REVERSE OSMOSIS PROCESS
FOR
WASTEWATER TREATMENT

by

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A thesis submitted in partial
fulfillment of the requirements
for the degree of
Master of Applied Science

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Ottawa, Canada
November, 1971

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ABSTRACT

The performance of Batch 316 type porous cellulose acetate membranes for the low pressure reverse osmosis separation of some organic pollutants in aqueous solutions and treatment of wastewater was illustrated. Experimental studies were carried out using six membranes whose $(D_{AM}/K\delta)_{NaCl}$ values at 300 psig were in the range 1.27×10^{-5} to 16.12×10^{-5} cm/sec. These membranes were capable of giving NaCl separations of 87% or above with the reference solution system 3500 mg/l $NaCl-H_2O$ when operated at 300 psig and a feed flow rate corresponding to $k_{NaCl} = 41 \times 10^{-4}$ cm/sec. The separation of glucose, sucrose, soluble starch, beef extract, glutamic acid, sodium stearate, ABS, and LAS in aqueous solutions at solute concentrations <2000 mg/l were studied using the same membranes under the above specified operating conditions. Solute separations were essentially independent of feed concentration and the average membrane fluxes were practically unaffected by the nature of the solute under the conditions indicated above; the average solute separation was >90% and the average membrane flux was in the range 15.8 to 40.6 gpd/sq.ft. The membranes were also capable of removing organics from wastewater and producing high quality water in the reverse osmosis treatment of wastewater. Reverse osmosis system analysis and process design were developed for the reference system $NaCl-H_2O$ and their utility was illustrated for wastewater treatment application.

ACKNOWLEDGEMENT

The author wishes to express his sincere appreciation to Dr. S. Sourirajan for his help, understanding, and never-failing guidance throughout this study, and to Dr. K. T. Brodersen for his advice and encouragement. Appreciation is also extended to Dr. T. Matsuura for his personal assistance.

The author is grateful to Dr. I. E. Puddington, Director of the Chemistry Division and Mr. W. S. Peterson, Head of the Chemical Engineering Section for granting permission to carry out most of this work in the Chemical Engineering Laboratories of the National Research Council.

Special thanks are also extended to Mr. R. Ironside for advice on analytical techniques; to Mr. L. Pageau, Mr. A. G. Baxter, and the staff of the analytical laboratories of the Ottawa City Pollution Control Centre for their help in this work; and to Mrs. W. Storto for her skillful typing of manuscripts.

The financial support of the National Research Council of Canada through the grant allocated to Dr. K. T. Brodersen and the financial assistance from the Department of Civil Engineering at the University of Ottawa, are greatly acknowledged.

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A	Pure water permeability constant, g.mole H_2O/sq.cm.sec.atm.
A_o	Value of A extrapolated to P = zero
ABS	Alkyl benzene sulfonate
B	Osmotic pressure proportionality constant, atm.
BOD	Biochemical oxygen demand, mg/l.
c	Molar density of solution, g.mole/cu.cm.
c_1, c_2, c_3	Molar density of feed solution, concentrated boundary solution, and membrane permeated product solution respectively, g.mole/cu.cm.
c_A	Solute concentration, g.mole/cu.cm. or mole fraction.
c_{A1}, c_{A2}, c_{A3}	Solute concentration in the bulk solution and the concentrated boundary solution on the high pressure side of the membrane, and in the product solution on the atmospheric pressure side of the membrane respectively, at any point (position or time) in the system, g.mole/cu.cm. or mole fraction.
$c^o_{A1}, c^o_{A2}, c^o_{A3}$	Values of c_{A1}, c_{A2}, c_{A3} respectively at time = zero in batch process, or at membrane entrance in the flow process.
c_{A3}	Average solute concentration in the membrane permeated product solution corresponding to given Δ, g.mole/cu.cm. or mole fraction.

C	(c_A/c^o_{A1}) or (X_A/X^o_{A1})
$C_1, C_2, C_3, \overline{C}_3$	Dimensionless quantities defined by Eqs. 2.12, 2.13, 2.14 and 2.15 respectively.
C^o_1, C^o_2, C^o_3	Values of C_1, C_2, C_3 respectively at time = zero in a batch process or at membrane entrance in a flow process.
COD	Chemical oxygen demand, mg/l.
$(D_{AM}/K\delta)$	Solute transport parameter, cm/sec.
$\overline{f}$	Average solute separation corresponding to given fraction (Δ) of product recovery = $1-\overline{C}_3$.
k	Mass transfer coefficient on high pressure side, cm/se
LAS	Linear alkylate sulfonate.
mμ	Millimicron.
M_B	Molecular weight of water.
N_B	Solvent water flux through membrane, g.mole/sq.cm.sec.
P	Operating pressure, atm.
(PR)	Rate of membrane permeated product solution per given area of membrane surface, g/hr.
(PWP)	Rate of pure water permeability through the membrane per given area of membrane surface, g/hr.
S	Area of membrane surface, sq.cm.
t	Time, sec.
TDS	Total dissolved solids, mg/l.
TOC	Total organic carbon, mg/l.

$\bar{u}$	Average fluid velocity parallel to membrane surface at any given position, cm/sec.
$\bar{u}^o$	Value of $\bar{u}$ at membrane entrance in a flow process, cm/sec.
v_w^*	Pure water permeation velocity through the membrane, cm/sec.
V_1	Quantity of solution on the high pressure side of the membrane at any time in batch process, cm^3.
V_1^o	Value of V_1 at time = zero in batch process, cm^3.
X	Nondimensional length parameter in flow process defined by Eq. 2.18.
X_A	Mole fraction of solute.
X_{A1}, X_{A2}, X_{A3}	Mole fraction of solute in bulk solution and concentrated boundary solution on the high pressure side of the membrane, and in membrane permeated product solution on the atmospheric pressure side of the membrane respectively, at any point (position or time) in the system.
$\bar{X}_{A3}$	Average value for the mole fraction of solute in the membrane permeated product solution corresponding to given fraction (Δ) of product recovery.
Z	$\int_1^{C_1} \frac{dC_1}{C_1-C_3}$
γ	Quantity defined by Eq. 2.8.

Δ	Fraction product recovery defined by Eq. 2.16.
η	Fraction power recovery.
θ	Quantity defined by Eq. 2.9.
λ	Quantity defined by Eq. 2.10.
τ	Quantity defined by Eq. 2.17.

CHAPTER 1

INTRODUCTION

Rapid urbanization has seriously contributed to today's water pollution problems. Pollution in many instances is due to the increased amounts of wastewater effluents reaching the receiving streams overloading the self-purification capacity of these running waters.

In the past, pollution control was concerned primarily with problems caused by domestic wastes and the simpler wastes of industry. Conventional waste treatment methods were efficient in treating such wastes and stopping their nuisance conditions. This is no longer true. Progress in abating pollution has been outdistanced by population growth, the increasing industrialization and technological developments which have placed additional burdens upon the wastewater treatment systems. Consequently, conventional treatment processes are already losing the battle against the modern-day tougher wastes. They are not able to prevent countless streams and lakes from receiving excessive amounts of suspended materials, oxygen consuming materials, nutrients and toxic wastes.

On the other hand, the unequal distribution of the total water supply of the world has left some areas having an abundance, others a deficiency. This condition is not new and has been recognized for many years but as population and demand for water increased, the situation is steadily getting worse and many areas are suffering from serious shortages in the water resource. In such cases, one of the important sources of good quality water, available in almost every country, is the wastewater discharged from many manufacturing operations and from sewage treatment plants. The renovation of such wastewater requires some special type of treatment to provide water of varying quality suitable for reuse in agricultural, industrial, recreational, ground water recharge or domestic applications. It should be pointed out that conventional waste treatment processes produce a water of limited reuse.

The need for water renovation shows the increasing requirement for developing more adequate treatment of wastes to provide better water pollution control and sufficient supply of clean waters for reuse. Such treatment is termed advanced waste treatment and refers to methods and processes that will remove more contaminants from wastewater than are usually removed in present day conventional treatment. It may modify or replace the various steps in conventional treatment, or it may be applied to the effluent from conventional secondary treatment. In the latter case it is also termed tertiary treatment.

The advanced treatment methods may be physical, chemical, or biological. The new physical and chemical separation techniques currently included are adsorption, electrodialysis, emulsion separation, evaporation, extraction, foaming, freezing, ion exchange, oxidation, and reverse osmosis. Many of these processes can accomplish complete water renovation. However, the challenge is to find the best and most economical way of achieving the goals. Some of the processes are routinely used in technologies other than the wastewater field such as in the conversion of saline water to fresh water.

A desalination process recently described as an advanced waste treating method is reverse osmosis. It is one of the few really new water-treating processes having substantial application where wastewater is to be processed to meet high effluent quality standards.

The osmosis phenomena describes the spontaneous passage of a pure solvent (e.g., water) through a semipermeable membrane into a solution, or from a less to a more concentrated solution. The semipermeable membrane is permeable to the solvent and impermeable to the solute. The normal flow of the solvent (water) continues until both the solutions in each side of the membrane are equally dilute or until sufficient pressure builds to create an osmotic equilibrium. Such a pressure is defined as the osmotic pressure of the solution having that particular concentration of solute. If a pressure greater than the osmotic pressure is exerted on the more concentrated solution, the solvent (water) will flow from the more concentrated solution through the semipermeable membrane to the less concentrated solution. The latter phenomena has been conveniently termed reverse osmosis. In both osmosis and reverse osmosis, under isothermal conditions, the preferential transport of material through the membrane is always in the direction of lower chemical potential which is a thermodynamic requirement. It seems necessary to point out that neither osmosis nor reverse osmosis is an explanation of the mechanism of the process involved. This mechanism is still an open question, and the distinction between the two terms is entirely one of arbitrary convention and popular usage.

Although the osmosis phenomena was first observed in 1748, it has been only recently that scientists realized the possibility of utilizing the reverse osmosis process as a general and widely applicable technique for the separation, concentration or fractionation of substances in fluid solutions. The process consists of letting the fluid mixture flow under pressure through an appropriate porous membrane, and withdrawing the membrane-permeated product which is enriched in one or more constituents of the mixture, leaving a concentrated solution of other constituents on the upstream side of the membrane. No heating of the membrane and no phase change in product recovery are involved in this separation process. The advantage of reverse osmosis lies in its potential to separate water from contaminants of many kinds rather than to remove one specific class of contaminants from the water.

The most significant current application of reverse osmosis process is in the field of saline water conversion. Although the important developments in this process have been in the last ten years, interest is rapidly increasing for new applications, one of which being in the field of wastewater treatment. The reason is the simplicity of the process, the high product quality, and the relatively low treatment cost that could result from further process development. For water purification applications, in general, porous cellulose acetate membranes have been found appropriate in the reverse osmosis process.

Perhaps the most significant problem that has arisen from the use of reverse osmosis for wastewater treatment rather than saline water demineralization is the one resulting from the presence of organic matter in wastewater. Higher levels of organic removal than those attainable by the biological treatment methods are required for more effective water pollution control and progressive water reuse. The well operated modern biological waste treatment plants can provide approximately 90 percent removal of organics measured as the biochemical oxygen demand (BOD). The possibility exists that more efficient removal of organics than that obtained by biological processes can be achieved through the use of reverse osmosis process. Accordingly, considerable research must be directed toward the investigation of removal of organics from wastewater by the reverse osmosis process.

This study was undertaken in order to investigate the performance of reverse osmosis membranes for the removal of organic materials from wastewater. A new improved type of porous cellulose acetate membranes operated at low pressure was used for this purpose.

CHAPTER 2

LITERATURE SURVEY

Reverse osmosis is a new term in the water purification technology but a subject of interest for world wide investigations. Its technical feasibility was only assured about ten years ago, mainly as a result of work carried out at the Universities of Florida and California. At UCLA, the development of the Loeb-Sourirajan (50,29) technique for making asymmetric porous cellulose acetate membranes for water desalination in 1960, has led to a rapid increase in our understanding of osmotic membranes and to major progress in the use of the reverse osmosis process. Since that time, extensive investigations have been initiated in several laboratories in search of superior membranes and new applications for the reverse osmosis process. Because the process was first developed for desalination purposes, most of the work in reverse osmosis subject has been carried out for saline water conversion, and emphasis therefore has been on the development of membranes and apparatus for this purpose. However, the developed technology appears to be appropriate for wastewater treatment and other water pollution control applications.

The reverse osmosis process is an interesting and promising development. In spite of the rapid advances which are being made in reverse osmosis applications, the process is still at its very early stages of development.

The following literature survey reviews the development of reverse osmosis theory, the advancement of reverse osmosis membrane technology, and the research accomplished on the application of the reverse osmosis process to wastewater treatment.

2.1 Reverse Osmosis Theory

The theory involved in reverse osmosis is concerned mainly with transport through the reverse osmosis membrane. Several approaches to the subject of transport in reverse osmosis are currently considered (40). All approaches are equally valid to the extent that they are consistent with the reverse osmosis experimental data and to the extent that they predict membrane performance. The approach which has proved most useful in the development of the reverse osmosis process is based on the preferential sorption-capillary flow mechanism suggested by Sourirajan (40) and is used in the Kimura-Sourirajan analysis (22) of reverse osmosis experimental data leading to the basic equations of transport through reverse osmosis membranes.

The development of basic transport equations together with the correlations of the reverse osmosis experimental data has led to the concepts of membrane specification and reverse osmosis system specification, and has offered the basis for reverse osmosis process design.

2.1.1 Preferential Sorption-Capillary Flow Mechanism

The preferential sorption-capillary flow mechanism describes the reverse osmosis separation as the combined result of an interfacial phenomenon and fluid transport under pressure through capillary pores, and describes the appropriate membrane as a porous medium at all levels of solute separation.

The mechanism (40) can be illustrated by a conceptual model for recovering fresh water from aqueous salt solutions (Figure 1-A). The model shows the solution in contact with a porous film whose surface is of such a chemical nature that it has a preferential sorption for water and/or preferential repulsion for the solute. Because of this preferential sorption for water, a multimolecular layer of preferentially sorbed pure water, called the pure water interface, exists at the film-solution interface. Above the pure water interface is found the bulk of the solution which consists of the aqueous salt solution. The continuous removal of the preferentially sorbed interfacial water is effected by letting it flow under pressure through the capillaries existing in the film. The above model also gives rise to the concept of a critical pore diameter for maximum separation and permeability. This critical pore diameter is twice the thickness of the preferentially sorbed interfacial water layer (Figure 1-B). The success of reverse osmosis separation process is therefore dependent upon the appropriate chemical nature of the film surface in contact with the solution and upon the existence

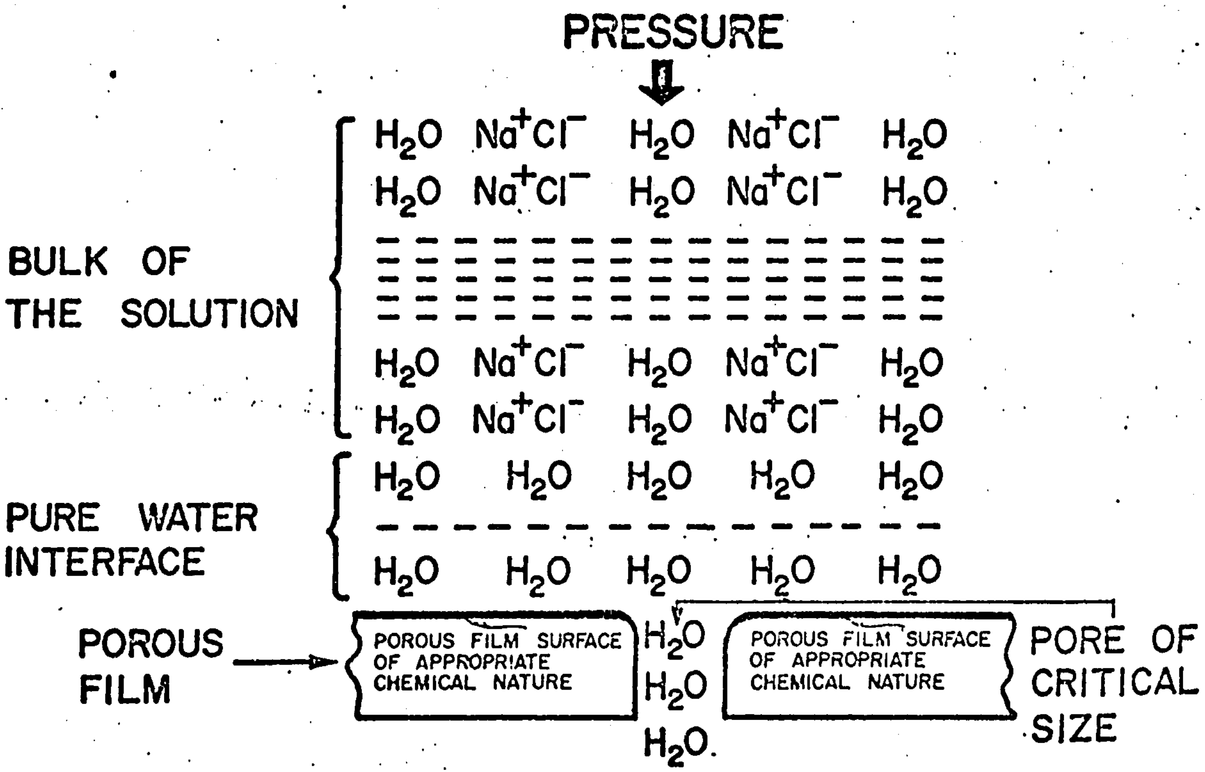


FIGURE 1-A SCHEMATIC REPRESENTATION OF PREFERENTIAL SORPTION-CAPILLARY FLOW MECHANISM (40).

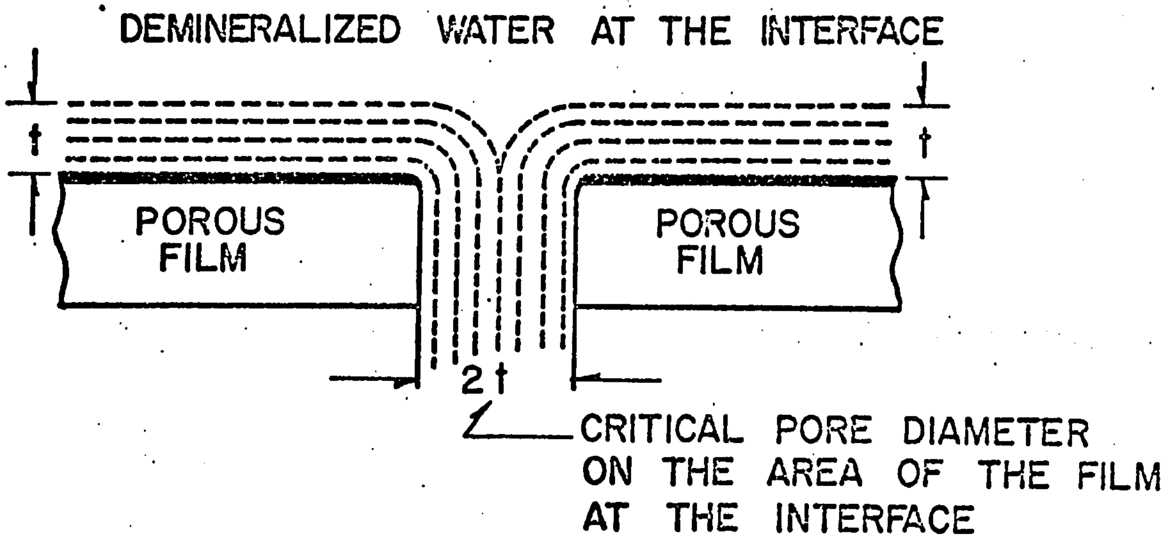


FIGURE 1-B CRITICAL PORE DIAMETER FOR MAXIMUM SEPARATION AND PERMEABILITY (40).

of pores of appropriate size on the area of the porous film at the film-solution interface.

2.1.2 Basic Transport Equations and Correlations

The preferential sorption-capillary flow mechanism discussed above provides the basis for the Kimura-Sourirajan analysis (22) of reverse osmosis experimental data by a consideration of viscous flow for water transport, pore diffusion for solute transport, and "film" theory for expressing the effect of mass transfer (on the high pressure side of the membrane) on total material transport across the membrane. In this analysis, it is considered that the total material transport across the membrane at any point in a given membrane-solution-operating system under steady state conditions is the vector sum of simultaneous water transport (proportional to effective pressure difference) and solute transport (proportional to effective concentration difference) across the membrane, both of which are controlled by the mass transfer situation on the high pressure side of the membrane. The Kimura-Sourirajan analysis, which is applicable to all levels of solute separation, gives rise to the following basic equations relating the pure water permeability through the membrane, the transport of both the solvent water and the solute through the membrane, and the mass transfer situation on the high pressure side of the membrane at any point (position or time) in the system under operating conditions of constant temperature and pressure.

The basic transport equations are:

$$A = \frac{(PWP)}{M_B \times S \times 3600 \times P} \tag{2.1}$$

$$N_B = A[P - \pi(X_{A2}) + \pi(X_{A3})] \tag{2.2}$$

$$= \left(\frac{D_{AM}}{K\delta}\right)\left(\frac{1-X_{A3}}{X_{A3}}\right)(c_2X_{A2}-c_3X_{A3}) \tag{2.3}$$

$$= kc_1(1-X_{A3}) \ln \left(\frac{X_{A2}-X_{A3}}{X_{A1}-X_{A3}}\right) \tag{2.4}$$

Equation 2.1 defines the pure water permeability constant, A, which is a fundamental quantity with respect to the membrane. The quantity A is a measure of the overall porosity of the membrane and is determined while operating the system with a pure water feed; it represents pure water transport in the absence of any concentration polarization on the upstream side of the membrane (absence of concentration gradient between the boundary solution and the bulk feed solution shown in Figure 2). The magnitude of A is independent of any solute under consideration.

The transport of solvent water N_B is defined by equations 2.2, 2.3, and 2.4. Equation 2.2 enables the calculation of the concentration X_{A2} of the boundary solution (Figure 2) on the high pressure side of the membrane under the reverse osmosis operating conditions. When the concentration X_{A3} of the membrane permeated product solution on the low pressure side of the membrane is less than the

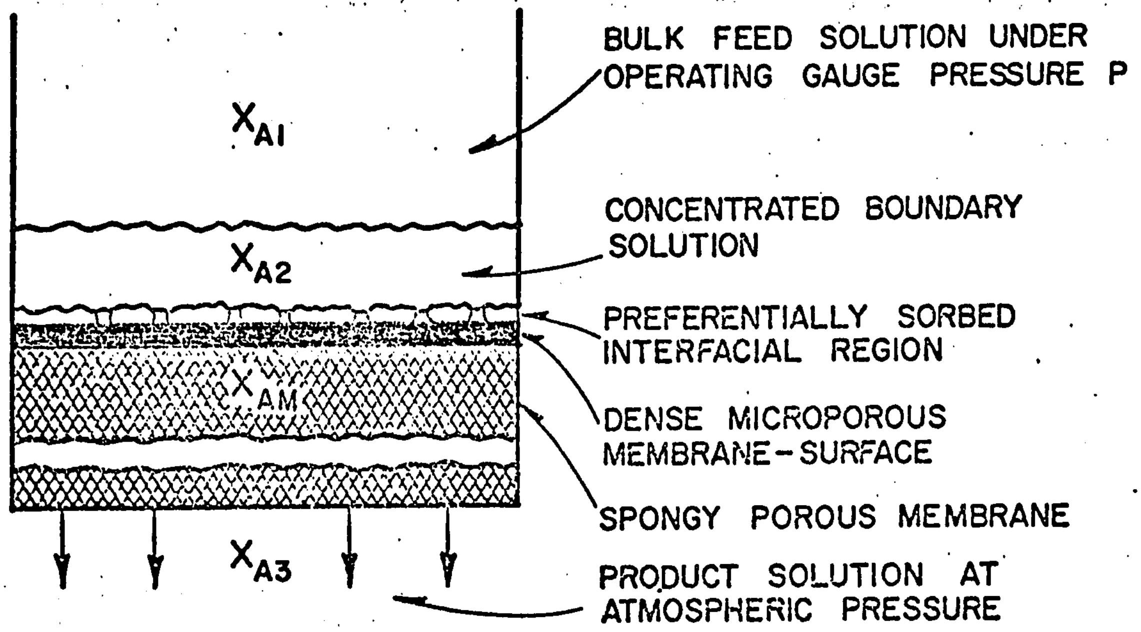


FIGURE 2 REVERSE OSMOSIS PROCESS UNDER STEADY-STATE CONDITIONS (40).

concentration X_{A1} of the bulk feed solution on the high pressure side of the membrane, X_{A2} is greater than X_{A1}.

Equation 2.3 defines the quantity $(D_{AM}/K\delta)$ for the transport of any given solute through the membrane. This quantity $(D_{AM}/K\delta)$, called the solute transport parameter, is a fundamental one for any given membrane-solution system. It is not a mere proportionality constant, but a combination of at least three distinct quantities all of which have important physical significance, but none of which can be precisely measured. |The above three quantities are the solute diffusivity through the membrane (D_{AM}), effective film thickness (δ), and the factor K which relates solute concentration in the membrane phase to that in the solution phase in equilibrium with the membrane surface|. From the point of view of reverse osmosis process design, it is sufficient to know the overall value of the parameter $(D_{AM}/K\delta)$ which plays the role of a mass transfer coefficient with respect to solute transport through the membrane; hence the parameter is treated as a single quantity for purposes of analysis.

Equation 2.4 defines the mass transfer coefficient k on the high pressure side of the membrane. The magnitude of k is a function of the nature of solute, concentration of the bulk feed solution and feed flow conditions on the high pressure side of the membrane. For a given membrane-solution-operating system, the obtainable value of k affects solute separation and membrane flux.

The basic transport Equations 2.1 to 2.4 are applicable for the analysis of any reverse osmosis separation process. In the reverse osmosis experimental data analysis the correlations of A, $(D_{AM}/K\delta)$, and k with operating conditions are of practical interest from the point of view of specifying membranes, and predicting membrane performance under different operating conditions. Extensive experimental studies (22,23,44) have shown that, at a given temperature, with respect to the Loeb-Sourirajan type porous cellulose acetate membranes and a number of solution systems such as sodium chloride-water, both A and $(D_{AM}/K\delta)$ are dependent on the porous structure of the membrane; hence they are different for different membranes. Both are functions of operating pressure (P), which may be expressed as follows:

$$A = A_o e^{-\alpha P} \tag{2.5}$$

and

$$(D_{AM}/K\delta) \propto P^{-\beta} \tag{2.6}$$

where α and β are constants, and A_o is the extrapolated value of A when $P = 0$. In addition, $(D_{AM}/K\delta)$ is dependent upon the chemical nature of solute, and, at a given operating pressure it is independent of feed concentration and feed flow rate for many membrane-solution systems. The mass transfer coefficient k on the high pressure side of the membrane is essentially a function of nature of solute, feed concentration, feed flow rate, and the geometry of the apparatus or membrane configuration. The effect of feed flow rate (Q) on k may be expressed by the relation:

$$k \propto Q^n \qquad (2.7)$$

where n is a constant characteristic of the solution system and the apparatus used.

All correlations of A, $(D_{AM}/K\delta)$, and k are required for reverse osmosis process design (40) and have to be experimentally determined and established for every specific membrane-solution-operating system. The basic transport equations themselves are independent of such correlations.

2.1.3 Membrane Specification

It has been shown (22,23,44) that the above observations lead to a useful method of specifying reverse osmosis membranes. For a given solution system and operating pressure, a membrane is completely specified in terms of its pure water permeability constant, A, and the solute transport parameter, $(D_{AM}/K\delta)$, at that pressure. The mass transfer coefficient, k, on the high pressure side of the membrane specifies the feed flow condition used. A single set of experimental data on pure water permeation rate (PWP), membrane permeated product (PR), and solute separation (f) at any feed concentration and feed flow condition at the given operating pressure is sufficient, following the Kimura-Sourirajan analysis, to obtain data on A and $(D_{AM}/K\delta)$ for the membrane-solution system at that pressure, and to get the value of k corresponding to the feed concentration and feed flow conditions used in the particular experiment. Sodium chloride-water system is chosen as a particular reference solution system for membrane specification because of its practical significance.

When a membrane is specified in terms of A and $(D_{AM}/K\delta)$ at a particular pressure, its performance at that pressure, expressed in terms of (PR) and f, can be predicted for the entire applicable range of feed concentrations and feed flow rates provided that the appropriate mass transfer coefficient data are available (44).

Further, during continuous operation of the reverse osmosis process under pressure, A is reduced but $(D_{AM}/K\delta)$ remains constant provided the surface pore structure of the membrane remains unchanged (23). This compaction effect of a membrane can be specified by an appropriate A-factor defined as the ratio of the A value for a membrane at any time to that at the start of the operation (or simply the ratio of the corresponding (PWP) values for this membrane).

For a given membrane-solution-operating reverse osmosis system, the specification of both the membrane and the feed flow condition of solution used in the system form the basis for the specification of this particular system as discussed in the following analysis (33).

2.1.4 Reverse Osmosis System Specification

In any reverse osmosis process, the membrane-solution-operating system may be specified by three fundamental non-dimensional parameters - γ, θ, and λ - defined as follows:

$$\gamma = \frac{\pi(X_{A1}^{o})}{P} = \frac{\text{osmotic pressure of initial feed solution}}{\text{operating pressure}} \qquad (2.8)$$

$$\theta = \frac{(D_{AM}/K\delta)}{V_W^*} = \frac{\text{solute transport parameter}}{\text{pure water permeation velocity}} \qquad (2.9)$$

and

$$\lambda = \frac{k}{(D_{AM}/K\delta)} = \frac{\text{mass transfer coefficient on the high pressure side of membrane}}{\text{solute transport parameter}} \qquad (2.10)$$

where

$$V_W^* = \frac{AP}{c} \qquad (2.11)$$

and the quantity (X_{A1}^{o}) refers to the osmotic pressure of the feed solution at membrane entrance in flow process or the start of operation in batch process.

For reverse osmosis systems so specified, the basic transport equations discussed previously (Eqs. 2.2 to 2.4) can be rewritten in forms suitable for a generalized approach to reverse osmosis process design. This is illustrated by the following analysis of Ohya and Sourirajan (33).

2.1.5 General Equations for Reverse Osmosis Process Design

The Ohya-Sourirajan analysis for reverse osmosis process design is applicable to reverse osmosis membranes for which $(D_{AM}/K\delta)$ is independent of solute concentration and feed flow rate, and for feed solutions whose osmotic pressure is directly proportional to mole fraction of solute (i.e., $\pi(X_A) = BX_A$), and whose molar density (c) may be assumed constant (these conditions are reasonably valid for aqueous sodium chloride solutions at least up to 2.0 M for purposes of engineering calculations). In this analysis the performance of a reverse osmosis system can be expressed in terms of the following dimensionless quantities:

$$C_1 = \frac{c_{A1}}{c^o_{A1}} \; ; \; C^o_1 = 1 \tag{2.12}$$

$$C_2 = \frac{c_{A2}}{c^o_{A1}} \; ; \; C^o_2 = \frac{c^o_{A2}}{c^o_{A1}} \tag{2.13}$$

$$C_3 = \frac{c_{A3}}{c^o_{A1}} \; ; \; C^o_3 = \frac{c^o_{A3}}{c^o_{A1}} \tag{2.14}$$

$$\bar{C}_3 = \frac{\bar{c}_{A3}}{c^o_{A1}} \tag{2.15}$$

$$\Delta = 1 - \frac{V_1}{V^o_1} \quad \text{or} \quad 1 - \frac{\bar{u}}{\bar{u}^o} \tag{2.16}$$

and

$$\tau = \frac{S v^*_w t}{V^o_1} \tag{2.17}$$

or

$$X = \frac{v_w^*}{\overline{u}^o} \frac{x}{h} \tag{2.18}$$

where

$$v_w^* = \frac{AP}{c} = \frac{A_o e^{-\alpha p_P}}{c} \tag{2.19}$$

The quantities C_1, C_2, and C_3 represent the concentrations (expressed in dimensionless terms) of the bulk solution and the concentrated boundary solution on the high pressure side of the membrane, and the membrane permeated product solution on the low pressure side of the membrane respectively at any point in the system. The quantity $\overline{C}_3$ is the average value of C_3 corresponding to a specified fraction (Δ) of product water recovery. The parameter τ is the dimensionless time for the batch-wise reverse osmosis process where S is the surface area of membrane and t is time. The value of τ is identical to that of X which is applicable for the reverse osmosis flow process where x is the longitudinal distance along the length of the membrane, $\overline{u}^o$ is the average velocity of feed solution at membrane entrance, and $\frac{1}{h}$ is membrane area per unit volume of fluid space in the reverse osmosis unit.

For any reverse osmosis system specified in terms of γ, θ, and λ, the precise relationships among the quantities C_1, C_2, C_3, $\overline{C}_3$, τ or X, and Δ constitute the fundamental aspect of membrane transport in the engineering science of reverse osmosis. These relationships are given by the Ohya-Sourirajan general system analysis in which the following relevant equations have been derived (24,33).

$$C_1^o = 1 = C_3^o \left[1 + \frac{1}{(\gamma C_3^o + \theta)} \quad \exp\{- \frac{1}{\lambda(\gamma C_3^o + \theta)} \} \right] \tag{2.20}$$

$$C_2^o = C_3^o \left[1 + \frac{1}{(\gamma C_3^o + \theta)} \right] \tag{2.21}$$

$$C_1 = C_3 \left[1 + \frac{1}{(\gamma C_3 + \theta)} \quad \exp\{- \frac{1}{\lambda(\gamma C_3 + \theta)} \} \right] \tag{2.22}$$

$$C_2 = C_3 \left[1 + \frac{1}{(\gamma C_3 + \theta)} \right] \tag{2.23}$$

$$z = \int_{C_3^o}^{C_3} \left[\frac{\gamma}{\lambda(\gamma C_3 + \theta)^2} - \frac{\gamma}{(\gamma C_3 + \theta)} + \{ \frac{(\gamma C_3 + \theta)}{C_3} \quad \exp \frac{1}{\lambda(\gamma C_3 + \theta)} \} + \frac{1}{C_3} \right] dC_3 \tag{2.24}$$

$$\begin{pmatrix} \tau \\ \text{or} \\ X \end{pmatrix} = \int_{C_3^o}^{C_3} \left[\frac{\gamma}{\lambda(\gamma C_3 + \theta)} - \gamma + \frac{(\gamma C_3 + \theta)}{C_3} \{ 1 + (\gamma C_3 + \theta) \exp \frac{1}{\lambda(\gamma C_3 + \theta)} \} \right] \frac{\exp(-z)}{\theta} \quad dC_3 \tag{2.25}$$

$$\Delta = 1 - \exp(-z) \tag{2.26}$$

$$1 = C_1(1 - \Delta) + \overline{C}_3 \; \Delta \tag{2.27}$$

It may be recalled that,

$$\tau(\text{for batch process}) = \frac{S v_w^* t}{V_1^o} \tag{2.17}$$

and

$$X(\text{for flow process}) = \frac{v_w^* x}{\overline{u}^o h} \tag{2.18}$$

The above equations show that for a specified reverse osmosis system, any one of the six quantities C_1, C_2, C_3, $\bar{C}_3$, τ or X, and Δ uniquely fixes all the other five quantities and that the performance of such specified system can be predicted to express any five of these quantities as a function of the sixth.

2.2 Reverse Osmosis Membrane Technology

Since the semipermeable membrane is the heart of the reverse osmosis process, the early work was initiated in search for such suitable membranes. The first serious effort to find reverse osmosis membranes suitable for desalination purposes was apparently that of Reid (37) and his associates at the University of Florida. They concluded that the cellulose acetate membranes tested by them could not be used as ultrafilters for purifying saline waters because their rate of filtration was too slow and their life was limited to only a few weeks, and consequently some other type of semipermeable membrane would probably be developed for the purpose. Soon after, in 1960, Loeb and Sourirajan (50,29) at UCLA developed their own technique for making porous cellulose acetate membranes capable of giving both high product water flows and high levels of solute separation from aqueous sodium chloride solutions.

The Loeb-Sourirajan film-making technique is essentially as follows: The film casting solution contains cellulose acetate dissolved in a solvent, such as acetone, to which is added a water soluble additive, such as an aqueous solution of magnesium perchlorate or pure formamide. With the above solution, membranes are cast on flat or tubular surfaces (for example, glass plates or metal tubes) at specified ambient conditions. On casting, part of the solvent is allowed to evaporate for a definite period of time from the surface of the membrane at the temperature of the casting atmosphere (evaporation step). This step is followed by immersion of the membrane in ice-cold water for at least an hour; during this time, the film sets to a gel from which the solvent and the additive are leached out (gelation step) leaving a tough solid porous film on the flat or the tubular surface from which it can be removed easily. Every detail relating to the casting solution composition, evaporation step, and the gelation step affects the subsequent performance of the membrane in reverse osmosis. Membranes made in the above manner have a relatively dense microporous structure (pore size probably <50Å) on an extremely thin layer (<0.25 micron) of the film surface which was exposed to the atmosphere during casting. This side of the film is called the "surface layer side" and the other side the "back side". The bulk of the film material underneath the thin surface layer is a spongy porous mass containing comparatively very big (∿4000-10,000Å) interconnecting pores. Consequently, the above films are said to have an asymmetric porous structure. It is the microporous structure of the dense surface layer which governs the level of solute separation in reverse osmosis, and hence the surface layer side is held facing the feed solution during reverse osmosis operation.

In the as-cast condition, the surface pores may be too big to give any significant solute separation. But the membrane can be shrunk and the pore size reduced by simply keeping the film immersed in hot water for about 10 minutes (temperature treatment). Thus, by adjusting the shrinkage temperature, different porosities on the surface layer side of the membrane capable of giving different levels of solute separation for a given solution system, can be obtained. In order to stabilize the overall porous structure of the film, it is necessary to subject the film to an initial pressure treatment, following the temperature treatment. The pressure treatment usually consists of pumping water past the surface of the film mounted in the reverse osmosis unit operating at a pressure about 15% higher than the maximum intended operating pressure for at least one hour.

It should be pointed out that the asymmetric porous structure and capability for solute separation under reverse osmosis operating conditions are the basic characteristics of Loeb-Sourirajan type membranes. The significant contribution involved in making such membranes has opened the way for more developments in reverse osmosis membrane technology and applications.

With respect to porous cellulose acetate membranes, two improvements have been reported in 1965, one by Sourirajan and Govidan (43) and the other by Manjikian, Loeb, and McCutchan (30), these membranes were designated as Batch 18 and Batch 47 type membranes respectively. For water desalination application, the performances of the latter two types of films were about the same, and very much better than the performance of Batch 25 type films developed earlier by Loeb and Sourirajan.

Although the latter two developments of new improved membranes have shown good performance, the need is still great for developing more productive reverse osmosis membranes for different applications.

During the past few years many materials have been tested to determine their usefulness as reverse osmosis membranes. Although some improved membranes especially those based upon cellulose are being developed, the cellulose acetate membranes, exhibiting the best performance, have been the most widely used in commercial or research and development reverse osmosis units for water-separation applications.

A new approach to the general problem of developing more productive reverse osmosis membranes has been stated recently (1970) by Kunst and Sourirajan (26,27). In this approach, the state or the structure of the casting solution and the rate of solvent evaporation during film formation together constitute an important interconnected variable governing the ultimate porous structure, and hence the performance of the resulting membranes in reverse osmosis. The work with the new approach has led to the development of two new types of porous cellulose acetate membranes designated as Batch 301 and Batch 316 type films.

In the preparation of Batch 301 porous cellulose acetate membranes, the temperature of the casting solution and that of the casting atmosphere are different and constitute separate variables. The latter feature in the process of making porous cellulose acetate membranes is the basis of a major improvement reported by Kunst and Sourirajan (26), and is different from the common feature in the process of making Batch 18 and 47 type membranes in which the above two temperatures are the same. The productivity of Batch 301 type membranes in reverse osmosis is twice that of Batch 18 and 47 type membranes at operating pressures of 250 psig or less (28). The film casting details with respect to Batch 25, 18, 47, 301, and 316 type membranes are given in Table 1.

TABLE 1

Film Casting Details - Summary of Literature Data

Film Casting Details	Film Type				
	Batch 25 (50)[a]	Batch 18 (43)	Batch 47 (30)	Batch 301 (26)	Batch 316 (25)
Casting Solution Composition, wt.%					
Cellulose acetate (acetyl content 39.8%)	22.2	17	25	17	17
Acetone	66.7	68	45	68	69.2
Magnesium perchorate	1.1	1.5	-	1.5	1.45
Water	10.0	13.5	-	13.5	12.35
Formamide	-	-	30	-	-
Casting Solution Temperature, ^oC	-10	-10	23-25	0	0
Temp. of Casting Atmosphere, ^oC	-10	-10	23-25	23-25	24
Casting Atmosphere	ambient air	ambient air	ambient air	ambient air in equilibirum with aqueous solution of 30 wt.% acetone	ambient air in equilibrium with aqueous solution of 30 wt. % acetone
Solvent Evaporation Time, min	4	4	½ to 1	2	6
Duration of Film-setting in					
Ice-cold water, hr	~1	~1	~1	~1	>1
Nominal Film Thickness, inch	0.010	0.004	0.004	0.004	0.005

[a] Reference Number

Extending their earlier investigations (26-28), Kunst and Sourirajan stated that the distinguishing feature of their new approach to reverse osmosis membranes research lies in the analysis of the usual film casting variables in terms of correlating membrane performance data with the casting solution structure, solvent evaporation rate during film formation, and the shrinkage temperature profile. On the basis of the above correlation, they have developed the Batch 316 type porous cellulose acetate membranes which are in some respects even better than the Batch 301 type membranes for low pressure reverse osmosis desalination and other applications (25).

The development of Batch 316 type membranes represents a significant contribution to the production of more productive-low pressure operated reverse osmosis membranes for the economical wide variety of reverse osmosis process applications.

2.3 Reverse Osmosis for Wastewater Treatment

Reverse osmosis is a general separation process. Its application to water desalination has been extensively studied. The possible application of the process to the treatment and reclamation of wastewaters, however, has relatively received little attention. It has been only recently that considerable research work has been carried out on the use of reverse osmosis for wastewater treatment and reclamation.

The first reported work on the application of reverse osmosis to wastewater reclamation was sponsored by the U.S. Public Health Service during 1964 (48). In this work, two membrane cells, each containing a single 500 sq cm sheet of membrane were operated at pressures of 50 and 100 atm, using municipal secondary sewage effluent. The process was found highly effective in removing many contaminants, but serious problems were encountered with product water flow rates.

At the Pomona Water Reclamation Facility, California, a series of experiments (5) using larger membrane areas and relatively lower system pressures were carried out with municipal secondary treated wastewater as a feed. After chlorination, the feed water was passed through a diatomaceous earth filter and then into a number of spiral-wound membrane modules operating in parallel. The cellulose acetate membranes used in the experiments were prepared by methods similar to those described by Loeb and Sourirajan (29) but with various permeabilities and selectivities. Reporting the experimental results (Table 2) Merten and Bray (32) showed the potentialities of using reverse osmosis for wastewater reclamation but recommended longer-term pilot-plant scale studies for the complete assessment of such potentialities. They found that high-selectivity membranes gave good retention of multivalent ions, but univalent ions were not so well retained. Finally, it was pointed out that there is the possibility of using less selective but more permeable membranes and of operating with less highly treated effluents. A laboratory study conducted by Ironside and Sourirajan (20) also indicated the possible practical use of the reverse osmosis porous cellulose acetate membrane technique for water pollution control and water renovation.

In a summary report (49) on advanced waste treatment research in the U.S., it was stated that reverse osmosis has the capability of removing inorganic and many of the organic materials in wastewater but there is the need for developing membranes with better capability of removing soluble organics from wastewater. It was recommended that membrane configuration, membrane fouling, pretreatment requirements, and membrane life are to be investigated. In another summary, Stenburg et al. (46) emphasized the importance of developing longer-lived membranes with high flux rates (20 gpd/sq ft to 30 gpd/sq ft) to improve the cost of wastewater treatment by reverse osmosis process. Improvements in the cost picture would also result from optimizing the pretreatment requirements, improving the product-to-waste ratio, and reducing membrane costs.

Parkhurst et al. (36) described the results of further investigations into the use of reverse osmosis for wastewater reclamation carried out on a pilot plant basis at Pomona, California, in early 1967. The system used was composed of eighty-six spiral wound modules encased in nine steel tubes. The modules contained 10 sq ft of cellulose acetate membranes in a 5,000 gpd reverse osmosis unit manufactured by Gulf General Atomic. Secondary treated sewage, after carbon treatment, was acidified to a pH of approximately 5.0 and used as feed to the reverse osmosis unit operated at 400 psi. Typical data on influent composition and removal efficiencies are presented in Table 2. The authors commented that reverse osmosis systems must be investigated at considerable length before a reasonable evaluation of its technical and economic potential can be ascertained. A similar study was reported by Sudak and Nusbaum (47) using the 5,000 gpd reverse osmosis unit at Pomona, California. The unit was operated at 400 psi on effluent from an activated carbon column. This effluent had previously been given secondary treatment by the activated sludge process. The experimental run was extended in order to recover 80 to 85% of the feed water as a product. During this run, a simple daily flushing procedure was established and was found effective in keeping the membranes free of deposits heavy enough to interfere with the transport of water. The data on influent water quality and percentage removal of the water pollutants by this reverse osmosis unit are given in Table 2. It was concluded that reverse osmosis could have a valuable role in upgrading wastewaters.

Hauck and Sourirajan (14) studied the performance of some porous cellulose acetate membranes for the removal of pollutants from wastewater. They used a typical reverse osmosis membrane of the type Batch 18 in a laboratory scale unit operating on raw sewage to recover 90% of the feed water as product. The reported average BOD removals were 85.8 and 80.8% at operating pressures of 1000 and 500 psig respectively. Under the conditions of the experiments made, the average separation of nitrates, alkylbenzene sulfonate, and phosphates were 50.3, 93 and >99% respectively. The average product rates were 32.7 and 18.3 gpd/sq ft at 1000 and 500 psig respectively. The membrane used was specified in terms of pure water permeability constant (A) and solute transport parameter $(\frac{D_{AM}}{K\delta})$ for sodium chloride. At operating pressure of 1000 psig, the reported values of (A) and $(\frac{D_{AM}}{K\delta})$ were $1.695 \times 10^{-6} \frac{\text{gm.mole } H_2O}{\text{sq.cm.sec.atm.}}$ and 3.457×10^{-5} cm/sec, respectively.

It must be noted that all of the above research was accomplished using actual wastewaters which had previously been given different levels of treatment. Since the composition and concentration of these wastewaters are different, it is important to study the removal of specific wastewater chemical constituents rather than solely investigate the removal of those substances found in a variety of treated wastewaters. The last concept has been recently emphasized by different workers.

Bennett et al. (3) and Hindin et al. (17) studied the permeation through modified cellulose acetate membranes of a variety of organic chemical species that may occur in wastewater. Using a laboratory scale reverse osmosis cell operated at a pressure differential of 100 atmospheres (about 1500 psi), they reported reductions in the range of 80 to 99 percent for those organic chemical species that existed primarily in the colloidal, aggregate, micelle, or macromolecular form, while removals of 50 to 80 percent were obtained for molecules in true solution for which the vapor pressure was appreciably greater than that of water. Therefore, it was claimed that the vapor-pressure theory might be involved in the transport of some organics across the reverse osmosis membrane. The observed membrane flux was in the range of 5 to 25 gpd/sq ft. The studies conducted with both aqueous solutions or dispersions of chemical species mixtures and wastewater effluents containing specific chemical species showed that the synergistic effect present in the multi-component aqueous solution was not present in sewage effluent. In another paper, Hindin and

Bennett (16) presented considerable data on the removal of specific inorganic species which are found in most treated wastewater effluents. Modified cellulose acetate membranes in a laboratory scale reverse osmosis cell were used. The authors also included results showing the permeation of the major inorganic constituents of secondary-treated sewage effluent. The latter results are presented in Table 2.

The reverse osmosis separation of some organic solutes in aqueous solution was investigated by Matsuura and Sourirajan (31) using the Loeb-Sourirajan type porous cellulose acetate membranes in a laboratory scale cell. The organic compounds were studied singly and in mixed systems. For the system glucose-water, the results showed 95% solute recovery with a processing capacity of 53 gallons of feed per day per sq ft of membrane surface obtained for the concentration of aqueous glucose solution from 0.1 to 1.0 M at an operating pressure of 1500 psig using a film capable of giving an average solute separation of 97.7%. It was found that the prediction technique developed previously for aqueous solution systems containing mixed inorganic solutes with a common ion was also applicable for the systems containing non-ionic mixed organic solutes. Also, the authors reported that the reverse osmosis transport characteristics for the systems glucose-water, maltose-water, lactose-water, ethylene glycol-water and propylene glycol-water in the feed concentration ranges 0.1 to 1.5 M, 0.03 to 0.11 M, 0.04 to 0.22 M, 0.2 to 1.5 M, and 0.2 to 0.8 M, respectively, were similar to those observed for the system sodium chloride-water.

The work with actual treated wastewater effluents in reverse osmosis treatment studies has been continued on both laboratory and pilot plant scales. Eden (9) published results of the treatment of raw sewage after chemical coagulation and aeration carried out at Stevenage, England, using a small recirculatory reverse osmosis system operated at 600 psi. The product was compared with Stevenage tap water and was found to be of rather better quality. These results are given in Table 2. Hardwick (13) also reported the previous work and described the work at Harwell, England, in which tests were running for many months with sewage effluent from settling tanks and percolating filters using a larger scale reverse osmosis system. The cellulose acetate membranes used showed satisfactory rejection of the effluent constituents and were still functioning well after 6000 hours exposure. The author discussed the difficulties attributable to bacterial attack and fouling of membranes but finally recommended reverse osmosis as a strong competitor for several applications.

The foregoing review has only presented the research in reverse osmosis conducted in the treatment and reclamation

TABLE 2

Reverse Osmosis for Municipal Wastewater Treatment - Summary of Literature Data

	(Ref. 48)			(Ref. 32)		
	Feed	% removal		Feed	% removal	
	Conc.	Test(1)	Test(2)	Conc.	Test(1)	Test(2)
Contaminants						
TDS mg/l	550	97	95	620	90	49
Conductivity µmho				1,034	94	45
Ammonia mg/l-N	25[a]	100	84	17	82	35
Nitrates mg/l-N	2[b]	100	100			
Phosphate mg/l	25[c]	100	100	39[c]	97	96
Sulphate mg/l-SO_4	200	99	98	62	>98	97
Chloride mg/l-Cl	65	69	64	128	95	23
Calcium mg/l-Ca	125	100	98			
Total hardness mg/l-$CaCO_3$				177	99	80
Total alkalinity mg/l-$CaCO_3$				226	94	62
COD mg/l	95	98	94	55.5	97	77
TOC mg/l						
Operating Pressure psi		750-1500			200	
Wastewater Feed	Filtered municipal secondary effluent			Chlorinated and diatomaceous-earth-filtered secondary effluent		
Remarks	Lab.-scale RO units.Results from 2 tests are given.Feed pH is adjusted at 5,prod. pH are 6.3 & 5.5 from tests 1 & 2 respectively.			Lab.-scale units.Prod.water after 150 hr of operation. Feed pH=7.2,prod.pH are 6.4 & 7.0 from tests 1 & 2 resp. Water perm.constants are 0.9 $x10^{-5}$ and $3.5x10^{-5}$gm/sq.cm sec.atm for the low & high permeability membranes used in tests 1 & 2 respectively.		

TABLE 2 Continued

	(Ref. 36)		(Ref. 47)	
	Feed Conc.	% removal	Feed Conc.	% remcval
Contaminants				
TDS mg/l	565	89.7	623	88
Conductivity µmho	825	89.1		
Ammonia mg/l-N	8.0	81.3	9.2	82
Nitrates mg/l-N	8.9	67.4	2.4	67
Phosphate mg/l	8.5d	99.6	30.9c	98.2
Sulphate mg/l-SO_4				
Chloride mg/l-Cl				
Calcium mg/l-Ca				
Total hardness mg/l-$CaCO_3$				
Total alkalinity mg/l-$CaCO_3$				
COD mg/l	10.3	97.1	10.8	84
TOC mg/l				
Operating Pressure psi	400		400	
Wastewater Feed	Activated sludge effluent after activated carbon treatment.		Activated sludge effluent after activated carbon treatment.	
Remarks	5,000 gpd RO pilot plant.		5,000 gpd RO pilot plant. Feed is acidified to a pH ≈ 5. 80-85% product recovery.	

TABLE 2 Continued

	(Ref. 15 & 16)		(Ref. 16)	
	Feed Conc.	% removal	Feed Conc.	% removal
Contaminants				
TDS mg/l				
Conductivity µmho	535	84		
Ammonia mg/l-N	9.9	96.1	18.2	98.4
Nitrates mg/l-N	8.2	53.2		
Phosphate mg/l	20.4^c	97.4	14.5^c	93.1
Sulphate mg/l-SO_4	21	90.5	19.0	90.0
Chloride mg/l-Cl	21.5	81.4	23.0	89.1
Calcium mg/l-Ca	21.1	92.4	24.0	>99.7
Total hardness mg/l-$CaCO_3$	124	87.1	125	94.9
Total alkalinity mg/l-$CaCO_3$	164	83.5	161	84.5
COD mg/l				
TOC mg/l				
Operating Pressure psi	1500		1500	
Wastewater Feed	Trickling filter effluent.		Activated sludge effluent.	
Remarks	Laboratory-scale RO unit. Feed pH=8.0 and prod. pH=7.5 product rate=20 gpd/sq.ft		Laboratory-scale RO unit. Feed pH=7.5, prod. pH=7.2.	

TABLE 2 Continued

	(Ref. 9)		(Ref. 4)	
	Feed Conc.	% removal	Feed Conc.	% removal
Contaminants				
TDS mg/l	570	89	585	96
Conductivity mho	945	85	720	96
Ammonia mg/l-N	14.0	86	9.7	86
Nitrates mg/l-N	1.8	44		
Phosphate mg/l	0.31[e]	87	1[e]	90
Sulphate mg/l-SO_4	65	97	54	98
Chloride mg/l-Cl	85	81	84	90
Calcium mg/l-Ca	50	98	73	98
Total hardness mg/l-$CaCO_3$	193	94	95	96
Total alkalinity mg/l-$CaCO_3$				
COD mg/l	276	98		
TOC mg/l	84	98	66	82
Operating Pressure psi	600		600-800[f]	
Wastewater Feed	Raw settled sewage after chemical coagulation, aeration, and neutralization.		Chemically clarified and aerated domestic sewage.	
Remarks	Laboratory-scale-RO unit. 80% product recovery. Feed pH=6.6, prod. pH=7.5		Commercial RO system 95% product recovery. Feed pH=6.0, prod. pH=6.1. Product rate[f] = 6 gpd/sq.ft.	

[a] Expressed as NH_4

[b] Expressed as NO_3

[c] Expressed as PO_4

[d] Expressed as P

[e] Soluble phosphate as P

[f] Figures presented in the 21st Canadian Chem. Engg. Conf.

of municipal wastewaters. However, reverse osmosis has also been studied and successfully used for the treatment, recovery of valuable by-products, and reuse of some industrial wastewaters. Golomb and Besik (12) in a review article discussed basic reverse osmosis designs and assessed the feasibility of employing reverse osmosis for treatment of spent liquors from the pulping industry, acid mine drainage waters, petrochemical complex wastewater, electroplating waste streams, cheese whey waste streams, corn processing effluents, and high-salinity irrigation return flows. A similar study was conducted by Channabasappa (6) who provided a general description of the reverse osmosis process for water reuse application and included several examples and economic analysis of the application of this process for wastewater renovation.

The literature has presented considerable data on removal of different wastewater pollutants by the reverse osmosis process but in many instances the emphasis was mainly on the rejection capability of cellulose acetate membranes without appreciable regard to the productivity of such membranes. Moreover, in most cases, the reverse osmosis membranes were not even specified so that the results can be used in scaling up the process or comparing the performance of different types of membranes.

Regarding the applicability of the reverse osmosis process to wastewater treatment, it can be seen that the low flux rates from the available reverse osmosis membranes as well as the compaction and fouling of these membranes are the interrelated major problems involved, accordingly more investigations should be directed toward solving these problems.

A possible solution to the low productivity of reverse osmosis membranes lies in the development and utilization of more productive-low pressure operated membranes, also required for the economic application of the reverse osmosis process. Because of the low concentrations of wastewater constituents, the osmotic pressures of many wastewaters are usually small. Consequently, the treatment of such wastewaters using reverse osmosis operation at low pressures is of particular interest. Since the recently developed Batch 316 porous cellulose acetate membranes have shown higher productivity in low pressure reverse osmosis desalination than any porous cellulose acetate membranes developed earlier, it seems reasonable to study the applicability of the Batch 316 type membranes to the low pressure reverse osmosis treatment of wastewaters. In this regard both membrane flux and membrane efficiency of separating wastewater contaminants are of major concern.

2.4 Purpose and Scope

The purpose of this study was to investigate the performance of the recently developed Batch 316 porous cellulose acetate membranes for the removal of organics from wastewater by the reverse osmosis process operated at low pressure.

Since the performance of membranes is affected by the character of their porous structure, the performance of six membranes with different surface pore structures (shrunk at different temperatures) was investigated. Also, in order to get firm significance of the performance data, the membranes used were first specified for the reference solution system sodium chloride-water and the reverse osmosis system performance was predicted for process design applications using the specified membranes. The applicability of the above membranes to wastewater treatment was tested and emphasis was mainly on the performance of membranes for the removal of organic compounds from wastewater rather than the inorganics because the latter have been studied for the use of reverse osmosis as a demineralization process using the Batch 316 type membranes.

Laboratory investigations were carried out using the flow-type reverse osmosis cells. The above six membranes (Batch 316 type) were used throughout the study and were tested at the same time in each experimental run. Essentially, the experimental programme consisted of three phases. The first phase was conducted using sodium chloride feed solutions. Eight wastewater organic contaminants were individually present in aqueous feed solutions to the reverse osmosis experimental cell during the second phase of the study. In the third phase chlorinated primary municipal wastewater effluent was fed to the reverse osmosis cell.

CHAPTER 3

EXPERIMENTAL DETAILS

The laboratory investigations of reverse osmosis process were carried out using Batch 316 porous cellulose acetate membranes and flow type cells. The films, (casting details are given in Table 3), were made in the laboratory in accordance with the general Loeb-Sourirajan method (29). Separation of sodium chloride, eight single organic compounds, and primary sewage effluent constituents were studied. In each experiment, pure water permeation rate, membrane permeated product rate, and solute separation were determined.

3.1 Apparatus

The main part of the apparatus used, the flow-type reverse osmosis cell (Figure 3-A), was made of Type 310 stainless steel and consisted of two detachable parts. The upper part was a high pressure chamber provided with inlet and outlet openings for the flow of the feed solution under pressure. The lower part was the membrane stand provided with an outlet opening through which the membrane-permeated product solution left. The effective area of the membrane in the cell was 7.6 cm^2. The wet preshrunk membrane was mounted on a stainless steel porous plate embedded in the lower part of the cell. Between the membrane and the porous plate was placed a wet Whatman filter paper to protect the membrane from

TABLE 3

FILM CASTING DETAILS

Film type: CA-NRC-316 (or simply, Batch 316)

Casting solution composition, weight %

Cellulose acetate (Eastman 398-3):	17
Acetone	69.2
Magnesium perchlorate	1.45
Water	12.35

Temperature of casting solution: 0°C

Temperature of casting atmosphere: 24°C

Casting atmosphere: ambient air in contact with aqueous solution of 30 wt.%acetone

Solvent evaporation time: 6 minutes

Duration of film-setting in ice cold water: >1 hr

Nominal film thickness: 0.005 inch

abrasion and also aid the flow of the product solution through the porous plate. The porous plate and the Whatman filter paper offered practically no resistance to fluid flow under operating conditions. The upper and lower parts of the cell were set in proper alignment with rubber O-ring contacts between the high pressure chamber and the wet membrane. A pressure-tight joint was obtained by clamping the cell tightly between two thick end-plates.

The flow diagram of the experimental set-up (Figures 3-B and 3-C) consisted of the following:

a. Motor-driven controlled-capacity diaphragm duplex pump which was used to pump the feed solution under pressure through the reverse osmosis cell.

b. Surge tank to minimize the pressure fluctuations in the cell.

c. Reverse osmosis cell as previously described.

d. Liquid-sealed Ashcroft pressure gauge to indicate the fluid pressure under operating conditions.

e. Purge valve to drain the system whenever needed.

f. Stainless steel Grove pressure regulator to maintain a constant operating pressure in the cell.

g. Commercial gas cylinder to supply nitrogen gas under pressure which was used to load the dome of the Grove regulator.

Monel metal high pressure tubings and HIP high pressure fittings made of Type 316 Stainless Steel were used throughout the system.

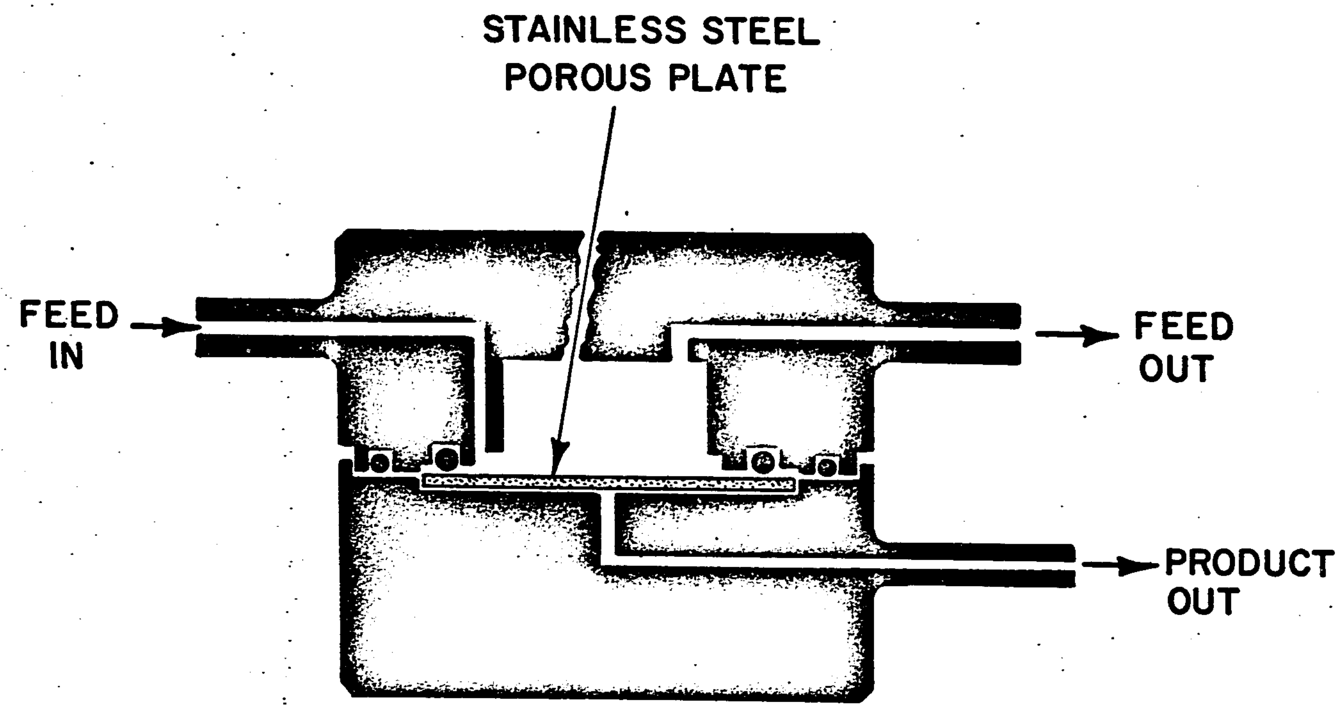


FIGURE 3-A REVERSE OSMOSIS CELL

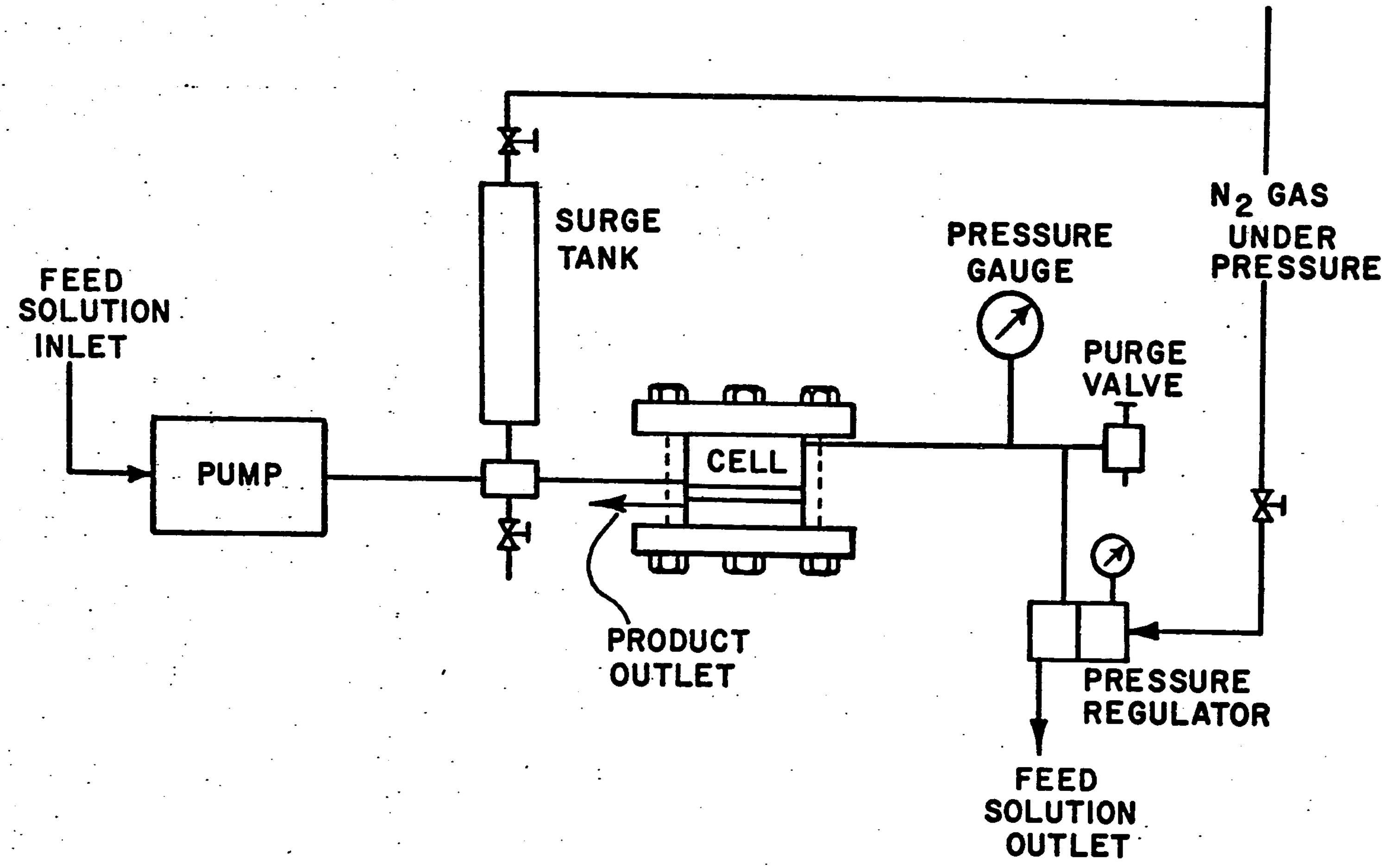


FIGURE 3-B FLOW DIAGRAM OF APPARATUS

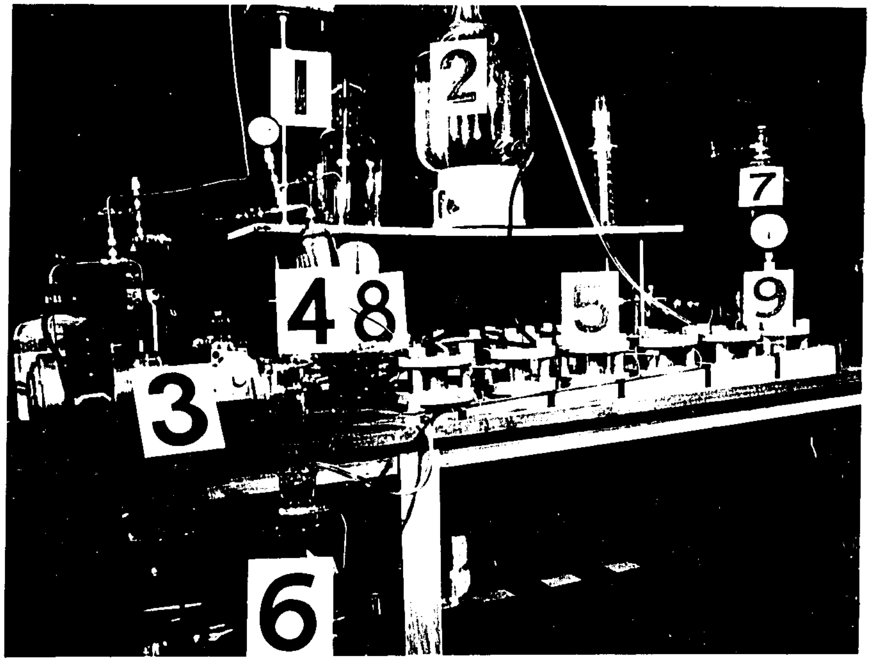


FIGURE 3-C EXPERIMENTAL SET-UP - PHOTOGRAPH

1. Distilled Water Line
2. Feed Water Container
3. Motor Driven Pump
4. Surge Tank
5. Reverse Osmosis Cells
6. Product Water Container
7. Nitrogen Gas Cylinder
8. Inlet Pressure Gauge
9. Outlet Pressure Gauge

3.2 Experimental Procedure

Six reverse osmosis cells were used in series as shown in Figure 3-C (only one cell is illustrated in Figure 3-A) so that six different membranes could be tested concurrently. Membranes preshrunk at different temperatures were used to give different surface pore structures and hence different levels of solute separation at a given set of operating conditions. Preliminary experimentation indicated that sufficient variation in surface pore structure was affected if the shrinking temperatures for the temperature treatment of membranes were 81, 77, 74, 72, 70 and 68°C; the corresponding treated membranes were therefore designated as films No. 1, 2, 3, 4, 5 and 6 respectively. The temperature treatment consisted of heating the film gradually in water from the laboratory temperature to the required temperature where it was kept for 10 minutes and subsequently cooled rapidly.

The shrunk membranes were initially subjected to a pressure treatment at 350 psig with pure water for 2 hours before subsequent use in reverse osmosis experiments at operating pressures of 300 psig or less.

All reverse osmosis experiments were carried out at the laboratory temperature ($\sim 24^{\circ}C$). Unless otherwise stated, the experiments were of the short-run type, each lasting for about 2 hours. In each experiment, the pure water permeating rate (PWP) and membrane permeated product rate (PR) in grams per hour per 7.6 cm^2 of effective film area, and solute separation (f) were determined at the preset operating conditions of pressure, feed concentration and feed flow rate. The reported product rates are those corrected to 25°C using the relative viscosity and density data for pure water. In all experiments, the term "product" and "product rate" refer to membrane permeated solutions. The accuracy of the product rate data is within 3% in all cases.

Because of the low solute concentrations involved, the data on solute separation, f, were calculated from the relation

$$f = \frac{\text{solute mg/l in feed} - \text{solute mg/l in product}}{\text{solute mg/l in feed}} \qquad (3.1)$$

The accuracy of the separation data is within 1% and 2% respectively for feed concentrations containing above and below 100 mg/l of solute.

3.3 Experimental Programme

The experimental programme was conducted through three main sets of experiments designed to study the following characteristics.

1. Specification and performance of the Batch-316-type porous cellulose acetate membranes which was intended to be used throughout the experimental study.

2. Reverse osmosis removal of some organic wastewater constituents individually present in aqueous solutions using the previously specified membranes.

3. Removal of organic compounds from actual, primary treated, chlorinated sewage by the same membranes in the reverse osmosis treatment.

3.3.1 Specifications of Membranes and their Performance

Using the feed system sodium chloride-water, experiments were carried out according to the scheme presented in Table 4.

The experimental (PWP), (PR), and f data obtained were used as follows:

1. To calculate the values of the pure water permeability constant, A, solute transport parameter, $(D_{AM}/K\delta)$ for NaCl, and the mass transfer coefficient, k. At a given operating pressure, the values of A, $(D_{AM}/K\delta)$, for sodium chloride (as a reference solute) specify a membrane, while k, on the high pressure side of the membrane for sodium chloride-water system (as a reference solution), specifies the feed flow condition used. It is necessary to provide such specifications for membranes and feed flow conditions used, in terms of parameters easily obtainable for a convenient reference feed solution system, so that we would be able to compare the relative performance of different membranes for feed solution systems for which osmotic pressure and other physico-chemical data are not readily available.

2. To show that the parameter $(D_{AM}/K\delta)$ is independent of feed concentration and feed flow rate. Hence, the

TABLE 4

Experimental Programme-Sodium Chloride Experiments

Experiment No.	Solute Concentration ppm NaCl	Feed Rate cm^3/min	Operating Pressure psig	Remarks
1	1500	400	300	Effect
2	2000	400	300	of
3	2500	400	300	Solute
4	3500	400	300	Concentration
5	3500	200	300	Effect
6	3500	300	300	of
7	3500	400	300	Feed
8	3500	500	300	Rate
9	1500	400	50	Effect
10	1500	400	100	of
11	1500	400	200	Operating
12	1500	400	250	Pressure
13	1500	400	300	

membrane specifications as obtained from the above analysis can be used for reverse osmosis system specification and process design following the general equations of Ohya and Sourirajan (33).

3. To illustrate the productivity levels of the membranes used in this study.

3.3.2 Removal of Single Organic Compounds

Since wastewaters and treated effluents are heterogenous mixtures, it is difficult to ensure the uniformity of their characteristics. In this investigation, a study was conducted to determine the permeation through reverse osmosis membranes of aqueous solutions of specific organic compounds most frequently found in abundance in wastewaters (1,18,19,21, 35). At least three different concentrations of each organic compound in aqueous solution were studied. The span of concentration selected for use was within the range of most wastewaters (1,18,19,21,35). Also, for each organic compound studied, this concentration span covered a wide range in which the average concentration value for a certain organic compound as present in wastewater treated effluents and a fold increase in concentration of ten times the average value (which is expected after 90% recovery of feed solution as product water), were both included.

Since proteins, carbohydrates and fats are the major organic groups present in sewage, six organic compounds were chosen to represent them in this study. Glucose (50 to 1000 mg/l), sucrose (50 to 2000 mg/l), soluble starch (10 to 100 mg/l), (Difco) beef extract (100 to 2000 mg/l), glutamic acid (10 to 550 mg/l), and sodium stearate (60 to 1000 mg/l) were used in aqueous feed solutions in the concentration ranges indicated to represent respectively monosaccharide, disaccharide, polysaccharide, complex protein, amino acid and salt of fatty acid. Since synthetic detergents are also abundant in many wastewaters, two surfactants; alkyl benzene sulfonate (5 to 100 mg/l ABS) and linear alkylate sulfonate (2.5 to 100 mg/l LAS) were used in aqueous feed solutions in the concentration ranges indicated. Because of the low concentrations of solutes, the corresponding osmotic pressures are small and the reverse osmosis treatment operated at low pressures is of practical interest. In this particular study, the operating pressure was 300 psig and the feed flow rate was 400 cm^3/min.

3.3.3 Removal of Organics from Wastewater

Since the detailed composition of municipal wastewaters generally is not known, the study of the reverse osmosis treatment of actual wastewaters is important to evaluate the effects of various components in such waters on the performance of the cellulose acetate membranes used in this treatment. For this purpose, reverse osmosis laboratory experiments were carried out using chlorinated wastewater effluent obtained from the Ottawa City primary sewage treatment plant. A grab sample was taken at 10 a.m. which, according to plant record, would give an indication of conditions that occur when the plant is receiving its maximum hourly load. Before applying it to the reverse osmosis apparatus, the wastewater sample was filtered through a coarse filter paper (Whatman No. 41H) using a vacuum pump to remove big suspended particles, after which pH was adjusted from 7.6 to 6.2 (to minimize the hydrolysis of membranes) using 1.2 ml of conc. NaOH solution per 20 liters of wastewater. Two reverse osmosis experiments were carried out using the wastewater feed. The first experiment was of the short-run type lasting for about 1 hour. The second one was run continuously until 90% of the feed water was recovered. In the latter experiment, permeated waters from the six membranes were combined together while in the first experiment permeated waters were collected separately from each membrane to evaluate PWP, PR and percentage organic material (measured as TOC) removed. In the continuous experiment, the average product rate from the six membranes was measured every 24 hrs while the product water collected throughout the whole run was analyzed to evaluate the average removal of the different wastewater constituents throughout the experiment.

3.4 Analytical Techniques

The concentrations of the feed and product solutions were measured in each case using the most suitable analytical technique. The analytical methods used in this study were selected on the basis of sensitivity at low concentrations, speed and ease of operation, and previous use in water and wastewater analyses. Measurements, at very low concentrations only, were carried out in duplicate or triplicate. The analytical techniques fall into three categories: those used to analyze the sodium chloride solutions; those used in the analysis of the organic solutions; those used in the analysis of the wastewater constituents. They will be outlined in this order.

3.4.1 Sodium Chloride Measurement

The concentration of sodium chloride was determined by specific resistance measurements using a conductivity bridge (Beckman Model RC 16B2).

3.4.2 Organic Solutions Analyses

These were carried out as follows. Carbohydrates, glutamic acid and anionic surfactants were determined spectrophotometrically using phenol-sulphuric acid (2,7), ninhydrin (38), and methylene blue (45) respectively as color developing agents. Coleman Junior Spectrophotometer (Model 6B) was used in these analyses. Nitrogen content of protein solutions were determined by the Kjeldahl method (45). Organic carbon content of sodium stearate solutions were measured using the Beckman Total Carbon Analyzer (Model 915).

3.4.2.1 Carbohydrate Determination

The concentration of carbohydrate in solution was measured using the phenol-sulphuric acid method (2,7). This method was followed in glucose, sucrose, and starch determinations.

With the phenol-sulphuric acid method, the sample to be analyzed is reacted with sulphuric acid and an 80% phenol solution. Under the acid conditions carbohydrates are converted to furfurals which in turn react with the phenol, yielding a characteristic stable yellow color, the intensity of which is proportional to the concentration of carbohydrate in the sample. The density of the color that developed was measured by making spectrophotometric readings at a wavelength of 485 mμ; and later converted to carbohydrate concentration.

3.4.2.2 Glutamic Acid Measurement

The amino acids in general react with the compound ninhydrin (triketohydrindene hydrate) $C_6H_4.(CO)_3.H_2O$. This reaction has been made the basis for several different types of quantitative methods for the determination of amino acids. One of these quantitative methods is based upon the color change which results from the reaction of ninhydrin with NH_2 (amino) groups to give diketohydrindylidene-diketohydrindamine.

A modified ninhydrin colorimetric method was followed as described by Rosen (38) for glutamic acid determination.

In this method, the sample to be analyzed is reacted with 3% ninhydrin solution in Methyl Cellosolve (ethylene glycol monomethyl ether) in the presence of cyanide-acetate buffer which maintains the reaction mixture at pH 5.0-5.2. After heating this mixture for 15 min. at 100°C, isopropyl alcohol-water dilutent is immediately added followed by vigorous shaking and cooling to room temperature. The intensity of the color that developed is proportional to the concentration of glutamic acid in the sample. This intensity was measured at 570 mμ using spectrophotometric readings, and later converted to glutamic acid concentration.

3.4.2.3 Anionic Surfactant Determination

The methylene blue method was used in the measurements of both alkyl benzene sulfonate (ABS) and linear alkylate sulfonate (LAS) surface active agents and was followed according to the standard procedures (45). This method depends on the formation of a blue-colored salt when methylene blue reacts with anionic surface active agents (surfactants). The salt is soluble in chloroform, and the intensity of color is proportional to the concentration of the surfactant used. The intensity is measured by making spectrophotometric readings in this solvent at a wavelength of 652 mμ, and later converted to surfactant concentration.

3.4.2.4 Protein Measurement

The elemental content of proteins include carbon, hydrogen, oxygen, nitrogen, and sometimes sulphur and phosphorus in the following approximate percentages: C-50 to 55, H-6 to 8, O-20 to 23, N-15 to 18, S-0 to 2.5, P-0 to 1.0. The protein content of a waste sample can therefore be estimated using any of the elemental percentages. However, since the average nitrogen content of most proteins is known to be 16 percent, the organic nitrogen content as determined using the Kjeldahl nitrogen method (45) was chosen as a measure of the protein content of beef extract solutions used in this study. If required, the protein content can be estimated by multiplying the Kjeldahl organic nitrogen value by 100/16, or 6.25. This procedure is often used by sanitary engineers to estimate the protein content of wastewaters and of sludges (39).

Through digestion with a sulphuric acid solution, the Kjeldahl nitrogen method converts organically bound nitrogen to ammonium bisulphate. The ammonium bisulphate is then made alkaline and the resulting ammonia is distilled into boric acid. The ammonium borate is then titrated with sulphuric acid, and the nitrogen content of the sample under consideration is determined using the value of this titration.

3.4.2.5 Organic Carbon Content Determination

One measurement of organic material that has proved to be convenient is the organic carbon content. The total organic carbon (TOC) content of sodium stearate solutions used in this study were determined using the Carbonaceous Analyzer Technique (49) which employs a dry oxidation system for this determination. The TOC determination has great accuracy due to the sensitivity of the measuring equipment and the very high degree of carbon oxidation that takes place.

Knowing the percentage carbon content of sodium stearate to be 68.3, the TOC values were converted to the corresponding sodium stearate concentrations.

3.4.3 Wastewater Analysis

In the reverse osmosis experiments using municipal wastewater, the compositions of the wastewater feed and the renovated water product were analyzed as follows. Total dissolved solids were determined by evaporation, chemical oxygen demand (COD) was determined using the dichromate reflux method, ammonia nitrogen was determined spectrophotometrically using the distillation method, total Kjeldahl nitrogen was measured using the Kjeldahl method and nitrate nitrogen was determined by the brucine method, according to the "Standard Methods" (45) given for each. Total organic carbon (TOC) content was measured using the Carbonaceous Analyzer Technique (49). Specific conductance measurements were carried out using a conductivity bridge. Orthophosate was determined using the FWPCA modified method (10). This method is based on reactions that are specific for the orthophosphate ion. The total phosphate content in water samples was determined following this method after treating the samples with strong acid solution and ammonium persulfate to convert available phosphorus into the orthophosphate form. In the FWPCA method, a solution of phosphorus is reacted with ammonium molybdate and potassium antimonyl tartrate in an acid medium to form an antimony-phosphate-molybdate complex. This complex is reduced to an intensely blue-colored complex by ascorbic acid. The intensity of the developed color is proportional to the phosphorus concentration. This intensity was measured spectrophotometrically at 880 mμ, and later converted to phosphorus concentration which was reported as P, mg/l. It should be pointed out that the phosphorus concentration expressed as (P), when multiplied by a factor of 3.1, gives the corresponding phosphorus content expressed as (PO_4).

CHAPTER 4

RESULTS AND ANALYSIS OF REVERSE OSMOSIS OPERATION WITH SODIUM CHLORIDE SOLUTIONS

Reverse osmosis short-run experiments using laboratory scale flow type cell and sodium chloride aqueous feed solutions were carried out following the scheme presented in Table 4. In each experiment, the pure water permeation rate (PWP) and membrane permeated product rate (PR) in gms per hour per given area of film surface (=7.6 sq cm in the cell used), and solute separation (f) were determined at the preset operating conditions for each of the six films used throughout the study. The experimental (PWP), (PR), and (f) data obtained were used, following the Kimura-Sourirajan analysis (discussed in Chapter 2), in specifying the membranes and the feed conditions used (for the reference solution system sodium chloride-water). The general equations of Ohya and Sourirajan (discussed in Chapter 2) were also used to specify and study the performance of reverse osmosis systems employing three of the previously specified membranes chosen to obtain a wide variation in system performance; and the dimensionless parameter $\lambda\theta$ was used as a design factor. An example was given to demonstrate the reverse osmosis process design applications in which both the membrane area and work requirements were considered.

4.1 Specifications of Membranes and Feed Flow Conditions

For each membrane used in this study, using the experimental (PWP), (PR), and (f) data obtained at preset operating conditions for the reference solution system sodium chloride-water containing 1500 to 3500 mg/l of salt and using the physical properties of the feed solution, the values of A, $(D_{AM}/K\delta)$ for NaCl, and k were calculated from Equations 2.1, 2.2, 2.3, and 2.4 assuming constant molar density (c) for feed and product solutions. The results of the computer solutions of these equations are given, for each membrane, in Appendix 1 along with the corresponding (PWP), (PR), and (f) data. These results were used in the reverse osmosis data analysis and correlation.

Table 5 presents some of the above data on membrane performance and membrane specification. Using a solution of 3500 mg/l sodium chloride operated at 300 psig, the solute separations obtained with films 1, 2, 3, 4, 5, and 6 were 97.6, 95.6, 93.7, 91.0, 88.3, and 87.7% respectively, and the corresponding membrane fluxes were 12.9, 19.5, 25.3, 29.1, 31.7, and 33.9 gpd/sq. ft. respectively. The above membranes were specified in terms of A and $(D_{AM}/K\delta)_{NaCl}$ at 300 psig. For films 1 to 6, the values of A were in the range 1.96×10^{-6} to 5.20×10^{-6} $\frac{\text{g mole H}_2\text{O}}{\text{sq.cm.sec.atm}}$ respectively, and the corresponding $(D_{AM}/K\delta)_{NaCl}$ values were in the range 1.27×10^{-5} to 16.12×10^{-5} cm/sec respectively. The data given in Table 5 illustrate the high productivity at high solute separation levels obtained with the Batch 316 type membranes operated at 300 psig.

TABLE 5

Specifications of Membranes and Some Performance Data[a]

Film No.	Film Shrinkage Temp., °C	$A \times 10^6$ $\frac{\text{g mole } H_2O}{\text{sq cm.sec.atm}}$	$(D_{AM}/K\delta)_{NaCl}$ $\times 10^5$ cm/sec	Feed: 3500 mg/l NaCl-H_2O k: 41×10^{-4} cm/sec	
				Solute Sepn., %	Product rate gpd/sq ft
1	81	1.96	1.27	97.6	12.9
2	77	2.99	3.36	95.6	19.5
3	74	3.88	6.25	93.7	25.3
4	72	4.47	9.96	91.0	29.1
5	70	4.86	14.34	88.3	31.7
6	68	5.20	16.12	87.7	33.9

[a] Film type: CA-NRC-316 (Batch 316)
Operating pressure: 300 psig
System: sodium chloride-water

Unless otherwise stated, the feed flow condition used in this study is that corresponding to a value of $k = 41 \times 10^{-4}$ cm/sec for the reference solution system 3500 mg/l NaCl-H_2O at 300 psig.

4.2 Analysis and Correlation of Reverse Osmosis Experimental Data

The correlations of A, $(D_{AM}/K\delta)$, and k with operating pressure, feed concentration, and feed flow rate are of practical interest from the point of view of specifying membranes and predicting membrane performance under different experimental conditions.

The effect of operating pressure on the pure water permeability constant is illustrated in Figure 4. The straight line plots of log A vs. P, for all films used, indicate that the relationship can be expressed as $A \propto e^{-\alpha P}$; where α is a constant expressing the compression effect on the pure water permeability characteristics of the particular film under pressure. The very small slope of each of the straight lines shown in Figure 4 illustrate the insignificant compression effect on the overall porosity characteristics of the Batch 316 type films under pressure which is an advantageous character of this particular type of membranes compared to the other types of porous cellulose acetate membranes developed earlier (40).

Figures 5 and 6 show that the parameter $(D_{AM}/K\delta)$ is independent of feed rate and feed concentration. The variation

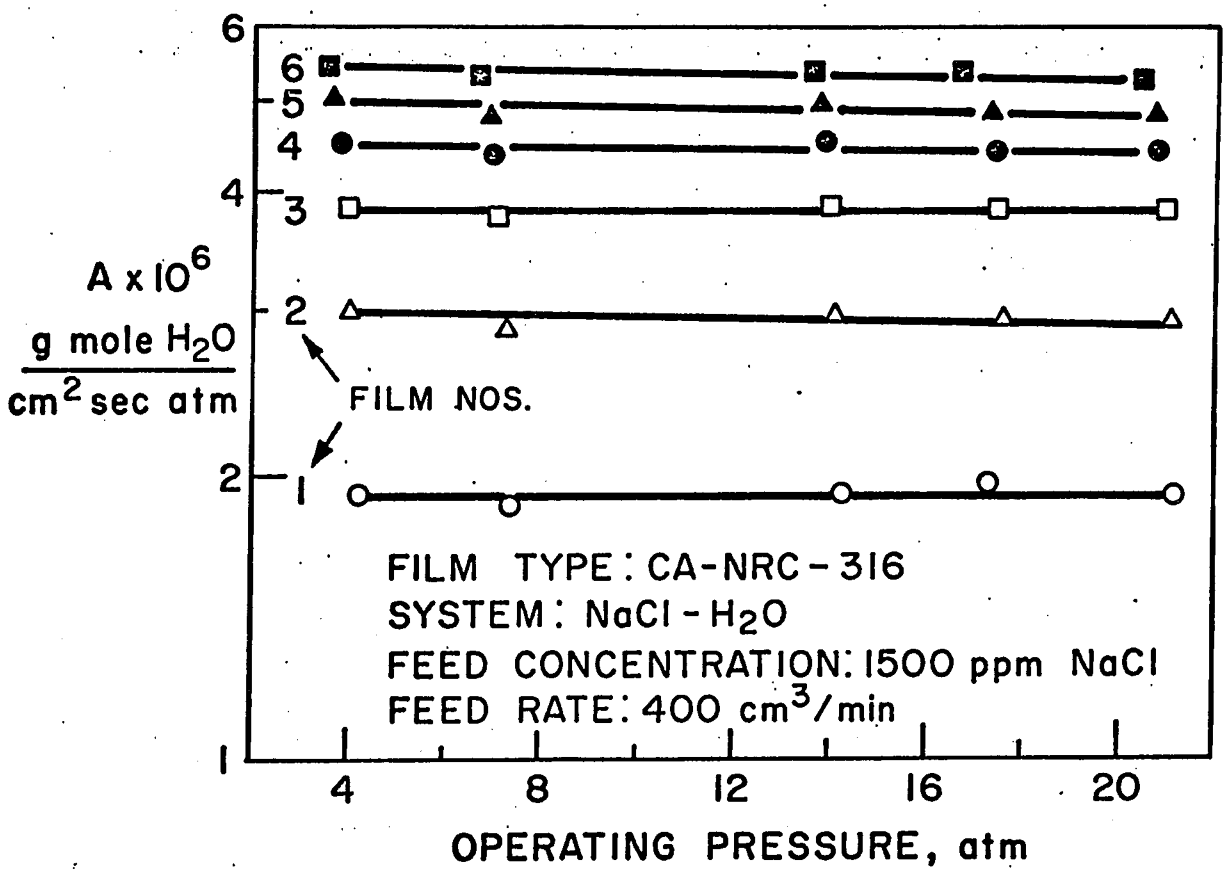


FIGURE 4 EFFECT OF OPERATING PRESSURE ON THE PURE WATER PERMEABILITY CONSTANT

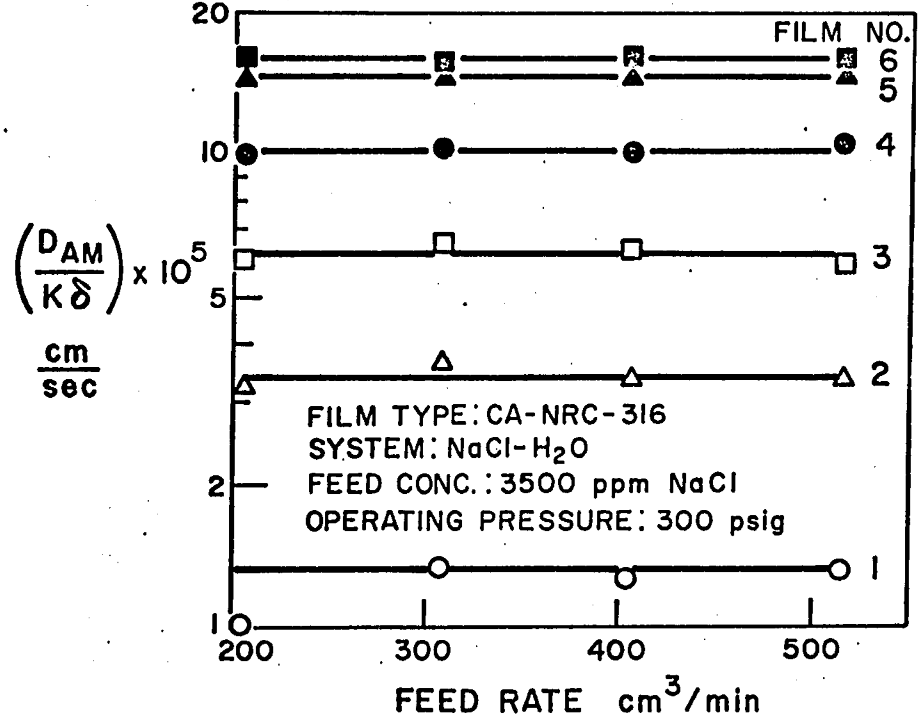


FIGURE 5 EFFECT OF FEED FLOW RATE ON THE SOLUTE TRANSPORT PARAMETER

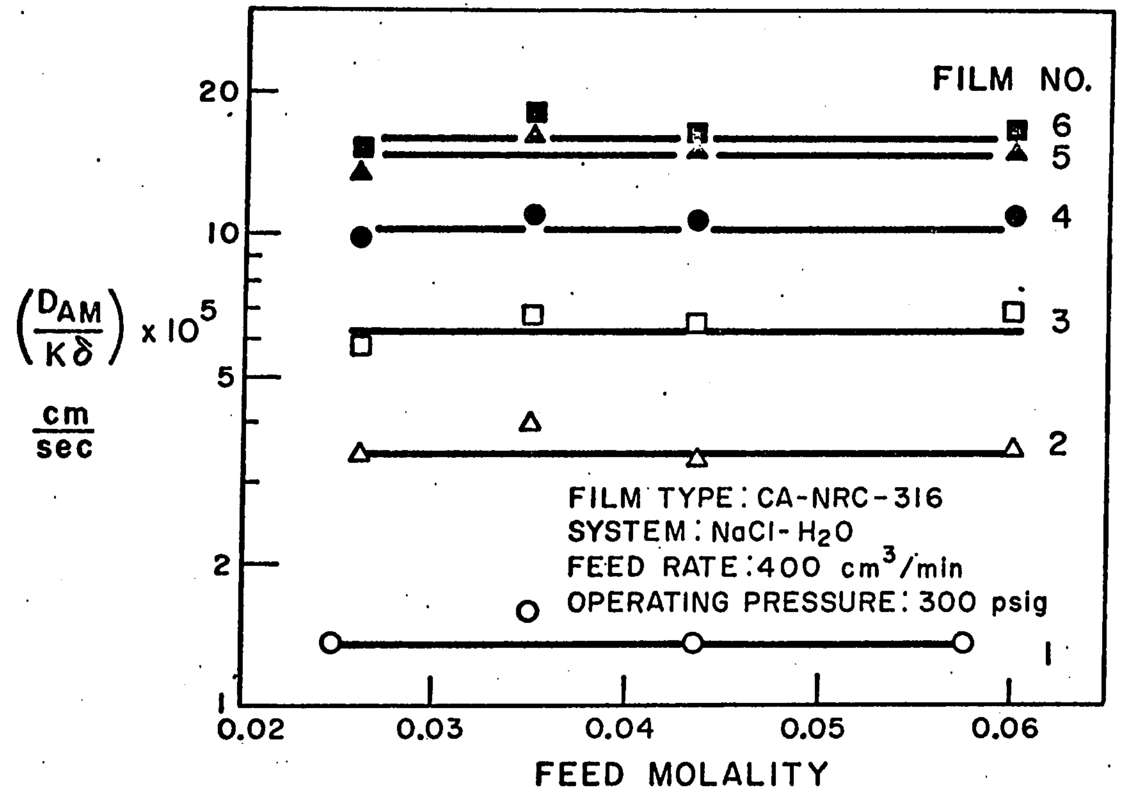


FIGURE 6 EFFECT OF FEED MOLALITY ON THE SOLUTE TRANSPORT PARAMETER

of $(D_{AM}/K\delta)$ with P is shown in Figure 7. The experimental data illustrated in Figure 7 observe general straight-line trends for the variation of $(D_{AM}/K\delta)$ with P exhibited by each film used indicating that such variation can be expressed by the relation $(D_{AM}/K\delta) \propto P^{-\beta}$ reported for other types of membranes developed earlier (discussed in Chapter 2); where β is a constant for the particular membrane for the given solute.

The effect of feed flow rate on k is shown in Figure 8 for solution of 3500 mg/l sodium chloride. The general trend observed by the experimental data in the log-log plot of k vs. feed rate could correlate by a single line all the data illustrated for the six films used with up to 30 percent variation in the value of k observed for the different films at a certain feed rate under the experimental conditions. A 30 percent variation in the experimental values of k does not appear to be too great from the point of view of establishing a practical criterion for calculating mass transfer coefficients (40). The specific correlation given in Figure 8 is for the particular apparatus used. Average values of k can be obtained from such correlation. Also, the effect of feed flow rate on k (Figure 8) can be expressed by the relation $k \propto Q^n$, where Q is the feed rate in cm^3 per second, and n is a constant characteristic of the solution system and the apparatus used.

It should be pointed out that experimental trends similar to those given above have been reported for other types

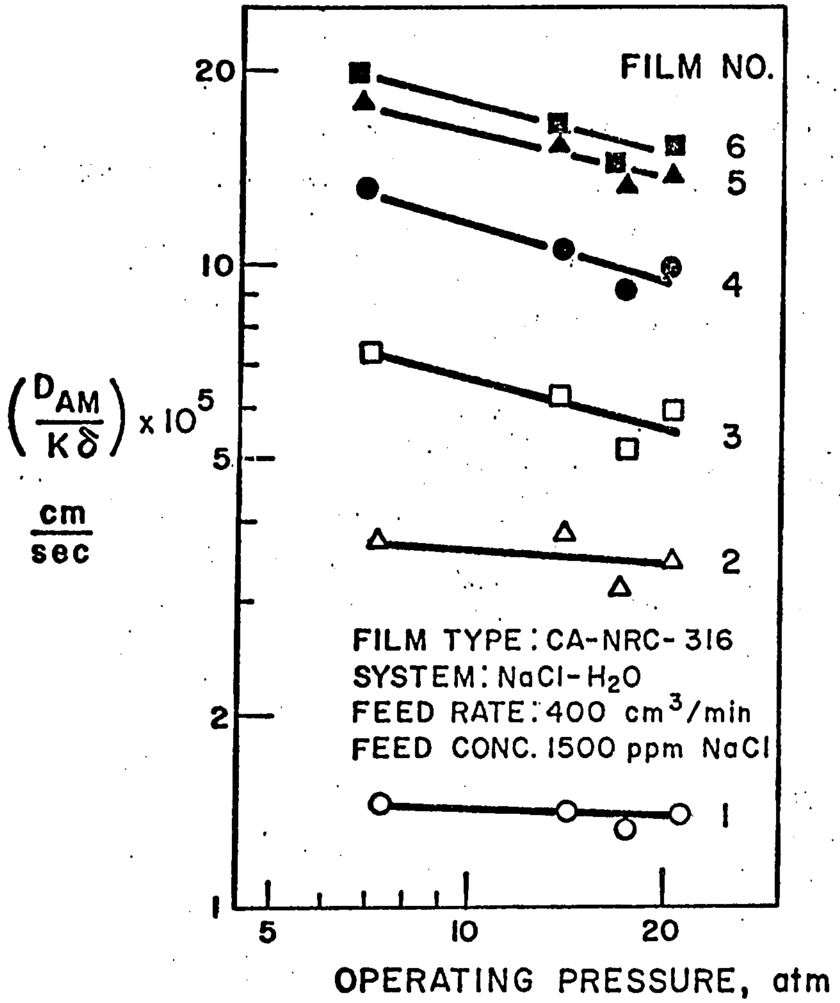


FIGURE 7 EFFECT OF OPERATING PRESSURE ON THE SOLUTE TRANSPORT PARAMETER

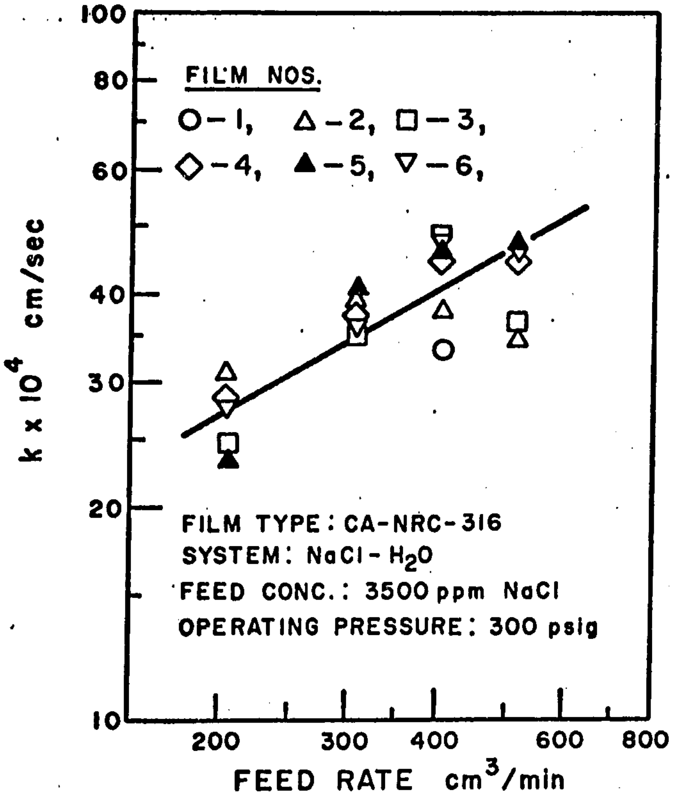


FIGURE 8 EFFECT OF FEED FLOW RATE ON THE MASS TRANSFER COEFFICIENT

of cellulose acetate membranes developed earlier (40). This is an important fact because it indicates that the membrane specifications obtained from the above analysis can be used for reverse osmosis process design, a definite requirement for the use of the general equations of Ohya and Sourirajan (33) as discussed in Chapter 2.

4.3 Performance of Reverse Osmosis System

Using the general equations for reverse osmosis process design (Eqs. 2.20 to 2.27), the performance data were obtained for several systems employing three reverse osmosis membranes (Nos. 1, 3, and 6) operating at three different pressures (100, 200, and 300 psi) for the feed sodium chloride-water system at three different concentrations (1500, 2500, and 3500 ppm). These systems were specified (by substituting in Eqs. 2.8, 2.9 and 2.11) for the different operating conditions as given in Table 6. The performance data, as tabulated in Appendix 2, are the results of computer solutions of the equations 2.20 to 2.27. For each specified reverse osmosis system, the performance data were arranged to give C_1, C_2, C_3, $\overline{C}_3$, τ or X as functions of Δ. The values of Δ were in steps of 0.05. The maximum value of Δ given in each table was determined only by the convenience of computation procedure used in this work and does not represent the maximum attainable value for the system.

TABLE 6

Reverse Osmosis System Specifications[a]

Film No. 1									
NaCl Concentration ppm	1500			2500			3500		
Pressure psi	100	200	300	100	200	300	100	200	300
$(D_{AM}/K\delta)\text{x}10^5$ cm/sec	1.452	1.401	1.340	1.452	1.401	1.340	1.452	1.401	1.340
$v_w^*\text{x}10^5$ cm/sec	24.67	49.66	72.93	24.67	49.66	72.93	24.67	49.66	72.93
γ	0.18	0.09	0.06	0.30	0.15	0.10	0.42	0.21	0.14
θ	0.059	0.028	0.018	0.059	0.028	0.018	0.059	0.028	0.018
$\lambda\theta$	0.5, 1, 2, 5, or ∞			0.5, 1, 2, 5, or ∞			0.5, 1, 2, 5 or ∞		

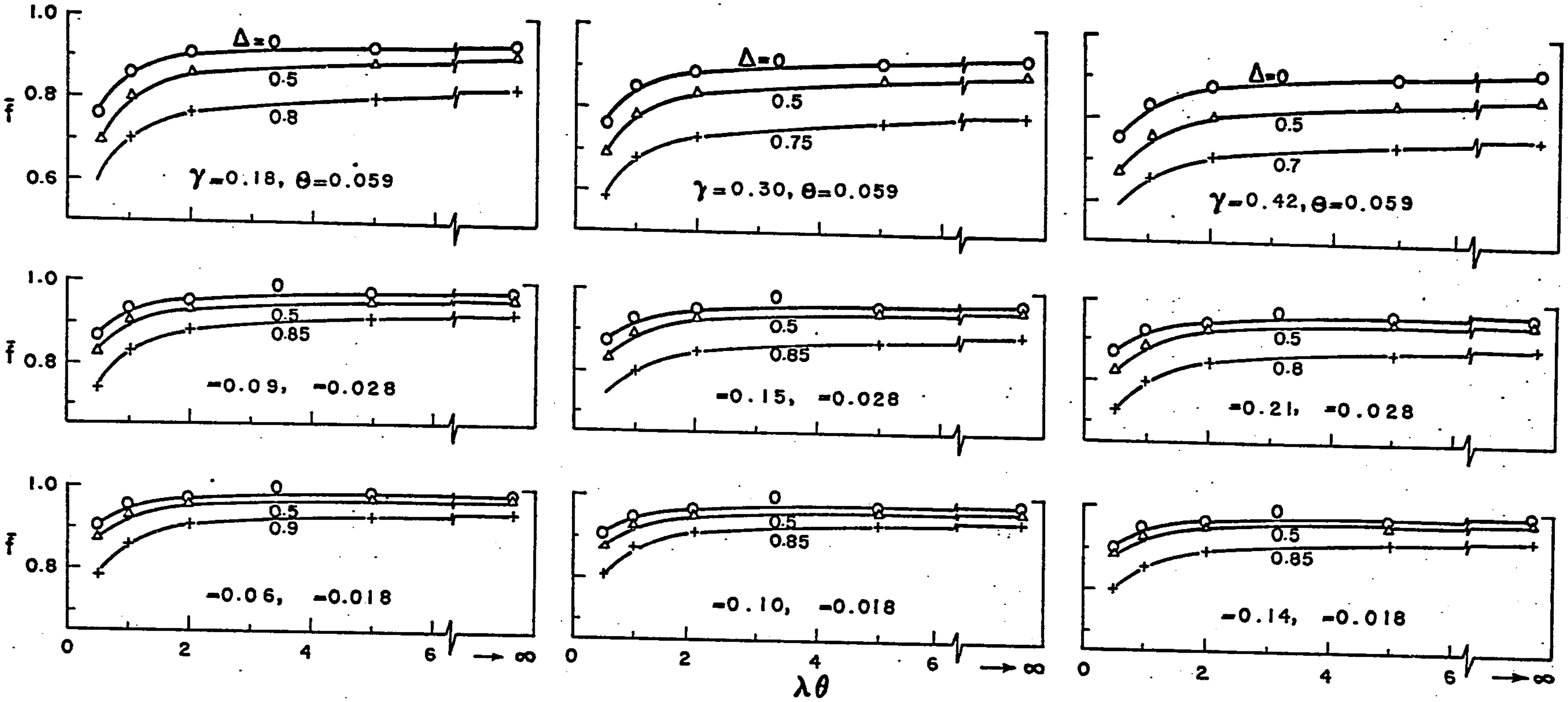
Film No. 3									
NaCl Concentration ppm	1500			2500			3500		
Pressure psi	100	200	300	100	200	300	100	200	300
$(D_{AM}/K\delta)\text{x}10^5$ cm/sec	7.321	6.176	6.358	7.321	6.176	6.358	7.321	6.176	6.358
$v_w^*\text{x}10^5$ cm/sec	48.33	98.47	145.69	48.33	98.47	145.69	48.33	98.47	145.69
γ	0.18	0.09	0.06	0.30	0.15	0.10	0.42	0.21	0.14
θ	0.151	0.063	0.044	0.151	0.063	0.044	0.151	0.063	0.044
$\lambda\theta$	0.5, 1, 2, 5, or ∞			0.5, 1, 2, 5 or ∞			0.5, 1, 2, 5 or ∞		

Film No. 6									
NaCl Concentration ppm	1500			2500			3500		
Pressure psi	100	200	300	100	200	300	100	200	300
$(D_{AM}/K\delta)\text{x}10^5$ cm/sec	19.890	16.530	16.00	19.890	16.530	16.000	19.890	16.530	16.000
$v_w^*\text{x}10^5$ cm/sec	63.84	131.55	194.33	63.84	131.55	194.33	63.84	131.55	194.33
γ	0.18	0.09	0.06	0.30	0.15	0.10	0.42	0.21	0.14
θ	0.312	0.126	0.082	0.312	0.126	0.082	0.312	0.126	0.082
$\lambda\theta$	0.5, 1, 2, 5 or ∞			0.5, 1, 2, 5 or ∞			0.5, 1, 2, 5 or ∞		

[a] Systems are specified in terms of γ, θ, and $\lambda\theta$ (which is the same as specifying γ, θ, and λ for small values of θ used in these systems) since $\lambda\theta$ $(=k/v_w^*)$ is a more useful design parameter than λ.

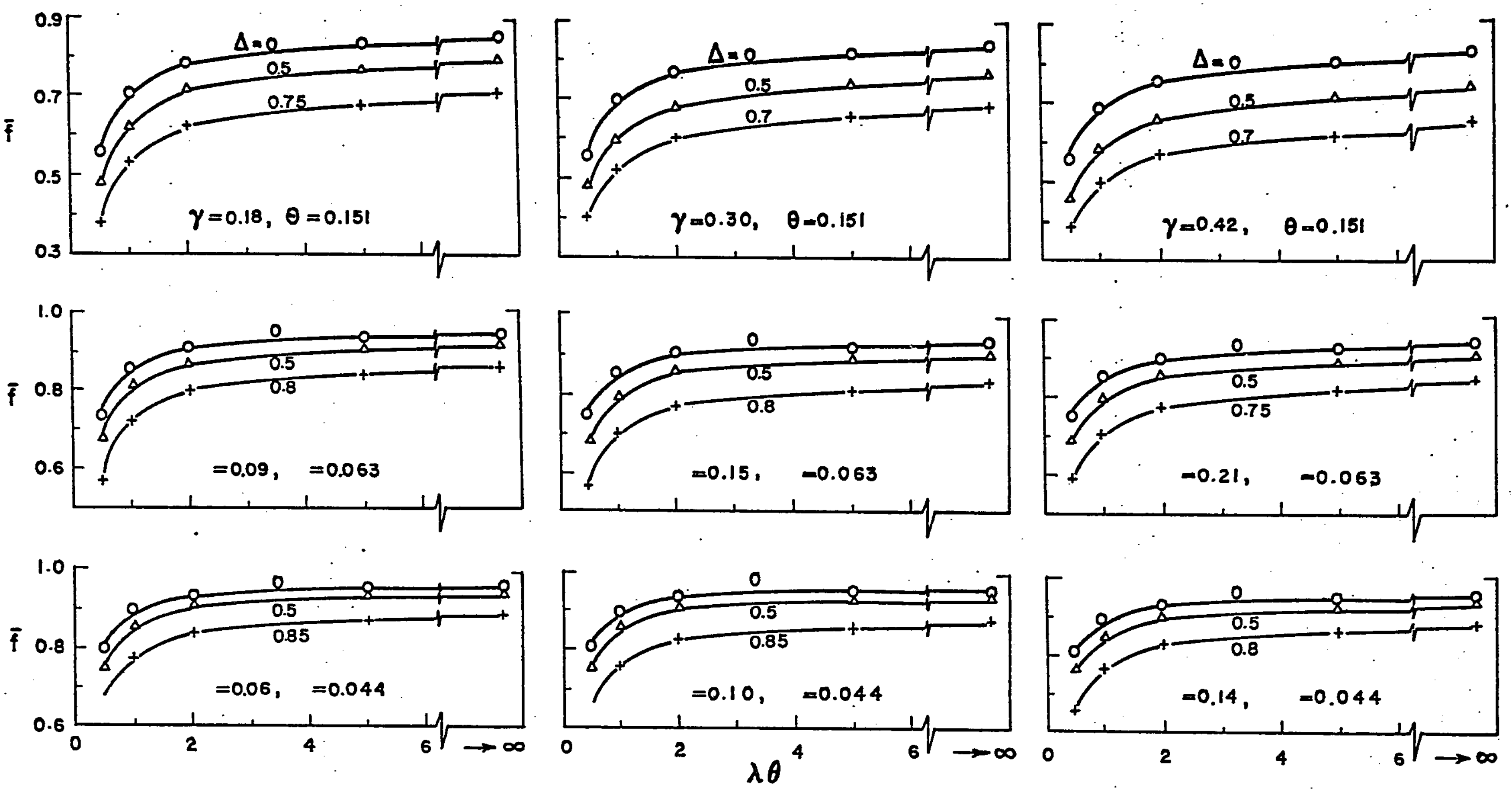
The reverse osmosis systems performance data (Appendix 2) are plotted in Figures 9, 10, and 11 which show the relationships between $\lambda\theta$ and $\overline{f}(=1 - \overline{C}_3)$ for the different combinations of γ and θ values at different values of Δ for the membranes Nos. 1, 3, and 6 respectively ($\overline{f}$ is the average fraction solute separation and Δ is the fraction product recovery). These plots are general, and they describe and predict the performance of any reverse osmosis system specified by the given values of γ, θ, and $\lambda\theta$.

Figures 9, 10, and 11 generally show that, for any specified reverse osmosis system studied at a certain level of product water recovery (Δ), the average solute separation ($\overline{f}$) was increased with both the operating pressure (represented by θ) and the $\lambda\theta$ value, under the specified conditions. This increase was more significant at low operating pressures (from 100 to 200 psi) and/or with $\lambda\theta < 2$. For $\lambda\theta = 5$ up to ∞, in the above specified system, the corresponding $\overline{f}$ values are almost the same at a certain level of product recovery. Meanwhile, for any specified system, the value of $\overline{f}$ was decreased with the increase in the fraction of recovered product (Δ). The latter effect is more significant at the higher levels of product recovered. In the range of solute concentration and operating pressure considered in this study, it was shown (Figures 9 to 11) that the relationship between $\overline{f}$ and $\lambda\theta$, at a given Δ, was more affected by the change in the operating pressure rather than the solute concentration.



FILM TYPE : CA-NRC-316
FILM NO.1
SYSTEM : $NaCl-H_2O$

FIGURE 9 RELATIONSHIP BETWEEN λθ AND AVERAGE SOLUTE SEPARATION-FILM 1

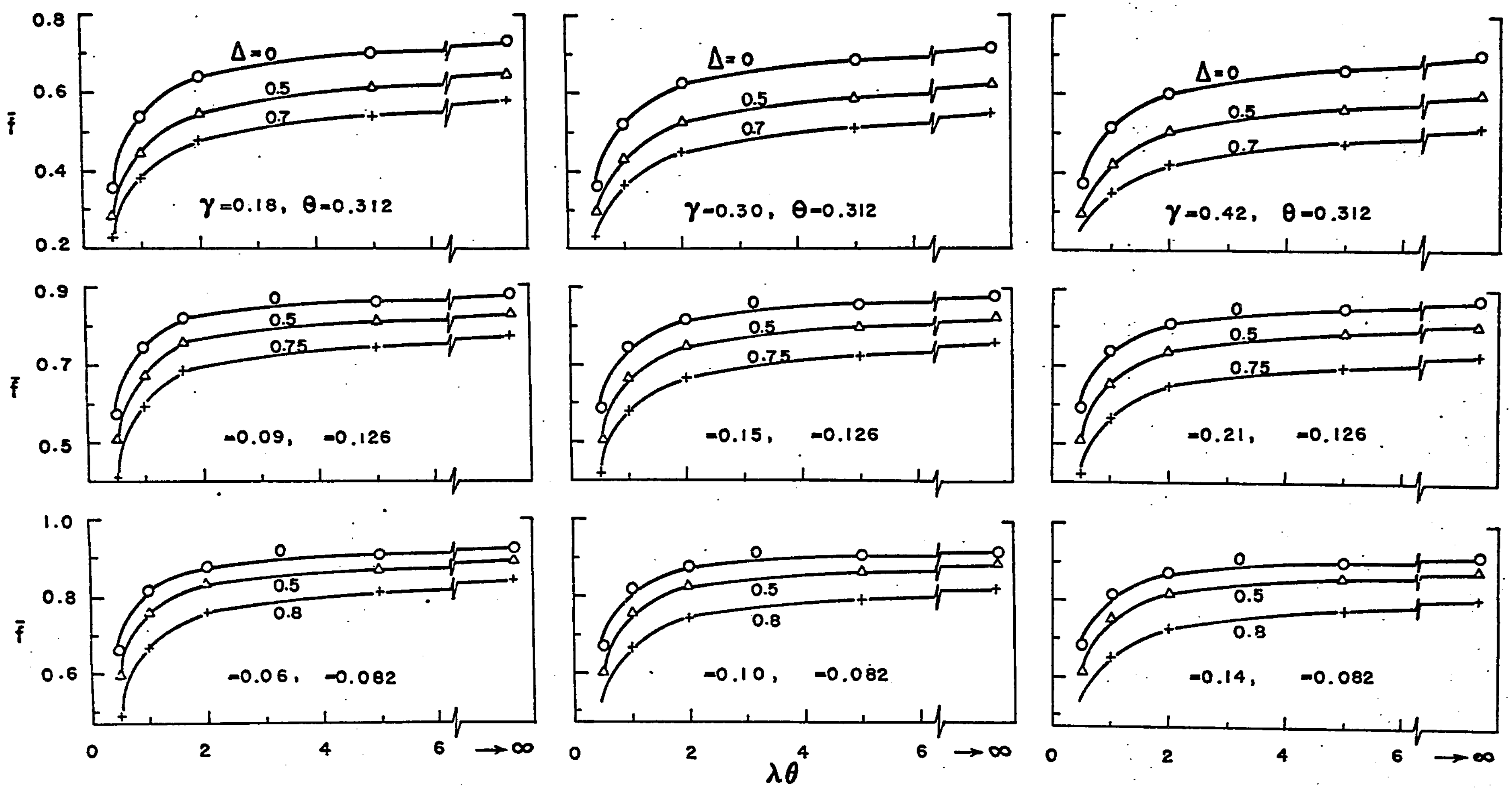


FILM TYPE : CA-NRC-316

FILM NO.3

SYSTEM : Na Cl - H_2O

FIGURE 10 RELATIONSHIP BETWEEN λθ AND AVERAGE SOLUTE SEPARATION-FILM 3



FILM TYPE : CA-NRC-316
FILM NO. 6
SYSTEM : $NaCl - H_2O$

FIGURE 11 RELATIONSHIP BETWEEN λθ AND AVERAGE SOLUTE SEPARATION-FILM 6

It must be pointed out that, upon comparing Figures 9, 10, and 11, any of the above reported effects was more evident with reverse osmosis systems employing more porous membranes (e.g., Film No. 6) rather than systems using less porous membranes (e.g., Film No. 3). This is an important consideration in reverse osmosis process design.

The results given in Figures 9 to 11 showed that, for reverse osmosis system operated at 300 psi with 1500 mg/l sodium chloride solution to recover 80% of the feed solution as product water, the average solute separations obtained with any of the studied films were greater than 76 and 81% at $\lambda\theta = 2$ and 5 respectively.

The reverse osmosis system analysis as stated above has shown useful results of practical significance in reverse osmosis process design.

4.4 Design Applications

From the reverse osmosis membrane specifications given in terms of A and $(D_{AM}/K\delta)$, the γ and θ parameters for the reverse osmosis system are calculated (using Eqs. 2.8, 2.9, and 2.11) for different operating pressures; for each assumed value of $\lambda\theta$, the reverse osmosis system is then completely specified in terms of γ, θ, and λ. From the feed and product concentrations, $\overline{C}_3$ is fixed; the corresponding Δ and τ (or X) values are then calculated from the general equations for process design (Eqs. 2.20 to 2.27). The work and membrane area requirements are calculated using the following relations (34):

$$\left.\begin{array}{l}\text{work required per unit}\\ \text{volume of product recovered}\end{array}\right\} = \frac{P\Delta+(1-\eta)P(1-\Delta)}{\Delta} \qquad (4.1)$$

where η represents fraction power recovery in a reverse osmosis unit. Since (1 atm.)(10^3 gallons) = 0.1066 kWh, the work requirements in terms of kWh per 1,000 gallons of product recovered can be calculated from Eq. (4.1) using the appropriate P, Δ, and η values for the particular membrane and operating conditions. The membrane area S is given by the relation:

$$S = \frac{\tau}{v_w^*} \; \frac{V_1^o}{t} \qquad (4.2)$$

where (V_1^o/t) = (product rate/Δ) and v_w^* is given by Eq. (2.11).

As an example using the above relations, the power and membrane area requirements were calculated as functions of $\lambda\theta$ for the already specified films 3 and 6 operating at 300 psi (20.4 atm.) with no power recovery (η=0) and A-factor = 1.0, for obtaining product water containing 500 ppm of salt from an aqueous sodium chloride feed solution of 3500 ppm (brackish water). The results, (Figure 12), showed that for reverse osmosis system employing a more porous membrane (e.g., Film No. 6), the required power per 1,000 gallons of the recovered product water was greater than the power that would be required if a less porous membrane (e.g., Film No. 3)

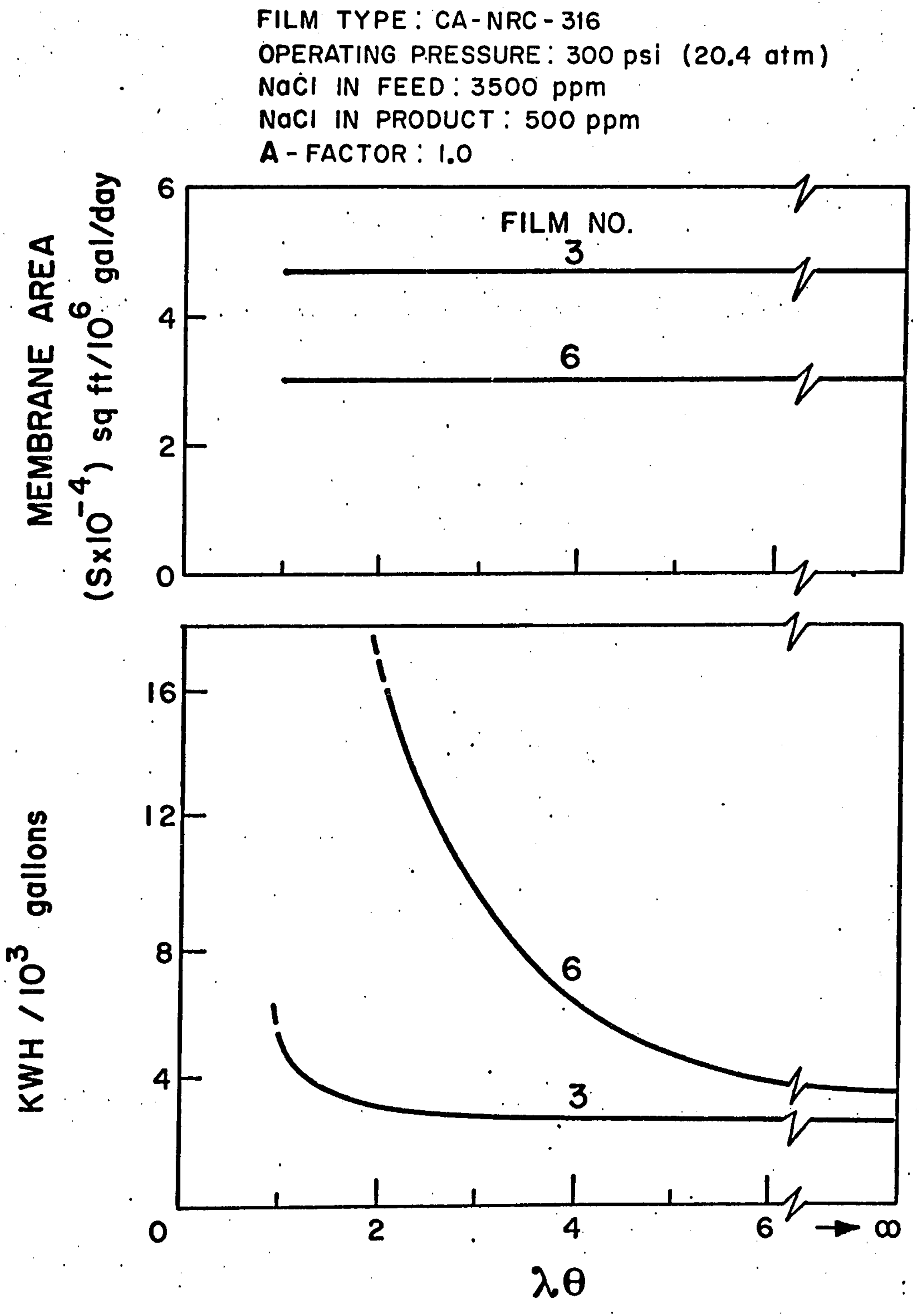


FIGURE 12 EFFECT OF λθ ON WORK AND MEMBRANE AREA REQUIREMENTS

was employed. However, the membrane area per 10^6 gallons of treated water per day required for the reverse osmosis system with more porous membrane was less than that area required for the system with less porous membrane. The difference in power requirements between the above two systems was more significant at low $\lambda\theta$ values and was no longer significant for high $\lambda\theta$ values (from $\lambda\theta = 5$ to ∞) while the required area for each system remained essentially unaffected by the change in $\lambda\theta$ values under the conditions specified in Figure 12. These results are of practical significance in reverse osmosis process design.

Figure 12 shows that, for the increase in $\lambda\theta$ value from 2 to 5, the required kWh/10^3 gallons was decreased by about 13 and 70% using films 3 and 6 respectively. The corresponding membrane area required was essentially constant in the range of $\lambda\theta$ values studied. This indicates the significant effect of $\lambda\theta$ value on the power requirements for systems employing more porous membranes. The results also showed that, under the specified operating conditions, for a practical value of $\lambda\theta = 2$, the work requirements were 3.1 and 16.7 kWh/1,000 gallons of product water recovered using films 3 and 6 respectively. The membrane area requirements, under the same conditions, were 4.7 and 3.0×10^4 ft^2/10^6 gallons of water treated per day with films 3 and 6 respectively. It should be pointed out that the above values for the work requirement do not take into consideration friction and other losses, and hence they present the practical minimum.

4.5 Summary

The membrane performance for low pressure (up to 300 psi) reverse osmosis separation of aqueous sodium chloride solutions in the concentration range of 1500 to 3500 mg/l-NaCl was studied under different operating conditions using films 1, 2, 3, 4, 5, and 6. The results showed the high productivity and solute separation capability of the type of membranes used in this study.

The reverse osmosis basic transport relationships (Eqs. 2.1 to 2.4) were used for calculating the values of A, $(D_{AM}/K\delta)$ for sodium chloride, and k for sodium chloride solution system from the experimental (PWP), (PR), and f data at preset operating conditions for films 1, 2, 3, 4, 5, and 6. At a given operating pressure, the values of A and $(D_{AM}/K\delta)$ specified the particular membrane for the reference solution system sodium chloride-water, and k specified the feed flow condition used.

Analysis and correlations of experimental data revealed that the membrane specifications obtained can be used in reverse osmosis process design following the general equations of Ohya and Sourirajan. The latter relationships (Eqs. 2.20 to 2.27) were used to specify different reverse osmosis systems and predict their performance with the reference solution system using films 1, 3, and 6.

In the range of solute concentration (1500 to 3500 mg/l NaCl) and operating pressure (100 to 300 psi) considered in this study, at a given level of product recovery, the reverse osmosis system performance was more affected by the change in the operating pressure (especially at the lower pressure range) than the solute concentration. The results showed that, for reverse osmosis system operated at 300 psi with 1500 mg/l sodium chloride solution and 80% product recovery, the average solute separations obtained with any of the films used in this study were greater than 76 and 81% at $\lambda\theta = 2$ and 5 respectively.

The results are of practical significance in reverse osmosis process design. An example for such design considerations was given, in which the power and membrane area requirements were calculated using membranes of higher porosities (film 6) and lower porosities (film 3). For the reference solution system sodium chloride-water containing 3500 mg/l of salt, the reverse osmosis process operated at 300 psi to obtain product water containing 500 mg/l of salt and with A-factor = 1, the minimum work requirements at a practical value of $\lambda\theta = 2$ were 3.1 and 16.7 kWh per 10^3 gallons of product water recovered using films 3 and 6 respectively. The corresponding area requirements were 4.7 and 3.0×10^4 ft^2 per 10^6 gallons of water treated per day with films 3 and 6 respectively.

CHAPTER 5

RESULTS OF REVERSE OSMOSIS OPERATION WITH ORGANIC WASTES

The previously specified membranes were used for the separation of single organic compounds individually presented in aqueous solutions, after which the membranes were applied to primary treated sewage to test their capability of removing the different wastewater constituents. All tests were performed at 300 psig using the continuous-flow laboratory-scale reverse osmosis cells. Unless otherwise stated, the feed flow condition used was that corresponding to a value of $k = 41 \times 10^{-4}$ cm/sec for the reference solution system 3500 mg/l $NaCl-H_2O$.

5.1 Separation of Single Organic Compounds

Glucose, sucrose and soluble starch were the carbohydrates chosen for this study. A protein rich compound - "Difco" beef extract, an amino acid-glutamic acid, and a sodium salt of a fatty acid-sodium stearate were also used. Anionic surface active agents were presented in this study by the non-biodegradable compound (ABS) and the more readily degradable one (LAS). Each of the foregoing organics, except the soluble starch, was individually present in aqueous solution at, at least, three different concentrations. With soluble starch, it was hard to ensure true aqueous solutions.

The pH of the glutamic acid solutions was adjusted at 6.0±0.1 by adding the required amounts of 1N NaOH solution.

The results from reverse osmosis short run experimentation obtained with films No. 1, 2, 3, 4, 5, and 6, for each organic solution system, (found in Appendix 3) are graphically presented in Figures 13 to 19. In these figures both the separation and product rate data are plotted against the feed concentration for the concentration range indicated for each studied system.

For the glucose-water system under the specified operating conditions, (Figure 13), the increase in feed concentration in the range ∿50 to 1000 mg/l slightly increased the solute separation while a 10-fold increase of feed concentration decreased the product rate by about 1.5%. The results showed that under the above specified conditions, glucose separation by reverse osmosis was >95.9% with average fluxes of 15.5, 23.6, 30.8, 35.0, 38.1, and 40.9 gpd/sq. ft could be obtained, respectively, with films 1, 2, 3, 4, 5, and 6.

Reverse osmosis experimental results showed that, for sucrose-water system under the specified operating conditions, (Figure 14), the solute separation was essentially constant in the feed concentration range of ∿50 to 2000 mg/l, and >99.3% solute separation could be obtained with films 1,

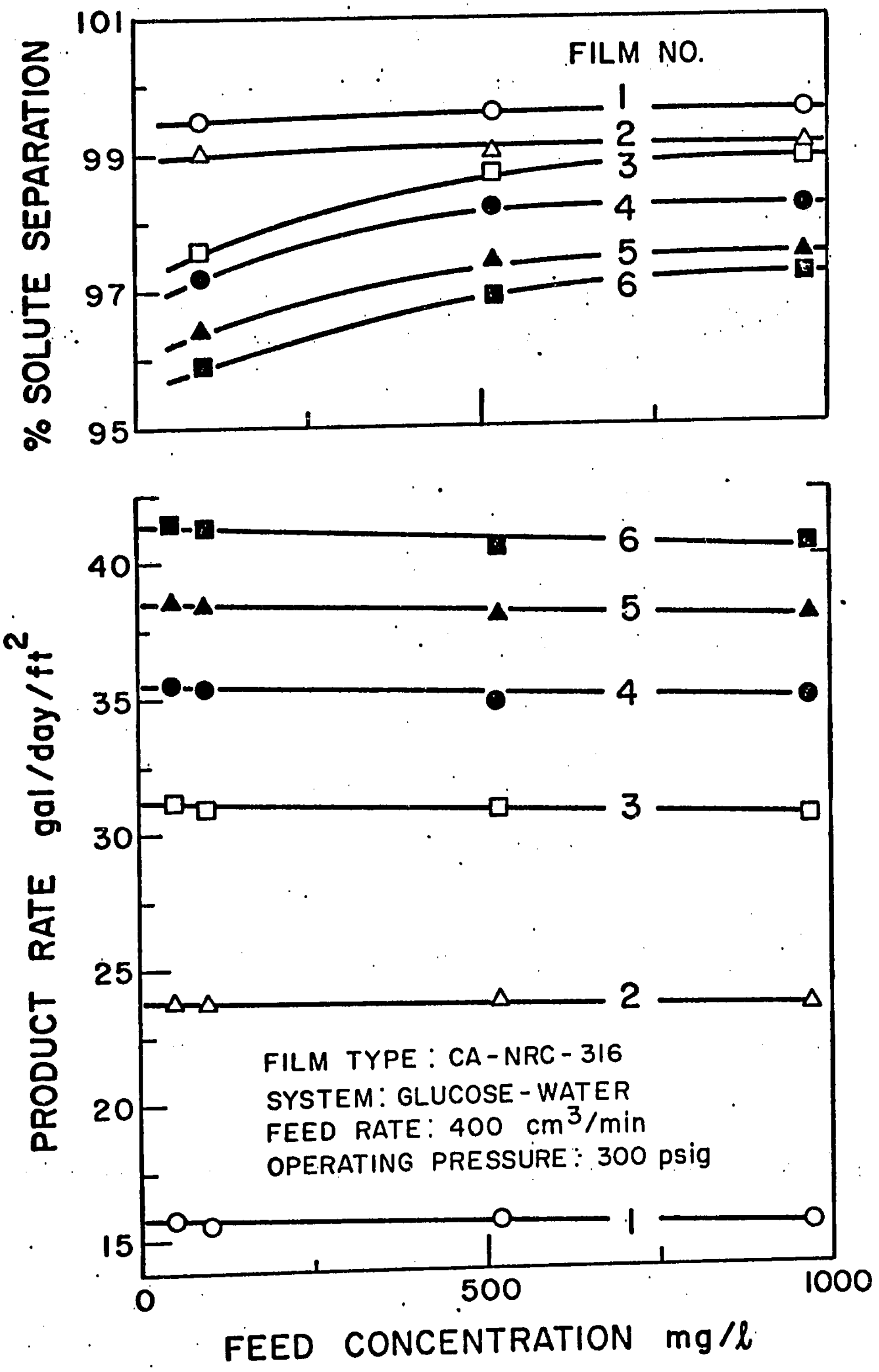


FIGURE 13 EFFECT OF SOLUTE CONCENTRATION ON PRODUCT RATE AND SOLUTE SEPARATION-GLUCOSE-WATER SYSTEM

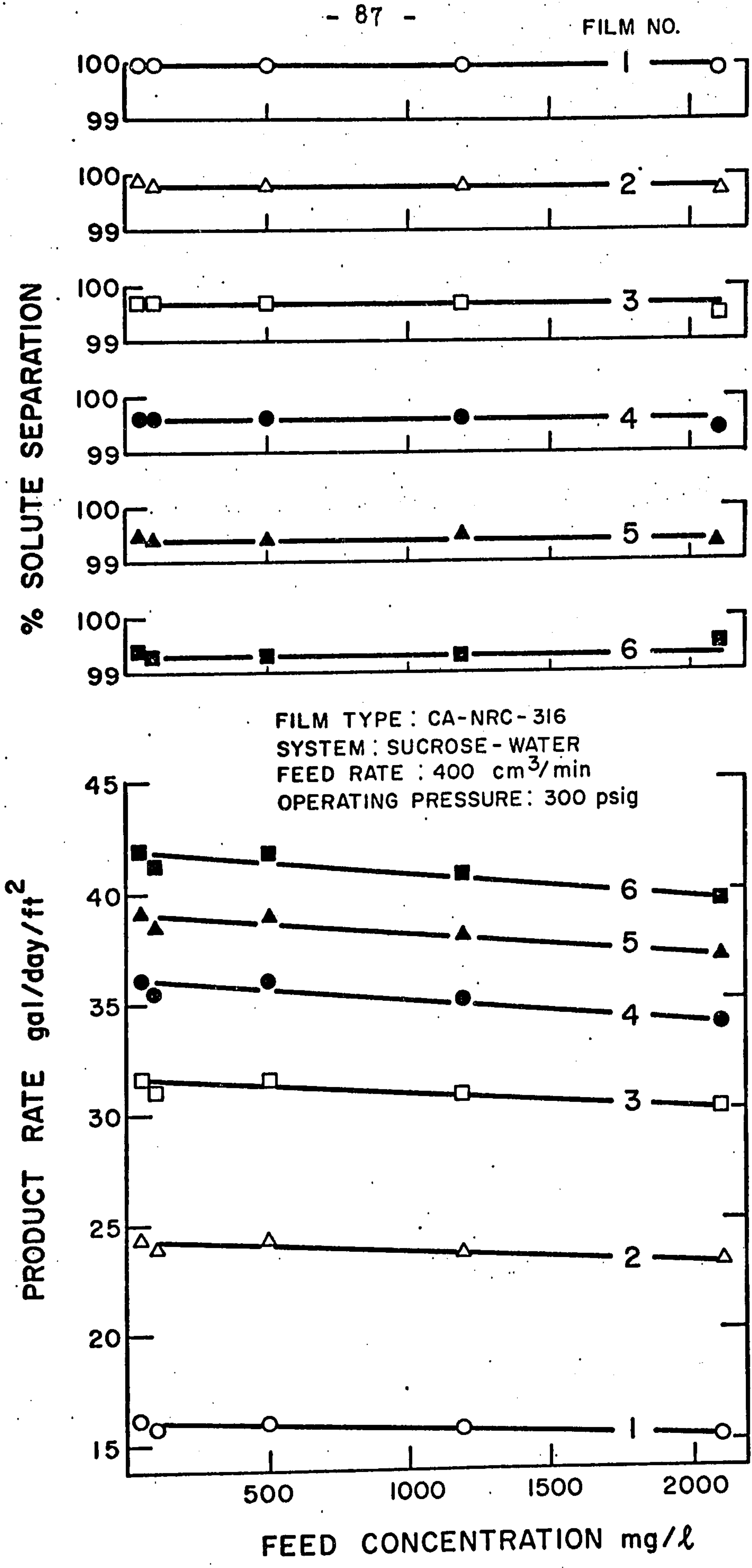


FIGURE 14 EFFECT OF SOLUTE CONCENTRATION ON PRODUCT RATE AND SOLUTE SEPARATION-SUCROSE-WATER SYSTEM

2, 3, 4, 5 and 6 with average fluxes of 15.6, 23.8, 31.0, 35.3, 38.3, and 41.0 gpd/sq. ft respectively. In the range of concentration studied, a 10-fold increase of feed concentration decreased the product rate by <2%.

Figure 15 shows that, under the specified operating conditions, the separation of soluble starch was increased with increasing feed concentration in the range of ∿10 to 100 mg/l. The corresponding product rate was practically unaffected by the increase in feed concentration because of the narrow range of feed concentration studied. The figure also shows that the average separation of soluble starch by reverse osmosis was about 97% which could be obtained with films 1, 2, 3, 4, 5, and 6, under the specified operating conditions, with average fluxes of 15.9, 24.0, 31.3, 35.1, 38.0, and 40.6 gpd/sq. ft respectively.

Using beef extract feed solutions in the range of ∿100 to 2000 mg/l (∿10 to 250 mg/l as organic nitrogen) under the specified operating conditions, (Figure 16), it was shown that a 10-fold increase of feed concentration decreased the product rate by about 5% while increased the percentage solute separation by a value of about 4. The average separation of beef extract by reverse osmosis, under the specified conditions, was about 93% with average fluxes of 15.7, 23.7, 30.5, 34.5, 37.4, and 39.9 gpd/ sq. ft could be obtained with films 1, 2, 3, 4, 5, and 6 respectively.

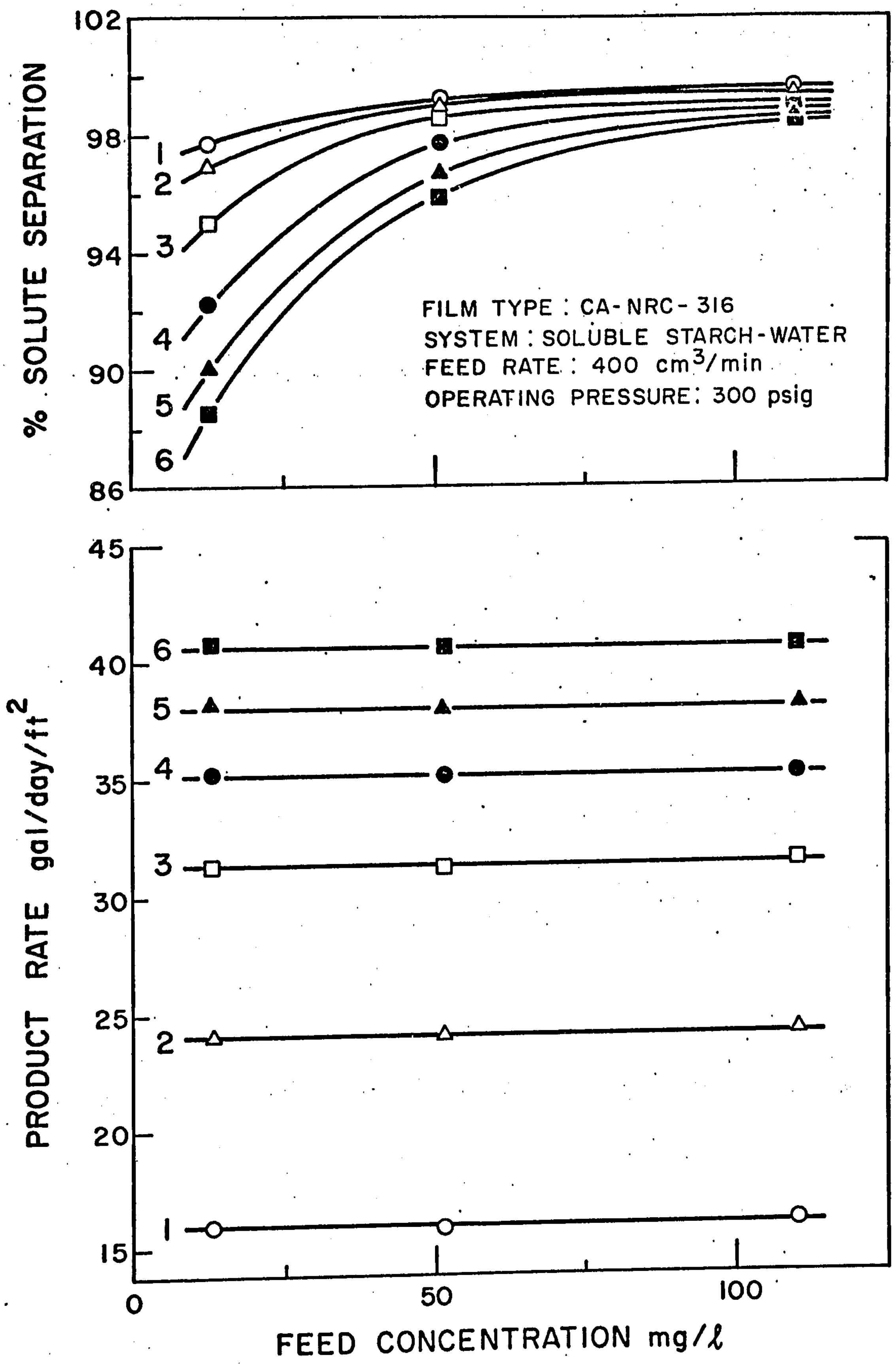


FIGURE 15 EFFECT OF SOLUTE CONCENTRATION ON PRODUCT RATE AND SOLUTE SEPARATION-SOLUBLE STARCH-WATER SYSTEM

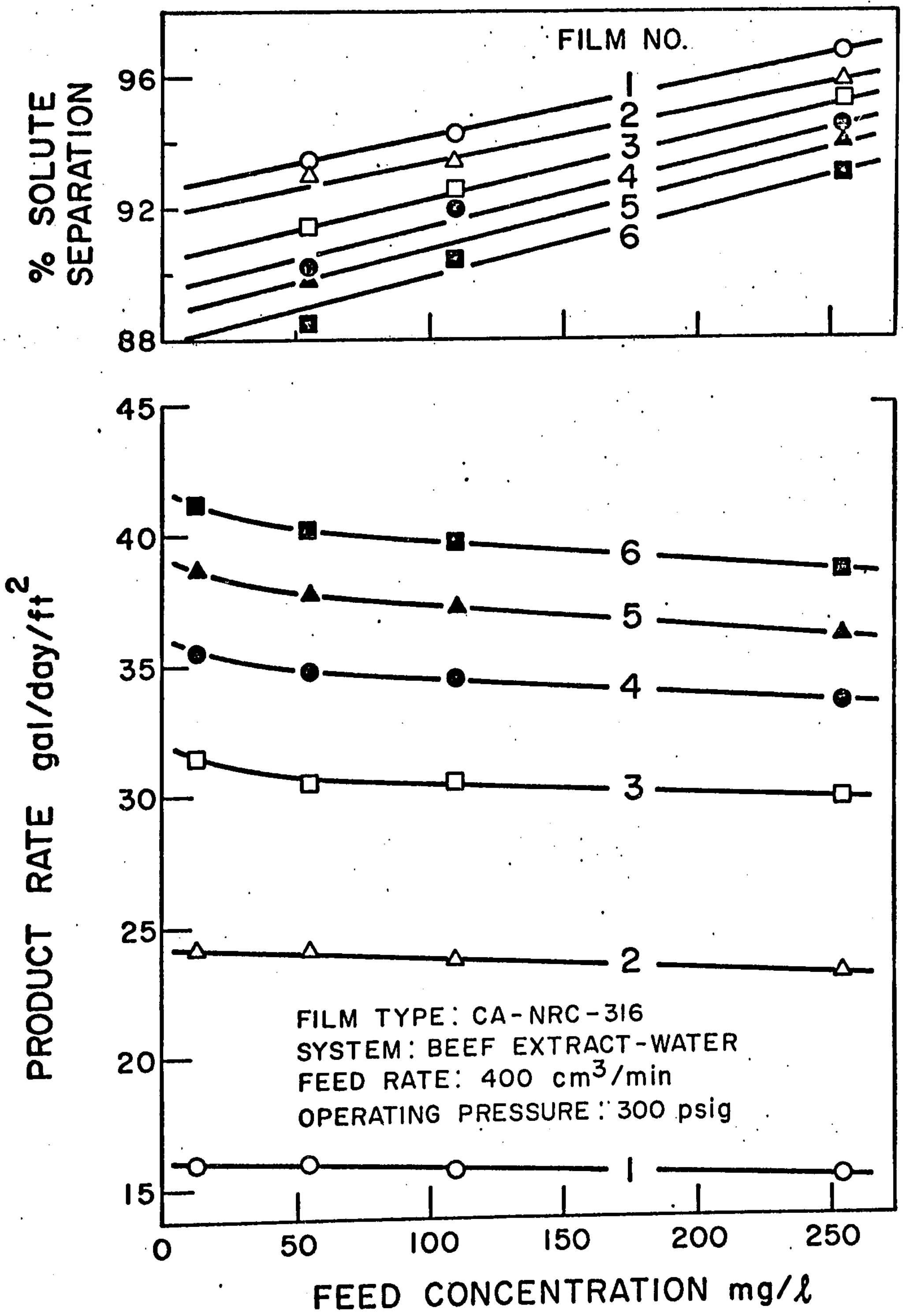


FIGURE 16 EFFECT OF SOLUTE CONCENTRATION ON PRODUCT RATE AND SOLUTE SEPARATION-BEEF EXTRACT-WATER SYSTEM

The results of reverse osmosis experiments with glutamic acid showed that, under the specified operating conditions (Figure 17), solute separation was slightly increased with feed concentration increase in the range of ∿10 to 500 mg/l, and >98.4% solute separation could be obtained with films 1, 2, 3, 4, 5, and 6 with average fluxes of 15.8, 24.0, 31.2, 35.3, 38.4, and 41.0 gpd/sq. ft respectively.

With respect to the sodium stearate-water system, Figure 18 shows that the solute separation was essentially unaffected by the increase in feed concentration in the range of ∿60 to 1000 mg/l while the product rate was decreased by about 2% for a 10-fold increase of feed concentration for films 4, 5, and 6. The product rates from films 1, 2, and 3 remained almost unchanged in the range of feed concentrations studied. Under the specified operating conditions, the separation of sodium stearate obtained was >97.5% with average fluxes of 15.8, 24.0, 31.2, 32.8, 37.4, and 38.5 gpd/sq. ft could be obtained with films 1, 2, 3, 4, 5, and 6 respectively.

Figure 19 gives the separation and product rate data obtained for the systems ABS-Water and LAS-Water as a function of the feed concentration in the range of ∿0 to 100 mg/l of total solutes. A 10-fold increase of feed concentration slightly increased the separation of both ABS and LAS with a corresponding decrease in the product rate of about

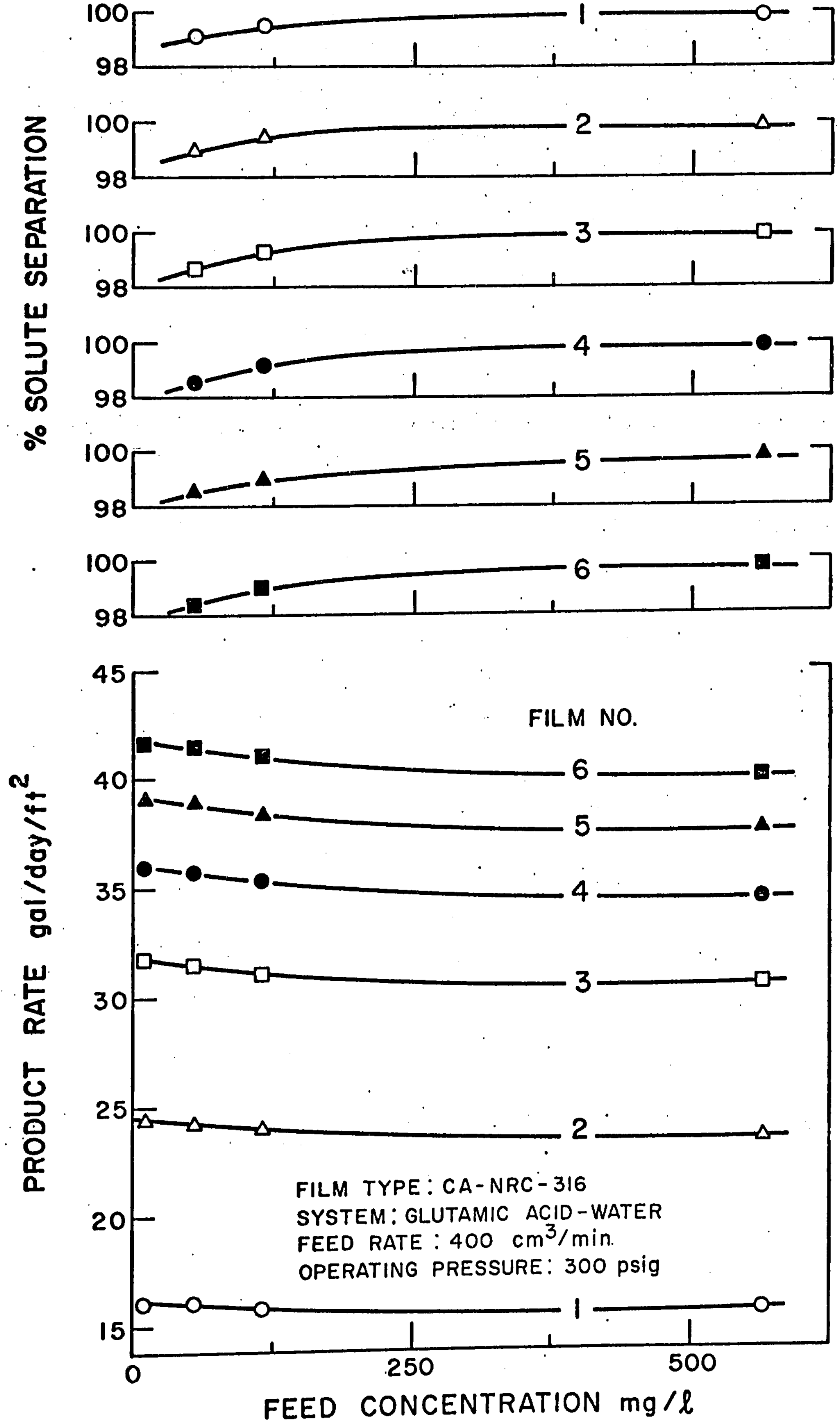


FIGURE 17 EFFECT OF SOLUTE CONCENTRATION ON PRODUCT RATE AND SOLUTE SEPARATION-GLUTAMIC ACID-WATER SYSTEM

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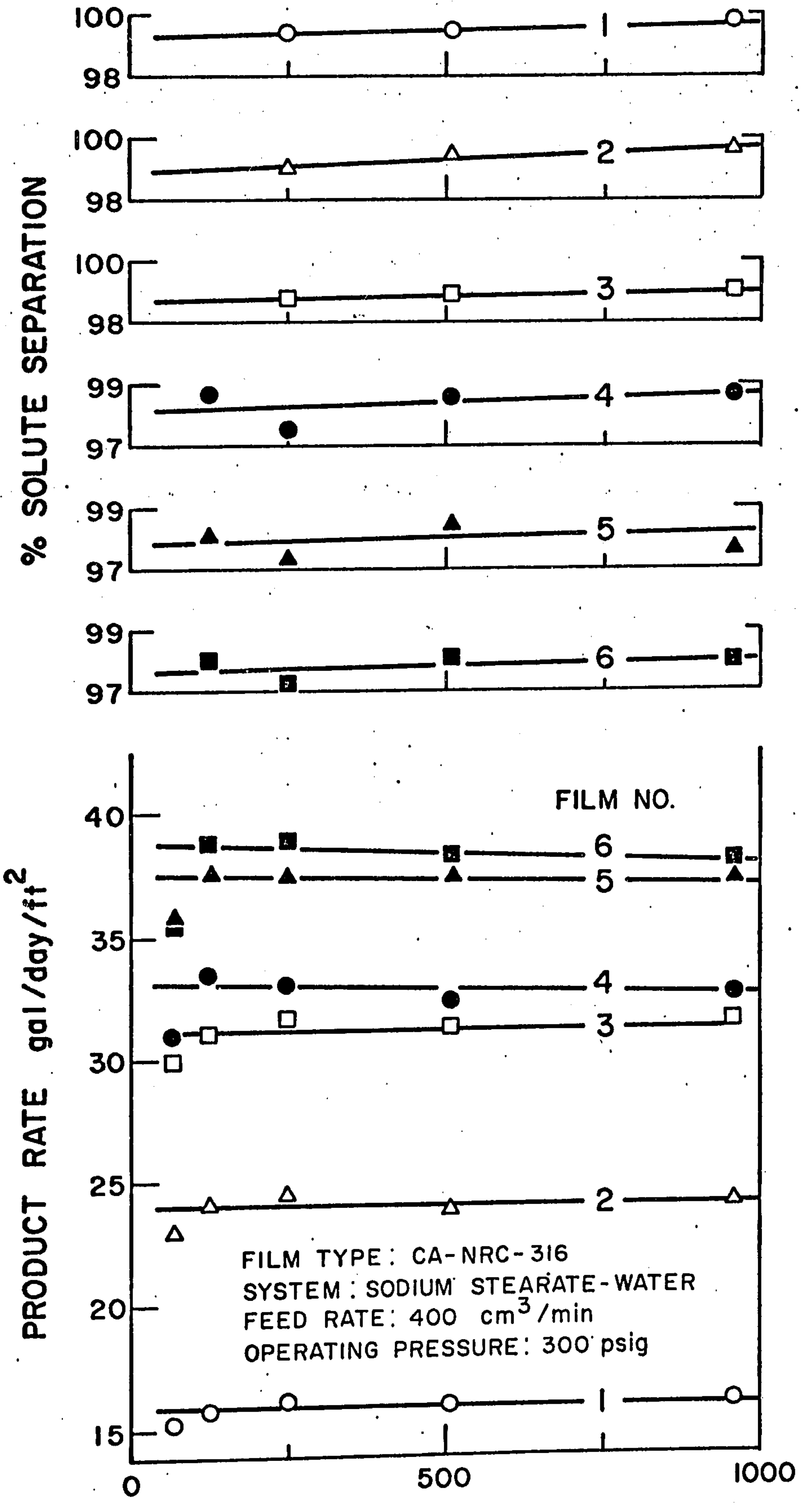

FIGURE 18 EFFECT OF SOLUTE CONCENTRATION ON PRODUCT RATE AND SOLUTE SEPARATION-SODIUM STEARATE-WATER SYSTEM

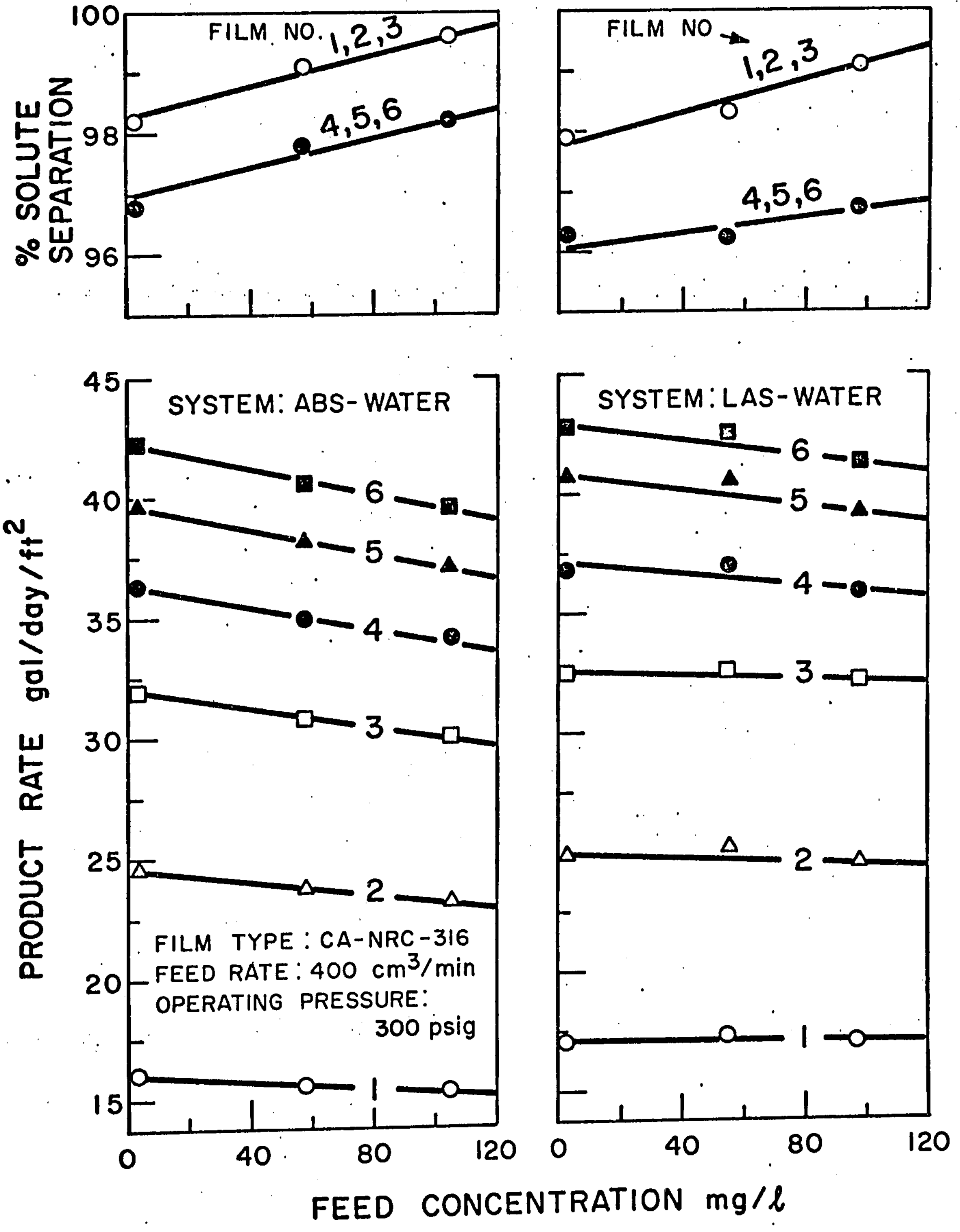

FIGURE 19 EFFECT OF SOLUTE CONCENTRATION ON PRODUCT RATE AND SOLUTE SEPARATION-ABS-WATER AND LAS-WATER SYSTEMS

5.7% and 2.5% respectively. The separation of ABS by reverse osmosis, operated under the specified conditions, was >96.8% with average fluxes of 15.6, 23.8, 31.0, 35.1, 38.3, and 40.8 gpd/sq. ft which could be obtained using films 1, 2, 3, 4, 5, and 6 respectively. Under the same operating conditions, LAS was separated >96.2% using films 1, 2, 3, 4, 5, and 6 with the corresponding product rates of 16.1, 24.8, 32.2, 36.5, 40.1, and 42.1 gpd/sq. ft respectively.

The effect of feed flow rate on the performance of reverse osmosis membranes operated at 300 psig was studied using sucrose solutions of 2000 mg/l. Figure 20 shows that, with films 1, 2, 3, 4, 5, and 6 under the above specified conditions, the solute separation was slightly increased with the feed rate increasing from 200 to 500 cu cm/min. The corresponding increase in the product rate was about 4%.

It should be pointed out that in this particular study, for each organic feed solution system, the concentration span studied covered both the average concentration of the specific organic compound as found in actual wastewaters and the expected concentration after 90% of feed solution recovery as product water (10-fold increase in feed concentration). Consequently, the presented results are of practical significance.

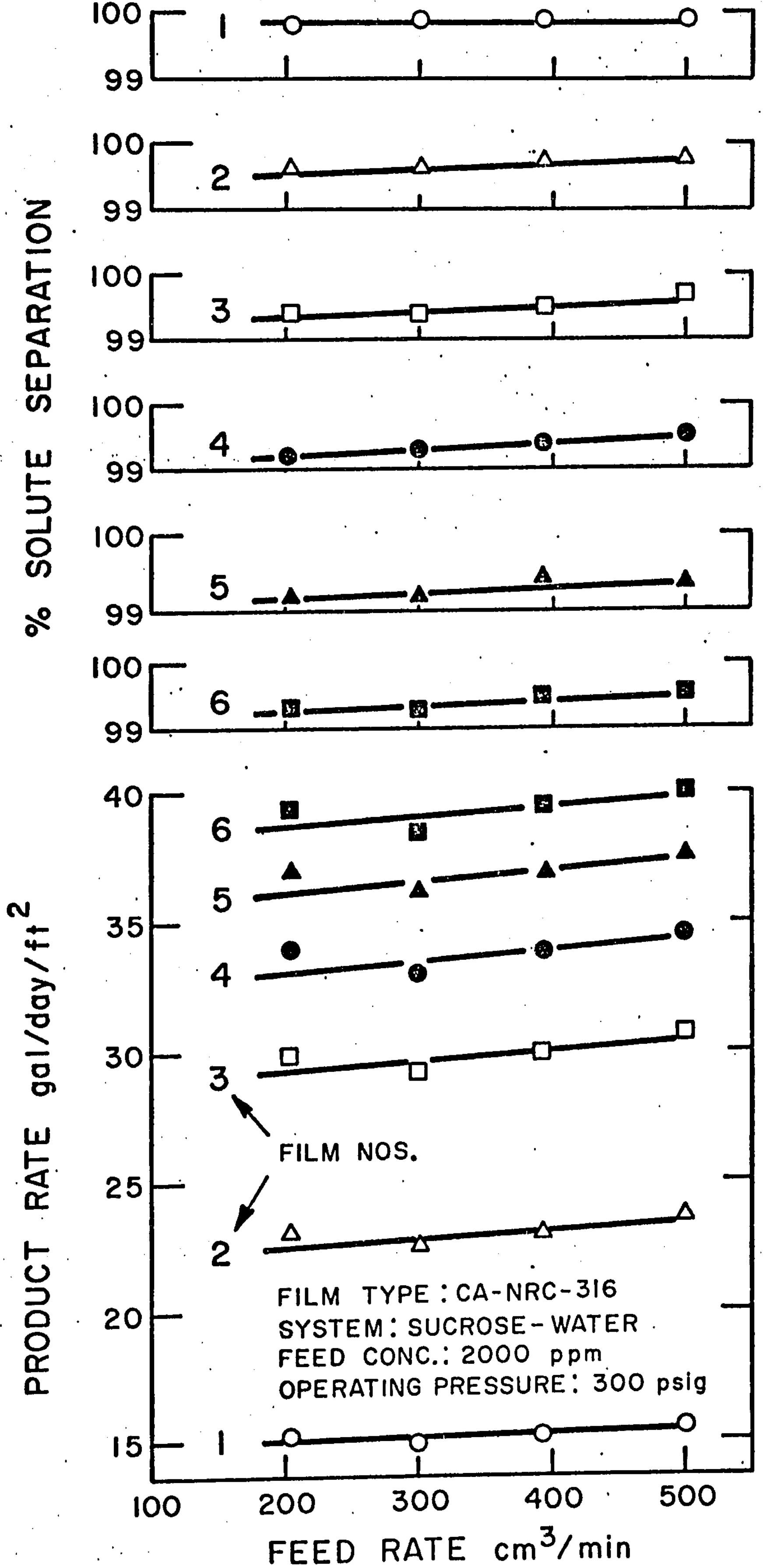

FIGURE 20 EFFECT OF FEED FLOW RATE ON PRODUCT RATE AND SOLUTE SEPARATION-SUCROSE-WATER SYSTEM

5.2 Relative Scales of Membrane Selectivity

Based on the foregoing experimental results, it was found that solute separation, at feed concentrations up to about 2000 mg/l, is essentially unaffected by changes in feed concentration for all solution systems studied at the feed flow conditions used in this work. Therefore relative membrane selectivity for different solutes may be expressed in terms of relative levels of solute separation. Such selectivity data are presented in Figure 21 for all the solutes considered in this study, using NaCl as the reference solute. The six relative levels of solute separation shown for each solution system are those given by the six films used in this work.

Figure 21 shows that membranes capable of giving NaCl separations of 87.5% or above could be used for the treatment of wastewaters by reverse osmosis. Consequently, the data on the performance of the six studied films are significant. The solute separation data available for the reference solution system NaCl-H_2O could be used to predict the relative solute separation data for the organic compounds using the relative selectivity scale shown in Figure 21.

5.3 Removal of Wastewater Constituents

Reverse osmosis laboratory experiments were carried out using the chlorinated wastewater effluent obtained from the Ottawa City primary sewage treatment plant. The operating pressure was 300 psig and the feed flow rate was 400 cu cm/min.

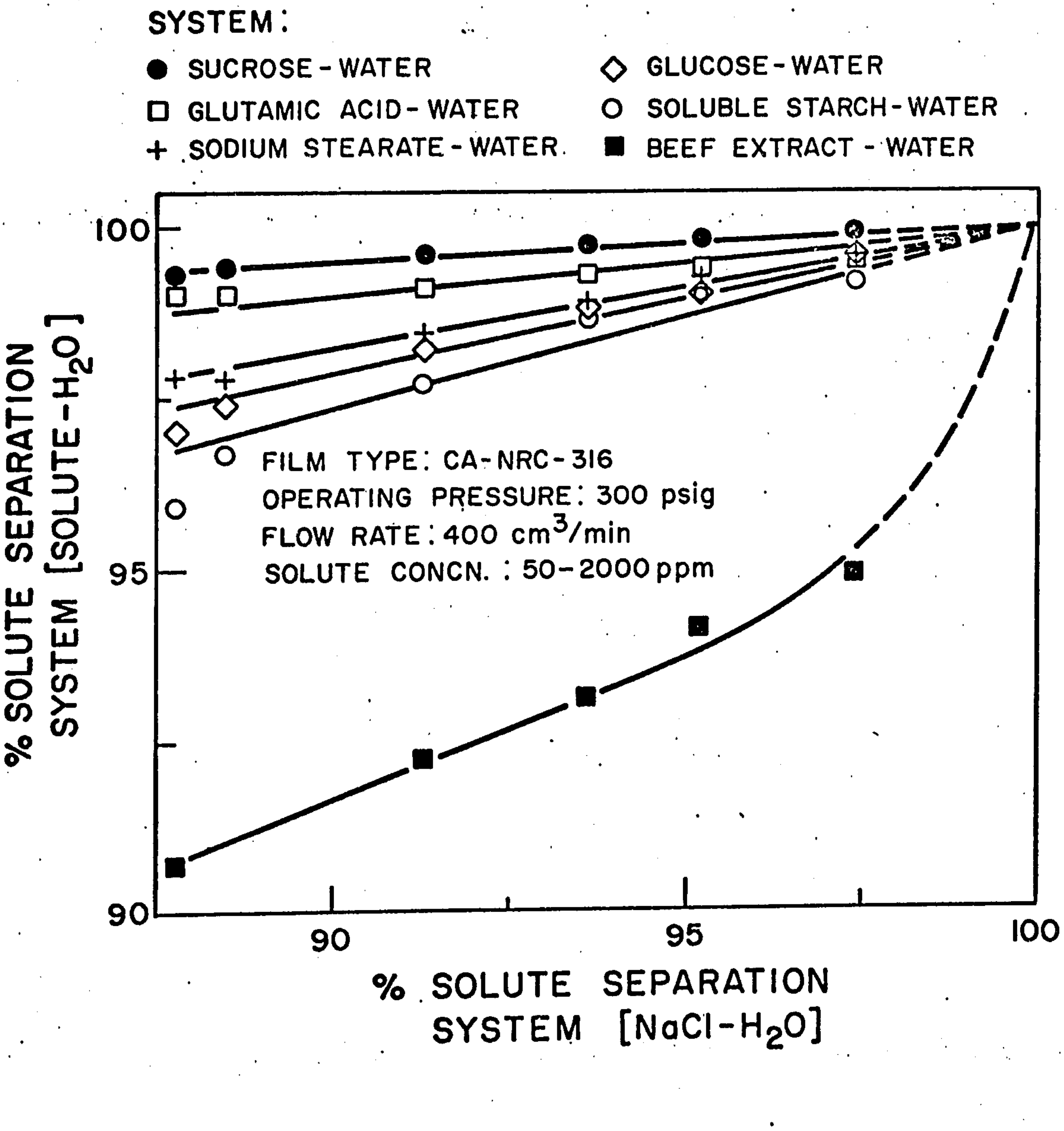

FIGURE 21 RELATIVE SCALES OF MEMBRANE SELECTIVITY

At first, a short-run type experiment lasting for about 1 hour, was carried out using the wastewater as a feed. The performance of each of the six membranes used throughout this work was studied with particular reference to the removal of organic wastewater constituents expressed in terms of their total organic carbon (TOC) content. Table 7 presents the results obtained with the typical films. Under the experimental conditions, the average removal of TOC was >96% with product rates of 14.5, 22.9, 29.1, 31.1, 33.4, and 33.9 gpd/sq. ft obtained with films 1, 2, 3, 4, 5, and 6 respectively.

Another experiment was running continuously with the wastewater feed in order to recover 90% of the feed as a product water in a single-stage reverse osmosis process. Table 8 gives the results obtained for different wastewater constituents. The membranes flux decreased with the progression of the experimental run. This was expected, especially with the primary sewage effluent feed which is an insufficiently treated type of waste. However, after finishing this experiment, when the precipitate on membranes surface was simply wiped off with distilled water, the membranes flux was recovered when tested with sodium chloride aqueous solutions.

The results given in Table 8 show that, using the improved type of reverse osmosis cellulose acetate membranes operated at 300 psig, the organic content of wastewater as

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TABLE 7

Membranes Performance Data with the Wastewater Feed[a]

Film No.	Film Shrinkage Temp. °C	TOC reduction, %	Product rate, gpd/sq. ft
1	81	>96.3	14.5
2	77	>96.3	22.9
3	74	>96.3	29.1
4	72	>96.3	31.1
5	70	96.3	33.4
6	68	96.3	33.9

[a] Film type: CA-NRC-316 (Batch 316)
Feed: chlorinated primary effluent
Feed TOC: 26.9 mg/l
Operating pressure: 300 psig
Feed rate: 400 cm^3/min

TABLE 8

Removal of Wastewater Contaminants by Reverse Osmosis[a]

Constituent	Feed Concn.	Product Concn.	% Removal
TDS mg/l	734	151	79.4
Conductivity μmho	1090	201	81.6
Ammonia mg/l-N	12.7	3.3	74.0
Nitrates mg/l-N	1.2	0.57	52.5
Orthophosphates mg/l-P	1.25	0	100
Total phosphates mg/l-P	1.63	0	100
COD mg/l	109	4	96.3
TOC mg/l	26.9	1	96.3
pH	6.2	7.0	-

[a] Film type: CA-NRC-316 (Batch 316)

Feed: Chlorinated primary sewage effluent (residual chlorine in feed is 0.5 ppm; feed pH was adjusted at 6.2)

Operating pressure: 300 psig
Feed rate: 400 cm^3/min
90% product recovery.

measured by the chemical oxygen demand and the total organic carbon content was reduced by about 96%. The dissolved solids in this wastewater as measured by the total dissolved solids content and specific conductivity were reduced by about 80%. Phosphates in both the form of ortho and total phosphate content were completely rejected by the type of cellulose acetate membranes used in this study. The removal of nitrates by these membranes was about 53% while, under the same conditions, 74% of ammonia was removed.

The above results illustrate the capability of improved reverse osmosis cellulose acetate membranes to give a high quality product water from wastewater feeds.

The initial average membrane flux was 27.3 gpd/sq.ft which shows the productivity of such membranes operated at low pressure in the reverse osmosis treatment of wastewaters. This flux was appreciably decreased with time apparently due to the presence of suspended and colloidal matter in the feed water. These results show the need for developing appropriate pretreatment techniques and membrane cleaning procedures for the application of reverse osmosis to municipal wastewater treatment. No attempt was made to develop such techniques and procedures in this work. The object of this particular study was simply to show the capability of removing wastewater contaminants, mainly the organics, by the new improved membranes used in the study.

5.4 Summary

In essence, the capability of the improved cellulose acetate membranes for the low pressure reverse osmosis organic treatment of wastewaters was demonstrated. This membrane process was found to be an efficient means of removing organics from wastewater and producing a high quality product water. In the concentration range of organics found in many wastewaters, the solute separation by the studied membranes was essentially unaffected by the concentration change, under the specified operating conditions used. Also, under the experimental specified conditions, the average product rates obtained with each membrane were essentially unaffected by the nature of the organic solute, for the organic materials in the range of concentrations used in this particular study. The average solute separation was >90% which could be obtained with all films used.

Based on the experimental results, a relative scale of membrane selectivity was given for each organic solute studied, in terms of relative solute separation using sodium chloride as the reference solute. Such a scale has practical significance in reverse osmosis process design for wastewater treatment applications using the relative data available for the reference solution system sodium chloride-water.

CHAPTER 6

DISCUSSION AND CONCLUSIONS

This study was conducted to investigate the performance of improved porous cellulose acetate membranes for low pressure reverse osmosis separation of organics from wastewater. The newly developed cellulose acetate membranes designated as CA-NRC-316, or simply Batch 316, type membranes operated at 300 psig were used for this purpose. Experimentation was accomplished using a laboratory-scale flow-type reverse osmosis cell. Membranes were first specified for the reference solution system sodium chloride-water and the obtained membrane specifications were used for reverse osmosis process design for the above system. The specified membranes were used for the reverse osmosis separation of some wastewater organic constituents and removal of different contaminants from chlorinated primary sewage effluent. In each experiment, the pure water permeation rate (PWP), the membrane permeated product rate (PR), and solute separation (f) were the parameters determined at the preset operating conditions.

6.1 Reverse Osmosis Membrane Specifications and Process Design

The correlations of experimental data on membrane specification for the reference solution system sodium chloride-water show that the solute transport parameter $(D_{AM}/K\delta)$, is independent of feed concentration and feed flow rate. The effect of operating pressure, P, on the parameter $(D_{AM}/K\delta)$ could be expressed by the relation $(D_{AM}/K\delta) \propto P^{-\beta}$; where β is a constant for the particular membrane for the given solute. The pure water permeability constant, A, is also pressure dependent following the relation $A \propto e^{-\alpha P}$ where α is a constant having very small values for Batch 316 type membranes under the conditions used.

For the reference solution system and the particular apparatus used, the effect of feed flow rate, Q, on the mass transfer coefficient, k, on the high pressure side of membrane could be expressed by the relation $k \propto Q^n$; where n is a constant.

The above experimental correlations similar to those reported for other types of cellulose acetate membranes developed earlier (40), are important because they show that the membrane specifications as obtained from the experimental analysis can be used for reverse osmosis process design following the general equations of Ohya and Sourirajan (33). For the same reason, such correlations need experimental confirmation for every new type of membrane developed. Also, it is necessary to provide specifications for membranes and feed flow conditions used in terms of parameters easily obtainable for a convenient reference feed solution system, in order to compare the relative performance of different membranes for feed solution systems for which the physicochemical data are not readily available.

Since the experimental data obtained by the six membranes used in this study showed that these membranes essentially exhibit a single relationship between k and Q, an average value of $k = 41 \times 10^{-4}$ cm/sec for the reference solution system 3500 mg/l $NaCl-H_2O$ was considered, for all membranes used, to express the feed flow condition used in this study, unless otherwise stated.

The data on performance of membranes in the low pressure reverse osmosis treatment of aqueous sodium chloride solutions containing 1500 to 3500 mg/l NaCl (Table 5 and Appendix 1) illustrate the high productivity of the type of membranes used in this work under the given operating conditions. These results are also of practical significance in brackish water desalination applications.

The membrane specifications as obtained from the experimental analysis were used for reverse osmosis system specification and process design following the Ohya-Sourirajan analysis (33). Different reverse osmosis systems were specified and their performances were studied using three different membranes (Films 1, 3 and 6). These membranes were chosen to be of different surface pores size and consequently reverse osmosis systems with different performances could be obtained using these films. With respect to the six membranes used throughout this investigation, film 6 (preshrunk at 68°C) had the biggest (average) surface pores and film 1 (preshrunk at 81°C) had the smallest (average) surface pores while film 3 (preshrunk at 74°C) was the one with intermediate pore size.

Under the specified conditions (Figures 9, 10, and 11) and at a given level of product recovery, the average separation in reverse osmosis is more affected by the change in the operating pressure (more appreciable in the range of 100 to 200 psi) than the solute concentration. This simply means that reverse osmosis systems using such membranes would be able to treat waters of varying quality without appreciable effect on the average separation of contaminants from these waters. But, because the average solute separation decreases with the increase in fraction product recovery, higher levels of product recovery could be obtained with reverse osmosis systems treating low concentrated solutions (without considerable effect on the average solute separation) than with systems treating high concentrated solutions. Since the average solute separation is not appreciably affected by the change of operating pressure in the range 200 to 300 psi, there is the possibility of using pressures less than 300 psi down to 200 psi. However, because the resulting productivity at the same levels of average solute separation and fraction product recovery will considerably decrease with the decrease in the operating pressure, the application of lower pressure, the application of lower pressure (<300 psi) may not be really advantageous from an economic standpoint. It should be pointed out that the foregoing discussion considers only single stage reverse osmosis systems for the sake of simplicity. Also, the above relations are considered for a certain value of $\lambda\theta$.

The results indicated that the average solute separation, for a given reverse osmosis system under the specified conditions and at a given level of product recovery, generally increased with the increase in the value of $\lambda\theta$ and this increase was more significant for $\lambda\theta < 2$ than for $\lambda\theta > 2$. Based on the experimental results, for the reverse osmosis apparatus used in this study under the specified conditions (at 300 psig), average values of $\lambda\theta = 5.5$, 2.9, and 2.1 could be obtained for the reference solution using films 1, 3, and 6 respectively. A value of $\lambda\theta = 2$ appears the most desirable from the point of view of maximum productivity.

It seems necessary to point out that all the above considered effects are more significant with reverse osmosis systems employing more porous membranes. This shows that with such systems, any required system performance could be significantly controlled by the appropriate operating conditions.

For reverse osmosis process design, it has been found that for the systems studied the work requirement is decreased with the increase in $\lambda\theta$ value while the area requirement remains unaffected by the change in $\lambda\theta$ value. The above work requirement for reverse osmosis systems using more porous membranes is greater than that required for systems with less porous membranes. The difference in power requirements by the above two systems is more considerable at low $\lambda\theta$ values and is essentially insignificant at high $\lambda\theta$ values (from $\lambda\theta = 5$ to ∞). Under the same operating conditions, more area is required by reverse osmosis systems using less porous membranes than systems with more porous membranes. The work and area requirements are of practical significance in reverse osmosis process design and economical consideration. The choice of either more porous or less porous membranes (when both give adequate separation) for reverse osmosis systems under certain operating conditions to produce water of a certain quality will be a matter of economical consideration for the cost of power and membranes required for this object. This is true for systems with low $\lambda\theta$ values and is no longer valid for systems with high $\lambda\theta$ values (a $\lambda\theta = 5$ is practically considered to be approaching the infinity as discussed before). In the latter case reverse osmosis systems using more porous membranes, with less area requirement and moreorless the same power requirement as systems with less porous membranes, will be preferred according to their higher productivities at any given level of solute separation. This is the case in many reverse osmosis applications to water pollution control for which it has been found that $\lambda\theta = 5$ is a practical value.

The results also illustrate the low power and area required for reverse osmosis systems using the type of membranes used in this study compared with other types of membranes under the same conditions (34). This is due to the high productivity, under low pressure operation, of the type of membranes used in this study. Consequently, the latter type of membranes is recommended in reverse osmosis for wastewater treatment and other water pollution control applications.

The reverse osmosis process design consideration as obtained for the reference solution system sodium chloride-water could be used for reverse osmosis process application to wastewater treatment. For instance, this may be used for reverse osmosis systems treating the organic solutions employed in this study by following the relative scales of membrane selectivity developed in the study (Figure 21). For a particular reverse osmosis system operated at a certain $\lambda\theta$ value to recover a certain fraction of product water using the reference solution system, the average solute separation obtained when referred to the relative selectivity scale of any of the organic solution systems studied will show the corresponding average organic solute separation value that would be expected to be obtained using this particular reverse osmosis system.

This also shows the practical significance of developing relative scales of membrane selectivity for different types of wastewaters.

6.2 Reverse Osmosis for Wastewater Treatment

The study of reverse osmosis removal of organics from wastewater was conducted with both organic synthetic wastewaters (single organics in aqueous solutions) and actual wastewaters using the previously specified membranes.

Eight organic compounds present individually in aqueous solutions in different concentrations up to 2000 mg/l were studied. Reagent grade glucose, sucrose, soluble starch, glutamic acid, and sodium stearate, and technical grade beef extract, ABS, and LAS were used. The work with these single compounds in aqueous solutions, under the operating conditions specified in Table 9, showed that the average removal of these organics by reverse osmosis was >90% and that the average product rates obtained with each membrane were essentially unaffected by the nature of the organic solute and the change in organic solute concentration. Moreover, solute separations were essentially independent of feed concentration under the same conditions. These results are significant to reverse osmosis process application for wastewater treatment and illustrate that the process could accommodate a wastewater feed stream that changes in composition and concentration continuously (e.g., domestic sewage). The results also illustrate the high productivity of the type of membranes used in this work.

The average solute separations for glucose, sucrose, soluble starch, beef extract, glutamic acid, sodium stearate, ABS, and LAS were greater than 95.9, 99.3, 97, 93, 98.4, 97.5, 96.8, and 96.2% respectively which could be obtained by all the films studied under the operating conditions used in this study. Since the solute separations of these organics covered only a narrow range, extensive comparisons or justification of differences are hardly possible. Moreover, the difference in the chemical nature of some of these organics makes any explanation more difficult. However, in general, it could be said that the effects of molecular structure, molecular weight, and molecular size are involved, as well as the physico-chemical criteria such as the polar effect of the solute molecule which governs reverse osmosis separation of organic solutes in aqueous solution. For the separation of glutamic acid, sodium stearate, ABS, and LAS, acidity (or the proton donating characteristic of the molecule) measured by the ease of hydrogen bond formation and/or the degree of dissociation of the molecule in solution, is probably the relevant expression of the polar effect of the molecules involved. In the case of glucose, sucrose, and soluble starch, the effect of molecular weight and molecular size could be significant, in addition to their structural characteristics.

The separation of beef extract which is a complex organic compound, might be governed by a combination of the above factors.

Although a relatively lower solute separation was reported for beef extract, being a nitrogen containing compound, the separation of glutamic acid which is also a nitrogen containing compound was very high. This can be explained by the higher degree of dissociation of glutamic acid especially when sodium hydroxide was added to glutamic acid solutions to adjust the pH of these feed solutions before application to reverse osmosis experimentation.

It should be pointed out that the slight increase in solute separation with the increase in solute concentration reported for some organic solution systems in the concentration range studied, is obviously due to agglomeration of solute particles.

While the results of this particular study, with chosen major organic compounds found in abundance in many wastewaters, show satisfactory high separations of these organics by reverse osmosis employing the type of membranes used in this work, there are some other organic compounds which might be present in wastewaters but could not be highly separated by these membranes. However, the presence of such compounds in wastewater feeds to reverse osmosis would not significantly affect the overall efficiency of organic removal by the reverse osmosis process since these compounds are, in many cases, present in very low concentrations and do not affect the final quality of the product water after the reverse osmosis treatment.

Since solute separation in reverse osmosis is a function of the extent of preferential sorption of water by the membrane material and the porous structure of the membrane surface (40), the separation data obtained with any of the membrane solution-operating systems used in this particular study do not represent the limiting values for the process.

The work with actual wastewaters showed that >96% separation of organics (measured as TOC and COD) could be obtained by the reverse osmosis treatment of these wastewaters using the type of membranes studied in this work under the operating conditions stated in Tables 7 and 8. This also confirms the results obtained with the synthetic organic wastes and illustrates the capability of these membranes for efficient removal of organics. The product rates from membranes 1, 2, 3, 4, 5, and 6 treating the primary sewage effluent were 14.5, 22.9, 29.1, 31,1, 33.4, and 33.9 gpd/sq. ft respectively which illustrate the high productivity of the type of membranes used when operated at low pressures.

A practical demonstration of the applicability of the reverse osmosis process to wastewater treatment was undertaken using the improved type of cellulose acetate membranes considered in this study which operated at 300 psig with chlorinated primary-treated municipal wastewater feed and 90% product recovery. Rejection of organics, nutrients, and dissolved salts was very good. This is illustrated by data presented in Table 8. The results obtained in this particular study as shown in Table 8, using the more-productive low-pressure operated membranes of the type studied are comparable to those reported by other workers for reverse osmosis treatment of wastewaters using another type of cellulose acetate membranes operated at higher pressures (Table 2). The high product water quality obtained from the reverse osmosis process, using the improved membranes used in this study, suggests the potentiality of such treatment process not only for pollution control applications but also to produce water which may be suitable for reuse with minor additional treatment. The latter application of the process could be important in making it economically feasible. Moreover, the use of more productive-low pressure operated reverse osmosis membranes of the type used in this study showed low power and area requirements obtainable at practical levels of product recovery. The use of such membranes in reverse osmosis treatment of wastewaters would result in great reductions in the treatment cost and consequently the economical feasibility of the process.

The initial average membrane flux from the films used in this study was 27.3 gpd/sq. ft which shows the high productivity of the type of membranes used when operated at 300 psig in the reverse osmosis treatment of primary sewage effluent under the conditions stated in Table 8. The rapid decrease in the membrane flux reported could be appreciably restored with the application of appropriate pretreatment techniques (which are also required for preventing the bacterial attack to cellulose acetate membranes), and membrane cleaning procedures.

For wastewater treatment applications, the use of more porous reverse osmosis membranes operated at low pressures could result in more significant reductions in the cost of treatment to produce water of high quality comparable to that obtained with less porous membranes of the same type.

Based on the results obtained in this study, it seems necessary to point out that the advantage of using reverse osmosis membrane treatment of wastewaters lies in the capability of membranes, by controlling the operating conditions, to produce waters of different quality levels to meet different quality standards required by municipalities and different water reuse. Moreover, the potentiality of membranes to remove different inorganic and organic wastewater constituents and successfully treat wastewaters of varying compositions and concentration without significant change in the membrane performance, make the reverse osmosis process an unique advanced wastewater treatment method competitor of different water pollution control applications.

The relative scale of membrane selectivity developed for organics used in this study (Figure 21) has practical significance in reverse osmosis process design for wastewater applications. Such a scale, would be able to predict, using membranes specified for the reference sodium chloride solution system, the performance of such membranes while treating those organics for which the scale is applicable. Moreover, reverse osmosis process design considerations for treatment of the above organic solutions could be related to and obtained from the design considerations developed for the treatment of the sodium chloride-water reference solution systems as previously discussed.

The experimental results indicate that, for the type of membranes used in this study under the conditions specified in Figure 21, membranes capable of giving sodium chloride separations of 87% or above are adequate for the treatment of sewage or similar wastewaters.

On the basis of the experimental findings in this study, a major divergence from the traditional concept of using reverse osmosis as a tertiary waste-treating method could be the direct application of the process to wastewater flows which have received special pretreatment sufficient to bring about an efficient reverse osmosis treatment. Obviously, such pretreatment of wastewater would be achieved by physical-chemical methods and its object is mainly the removal of particles in the suspended and colloidal form, some organic materials, and bacteria. In some cases it could simply be pH adjustment. A proposed reverse osmosis single-stage treatment might be carried out in steps until a final concentrated solution is obtained which is more suitable to handle for ultimate disposal leaving a high quality product water recovered. Using the type of membranes employed in this study, reverse osmosis units in series could be used in order to recover a certain fraction of product water as required. In such reverse osmosis series, the first step units would be operated at low pressure (e.g., 300 psi) and could be physically big. This step is followed by another one of physically smaller reverse osmosis units operated at relatively higher pressure (e.g., 600 psi) with the concentrated feed solution from the first step. The third step reverse osmosis units could be physically of still smaller size than that used in the foregoing steps and operated at high pressures (e.g., 1000 psi or more) with the concentrated feed solution from the foregoing step. Three reverse osmosis treatment steps as discussed above are thought to be sufficient to concentrate the wastewater feed and recover big product water volume of a high quality, which is collected through the different reverse osmosis treatment steps. The concentrated feed solution from any of these steps could be bypassed and disposed of depending upon the required fraction of product recovery. Also, membranes with different (average) surface pore sizes could be used through the different reverse osmosis treatment steps. Membranes with bigger (average) surface pores would be used for the treatment of less concentrated wastewater feeds and operated at lower pressures while membranes with smaller (average) surface pores could be used to treat more concentrated feeds and operated at higher pressures.

6.3 Conclusions

1. The experimental data presented illustrate the applicability of more productive low-pressure operated reverse osmosis membranes for the removal of organics from wastewater using the newly developed Batch 316 type porous cellulose acetate membranes operated at 300 psi.

2. The analysis and correlation of reverse osmosis experimental data obtained using Batch 316 type membranes are similar to those reported earlier for other types of porous cellulose acetate membranes. The analysis developed by Sourirajan and his associates for reverse osmosis membranes and membrane-solution operating systems are appropriate for systems using the Batch 316 type membranes.

3. In the application of Batch 316 type membranes for separation of some organic solutes from aqueous solutions under the conditions stated in Table 9, the solute separation is essentially independent of feed concentration and the average membrane flux is practically unaffected by the nature of the

solute. The average organic solute separation is >90% and the average membrane flux is in the range 15.8 to 40.6 gpd/sq.ft.

4. Relative scales of membrane selectivity developed for the organic solution systems, using NaCl-H_2O as a reference solution system, could be used in reverse osmosis design for wastewater treatment applications.

5. Reverse osmosis treatment of wastewater employing Batch 316 type membranes could remove >96% of organics (measured as COD or TOC) from the wastewater producing a high quality water.

APPENDIX 1

MEMBRANE SPECIFICATIONS AND PERFORMANCE DATA*

*PRESS = operating pressure (atm)

FEED ML = feed molality

FEED RT = feed flow rate (cm^3/min)

PWP = rate of pure water permeability per 7.6 cm^2 of membrane surface (g/hr) at 25°C

PROD RT = rate of membrane permeated product solution per 7.6 cm^2 of membrane surface (g/hr) at 25°C

SEPRTN = solute separation

D_{AM} = solute transport parameter (cm/sec)

MASS TRANS = mass transfer coefficient on high pressure side of membrane (cm/sec)

A = pure water permeability constant, g.mole H_2O/sq.cm. sec. atm.

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NO. 1                        SODIUM CHLORIDE

NO     PRESS          FEEDML        FEEDRT          PWP           PRODRT        SEPRTN

 1     20.40000       0.05740      390.00000      19.96000      16.91000      0.97410
 2     20.40000       0.02630      386.00000      19.89000      18.67000      0.97110
 3     17.70000       0.02720      415.00000      17.34000      15.56000      0.96540
 4     20.40000       0.02610      406.00000      19.70000      18.07000      0.97400
 5     20.40000       0.03490      410.00000      20.06000      18.27000      0.97300
 6     20.40000       0.04340      403.00000      19.29000      17.14000      0.97530
 7     20.40000       0.06010      404.00000      19.81000      16.85000      0.97470
 8     20.40000       0.06010      206.00000      19.68000      15.96000      0.97550
 9     20.40000       0.05820      308.00000      19.80000      16.54000      0.97220
10     20.40000       0.06070      405.00000      19.71000      16.58000      0.97560
11     20.40000       0.05710      516.00000      20.90000      17.51000      0.97420
12      4.30000       0.02550      400.00000       4.04000       2.92000      0.89950
13      7.40000       0.02530      405.00000       6.73000       5.78000      0.93800
14     14.30000       0.02580      403.00000      13.55000      12.31000      0.96550
15     17.20000       0.02520      403.00000      16.78000      15.11000      0.97160
16     21.20000       0.02470      402.00000      19.90000      18.55000      0.97400

NNO     DAM           MASSTRANS      A              XA1            XA2            XA3

 1    0.1350D-04    0.3086D-02    0.1985D-05    0.1033D-02    0.1257D-02    0.2679D-04
 2    0.1897D-04    0.9614D-02    0.1978D-05    0.4736D-03    0.5075D-03    0.1370D-04
 3    0.1310D-04    0.1373D-02    0.1987D-05    0.4898D-03    0.7334D-03    0.1647D-04
 4    0.1216D-04    0.1766D-02    0.1959D-05    0.4700D-03    0.6782D-03    0.1223D-04
 5    0.1583D-04    0.4153D-02    0.1994D-05    0.6284D-03    0.7353D-03    0.1698D-04
 6    0.1352D-04    0.3845D-02    0.1918D-05    0.7813D-03    0.9165D-03    0.1932D-04
 7    0.1407D-04    0.4683D-02    0.1970D-05    0.1082D-02    0.1220D-02    0.2740D-04
 8    0.1020D-04    0.1598D-02    0.1957D-05    0.1082D-02    0.1548D-02    0.2653D-04
 9    0.1333D-04    0.2300D-02    0.1969D-05    0.1047D-02    0.1354D-02    0.2915D-04
10    0.1268D-04    0.3341D-02    0.1960D-05    0.1092D-02    0.1305D-02    0.2669D-04
11    0.1305D-04    0.2416D-02    0.2078D-05    0.1028D-02    0.1332D-02    0.2655D-04
12    0.1050D-04    0.8235D-03    0.1906D-05    0.4592D-03    0.5165D-03    0.4613D-04
13    0.1452D-04   -0.5872D-02    0.1845D-05    0.4556D-03    0.4405D-03    0.2827D-04
14    0.1401D-04    0.3204D-02    0.1922D-05    0.4646D-03    0.5324D-03    0.1604D-04
15    0.1051D-04    0.1281D-02    0.1967D-05    0.4538D-03    0.6922D-03    0.1290D-04
16    0.1386D-04    0.2518D-02    0.1904D-05    0.4448D-03    0.5791D-03    0.1157D-04
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ILM NO. 2                  SODIUM CHLORIDE

RUNNO  PPRESS       FEEDML     FEEDRT       PWP         PRODRT      SEPRTN

  1    20.40000     0.05740    390.00000    29.36000    24.97000    0.95890
  2    20.40000     0.02630    386.00000    29.61000    27.66000    0.95250
  3    17.60000     0.02720    415.00000    25.43000    22.84000    0.94930
  4    20.40000     0.02610    406.00000    29.58000    27.14000    0.95160
  5    20.40000     0.03490    410.00000    30.10000    27.37000    0.95390
  6    20.40000     0.04340    408.00000    28.85000    25.73000    0.96060
  7    20.40000     0.06010    404.00000    30.02000    25.36000    0.95650
  8    20.40000     0.06010    206.00000    29.64000    24.12000    0.94960
  9    20.40000     0.05820    308.00000    29.79000    25.09000    0.95260
 10    20.40000     0.06070    405.00000    30.03000    25.07000    0.95550
 11    20.40000     0.05710    516.00000    31.44000    26.35000    0.95710
 12     4.10000     0.02550    400.00000     6.07000     4.49000    0.81840
 13     7.30000     0.02530    405.00000    10.29000     8.79000    0.89380
 14    14.10000     0.02580    403.00000    20.54000    18.54000    0.93450
 15    17.20000     0.02520    403.00000    25.56000    22.98000    0.94480
 16    21.10000     0.02470    402.00000    30.15000    28.18000    0.95100


RUNNO  CAM          MASSTRANS    A            XA1          XA2          XA3

  1    0.3231D-04   0.4690D-02   0.2919D-05   0.1033D-02   0.1246D-02   0.4251D-04
  2    0.4302D-04   0.6258D-02   0.2944D-05   0.4736D-03   0.5529D-03   0.2251D-04
  3    0.2940D-04   0.1994D-02   0.2931D-05   0.4898D-03   0.7323D-03   0.2485D-04
  4    0.3408D-04   0.2515D-02   0.2941D-05   0.4700D-03   0.6870D-03   0.2276D-04
  5    0.3981D-04   0.5073D-02   0.2993D-05   0.6284D-03   0.7594D-03   0.2899D-04
  6    0.3336D-04   0.6340D-02   0.2868D-05   0.7813D-03   0.9016D-03   0.3081D-04
  7    0.3502D-04   0.4912D-02   0.2985D-05   0.1082D-02   0.1297D-02   0.4711D-04
  8    0.3217D-04   0.2335D-02   0.2947D-05   0.1082D-02   0.1554D-02   0.5453D-04
  9    0.3597D-04   0.3805D-02   0.2962D-05   0.1047D-02   0.1320D-02   0.4971D-04
 10    0.3362D-04   0.3791D-02   0.2986D-05   0.1092D-02   0.1379D-02   0.4367D-04
 11    0.3269D-04   0.3428D-02   0.3126D-05   0.1028D-02   0.1348D-02   0.4414D-04
 12    0.3259D-04   0.1440D-02   0.3003D-05   0.4592D-03   0.5047D-03   0.8344D-04
 13    0.3712D-04   0.1034D-01   0.2859D-05   0.4556D-03   0.4685D-03   0.4341D-04
 14    0.3817D-04   0.3063D-02   0.2955D-05   0.4646D-03   0.5724D-03   0.3045D-04
 15    0.3080D-04   0.1796D-02   0.3014D-05   0.4538D-03   0.7103D-03   0.2507D-04
 16    0.4133D-04   0.4096D-02   0.2898D-05   0.4448D-03   0.5661D-03   0.2181D-04
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FILM NO. 3 SODIUM CHLORIDE

RUNNO	PRESS	FEEDML	FEEDRT	PWP	PRODPT	SEPRTN
1	20.40000	0.05740	390.00000	39.12000	32.46000	0.94040
2	20.40000	0.02630	386.00000	38.85000	35.93000	0.94410
3	17.50000	0.02720	415.00000	33.20000	29.81000	0.93330
4	20.40000	0.02610	406.00000	38.88000	35.63000	0.93570
5	20.40000	0.03490	410.00000	39.62000	35.80000	0.93630
6	20.40000	0.04340	408.00000	38.01000	33.68000	0.93390
7	20.40000	0.06010	404.00000	39.22000	33.10000	0.93450
8	20.40000	0.06010	206.00000	38.79000	31.18000	0.92320
9	20.40000	0.05820	303.00000	39.26000	32.61000	0.92970
10	20.40000	0.06070	405.00000	38.97000	32.61000	0.93710
11	20.40000	0.05710	516.00000	41.15000	34.20000	0.93820
12	4.00000	0.02550	400.00000	7.61000	5.85000	0.75670
13	7.10000	0.02530	405.00000	13.19000	11.38000	0.85180
14	14.00000	0.02580	403.00000	26.87000	24.22000	0.91840
15	17.10000	0.02520	403.00000	33.38000	29.94000	0.93000
16	21.00000	0.02470	402.00000	39.76000	36.93000	0.93770

RUNNO	DAM	MASSTRANS	A	XA1	XA2	XA3
1	0.5347D-04	0.3452D-02	0.3890D-05	0.1033D-02	0.1433D-02	0.6164D-04
2	0.5762D-04	0.4345D-02	0.3863D-05	0.4736D-03	0.6318D-03	0.2649D-04
3	0.5062D-04	0.2513D-02	0.3848D-05	0.4898D-03	0.7381D-03	0.3269D-04
4	0.5864D-04	0.3066D-02	0.3866D-05	0.4700D-03	0.7035D-03	0.3024D-04
5	0.6766D-04	0.4723D-02	0.3939D-05	0.6284D-03	0.8167D-03	0.4006D-04
6	0.6426D-04	0.5510D-02	0.3779D-05	0.7813D-03	0.9654D-03	0.4773D-04
7	0.6843D-04	0.5560D-02	0.3899D-05	0.1082D-02	0.1328D-02	0.7093D-04
8	0.6017D-04	0.2438D-02	0.3857D-05	0.1082D-02	0.1663D-02	0.8317D-04
9	0.6456D-04	0.3542D-02	0.3903D-05	0.1047D-02	0.1433D-02	0.7372D-04
10	0.6253D-04	0.4776D-02	0.3875D-05	0.1092D-02	0.1383D-02	0.6830D-04
11	0.5858D-04	0.3640D-02	0.4091D-05	0.1028D-02	0.1424D-02	0.6358D-04
12	0.6559D-04	0.4272D-02	0.3859D-05	0.4592D-03	0.4771D-03	0.1118D-03
13	0.7221D-04	-0.4942D-01	0.3768D-05	0.4556D-03	0.4523D-03	0.6756D-04
14	0.6176D-04	0.3621D-02	0.3893D-05	0.4646D-03	0.5831D-03	0.3794D-04
15	0.5013D-04	0.2165D-02	0.3959D-05	0.4538D-03	0.7274D-03	0.3179D-04
16	0.6359D-04	0.3838D-02	0.3840D-05	0.4448D-03	0.6179D-03	0.2773D-04

FILM NO. 4 SODIUM CHLORIDE

RUNNO	PRESS	FEEDML	FEEDRT	PWP	PRODRT	SEPRTN
1	20.40000	0.05740	390.00000	44.08000	37.53000	0.91240
2	20.40000	0.02630	386.00000	44.83000	41.49000	0.91480
3	17.40000	0.02720	415.00000	38.08000	34.42000	0.90380
4	20.40000	0.02610	406.00000	44.52000	40.92000	0.90950
5	20.40000	0.03490	410.00000	45.31000	41.12000	0.91130
6	20.40000	0.04340	408.00000	43.78000	38.71000	0.90930
7	20.40000	0.06010	404.00000	45.12000	38.03000	0.90750
8	20.40000	0.06010	206.00000	44.43000	35.95000	0.89330
9	20.40000	0.05820	308.00000	45.07000	37.45000	0.90380
10	20.40000	0.06070	405.00000	44.94000	37.42000	0.91040
11	20.40000	0.05710	516.00000	46.61000	39.19000	0.90350
12	3.90000	0.02550	400.00000	8.62000	6.76000	0.68630
13	7.00000	0.02530	405.00000	15.08000	13.19000	0.79500
14	13.90000	0.02580	403.00000	30.92000	27.94000	0.88400
15	16.90000	0.02520	403.00000	38.26000	34.39000	0.90080
16	20.80000	0.02470	402.00000	45.20000	42.10000	0.91350

RUNNO	DAM	MASSTRANS	A	XA1	XA2	XA3
1	0.1040D-03	0.5741D-02	0.4383D-05	0.1033D-02	0.1288D-02	0.9060D-04
2	0.1022D-03	0.4659D-02	0.4457D-05	0.4736D-03	0.6408D-03	0.4033D-04
3	0.9003D-04	0.3151D-02	0.4439D-05	0.4898D-03	0.7077D-03	0.4715D-04
4	0.9794D-04	0.3556D-02	0.4426D-05	0.4700D-03	0.6942D-03	0.4256D-04
5	0.1121D-03	0.5727D-02	0.4505D-05	0.6284D-03	0.8008D-03	0.5547D-04
6	0.1078D-03	0.5195D-02	0.4353D-05	0.7813D-03	0.1004D-02	0.7093D-04
7	0.1102D-03	0.5465D-02	0.4486D-05	0.1082D-02	0.1367D-02	0.1002D-03
8	0.9894D-04	0.2832D-02	0.4417D-05	0.1082D-02	0.1654D-02	0.1155D-03
9	0.1016D-03	0.3767D-02	0.4491D-05	0.1047D-02	0.1463D-02	0.1009D-03
10	0.9963D-04	0.4501D-02	0.4468D-05	0.1092D-02	0.1446D-02	0.9800D-04
11	0.1054D-03	0.4517D-02	0.4634D-05	0.1028D-02	0.1377D-02	0.9414D-04
12	0.1074D-03	0.4612D-02	0.4483D-05	0.4592D-03	0.4766D-03	0.1441D-03
13	0.1302D-03	-0.1111D-01	0.4370D-05	0.4556D-03	0.4402D-03	0.9345D-04
14	0.1043D-03	0.4037D-02	0.4512D-05	0.4646D-03	0.5831D-03	0.5393D-04
15	0.8408D-04	0.2512D-02	0.4592D-05	0.4538D-03	0.7201D-03	0.4505D-04
16	0.1054D-03	0.4715D-02	0.4408D-05	0.4448D-03	0.6021D-03	0.3850D-04

ILM NO. 5 SODIUM CHLORIDE

UNNO	PRESS	FEEDML	FEEDRT	PWP	PRODRT	SEPRTN
1	20.40000	0.05740	390.00000	48.53000	41.43000	0.88440
2	20.40000	0.02630	386.00000	49.22000	45.21000	0.89530
3	17.30000	0.02720	415.00000	41.06000	37.22000	0.87550
4	20.40000	0.02610	406.00000	48.24000	44.34000	0.88520
5	20.40000	0.03490	410.00000	49.04000	44.70000	0.88730
6	20.40000	0.04340	408.00000	47.37000	42.07000	0.88690
7	20.40000	0.06010	404.00000	49.09000	41.37000	0.88280
8	20.40000	0.06010	206.00000	48.17000	39.30000	0.86280
9	20.40000	0.05820	308.00000	49.06000	40.96000	0.87560
10	20.40000	0.06070	405.00000	48.83000	40.76000	0.88300
11	20.40000	0.05710	516.00000	50.26000	42.37000	0.88490
12	3.70000	0.02550	400.00000	9.12000	7.32000	0.62870
13	6.90000	0.02530	405.00000	16.28000	14.23000	0.74420
14	13.70000	0.02580	403.00000	33.35000	30.24000	0.84950
15	16.80000	0.02520	403.00000	41.26000	37.18000	0.86890
16	20.70000	0.02470	402.00000	49.32000	45.96000	0.88510

UNNO	DAM	MASSTRANS	A	XA1	XA2	XA3
1	0.1538D-03	0.5935D-02	0.4825D-05	0.1033D-02	0.1299D-02	0.1196D-03
2	0.1252D-03	0.3797D-02	0.4894D-05	0.4736D-03	0.7063D-03	0.4962D-04
3	0.1302D-03	0.3414D-02	0.4814D-05	0.4898D-03	0.7003D-03	0.6102D-04
4	0.1346D-03	0.3619D-02	0.4796D-05	0.4700D-03	0.7057D-03	0.5399D-04
5	0.1626D-03	0.6630D-02	0.4876D-05	0.6284D-03	0.7846D-03	0.7087D-04
6	0.1511D-03	0.5837D-02	0.4710D-05	0.7813D-03	0.9907D-03	0.8844D-04
7	0.1517D-03	0.5339D-02	0.4881D-05	0.1082D-02	0.1395D-02	0.1269D-03
8	0.1441D-03	0.3099D-02	0.4789D-05	0.1082D-02	0.1633D-02	0.1486D-03
9	0.1470D-03	0.4026D-02	0.4878D-05	0.1047D-02	0.1462D-02	0.1304D-03
10	0.1434D-03	0.4620D-02	0.4955D-05	0.1092D-02	0.1460D-02	0.1280D-03
11	0.1452D-03	0.4694D-02	0.4997D-05	0.1028D-02	0.1384D-02	0.1184D-03
12	0.1585D-03	0.1913D+01	0.4999D-05	0.4592D-03	0.4592D-03	0.1706D-03
13	0.1769D-03	0.3892D-01	0.4786D-05	0.4556D-03	0.4602D-03	0.1165D-03
14	0.1535D-03	0.4494D-02	0.4937D-05	0.4646D-03	0.5749D-03	0.6396D-04
15	0.1236D-03	0.2673D-02	0.4981D-05	0.4538D-03	0.7159D-03	0.5953D-04
16	0.1545D-03	0.4841D-02	0.4832D-05	0.4448D-03	0.6086D-03	0.5114D-04

ILM NO. 6 SODIUM CHLORIDE

UNNO	PRESS	FEEDML	FEEDRT	PWP	PRODRT	SEPRTN
1	20.40000	0.02610	406.00000	51.90000	47.51000	0.87410
2	20.40000	0.03490	410.00000	52.76000	47.84000	0.87750
3	20.40000	0.04340	408.00000	51.28000	45.18000	0.87930
4	20.40000	0.06010	404.00000	52.75000	44.25000	0.87590
5	20.40000	0.06010	206.00000	52.32000	42.04000	0.84690
6	20.40000	0.05820	308.00000	53.07000	43.74000	0.86910
7	20.40000	0.06070	405.00000	52.28000	43.61000	0.87660
8	20.40000	0.05710	516.00000	53.90000	45.21000	0.87830
9	3.60000	0.02550	400.00000	9.66000	7.76000	0.60790
10	5.70000	0.02530	405.00000	17.42000	15.18000	0.73270
11	13.60000	0.02580	403.00000	35.90000	32.37000	0.84150
12	16.70000	0.02520	403.00000	44.34000	39.98000	0.86290
13	20.50000	0.02470	402.00000	53.03000	49.36000	0.87960

RUNNO	CAM	MASSTRANS	A	XA1	XA2	XA3
1	0.1511D-03	0.3434D-02	0.5160D-05	0.4700D-03	0.7412D-03	0.5921D-04
2	0.1795D-03	0.5645D-02	0.5246D-05	0.6284D-03	0.8291D-03	0.7703D-04
3	0.1622D-03	0.4966D-02	0.5099D-05	0.7813D-03	0.1053D-02	0.9400D-04
4	0.1677D-03	0.5128D-02	0.5245D-05	0.1082D-02	0.1434D-02	0.1344D-03
5	0.1611D-03	0.2810D-02	0.5202D-05	0.1082D-02	0.1750D-02	0.1658D-03
6	0.1552D-03	0.3618D-02	0.5277D-05	0.1047D-02	0.1555D-02	0.1373D-03
7	0.1612D-03	0.4732D-02	0.5193D-05	0.1092D-02	0.1472D-02	0.1350D-03
8	0.1595D-03	0.4539D-02	0.5359D-05	0.1028D-02	0.1425D-02	0.1252D-03
9	0.1829D-03	0.9211D-01	0.5443D-05	0.4592D-03	0.4601D-03	0.1801D-03
10	0.1589D-03	0.2703D-01	0.5274D-05	0.4556D-03	0.4625D-03	0.1213D-03
11	0.1653D-03	0.3932D-02	0.5354D-05	0.4646D-03	0.6022D-03	0.7362D-04
12	0.1407D-03	0.2900D-02	0.5335D-05	0.4538D-03	0.7111D-03	0.6230D-04
13	0.1723D-03	0.5021D-02	0.5247D-05	0.4448D-03	0.6144D-03	0.5359D-04

APPENDIX 2

REVERSE OSMOSIS SYSTEM ANALYSIS DATA*

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*Some data are given - these are for systems having actual values of $\lambda\theta$ that could be obtained using the reference solution system NaCl-Water and the apparatus employed in this study.

FILM NO. 1

γ = 0.060 θ = 0.0180 λθ = 5.000 PEB=INFINITE

Δ	C1	C2	C3	$\bar{C}3$	τ OR X
0.0	1.000	1.200	0.2279E-01	0.2279E-01	0.0
0.5000E-01	1.051	1.260	0.2404E-01	0.2350E-01	0.5390E-01
0.1000E+00	1.108	1.328	0.2543E-01	0.2411E-01	0.1080
0.1500	1.172	1.403	0.2699E-01	0.2482E-01	0.1624
0.2000	1.244	1.487	0.2876E-01	0.2558E-01	0.2170
0.2500	1.325	1.582	0.3078E-01	0.2642E-01	0.2720
0.3000	1.417	1.690	0.3311E-01	0.2734E-01	0.3273
0.3500	1.523	1.814	0.3583E-01	0.2835E-01	0.3830
0.4000	1.647	1.958	0.3904E-01	0.2948E-01	0.4393
0.4500	1.793	2.128	0.4289E-01	0.3075E-01	0.4961
0.5000	1.968	2.330	0.4759E-01	0.3219E-01	0.5536
0.5500	2.181	2.574	0.5346E-01	0.3385E-01	0.6120
0.6000	2.446	2.878	0.6101E-01	0.3578E-01	0.6715
0.6500	2.786	3.263	0.7107E-01	0.3809E-01	0.7325
0.7000	3.238	3.769	0.8514E-01	0.4091E-01	0.7955
0.7500	3.866	4.465	0.1062	0.4450E-01	0.8613
0.8000	4.804	5.482	0.1416	0.4935E-01	0.9316
0.8500	6.349	7.111	0.2110	0.5655E-01	1.010
0.9000	9.375	10.15	0.4221	0.6964E-01	1.109

FILM NO. 1

γ = 0.100 θ = 0.0180 $\lambda\theta$ = 5.000 PEB=INFINITE

Δ	C1	C2	C3	$\overline{C3}$	τ OR X
0.0	1.000	1.189	0.2373E-01	0.2373E-01	0.0
0.5000E-01	1.051	1.248	0.2509E-01	0.2449E-01	0.5678E-01
0.10000E+00	1.108	1.314	0.2660E-01	0.2516E-01	0.1140
0.1500	1.172	1.388	0.2832E-01	0.2593E-01	0.1716
0.2000	1.243	1.470	0.3027E-01	0.2676E-01	0.2297
0.2500	1.324	1.563	0.3251E-01	0.2769E-01	0.2884
0.3000	1.416	1.668	0.3512E-01	0.2871E-01	0.3478
0.3500	1.522	1.789	0.3819E-01	0.2984E-01	0.4080
0.4000	1.646	1.929	0.4186E-01	0.3111E-01	0.4691
0.4500	1.792	2.092	0.4631E-01	0.3254E-01	0.5313
0.5000	1.966	2.287	0.5183E-01	0.3418E-01	0.5949
0.5500	2.178	2.523	0.5885E-01	0.3609E-01	0.6602
0.6000	2.442	2.813	0.6811E-01	0.3835E-01	0.7278
0.6500	2.781	3.180	0.8085E-01	0.4110E-01	0.7984
0.7000	3.229	3.660	0.9952E-01	0.4456E-01	0.8733
0.7500	3.853	4.312	0.1294	0.4912E-01	0.9547
0.8000	4.777	5.253	0.1850	0.5566E-01	1.047
0.8500	6.290	6.734	0.3221	0.6657E-01	1.164

FILM NO. 1

γ = 0.140 θ = 0.0180 $\lambda\theta$ = 5.000 PEB=INFINITE

Δ	C1	C2	C3	$\bar{C}3$	τ OR X
0.0	1.000	1.178	0.2476E-01	0.2476E-01	0.0
0.5000E-01	1.051	1.237	0.2623E-01	0.2560E-01	0.5991E-01
0.1000E+00	1.108	1.301	0.2789E-01	0.2632E-01	0.1204
0.1500	1.172	1.373	0.2978E-01	0.2716E-01	0.1816
0.2000	1.243	1.453	0.3194E-01	0.2809E-01	0.2436
0.2500	1.324	1.544	0.3445E-01	0.2911E-01	0.3065
0.3000	1.416	1.647	0.3739E-01	0.3024E-01	0.3705
0.3500	1.521	1.764	0.4088E-01	0.3150E-01	0.4357
0.4000	1.645	1.900	0.4510E-01	0.3293E-01	0.5024
0.4500	1.790	2.059	0.5030E-01	0.3456E-01	0.5709
0.5000	1.964	2.247	0.5686E-01	0.3645E-01	0.6417
0.5500	2.175	2.474	0.6541E-01	0.3867E-01	0.7154
0.6000	2.438	2.753	0.7700E-01	0.4135E-01	0.7929
0.6500	2.774	3.103	0.9361E-01	0.4469E-01	0.8759
0.7000	3.219	3.558	0.1193	0.4903E-01	0.9669
0.7500	3.835	4.172	0.1643	0.5505E-01	1.071
0.8000	4.742	5.048	0.2612	0.6446E-01	1.201
0.8500	6.196	6.406	0.5726	0.8311E-01	1.399

FILM NO. 3

γ = 0.060 θ = 0.0440 $\lambda\theta$ = 2.000 PEB=INFINITE

Δ	C1	C2	C3	$\bar{C}3$	τ OR X
0.0	1.000	1.537	0.7073E-01	0.7073E-01	0.0
0.5000E-01	1.049	1.609	0.7435E-01	0.7261E-01	0.5494E-01
0.1000E+00	1.103	1.688	0.7838E-01	0.7448E-01	0.1101
0.1500	1.163	1.775	0.8289E-01	0.7653E-01	0.1656
0.2000	1.230	1.873	0.8797E-01	0.7874E-01	0.2215
0.2500	1.306	1.983	0.9374E-01	0.8115E-01	0.2777
0.3000	1.393	2.107	0.1004	0.8379E-01	0.3343
0.3500	1.492	2.248	0.1080	0.8669E-01	0.3914
0.4000	1.607	2.410	0.1170	0.8990E-01	0.4490
0.4500	1.742	2.599	0.1276	0.9348E-01	0.5074
0.5000	1.902	2.821	0.1405	0.9753E-01	0.5665
0.5500	2.097	3.087	0.1565	0.1021	0.6266
0.6000	2.339	3.412	0.1766	0.1075	0.6879
0.6500	2.646	3.817	0.2030	0.1138	0.7508
0.7000	3.050	4.338	0.2391	0.1213	0.8158
0.7500	3.608	5.034	0.2917	0.1308	0.8838
0.8000	4.427	6.015	0.3751	0.1432	0.9563
0.8500	5.754	7.515	0.5290	0.1609	1.036

FILM NO. 3

γ = 0.100 θ = 0.0440 $\lambda\theta$ = 2.000 PEB=INFINITE

Δ	C1	C2	C3	$\bar{C}3$	τ OR X
0.0	1.000	1.496	0.7302E-01	0.7302E-01	0.0
0.5000E-01	1.049	1.564	0.7688E-01	0.7502E-01	0.5851E-01
0.1000E+00	1.103	1.639	0.8119E-01	0.7700E-01	0.1175
0.1500	1.162	1.721	0.8603E-01	0.7920E-01	0.1770
0.2000	1.230	1.813	0.9151E-01	0.8158E-01	0.2371
0.2500	1.305	1.916	0.9776E-01	0.8418E-01	0.2978
0.3000	1.391	2.031	0.1050	0.8703E-01	0.3593
0.3500	1.490	2.162	0.1134	0.9017E-01	0.4217
0.4000	1.604	2.312	0.1233	0.9368E-01	0.4851
0.4500	1.738	2.485	0.1352	0.9761E-01	0.5498
0.5000	1.898	2.688	0.1497	0.1021	0.6160
0.5500	2.091	2.930	0.1679	0.1072	0.6840
0.6000	2.330	3.222	0.1913	0.1132	0.7543
0.6500	2.634	3.583	0.2227	0.1203	0.8278
0.7000	3.032	4.042	0.2668	0.1291	0.9054
0.7500	3.579	4.646	0.3337	0.1404	0.9892
0.8000	4.378	5.485	0.4468	0.1556	1.083
0.8500	5.655	6.738	0.6770	0.1786	1.195

FILM NO. 3

γ = 0.140 θ = 0.0440 $\lambda\theta$ = 2.000 PEB=INFINITE

Δ	C1	C2	$\dot{C}3$	$\bar{C}3$	τ OR X
0.0	1.000	1.459	0.7550E-01	0.7550E-01	0.0
0.5000E-01	1.049	1.524	0.7963E-01	0.7765E-01	0.6233E-01
0.1000E+00	1.102	1.594	0.8425E-01	0.7977E-01	0.1254
0.1500	1.162	1.672	0.8947E-01	0.8213E-01	0.1892
0.2000	1.229	1.759	0.9540E-01	0.8469E-01	0.2539
0.2500	1.304	1.855	0.1022	0.8750E-01	0.3196
0.3000	1.390	1.963	0.1101	0.9060E-01	0.3864
0.3500	1.488	2.085	0.1194	0.9403E-01	0.4547
0.4000	1.601	2.225	0.1304	0.9786E-01	0.5245
0.4500	1.735	2.386	0.1438	0.1022	0.5963
0.5000	1.893	2.573	0.1603	0.1072	0.6704
0.5500	2.084	2.794	0.1813	0.1129	0.7475
0.6000	2.320	3.060	0.2088	0.1197	0.8285
0.6500	2.620	3.387	0.2466	0.1279	0.9145
0.7000	3.011	3.799	0.3015	0.1382	1.008
0.7500	3.545	4.336	0.3883	0.1517	1.112
0.8000	4.316	5.074	0.5446	0.1709	1.235
0.8500	5.525	6.170	0.8907	0.2015	1.395

FILM NO. 6

γ = 0.060 θ = 0.0820 $\lambda\theta$ = 2.000 PEB=INFINITE

Δ	C1	C2	C3	$\bar{C}3$	τ CR X
0.0	1.000	1.510	0.1239	0.1239	0.0
0.5000E-01	1.046	1.576	0.1299	0.1269	0.5464E-01
0.1000E+00	1.097	1.649	0.1364	0.1300	0.1095
0.1500	1.153	1.730	0.1438	0.1334	0.1646
0.2000	1.216	1.820	0.1520	0.1370	0.2200
0.2500	1.286	1.921	0.1613	0.1409	0.2758
0.3000	1.366	2.034	0.1719	0.1452	0.3319
0.3500	1.458	2.163	0.1841	0.1498	0.3884
0.4000	1.563	2.311	0.1983	0.1550	0.4454
0.4500	1.687	2.482	0.2151	0.1607	0.5029
0.5000	1.833	2.683	0.2352	0.1671	0.5611
0.5500	2.009	2.922	0.2598	0.1744	0.6202
0.6000	2.226	3.213	0.2906	0.1828	0.6802
0.6500	2.500	3.575	0.3304	0.1925	0.7416
0.7000	2.857	4.038	0.3838	0.2042	0.8046
0.7500	3.345	4.654	0.4597	0.2185	0.8699
0.8000	4.052	5.518	0.5759	0.2369	0.9387

FILM NO. 6

γ = 0.100 $\dot{\theta}$ = 0.0820 $\lambda\theta$ = 2.000 PEB=INFINITE

Δ	C1	C2	C3	$\bar{C}3$	τ CR X
0.0	1.000	1.472	0.1274	0.1274	0.0
0.5000E-01	1.046	1.536	0.1337	0.1306	0.5796E-01
0.1000E+00	1.096	1.605	0.1407	0.1339	0.1163
0.1500	1.152	1.681	0.1484	0.1374	0.1751
0.2000	1.215	1.766	0.1572	0.1413	0.2344
0.2500	1.285	1.860	0.1671	0.1454	0.2943
0.3000	1.364	1.966	0.1785	0.1500	0.3549
0.3500	1.455	2.086	0.1917	0.1550	0.4161
0.4000	1.560	2.223	0.2071	0.1605	0.4783
0.4500	1.682	2.381	0.2254	0.1667	0.5414
0.5000	1.826	2.566	0.2475	0.1736	0.6058
0.5500	2.000	2.784	0.2747	0.1815	0.6717
0.6000	2.214	3.048	0.3093	0.1907	0.7395
0.6500	2.483	3.372	0.3544	0.2014	0.8097
0.7000	2.833	3.783	0.4163	0.2144	0.8831
0.7500	3.308	4.323	0.5062	0.2307	0.9610
0.8000	3.992	5.068	0.6492	0.2520	1.046

FILM NO. 6

γ = 0.140 θ = 0.0820 $\lambda\theta$ = 2.000 PEB=INFINITE

Δ	C1	C2	C3	$\bar{C}3$	τ OR X
0.0	1.000	1.438	0.1312	0.1312	0.0
0.5000E-01	1.046	1.498	0.1378	0.1346	0.6148E-01
0.1000E+00	1.096	1.564	0.1452	0.1380	0.1236
0.1500	1.151	1.637	0.1535	0.1418	0.1863
0.2000	1.214	1.717	0.1628	0.1459	0.2498
0.2500	1.283	1.806	0.1735	0.1503	0.3141
0.3000	1.362	1.905	0.1857	0.1552	0.3795
0.3500	1.452	2.018	0.1999	0.1605	0.4459
0.4000	1.556	2.146	0.2167	0.1665	0.5137
0.4500	1.677	2.292	0.2367	0.1731	0.5830
0.5000	1.819	2.463	0.2611	0.1806	0.6542
0.5500	1.991	2.665	0.2914	0.1893	0.7277
0.6000	2.201	2.906	0.3303	0.1993	0.8042
0.6500	2.465	3.201	0.3819	0.2113	0.8845
0.7000	2.806	3.572	0.4538	0.2259	0.9700
0.7500	3.267	4.054	0.5607	0.2444	1.063
0.8000	3.923	4.714	0.7362	0.2691	1.168

APPENDIX 3

SEPARATION OF SOME ORGANIC COMPOUNDS BY REVERSE OSMOSIS-EXPERIMENTAL DATA

TABLE 9

Separation of Some Organic Compounds[a]

System	Feed Concn. %	Film No. 1	
		Solute Sepn. %	Prod. Rate gpd/sq. ft
Glucose-Water	49.5	-[b]	15.7
	97.5	99.5	15.5
	520	99.6	15.6
	972	99.6	15.3
Sucrose-Water	46.3	>99.9	16.1
	101	>99.9	15.7
	500	>99.9	16.0
	1190	>99.9	15.5
	2110	99.9	15.2
Soluble Starch-Water	13	97.7	15.9
	51.5	99.2	15.8
	110.5	99.6	15.9
Beef Extract-Water	13.2[c]	-[b]	16.0
	55	93.5	15.9
	109	94.3	15.6
	255	96.8	15.3
Glutamic Acid-Water[d]	10	-[b]	16.1
	54.7	99.1	16.0
	116	99.5	15.7
	564	99.9	15.4
Sodium Stearate-Water	61.8	>97.6	15.2
	125	>98.8	15.7
	248	99.4	16.1
	508	99.5	15.9
	961	99.8	16.0
ABS-Water	4.1	98.2	16.1
	57.8	99.1	15.5
	105	99.6	15.2
LAS-Water	2.6	97.9	16.1
	55.2	98.3	16.2
	97.5	99.1	15.9

TABLE 9 Continued

System	Feed Concn. mg/l	Film No. 2	
		Solute Sepn. %	Prod. Rate gpd/sq. ft
Glucose-Water	49.5	-[b]	23.8
	97.5	99.0	23.7
	520	99.0	23.6
	972	99.1	23.3
Sucrose-Water	46.3	99.9	24.3
	101	99.8	23.8
	500	99.8	24.2
	1190	99.8	23.5
	2110	99.7	23.0
Soluble Starch-Water	13	96.9	24.0
	51.5	99.0	24.0
	110.5	99.1	24.1
Beef Extract-Water	13.2[c]	-[b]	24.1
	55	93.0	24.1
	109	93.4	23.6
	255	95.9	22.9
Glutamic Acid-Water[d]	10	-[b]	24.4
	54.7	98.9	24.2
	116	99.4	23.9
	564	99.9	23.3
Sodium Stearate-Water	61.8	>97.6	22.9
	125	>98.8	24.0
	248	99.0	24.5
	508	99.4	23.8
	961	99.6	24.0
ABS-Water	4.1	98.2	24.6
	57.8	99.1	23.7
	105	99.6	23.0
LAS-Water	2.6	97.9	24.8
	55.2	98.3	25.1
	97.5	99.1	24.4

TABLE 9 Continued

System	Feed Concn. mg/l	Film No. 3	
		Solute Sepn. %	Prod. Rate gpd/sq. ft
Glucose-Water	49.5	-[b]	31.2
	97.5	97.6	30.9
	520	98.7	30.8
	972	98.9	30.4
Sucrose-Water	46.3	99.7	31.7
	101	99.7	31.1
	500	99.7	31.6
	1190	99.7	30.7
	2110	99.5	30.0
Soluble Starch-Water	13	95.0	31.3
	51.5	98.6	31.2
	110.5	99.2	31.4
Beef Extract-Water	13.2[c]	-[b]	31.5
	55	91.5	30.6
	109	92.5	30.5
	255	95.3	29.7
Glutamic Acid-Water[d]	10	-[b]	31.7
	54.7	98.7	31.5
	116	99.3	31.1
	564	99.9	30.4
Sodium Stearate-Water	61.8	>97.6	29.9
	125	>98.8	31.1
	248	98.8	31.7
	508	98.9	31.2
	961	99.0	31.4
ABS-Water	4.1	98.2	32.0
	57.8	99.1	30.8
	105	99.6	30.1
LAS-Water	2.6	97.9	32.5
	55.2	98.3	32.6
	97.5	99.1	31.6

TABLE 9 Continued

System	Feed Concn. mg/l	Film No. 4	
		Solute Sepn. %	Prod. Rate gpd/sq. ft
Glucose-Water	49.5	-[b]	35.5
	97.5	97.2	35.3
	520	98.2	34.7
	972	98.2	34.7
Sucrose-Water	46.3	99.6	36.1
	101	99.6	35.5
	500	99.6	36.0
	1190	99.6	35.1
	2110	99.4	33.9
Soluble Starch-Water	13	92.3	35.2
	51.5	97.7	35.1
	110.5	98.8	35.1
Beef Extract-Water	13.2[c]	-[b]	35.6
	55	90.2	34.8
	109	92.0	34.4
	255	94.5	33.4
Glutamic Acid-Water[d]	10	-[b]	35.9
	54.7	98.4	35.7
	116	99.1	35.3
	564	99.8	34.3
Sodium Stearate-Water	61.8	>97.6	31.0
	125	98.7	33.5
	248	97.5	33.0
	508	98.5	32.3
	961	98.6	32.6
ABS-Water	4.1	96.8	36.3
	57.8	97.8	34.9
	105	98.2	34.1
LAS-Water	2.6	96.3	36.8
	55.2	96.2	36.9
	97.5	96.7	35.8

TABLE 9 Continued

System	Feed Concn. mg/l	Film No. 5 Solute Sepn. %	Film No. 5 Prod. Rate gpd/sq. ft
Glucose-Water	49.5	-[b]	38.5
	97.5	96.4	38.4
	520	97.4	37.9
	972	97.5	37.8
Sucrose-Water	46.3	99.5	39.1
	101	99.4	38.4
	500	99.4	39.0
	1190	99.5	38.0
	2110	99.3	36.9
Soluble Starch-Water	13	90.0	38.2
	51.5	96.7	37.9
	110.5	98.6	37.9
Beef Extract-Water	13.2[c]	-[b]	38.6
	55	89.9	37.7
	109	-[b]	37.2
	255	94.0	36.1
Glutamic Acid-Water[d]	10	-[b]	39.0
	54.7	98.5	38.8
	116	98.9	38.3
	564	99.7	37.4
Sodium Stearate-Water	61.8	>97.6	35.8
	125	98.0	37.5
	248	97.3	37.4
	508	98.4	37.4
	961	97.5	37.2
ABS-Water	4.1	96.8	39.7
	57.8	97.8	38.1
	105	98.2	37.1
LAS-Water	2.6	96.3	40.6
	55.2	96.2	40.5
	97.5	96.7	39.1

TABLE 9 Continued

System	Feed Concn. mg/l	Film No. 6 Solute Sepn. %	Film No. 6 Prod. Rate gpd/sq. ft
Glucose-Water	49.5	-[b]	41.4
	97.5	95.9	41.3
	520	96.9	40.4
	972	97.2	40.5
Sucrose-Water	46.3	99.4	41.9
	101	99.3	41.2
	500	99.3	41.8
	1190	99.3	40.8
	2110	99.5	39.5
Soluble Starch-Water	13	88.5	40.8
	51.5	95.9	40.6
	110.5	98.5	40.6
Beef Extract-Water	13.2[c]	-[b]	41.2
	55	88.5	40.3
	109	90.5	39.7
	255	93.1	38.6
Glutamic Acid-Water[d]	10	-[b]	41.6
	54.7	98.4	41.4
	116	99.0	41.0
	564	99.7	39.9
Sodium Stearate-Water	61.8	>97.6	35.6
	125	98.0	38.8
	248	97.3	38.8
	508	98.1	38.3
	961	97.9	38.0
ABS-Water	4.1	96.8	42.3
	57.8	97.8	40.6
	105	98.2	39.6
LAS-Water	2.6	96.3	42.7
	55.2	96.2	42.5
	97.5	96.7	41.2

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[a] Film type: CA-NRC-316 (Batch 316)
Operating pressure: 300 psig; Feed flow rate: 400 cm^3/min

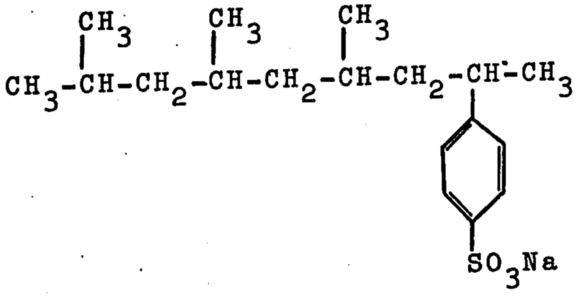
[b] Not determined.

[c] As organic nitrogen.

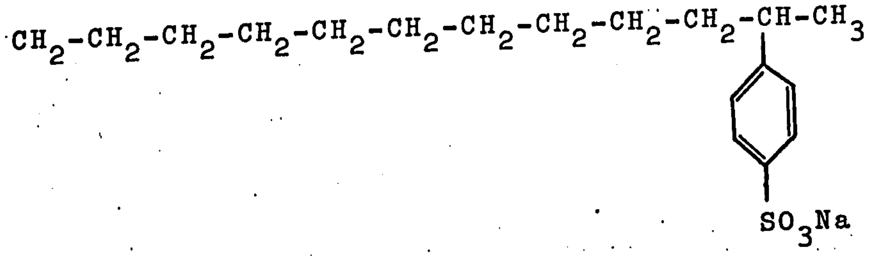
[d] Feed pH adjusted at 6.0 ± 0.1

TABLE 10

Structural Formulae of Some Organic Compounds Studied

Compound	Formula
Glucose	$C_6H_{12}O_6$
Sucrose	$C_{12}H_{22}O_{11}$
Soluble starch	$(C_6H_{10}O_5)_x$ x ≈ 100 to 1000
Glutamic acid	$CHNH_2(CH_2)_2(COOH)_2$
Sodium stearate	$C_{18}H_{35}COONa$
Alkyl benzene sulfonate (ABS)	

Linear alkylate sulfonate (LAS)



Chain lengths in typical LAS: $C_{11}-C_{14}$

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