
<tr>
<th>Molecular Wt. (kD)</th>
<th>Conc. (mg/L)</th>
</tr>
</thead>
<tbody>
<tr>
<td>[ 1]</td>
<td>2.750</td>
</tr>
<tr>
<td>[ 2]</td>
<td>.300</td>
</tr>
<tr>
<td>[ 3]</td>
<td>3.300</td>
</tr>
<tr>
<td>[ 4]</td>
<td>1.250</td>
</tr>
<tr>
<td>[ 5]</td>
<td>.700</td>
</tr>
<tr>
<td>[ 6]</td>
<td>1.700</td>
</tr>
</tbody>
</table>

Fig. A-4.2.11 Format of a sample TOC molecular weight distribution file
### Fig. A-4.2.12 Format of the cost data file (page1)

**MS-COST.DAT**

<table>
<thead>
<tr>
<th>Cost Parameters</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Design Capacity (cu.m/hr)</td>
<td>200.9200</td>
</tr>
<tr>
<td>Module Packing Density (%)</td>
<td>13.7000</td>
</tr>
<tr>
<td>Module Diameter (cm)</td>
<td>30.0000</td>
</tr>
<tr>
<td>Module Cost ($/module)</td>
<td>8000.0000</td>
</tr>
<tr>
<td>Operating Period (min)</td>
<td>2.0000</td>
</tr>
<tr>
<td>Fastflush Period (sec)</td>
<td>15.0000</td>
</tr>
<tr>
<td>Fastflush Velocity (cm/s)</td>
<td>111.0000</td>
</tr>
<tr>
<td>Backflush Period (sec)</td>
<td>1.0000</td>
</tr>
<tr>
<td>Backflush Flux (L/m²/hr)</td>
<td>387.0000</td>
</tr>
<tr>
<td>Backflush Pressure (bars)</td>
<td>2.0000</td>
</tr>
<tr>
<td>Coagulant Dose (mg/L)</td>
<td>.0000</td>
</tr>
<tr>
<td>Coagulant Cost ($/1000kg)</td>
<td>130.0000</td>
</tr>
<tr>
<td>Energy Cost ($/kw-hr)</td>
<td>0.0700</td>
</tr>
<tr>
<td>Plant Life (yrs)</td>
<td>20.0000</td>
</tr>
</tbody>
</table>

### Fig. A-4.2.13 Format of the cost data file (page2)

**MS-COST.DAT**

<table>
<thead>
<tr>
<th>Cost Parameters</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Membrane Life (yrs)</td>
<td>5.0000</td>
</tr>
<tr>
<td>Borrowing Interest Rate (%)</td>
<td>10.0000</td>
</tr>
<tr>
<td>Earning Interest Rate (%)</td>
<td>8.0000</td>
</tr>
<tr>
<td>Feed Pump Efficiency (%)</td>
<td>70.0000</td>
</tr>
<tr>
<td>Recycle Pump Efficiency (%)</td>
<td>70.0000</td>
</tr>
<tr>
<td>Backflush Pump Efficiency (%)</td>
<td>70.0000</td>
</tr>
</tbody>
</table>
A-4.2.2.2 File Management Using the OPTIONS MENU

Various operations can be performed on the file currently displayed by the DATA MANAGER by using the OPTIONS MENU. This menu is operated by pressing the F6 function key when it is shown to be active in the status bar. The options listed in the OPTIONS MENU depend on the file displayed. The options available for the "global data" and "cost data" files include LOAD, EDIT, SAVE, SAVE AS, PRINT and DEFAULT (Fig. A-4.2.14). The Particle Size Distribution and TOC distribution files have the additional options of CREATE, ADD AND DELETE for creating new files and adding/deleting records (Fig. A-4.2.15). The various options available for file management are discussed in the following sections in a stepwise manner.

Fig. A-4.2.14 Format of the OPTIONS MENU for global data and cost data files

A-4.2.2.2.1 LOAD Option

The user can load previously saved files from the disk by choosing the LOAD option from the OPTIONS MENU. To load a particular file, the FILE MENU (Section A-4.2.2.1) should first be utilized to select the file type. The procedure to load a file is explained in the following steps:
STEP 1: Select the file type to be loaded by using the FILE MENU
STEP 2: Activate the OPTIONS MENU (by pressing F6) and select LOAD
STEP 3: Enter the name of the file to be loaded at the prompt. Note that the extension of the data file should not be specified. Only the filename (maximum 8 characters) is expected by MEMSYS since it only opens data files which have a '.DAT' extension.

<table>
<thead>
<tr>
<th>Particle Size Distribution</th>
<th>Number/mL</th>
</tr>
</thead>
<tbody>
<tr>
<td>Size Class (μm)</td>
<td></td>
</tr>
<tr>
<td>[ 1] 0.380</td>
<td>46000000.</td>
</tr>
<tr>
<td>[ 2] 0.590</td>
<td>34000000.</td>
</tr>
<tr>
<td>[ 8] 4.000</td>
<td>10000.</td>
</tr>
</tbody>
</table>

![Fig. A-4.2.15 Format of the OPTIONS MENU for particle and TOC distribution files](image)

Any of the following conditions can lead to an error message being invoked in trying to load a file:

- the specified file cannot be found by the program
- the specified file does not have the appropriate number of fields
- the specified file has non-numeric data or is corrupted

All the above errors are non-fatal and are trapped by MEMSYS to avoid program termination. Hitting the Esc key clears the error message and returns the user to the previously active file.

### A-4.2.2.2.2 EDIT Option

The value of different variables in a data file can be modified using this option. To edit a variable(s) or record(s) in a particular file, the FILE MENU (Section A-4.2.2.1)
should first be utilized to select the file type. The procedure to edit a variable or record is explained in the following steps:

**STEP 1:** Select the file type to be edited by using the FILE MENU
**STEP 2:** Activate the OPTIONS MENU (by pressing F6) and select EDIT
**STEP 3:** Enter the item number to be edited at the prompt
**STEP 4:** Enter the new value of the variable at the prompt. It should be noted that for the particle size and TOC files a whole new record, comprising the size class and the concentration, has to be entered. MEMSYS displays the new value in the DISPLAY WINDOW.
**STEP 5:** Repeat STEPS 1-4 till all required fields /records have been edited. The EDIT mode can then be deactivated by entering a zero at the prompt for the item number.

A non-numeric edition by the user can lead to an error message being invoked. The error is non-fatal and can be recovered by pressing the Esc key. However, the EDIT mode is automatically deactivated when such a condition occurs.

The user should save the file, if required, after editing it by using the file saving options (Sections A-4.2.2.2.3 and A-4.2.2.2.4).

### A-4.2.2.2.3 SAVE Option

The user can save a data file under the current filename by choosing the SAVE option from the OPTIONS MENU. To save a particular file, the FILE MENU (Section A-4.2.2.1) should first be utilized to select the file type. The procedure to save a file is explained in the following steps:

**STEP 1:** Select the file type to be saved by using the FILE MENU
**STEP 2:** Activate the OPTIONS MENU (by pressing F6) and select SAVE
   The file will be saved under its current filename

### A-4.2.2.2.4 SAVE AS Option

The user can save a data file under a different filename by choosing the SAVE AS option from the OPTIONS MENU. To save a particular file, the FILE MENU (Section A-4.2.2.1) should first be utilized to select the file type. The procedure to save a file under a different name is explained in the following steps:

**STEP 1:** Select the file type to be saved by using the FILE MENU
**STEP 2:** Activate the OPTIONS MENU (by pressing F6) and select SAVE AS
**STEP 3:** Enter the name of the file to be saved at the prompt. The file will be saved under the new filename. Note that a maximum of 8 characters is allowed for the filename. The extension should *not* be specified as a '.DAT' extension is automatically appended to the filename by MEMSYS
A-4.2.2.2.5 CREATE Option

This option is used to create new files for describing the particle size and TOC molecular weight distributions. It should be noted that the "global data" and "cost data" files contain a fixed number of variables and the values assigned to these variables can be changed over different simulations by using the EDIT option (Section A-4.2.2.2.2). As such the CREATE option is limited to only the distribution data files. The distribution files can contain a variable number of records (up to a maximum of 50) and the user can utilize the CREATE option to create new files describing the distributions.

To create a distribution file, the FILE MENU (Section A-4.2.2.1) should first be utilized to select the distribution file type. The procedure to create a distribution file is explained in the following steps:

STEP 1: Select the file type to be created by using the FILE MENU
STEP 2: Activate the OPTIONS MENU (by pressing F6) and select CREATE
STEP 3: Enter the number of items in the distribution file to be created at the prompt. A distribution file can contain a maximum of up to 50 items or records. MEMSYS will not except a number which is not between 0 to 50
STEP 4: Enter the record (constituted of the size field and the concentration filed) corresponding to the item number displayed. MEMSYS adds the new record in the distribution file displayed in the DISPLAY WINDOW.
STEP 5: Repeat STEP 4 till all required items or records have been entered. The CREATE mode will be automatically deactivated after all items have been entered.

A non-numeric edition by the user (in any field) can lead to an error message being invoked. The error is non-fatal and can be recovered by pressing the Esc key. However, the CREATE mode is automatically deactivated when such a condition occurs and the number of records in the file is initialized to the number of records defined up till and excluding the point where the error was encountered. Note that this can be used as a strategy to escape from the CREATE mode in the middle of a creation if the user decides to truncate the number of records defined in the beginning.

The user should save the file, if required, after creating it by using the file saving options (Sections A-4.2.2.2.3 and A-4.2.2.2.4).

A-4.2.2.2.6 ADD option

This option is used to add or insert new records to the particle size and TOC molecular weight distribution data files. It should be noted that the "global data" and "cost data" files contain a fixed number of variables. As such the ADD option is limited only to the distribution data files. The distribution files can contain a variable number of records (up to a maximum of 50) and the user can utilize the ADD option to insert new records in the distribution files.

To add record(s) to a distribution file, the FILE MENU (Section A-4.2.2.1) should first be utilized to select the distribution file type. The procedure to add records to a distribution file is explained in the following steps:
STEP 1: Select the file type to be edited by using the FILE MENU
STEP 2: Activate the OPTIONS MENU (by pressing F6) and select ADD
STEP 3: Enter the number of the item or record after which the new record is to be inserted at the prompt. MEMSYS will not accept a number which is not between zero and the number of records in the file
STEP 4: Enter the record (constituted of the size field and the concentration field). MEMSYS inserts the new record in the distribution file displayed in the DISPLAY WINDOW.
STEP 5: Repeat STEP 4 till the required number of items or records have been inserted. The ADD mode can then be deactivated by entering a zero at the prompt for the item number (NOTE: Due to this escape code for the ADD option a record cannot be inserted before the first record in the distribution file).

A non-numeric edition by the user can lead to an error message being invoked. The error is non-fatal and can be recovered by pressing the Esc key. However, the ADD mode is automatically deactivated when such a condition occurs.

The user should save the file, if required, after the addition operation by using the file saving options (Sections A-4.2.2.2.3 and A-4.2.2.2.4).

A-4.2.2.2.7 DELETE Option

This option is used to delete or remove records from the particle size and TOC molecular weight distribution data files. It should be noted that the "global data" and "cost data" files contain a fixed number of variables. As such the DELETE option is limited only to the distribution data files. The distribution files can contain a variable number of records (up to a maximum of 50) and the user can utilize the DELETE option to remove records from the distribution files.

To delete a record(s) from a distribution file, the FILE MENU (Section A-4.2.2.1) should first be utilized to select the distribution file type. The procedure to delete records from a distribution file is explained in the following steps:

STEP 1: Select the file type to be edited by using the FILE MENU
STEP 2: Activate the OPTIONS MENU (by pressing F6) and select DELETE
STEP 3: Enter the number of the item or record to be deleted at the prompt. MEMSYS will not accept a number which is not between zero and the number of records in the file. MEMSYS deletes the record from the distribution file displayed in the DISPLAY WINDOW.
STEP 4: Repeat STEP 4 till the required number of items or records have been deleted. The DELETE mode can then be deactivated by entering a zero at the prompt for the item number.

A non-numeric edition by the user can lead to an error message being invoked. The error is non-fatal and can be recovered by pressing the Esc key. However, the DELETE mode is automatically deactivated when such a condition occurs. No change is made in the distribution file.

The user should save the file, if required, after the deletion operation by using the file
saving options (Sections A-4.2.2.2.3 and A-4.2.2.2.4).

A-4.2.2.2.8 PRINT Option

This option is used to print the input data files in a formatted form. The user should ensure printing compatibility by following the instructions given in Section A-1.5.4.

To print a data file, the FILE MENU (Section A-4.2.2.1) should first be utilized to select the file type. The procedure to print a file is explained in the following steps:

<table>
<thead>
<tr>
<th>STEP 1:</th>
<th>Select the file type to be printed by using the FILE MENU</th>
</tr>
</thead>
<tbody>
<tr>
<td>STEP 2:</td>
<td>Activate the OPTIONS MENU (by pressing F6) and select PRINT</td>
</tr>
<tr>
<td></td>
<td>MEMSYS exports the file to the PRN device</td>
</tr>
</tbody>
</table>

A-4.2.2.2.9 DEFAULT Option

This option is used to load default data files. The default data files are loaded automatically at the start of the program (Section A-4.1.2) However, if the user has loaded some other data file using the LOAD option and now needs to revert back to a default data file then he can do so by using this option. To load a particular default file, the FILE MENU (Section A-4.2.2.1) should first be utilized to select the file type. The procedure to load a default file is explained in the following steps:

<table>
<thead>
<tr>
<th>STEP 1:</th>
<th>Select the file type to be loaded by using the FILE MENU</th>
</tr>
</thead>
<tbody>
<tr>
<td>STEP 2:</td>
<td>Activate the OPTIONS MENU (by pressing F6) and select DEFAULT</td>
</tr>
</tbody>
</table>

Note that the conditions which can lead to an error message being invoked in the case of loading a default file are similar to the ones detailed in the case of LOADing a file (Section A-4.2.2.2.1) and identical to the ones detailed in case of autoload (Section A-4.1.2).

A-4.2.3 Estimating the permeate quality and flux using the PERFORMANCE ANALYZER

Choosing PERFORMANCE ANALYZER from the MAIN MENU brings up the PERFORMANCE MENU (Fig. A-4.2.16). The user can select the option for estimating the permeate flux, permeate quality with respect to particles or permeate quality with respect to TOC.

A-4.2.3.1 PERMEATE FLUX Option

Actually the PERMEATE FLUX option also estimates the permeate quality before proceeding with estimating the permeate flux but no detailed results are provided except for the total feed, bulk and permeate volume fractions and total removal. The progress of the subroutine which solves for the root of a system of nonlinear equations to estimate the permeate flux at discretized points along the membrane length is displayed
on the screen (Fig. A-4.2.17). The maximum number of iterations allowed for convergence to a root in the subroutine is 100. Failure to converge within 100 iterations results in an error message being displayed and abortion of the permeate flux estimation subroutine. This can probably happen if the problem is not well posed. Under such conditions the user should check all input variables to make sure that all are properly defined and do not have unrealistic values.

The results of the permeate flux analysis include parameters detailed in Table A-3.3. Fig. A-4.2.18 shows the format of the results. There are two extreme cases encountered in permeate flux estimation, viz., when no cake growth is predicted and when the membrane element is predicted to be plugged with cake. In the former case the average permeate flux equals the initial flux. When plugging occurs the permeate flux from the point of plugging to the exit of the membrane element is assumed to be zero. Messages pertaining to both these conditions are displayed when they occur. It should be noted that the model does not actually hold in a plugging scenario (Section A-1.2). The results of the permeate flux analysis can be saved in a disk file or exported to the printer. The status bar at the bottom of the screen displays all the options available to the user:

- **F2** - used to save the permeate quality results onto the default file (MS-FLUX.OUT)
- **Shift + F2** - used to save the permeate results onto a user defined filename
- **F4** - used to export the results to the printer
- **F10** - used to return to the main menu
Fig. A-4.2.17 Progress of the evaluation of the permeate flux along the length of the membrane element

Fig. A-4.2.18 Format of the PERMEATE FLUX results screen
A-4.2.3.2 PERMEATE QUALITY --> PARTICLE and PERMEATE QUALITY --> TOC Options

Detailed permeate quality analysis with respect to particle removal and TOC removal can be obtained by choosing the PERMEATE QUALITY --> PARTICLE and PERMEATE QUALITY --> TOC options respectively from the PERFORMANCE MENU. Results include the feed, bulk and permeate concentrations along with percent removals for each size class and also in total. The format of the permeate quality results with respect to particles and TOC is depicted in Figs. A-4.2.19 and A-4.2.20 respectively.

![Table of Permeate Quality (Particles)]

<table>
<thead>
<tr>
<th>SIZE(µm)</th>
<th>FEED</th>
<th>BULK</th>
<th>PERMEATE</th>
<th>REMOVAL(%)</th>
</tr>
</thead>
</table>

**Fig. A-4.2.19 Format of the PERMEATE QUALITY --> PARTICLES results screen**

The results can be saved in a disk file or exported to the printer. The status bar at the bottom of the screen displays all the options available to the user:

- **PgUp** - used to scroll up a page
- **PgDn** - used to scroll down a page
- **F2** - used to save the permeate quality results onto the default file (MS-PART.OUT or MS-TOC.OUT)
- **Shift+F2** - used to save the permeate results onto a user defined filename
- **F4** - used to export the results to the printer
- **F10** - used to return to the main menu
A-4.2.4 Estimating the Capital and Operating Costs Using the COST ESTIMATOR

Choosing COST ESTIMATOR from the MAIN MENU executes the algorithm for estimating the capital and operating costs. It should be noted that the cost model requires the permeate flux through the membrane as an input parameter. Therefore the user should have either estimated the permeate flux using the PERFORMANCE ANALYZER (Section A-4.2.3.1) or specified a user defined value of the permeate flux (Section A-4.2.1.1) to proceed with the cost analysis. Failure to do so results in an appropriate error message being displayed by MEMSYS.

The results of the cost analysis include parameters detailed in Table A-3.4. Fig. A-4.2.21 shows the format of the results. The status bar at the bottom of the screen displays all the options available to the user:

- **F2**: used to save the permeate quality results onto the default file (MS-COST.OUT)
- **Shift+F2**: used to save the permeate results onto a user defined filename
- **F4**: used to export the results to the printer
- **F10**: used to return to the main menu
### A-4.2.5 Displaying Latest Simulation Results Using the RESULTS DISPLAY

The RESULTS DISPLAY option of the MAIN MENU can be utilized to review the permeate flux and cost results from the latest simulation. Also, if the user had forgotten to save and/or print the simulation results from within the PERFORMANCE ANALYZER or the COST ESTIMATOR, then the RESULTS DISPLAY can be utilized to do so anytime without having to actually perform the simulation again. These options of the RESULTS DISPLAY MENU are illustrated in Fig. A-4.2.22. The standard options for saving and printing available from within the PERFORMANCE ANALYZER and the COST ESTIMATOR are available here too:

- **F2**: used to save the permeate quality results onto the default file (MS-FLUX.OUT or MS-COST.OUT)
- **Shift+F2**: used to save the permeate results onto a user defined filename
- **F4**: used to export the results to the printer
- **F10**: used to return to the main menu

### A-4.2.6 Viewing Dimensionless Plots Using the GRAPH VIEW

Choosing GRAPH VIEW from the MAIN MENU allows the user to view dimensionless graphical profiles of the permeate flux, cake thickness and shear-stress at the wall as they vary along the length of the membrane. GRAPH VIEW can only be
executed after the PERFORMANCE ANALYZER has been run, otherwise the program displays an error message. The maximum, minimum and length-averaged values of each parameter are also displayed along with the graphs.

<table>
<thead>
<tr>
<th>PARAMETER</th>
<th>NON-DIMENSIONING PARAMETER</th>
</tr>
</thead>
<tbody>
<tr>
<td>Permeate flux</td>
<td>Initial permeate flux (&quot;Clean&quot; water flux)</td>
</tr>
<tr>
<td>Cake thickness</td>
<td>Maximum cake thickness</td>
</tr>
<tr>
<td>Shear stress at wall</td>
<td>Maximum shear stress at wall</td>
</tr>
</tbody>
</table>

A-4.2.7 Printing from the MAIN MENU using the PRINT MENU

All input/output files can individually be printed from within the DATA MANAGER, PERFORMANCE ANALYZER and COST ESTIMATOR. As an added facility the user can print all inputs and outputs or just all inputs and just all outputs at the touch of a single keystroke by utilizing the PRINT option of the MAIN MENU. These options of the PRINT MENU are illustrated in Fig. A-4.2.23. An input or output file corresponding to the particles size and TOC molecular weight distribution is printed only if it contains some records, i.e. only if a distribution has been defined. Similarly an output file from the permeate flux model or cost model is printed only if the corresponding model has been run.
A-4.2.8 Executing DOS Commands Using the DOS SHELL Option

The DOS SHELL option of the MAIN MENU allows the user to execute DOS command(s). This option allows all MEMSYS parameters to be retained in the computer memory (RAM) while allowing the user to carry out DOS commands. On choosing the DOS SHELL option the user is prompted with the following:

- Enter a blank line to return to Memsys, or
- Enter a DOS command, or
- Enter ‘Command’ to escape to DOS shell
  Typing 'Exit' from shell returns to Memsys

The user is thus essentially given three options:

(1) Return to MEMSYS without doing anything by pressing Enter (or a blank line).
(2) Execute a DOS command by typing the command and pressing Enter. The command is executed and then control is returned to MEMSYS immediately.
(3) Execute a series of DOS commands by typing the word 'COMMAND' (in lowercase or uppercase) and pressing Enter. This takes the user to the DOS prompt. To return control to MEMSYS the user must enter 'EXIT' (uppercase or lowercase) at the DOS prompt after all required DOS commands have been performed.
A-4.3 Error Handling in MEMSYS

MEMSYS employs extensive error-handling for various input-output operations to avoid the frustration of unwanted program termination. An "error" refers to a run time error and/or a "mistake" on the part of the user. Common examples are attempting to open a non-existent file, incorrectly typing and entering non-numeric data when numeric data is expected, trying to delete a non-existent record in a data file, etc. It has been attempted to trap various kinds of common errors, but it is not possible to either anticipate or attempt to program for all kinds of errors which might occur while running a program, so under certain cases the user might be faced with program termination due to a run-time error.

Error-handling in MEMSYS is done under two categories:

<table>
<thead>
<tr>
<th>ERROR TYPE</th>
<th>COMMON EXAMPLES</th>
<th>ACTION</th>
</tr>
</thead>
<tbody>
<tr>
<td>Run time errors</td>
<td>Trying to open a non-existent file, entering non-numeric data in a numeric field, etc.</td>
<td>An error message is displayed informing the user of the error that occurred and the recourse to the error. The user can clear the error message by pressing the Esc key.</td>
</tr>
<tr>
<td>Logical errors</td>
<td>Trying to delete a non-existent record, trying to run the COST ESTIMATOR without estimating or defining the permeate flux, etc.</td>
<td>An error message is sometimes displayed as in the case of run time errors. In other cases the user is prompted repeatedly for input till an acceptable value is entered.</td>
</tr>
</tbody>
</table>

Error Messages

Following is the list of the error messages displayed by MEMSYS along with the particular situation under which they occur:

1. File LOAD failed: Cannot open file
   File does not exist

   This message is displayed when the user tries to load an input data file using MEMSYS and MEMSYS cannot find the specified file in the current directory.

2. File READ failed: Inconsistent data
   File incompatible or corrupted

   This message is displayed when the user tries to load an input data file using MEMSYS
and MEMSYS finds either a discrepancy in the data type (non-numeric data) or the number of fields in the file is less than expected.

3.

Cannot run COST ESTIMATOR: Flux unevaluated
Run PERFORMANCE ANALYZER or specify USER DEFINED FLUX first

This message is displayed when the user tries to run the cost model (the COST ESTIMATOR) without either running the PERMEATE FLUX option of the performance model (the PERFORMANCE ANALYZER) or defining a user defined value for the permeate flux.

4.

READ failed: Invalid Input
A valid NUMERIC input expected

This message is displayed when MEMSYS receives a non-numeric input from the user where numeric input is expected or a real number where an integer is expected.

5.

Could not converge to a root in 100 iterations
Problem might be ill posed. Check all input data

This message is displayed when the permeate flux evaluation model (PERMEATE FLUX option of the PERFORMANCE ANALYZER MENU) fails to converge to a root of the system of governing equations at a grid point on the membrane element. As suggested in the message this can happen if the input data is not well defined for the problem to be well posed (for example: negative values for some input parameter(s), an non initialized value or a value of zero for some input parameter(s)).

6.

ADD Failed: Cannot add to file
File has maximum records (50)

This message is displayed when the user tries to add a 51th record to a input data distribution file (particle size or TOC molecular weight distribution file). The maximum size of a distribution file is limited to 50 records.

7.

Cannot run GRAPH VIEW: Flux unevaluated
Run PERFORMANCE ANALYZER first

This message is displayed when the user tries to run the GRAPH VIEW without running the PERMEATE FLUX option of the performance model (the PERFORMANCE ANALYZER).
INFEASIBLE CONFIGURATION: Net permeate flux negative
Reduce BACKFLUSH DURATION and/or BACKFLUSH FLUX

This message is displayed when the cost model (COST ESTIMATOR) estimates that the current system configuration utilizes more water for hydrodynamic cleaning (backflushing) than is being produced. This results in a negative value for the net permeate flux. To correct the situation the user should reduce the quantity of water utilized in backflushing (by reducing either or both of the backflush duration and the backflush flux) and/or adopt a configuration which results in an increase in the permeate flux.
Appendix B

Notation

\[ a_{ke} \]  
kinetic energy coefficient

\[ a_p \]  
particle radius

\[ A_{mem} \]  
membrane area required to produce design flow

\[ A_{mod} \]  
membrane area per module

\[ A_x \]  
cross-sectional area of membrane element

\[ A / F \]  
uniform series sinking fund factor

\[ A / P \]  
uniform series capital recovery factor

\[ c \]  
concentration

\[ c_b \]  
bulk suspension concentration

\[ c_{kw} \]  
cost of one kilowatt-hour of electricity

\[ c_{mod} \]  
cost of one membrane module

\[ c_o \]  
feed suspension concentration

\[ c_w \]  
waei concentration

\[ C_{chemical} \]  
cost of chemicals per volume of water produced

\[ C_{disposal} \]  
cost of concentrate disposal per volume of water produced

\[ C_{energy} \]  
cost of energy consumed per volume of water produced

\[ C_{membrane} \]  
amortized capital cost of membranes

\[ C_{mr} \]  
cost of membrane replacement per volume of water produced

\[ C_{plant} \]  
amortized capital cost of pipes and pumps

\[ CH_c \]  
cost of bulk coagulant

\[ CH_d \]  
coagulant dose

\[ d_h \]  
hydraulic diameter

\[ d_o \]  
membrane pore diameter

\[ d_p \]  
particle diameter

\[ D \]  
diffusion coefficient

\[ D_{brn} \]  
Brownian diffusion coefficient

\[ D_c \]  
diameter of membrane element

\[ D_f \]  
fractal dimension

\[ D_m \]  
diameter of membrane module
\(D_{sh}\)  shear-induced diffusion coefficient
\(\hat{D}_{sh}\)  dimensionless shear-induced diffusion coefficient
\(DL\)  design life of plant
\(E_{bf}\)  rate of energy consumption of backflush
\(E_f\)  rate of energy consumption of feed pump
\(E_{ff}\)  rate of energy consumption of fastflush
\(E_r\)  rate of energy consumption of recycle pump
\(f_f\)  Fanning friction factor
\(F_{drag}\)  drag force
\(g\)  acceleration due to gravity
\(G\)  lag factor
\(i_c\)  cost of capital, as interest rate
\(i_f\)  cost of sinking fund, as interest rate
\(k\)  Boltzmann’s constant
\(k'\)  fitting coefficient
\(K\)  mass transfer coefficient
\(L_e\)  length of membrane element
\(ML\)  expected average membrane life
\(MW\)  molecular weight
\(MWCO\)  molecular weight cut-off of membrane
\(N\)  grid size
\(N_{f}\)  number of elements (hollow fibers or tubes) per module
\(N_{mod}\)  number of membrane modules required to produce design flow
\(P_{bf}\)  backflush pressure
\(P_d\)  volumetric packing density
\(P_f\)  feed pressure
\(P_r\)  axial pressure drop across module
\(Q\)  excess particle flux
\(Q_{bf}\)  backflush flow rate
\(Q_{cr}\)  critical excess particle flux
\(Q_{des}\)  plant design capacity
\(Q_f\)  feed flow rate
\(Q_{ff}\)  fastflush flow rate
$Q_p$  permeate flow rate
$Q_r$  recycle flow rate
$Q_t$  total flow rate
$Q_w$  waste flow rate
$r_o$  membrane pore radius
$R$  effective element radius
$R_f$  fouling resistance
$R_{fl}$  flowing particle layer specific resistance
$R_m$  membrane resistance
$R_o$  clean element radius
$R_p$  polarization resistance
$R_{rec}$  recovery ($Q_p / Q_f$)
$R_{sl}$  stagnant cake layer specific resistance
$Re$  Reynolds number
$Sc$  Schmidt number
$Sh$  Sherwood number
$t$  time
$t_{bf}$  backflush duration
$t_{ff}$  fastflush duration
$t_o$  operating time between two flux enhancement cycles
$t_{tot}$  total time for one complete operating and flux enhancement cycle
$T$  absolute temperature
$u$  axial velocity
$\bar{U}$  average cross-flow velocity of bulk suspension
$U_{ff}$  fastflush velocity
$\bar{U}_m$  average cross-flow velocity at mid-point of element
$U_o$  cross-flow velocity at channel inlet
$v$  transverse velocity
$V_{bf}$  backflush flux
$V_s$  solute flux
$V_w$  permeate flux
$V_{wo}$  permeate flux before formation of stagnant cake
$x$  axial coordinate
Greek Symbols

\( x_{cr} \)  
\( y \)

critical length
transverse coordinate

\( \alpha \)
fitting coefficient

\( \beta \)
dimensionless radial position within pore \((y / r_o)\)

\( \kappa \)
sieve coefficient

\( \delta \)
particle layer thickness

\( \delta_f \)
flowing layer thickness

\( \delta_{st} \)
stagnant layer thickness

\( \Delta l \)
membrane skin thickness

\( \Delta P \)
transmembrane pressure

\( \varepsilon \)
porosity

\( \phi \)
particle volume fraction

\( \bar{\phi} \)
average particle volume fraction in cake

\( \phi_b \)
bulk suspension particle volume fraction

\( \phi_o \)
feed suspension particle volume fraction

\( \phi_p \)
permeate suspension particle volume fraction

\( \phi_{\text{max}} \)
maximum particle volume fraction

\( \phi_w \)
particle volume fraction at wall

\( \dot{\gamma} \)
shear rate

\( \eta \)
relative viscosity \( (\mu(\phi) / \mu_o)) \)

\( \eta_{bf} \)
efficiency of backflush pump

\( \eta_f \)
efficiency of feed pump

\( \eta_r \)
efficiency of recycle pump

\( \lambda \)
solute to pore size ratio \((a_p / r_o)\)

\( \mu \)
dynamic viscosity

\( \mu_o \)
dynamic viscosity of particle-free fluid

\( \rho \)
fluid density

\( \nu \)
kinematic viscosity

\( \tau_w \)
wall shear stress

\( \tau_{wo} \)
wall shear stress before formation of stagnant cake
List of References


