INSTRUMENTATION IN THE FOOD AND BEVERAGE INDUSTRY VOLUME 2

VOLUME 2
1973

INSTRUMENTATION IN THE FOOD AND BEVERAGE INDUSTRY VOLUME 2

Proceedings of the Second International ISA
Food Instrumentation Division Symposium
June 4-6, 1973, Montreal, Canada

Sponsored by the
Montreal Section, Instrument Society of America

Edited by

David S. Harding

Taylor Instrument Company Ltd.

INSTRUMENT SOCIETY OF AMERICA
Pittsburgh

PUBLICATION POLICY

Technical papers may not be reproduced in any form without written permission from the Instrument Society of America. The Society reserves the exclusive right of publication in its periodicals of all papers presented at the Annual ISA Conference, at ISA Symposia, and at meetings co-sponsored by ISA when the Society acts as publisher. Papers not selected for such publication will be released to authors upon request.

In any event, following oral presentation, other publications are urged to publish up to 300-word excerpts of ISA papers, provided credits are given to the author, the meeting, and the Society using the Society's name in full, rather than simply "ISA".

For further information concerning publications policy write to:

Publications Department
Instrument Society of America
400 Stanwix Street
Pittsburgh, Pa. 15222

Library of Congress Catalog Card No. 72-90018
ISBN 87664-205-9

FOREWORD

Now, more than any other time in history, food is in the news as a subject of concern to all people. It has been said that there will always be a demand for food; but, with recent developments, it is not inconceivable that processed foods could become a luxury to many. There is, then, a real necessity to improve the efficiency of food processing through the use of modern technology. In presenting these Food and Beverage Instrumentation symposia, the Food Industry Division of the Instrument Society of America hopes to promote the exchange of ideas, to disseminate information, and to increase interest in the ISA among food processors. We feel confident that the excellent atmosphere and content of the 1973 symposium went a long way toward achieving these goals. Some in the Food and Beverage industries have been using sophisticated automatic control systems for many years. There are still, however, large areas of potential for use of control and automation techniques. We hope that this symposium has contributed to the exploration of new applications of instrumentation to aid the industry in improving product quality and uniformity while at the same time reducing costs.

The Second Annual Food and Beverage Instrumentation Symposium was hosted by the Montreal Chapter of the Instrument Society of America. The Committee, chaired by Phil McAsey (Beckman Instruments Inc.), turned in its second fine performance in as many years. Thanks are extended to Don Deemer (Campbell Soup Co.), Ed Nobrega (Foxboro Co.), Oscar Soroko (Fischer & Porter Co.), and Irwin Wecker (Amstar Corp.), who, as session developers, worked hard and long at obtaining speakers and seeing to the smooth running of the proceedings. Also a great help and deserving of our thanks was Al Papaionnou (Masonielan International Inc.), Director of the Food Industry Division of ISA.

David S. Harding
Program Editor

CONTENTS

MEMBRANE PROCESSING IN THE FOOD INDUSTRY

Michael E. Matthews
Research Assistant
Department of Food Science

Clyde H. Amundson
Professor
Department of Food Science

Simeon E. So
Research Specialist
Department of Food Science

Charles G. Hill, Jr.
Professor
Department of Chemical Engineering

University of Wisconsin - Madison

ABSTRACT

Ultrafiltration and reverse osmosis are pressure-driven membrane processes which effect fractionation or concentration of multicomponent liquid feed streams on the basis of molecular size. Synthetic, anisotropic membranes cast against a porous backing permit the passage of solvent and low molecular weight solutes to an extent dictated by membrane pore size. This may be varied by the manufacturer according to process application. Developed from desalination technology, membranes today can be utilized in several areas of interest to the waste systems engineer.

Work at the University of Wisconsin - Madison has essentially been concerned with concentrating cheese wheys and skim milk and recovering nutritionally valuable whey proteins. Data describing these processes are presented, together with recent information relating to changes in undenatured whey protein contents of cheese wheys during ultrafiltration.

Rapid monitoring of whey composition by an ultrasonic method has been investigated and appears promising. Equations permitting precise determinations of total lipid, solids-not-lipid, and protein contents of wheys on the basis of two simple triggering frequency (indirect sound velocity) measurements have been developed.

INTRODUCTION

Modern membrane processing technology has developed from pioneering work during the last fifteen years in desalination. On the basis of interface adsorption theory, it has been known for many years that in aqueous sodium chloride solutions, there exists a monomolecular layer of pure water at the air - solution interface, owing to the negative adsorption characteristic of the solute. It was first suggested in 1956 that skimming this layer of water from a salt solution might form the basis of a desalination technique. The Gibbs equation relating interfacial tension of a solution and the adsorption of solutes at interfaces indicated that the nature of the interface itself determined in part the thickness of the pure water layer. Hence, if an interfacial material were utilized which combined preferential solute rejection with water sorption, the problem of pure water recovery would be simplified.

This essentially was the rationale of the reverse osmosis research program established at the University of California - Los Angeles in 1956[14]. The concept is schematically depicted in Figure 1 in which a salt solution under pressure is in contact with a thin, porous film (the interfacial material) which has pores of some critical diameter for removal of water with minimum entrainment of solute. For maximum desalination rate this film or membrane should possess

a) preferential solute rejection
b) preferential water sorption
c) maximum number of pores of critical diameter per unit area
d) minimum practical resistance to transmembrane fluid flow

Anisotropism was therefore considered a desirable characteristic. Ideally, the solution - side surface of the membrane containing the pores should be very thin with the remainder of the film being spongy so as to offer little resistance to fluid flow.

Extensive research was conducted into types of film materials which might meet these criteria[15]. Of those tested, cellulose acetate membranes were found to possess the most efficient solute - solvent separation capacity, commensurate with acceptable product flow rates. From a commercial standpoint, the major development was considered to be the production of a membrane cast from an aqueous preparation of cellulose acetate and acetone, with magnesium perchlorate added to assist in attaining any desired degree of porosity. Such a film was known as the Loeb and Sourirajan mem-

Superior numbers refer to similarly-numbered references at the end of this paper.

brane[15] after its developers and essentially provided the starting point for commercial development of membrane processing systems. Numerous other materials have been tested, including cellophane, cellulose propionate, and cellulose acetate butyrate, with other solvents and chemical additives. However, none has proved as effective as cellulose acetate which to date remains the material of choice in commercial systems.

REVERSE OSMOSIS AND ULTRAFILTRATION

Early workers recognized that membranes offered wider promise than use in desalination, although potable water production has been a major interest to date and research in this field has provided much of the basic technology. However, the modern approach is to regard membrane processing as a discrete unit operation of multi - functional application.

Reverse osmosis and ultrafiltration are essentially variations on the same theme. As noted, pore size is variable at the time of manufacture and this has led to the concept of "tight" versus "loose" membranes, as depicted in Figure 2. "Tight" or reverse osmosis membranes have low porosity and are designed to remove water only (usually with trace quantities of solute) whereas "loose" or ultrafiltration membranes permit the passage of low molecular weight solutes such as mono- and disaccharides. Ultrafiltration is therefore considered more as a fractionation method. Generally, it is customary to specify membrane separation ability in terms of a molecular weight cut - off range which will indicate the approximate minimum size of molecule rejected. It is important to note, however, that membrane separations are not based solely on pore size. Solute rejection, especially in the case of reverse osmosis, is significant as separation capacity cannot be explained simply by differences in membrane porosity.

Consideration of basic flux equations also provides a means of differentiating reverse osmosis and ultrafiltration. Transmembrane solvent and solute fluxes, J_1 and J_2, may be expressed by[1]

$$J_1 = \frac{\bar{P}_1}{t_m} (\Delta P - \Delta \pi) \qquad (1)$$

$$J_2 = \frac{\bar{P}_2}{t_m} (C_B - C_P) \qquad (2)$$

where $\bar{P}_1$ and $\bar{P}_2$ are the specific permeabilities of the membrane to solvent and solute, respectively; ΔP is the hydraulic pressure drop across the membrane; $\Delta \pi$ is the osmotic pressure difference between the upstream and permeate sides of the membrane; t_m is the membrane thickness; C_B is the solute concentration on the upstream side; and C_P is the solute concentration in the permeate. Mass balance also requires that

$$J_2 = J_1 C_P \qquad (3)$$

Rearranging (1), (2), and (3) yields

$$\sigma = 1 - \frac{C_P}{C_B} = \frac{[\bar{P}_1/\bar{P}_2 (\Delta P - \Delta \pi)]}{[1 + \bar{P}_1/\bar{P}_2 (\Delta P - \Delta \pi)]} \qquad (4)$$

where σ is the solute rejection coefficient.

From Equation (1), ΔP must be sufficient to overcome $\Delta \pi$ in order to have solvent flux and it is this requirement which has led to the commonly used term, reverse osmosis. For sea water, $\Delta \pi$ is approximately 350 psia so that substantially greater hydraulic pressure drops must be employed to effect desalination. In ultrafiltration, however, $\Delta \pi$ values are generally much lower. For example, when fractionating whey, the permeate contains approximately the same concentrations of lactose and inorganic salts as the upstream fluid so that the protein concentration gradient is the primary contributor to the osmotic pressure difference. Hence, as the osmotic pressure of protein solutions is usually very low, $\Delta \pi$ is low and in some ultrafiltration equipment, ΔP need be only 40 psig to effect fractionation.

Equation (4) implies that solute rejection is hyperbolically related to ΔP. Provided that solute rejection is considerably greater than solvent rejection at quite low pressures, then $\sigma \rightarrow 1.0$ as ΔP increases, or solute rejection is high.

CONCENTRATION POLARIZATION

From Equation (1), transmembrane flux would be expected to increase linearly with applied hydraulic pressure. However, this is not the case, especially during the membrane processing of protein-containing fluids. The rate of increase in flux is observed to decline with increasing ΔP to a point beyond which flux is relatively independent of applied pressure. The phenomenon responsible is termed concentration polarization and it is schematically depicted in Figure 3. Solute, such as protein, has a tendency to gel in the region of the membrane, in part because back diffusion of the large, low - diffusivity molecules is slower than their rate of movement to the wall. A gel layer is developed, the thickness of which is in equilibrium with the upstream solids concentration, C_B, and the applied hydraulic pressure, ΔP. This gel layer contributes to the overall resistance to mass transfer and may, with concentrated upstream fluids, be the greatest component of resistance. It is the primary cause of the normally - observed decline in flux rate when concentrating cheese whey by reverse osmosis and as such is a major factor in operating economics.

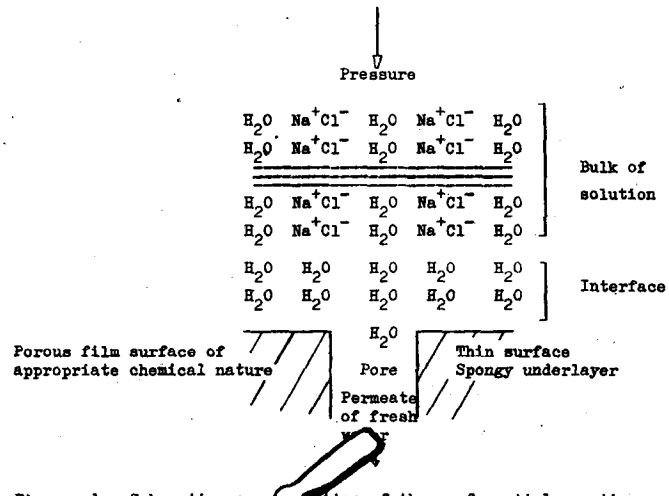

Figure 1. Schematic representation of the preferential sorption - capillary flow mechanism. (Sourirajan[14])

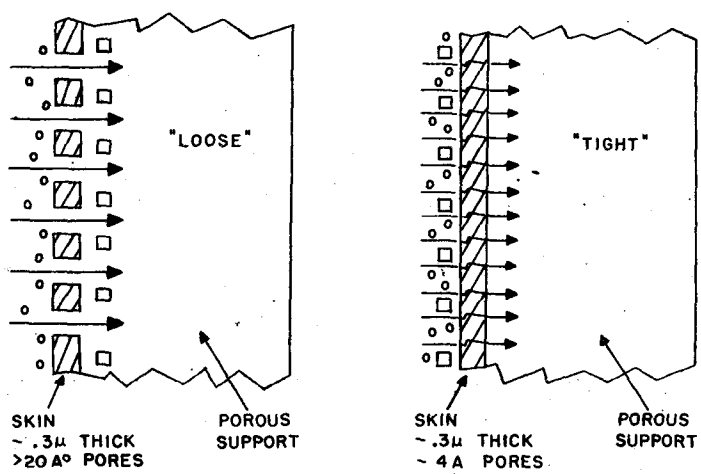

Figure 2. Schematic representation of ultrafiltration ("loose") and reverse osmosis ("tight") membranes.[6]

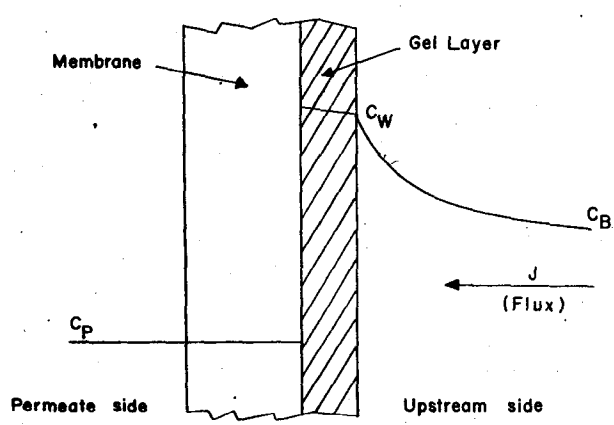

Figure 3. Schematic representation of concentration polarization. Symbols C_B, C_W, and C_P refer to solute concentrations in the bulk upstream fluid, at the gel layer, and in the permeate, respectively.

Minimizing concentration polarization is an important consideration in equipment design and it is generally achieved by promoting turbulent flow over the membrane surface. High pumping rates may be employed but in several systems turbulence promoters are installed. For example, in the tubular configuration utilized by Havens, a spirally constructed nylon rod is inserted into each membrane-containing tube. American Standard reverse osmosis units contain spherical nylon balls within the tubes to promote turbulence.

Lowering feed stream viscosity will also increase Reynold's Numbers as well as increase molecular diffusivities. The Abcor whey processing equipment at Crowley Foods, Inc., at LaFargeville, New York[7] processes cottage cheese whey received, without cooling, from the cottage cheese cooking operation. For cellulose acetate membranes, however, there is a practical operating temperature limit of 130F on account of higher rates of ester hydrolysis and concomitantly reduced membrane lifetimes at higher temperatures.

The causes and control of concentration polarization have been well described by Blatt *et al*.[1]

APPLICATIONS OF MEMBRANE PROCESSING

Numerous suggestions have been made for applications of membranes to separations problems. An approximate categorization of these proposals is given below[3].

a) Biological. Recovery of products such as pharmaceuticals and enzymes from fermentation media and bacterial cultures; ultrafiltration of human blood plasma.

b) Waste treatment. Reduction in chemical - oxygen demand (COD) of waste streams from food processing plants *e.g.* dairy, potato starch.

   Tertiary sewage treatment for reduction of inorganic compound levels.

c) Desalination. Obtaining potable water from sea and brackish water.

d) Food and beverage. Concentration and/or fractionation of fruit juices, maple sap, egg white, cheese wheys, and skim milk. Enzyme processing.

Specific examples

Reverse osmosis concentration of potato starch industry waste streams

Porter *et al*.[13] described pilot scale experiments in which dilute, secondary waste water (0.5% solids, wet basis) from a potato starch process was concentrated four - fold by reverse osmosis. Using medium porosity membranes in a Havens "Osmotik" unit, a 98% reduction in COD was reported. Heat and acid coagulation were proposed as means of recovering protein from the concentrated material, thereby providing a potentially saleable product to partially offset processing costs.

The COD of the secondary waste from a 30 ton-per-day starch plant is equivalent to that of a city of 85,000 people[13]. Current disposal procedures typically involve lagooning as the secondary waste treatment, with a COD reduction of approximately 80%. Potato solids however are comparatively resistant to degradation so that large lagoon areas are required with concomitant odor problems.

Reverse osmosis was considered by the authors to be a future possibility for waste treatment. However, using 1970 costs, they estimated that the capital cost of the membranes alone to process the 432,000 U.S. gallons-per-day of waste water from a 30 ton-per-day starch plant would be over $300,000. Membrane replacement costs, valves, pumps, membrane support frames, and piping were not included in this figure.

The process, then, was considered functional but expensive. It was predicted that lower capital costs and interest in other areas of the potato industry would make investment more attractive.

Maple sap concentration

Maple syrup is produced from maple sap by open pan evaporation. To produce one gallon of syrup, thirty three gallons of water must be removed by boiling. Fuel costs are therefore high, typically being greater than 50 cents per gallon of syrup. Heating is essential, however, to develop the characteristic flavor and texture of the finished product.

Studies conducted by the Eastern Utilization Research and Development Division, USDA, indicated that the water removal cost could be approximately halved by halving the initial volume of sap by reverse osmosis prior to thermal processing without adversely affecting flavor[16]. The major advantage related to respective energy costs in removing water. By thermal distillation, the energy cost was determined to be 1.50¢ per gallon of water removed. By contrast, the cost to remove the same amount of water by reverse osmosis was 0.06¢ which resulted from the electrical requirement in operating the membrane system feed pump. Other operating costs such as membrane replacement were not given but the overall prediction for the method's utility was favorable.

Egg white concentration

Egg white albumen coagulates rapidly at temperatures greater than 140F which precludes efficient vacuum pan concentration prior to spray drying. The egg white market is large, however, its value in the U.S being greater than $30 million annually.

Approximately 65% of this is marketed as the dried product[10].

Ultrafiltration has been investigated as a means for concentrating egg white protein with simultaneous partial glucose removal[10]. The low pressures involved minimize protein damage during depressurizing and, provided shear stress induced by pumping is low, the concentrated egg white proteins are essentially undenatured. Concentration to 30% solids, wet basis, has been achieved without affecting important functional properties.

ULTRAFILTRATION AND REVERSE OSMOSIS OF DAIRY PRODUCTS

The magnitude of the cheese whey disposal problem has been frquently cited. World - wide production of whey has been estimated at 50 billion pounds[2] and the Office of Saline Water has reported that approximately 20 billion pounds of liquid whey are currently disposed of annually into waste receiving systems in the United States[9]. The high biological oxygen demand (BOD) of this fluid (typically 50,000 to 60,000 ppm) makes it a serious source of water pollution and it is this realization which in part has led to extensive investigations of systems designed to reduce the BOD of cheese plant effluents while recovering potentially useful whey components.

The application of membrane processes to the concentration and fractionation of whey has been extensively researched during the last six years[4, 6, 11]. Progress has been such that commercial scale processing is now considered feasible. Crowley Foods, Inc., at LaFargeville, New York, has a plant designed to process 300,000 pounds of cottage cheese whey per day[7]. Overseas, the New Zealand Co-operative Dairy Company has commissioned an Abcor system designed ultimately to process 500,000 US gallons of acid casein whey daily using sanitary CIP equipment[8].

Characteristics of membrane systems designed to process dairy fluids

Whey

Cheese whey typically contains 5 - 7% total solids (wet basis) which is composed of 12 - 15% protein (%N x 6.38), 4 - 9% incombustible matter (mostly inorganic salts), 76 - 79% lactose, and, in the case of whey from whole milk cheeses, approximately 4% fat in the dry matter. This fat is normally separated as cream from the whey after cheese making, owing to its high value and ease of separation, so that whey for subsequent processing is essentially defatted and also free of cheese "fines". These latter characteristics are desirable as any unemulsified fat and insoluble particulate matter such as precipitated casein will interfere with membrane processing by blocking the pores.

Ultrafiltration

Ultrafiltration of cheese whey will yield, with time, a retentate which is concentrated in protein but reduced, on a solids basis, in those components which are free to pass through the membrane, namely, inorganic ions and lactose. A typical plot depicting component concentrations versus time is given in Figure 4. Total solids and protein concentration rise sharply as concentration proceeds whereas lactose and ash contents, as fractions of the total solids, decline. This is a well established pattern and has been obtained in our work with both Havens and Abcor pilot scale equipment (up to 80 square feet of membrane area).

Increasing concentration polarization from the steadily increasing feed solids causes a decline in permeate flux rate, as illustrated in Figure 5. This particular plot also depicts the concomitant rise in inlet pressure which occurs in a Havens unit when the outlet pressure is held constant.

Using both Havens and Abcor pilot scale equipment, we have regularly produced concentrates which contain 80 - 85% protein (%N x 6.38) in the dry matter from cottage, brick, and cheddar cheese wheys. Over nine months' operation we have been able to restore initial water flux rates by routinely following manufacturers' recommended cleaning and sanitizing procedures.

On a commercial scale, Abcor engineers assume a satisfactory concentration of protein to be 20 - fold, for which operating costs would be approximately 10 - 20 ¢ per pound of protein, depending on plant capacity[8].

The whey protein concentrates obtained from both Abcor and Havens pilot scale units contain negligible quantities of denatured protein, as measured by both acid precipitation and polyacrylamide gel electrophoresis (PAGE). Figure 6 depicts the change in undenatured whey protein nitrogen (as a fraction of total non - casein nitrogen) with time, as measured by the acid precipitation technique of Wyeth[18]. The linear plot indicates that no appreciable denaturation was occurring, despite repeated exposure of the protein molecules to the shearing action of the feed pump. The linear increase results from a steady decline in the levels of those nitrogenous compounds which are free to pass through the membrane. In cheddar cheese whey, nitrogen - containing compounds which are soluble in 12% trichloracetic acid may amount to over 30% of the total nitrogen.

Corroboration of minimum protein damage was provided by PAGE patterns of whey protein concentrates from both processing units. The gel patterns in Figure 7 (obtained using the discontinuous gel system of Melachouris[12] as adapted for a vertical slab apparatus) indicate good retention of protein separation patterns during ultrafiltration. By contrast, the PAGE pattern of protein from a conventionally produced spray dried whey powder

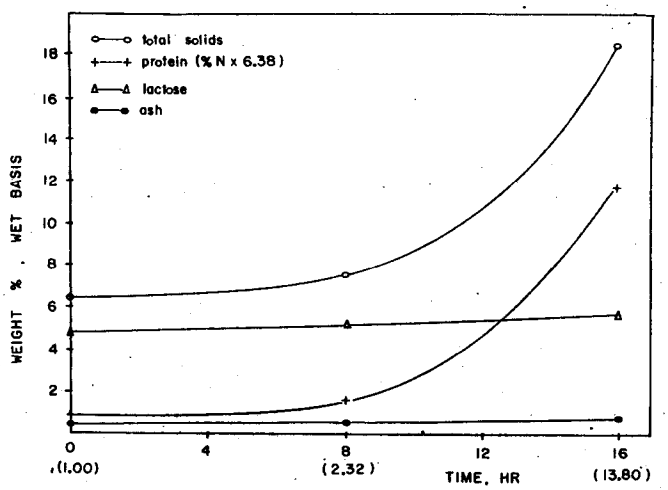

Figure 4. Ultrafiltration of cheddar cheese whey. Changes in retentate composition with time (hr). Volumetric concentration ratios corresponding to the times shown are in parentheses.

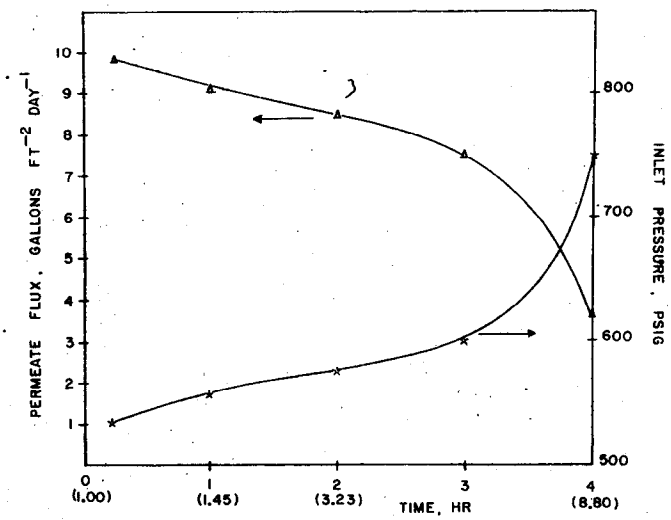

Figure 5. Ultrafiltration of Brick cheese whey. Changes in permeate flux rate (gallons per square foot per day) and inlet pressure (psig) ~~versus~~ time (hr). Volumetric concentration ratios corresponding to the times shown are in parentheses.

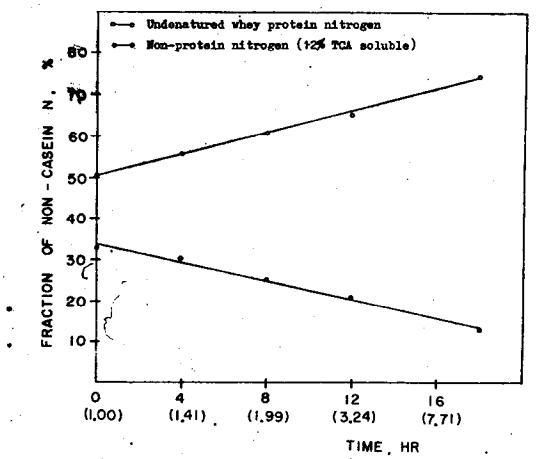

Figure 6. Changes in undenatured whey protein nitrogen and non-protein nitrogen (12% TCA soluble) with time during ultrafiltration of cheddar cheese whey. Nitrogen values are expressed as fractions of the total non-casein nitrogen. Numbers in parentheses are volumetric concentration ratios corresponding to the times shown.

(for which fluid whey was preconcentrated to 35% solids in a rising film, vacuum pan evaporator prior to spray drying) exhibits some loss of resolution on account of heat damage.

Reverse osmosis

Reverse osmosis has been proposed as a means for concentrating whey at small cheese factories in preparation for shipment to larger, centralized whey processing plants. Current economics indicates that concentration of whey by reverse osmosis is cheaper than vacuum pan evaporation up to approximately 12% total solids, or to halving the initial whey volume. Trucking costs would be reduced and the small plant would no longer be forced to directly dispose of its whey as waste.

The Crowley Foods, Inc., plant utilizes reverse osmosis as the second stage of their two - stage process. Permeate from the ultrafiltration step is high in BOD, on account of the unimpeded transmembrane movement of lactose. The purpose of the first step is to recover nutritionally valuable whey protein rather than to effect BOD reduction. A "tight" membrane system is used as the second stage to concentrate permeate from the ultrafiltration step in preparation for lactose recovery. The absence of protein in the permeate reduces the extent of serious concentration polarization during reverse osmosis(7).

Table 1 summarizes composition data from such a two - stage sequence(3).

Skim milk

Less attention has been given to skim milk processing than to whey, although research into operating characteristics has been conducted(5). The high initial solids level (approximately 9%) and large concentration of micellar casein (usually 2.5%) led to early predictions of greater concentration polarization with attendant decline in permeate flux rate. However, useful products can be produced with less volumetric concentration than is required for whey because of the higher initial protein level(8). Based on extensive work, Fenton-May et al. stated that skim milk could be concentrated to 22% total solids by reverse osmosis or fractionated to give 50 - 80% protein (dry basis) liquid concentrates by ultrafiltration(5).

In recent, unpublished work at the University of Wisconsin, Matthews concentrated 850 pounds of skim milk by ultrafiltration to produce a concentrate containing 53.0% protein (%N x 6.38), dry basis. Flux rates and inlet pressure data are shown in Figure 8. Initial flux rate was comparatively low at 4.37 gallons per square foot of membrane surface per day (gfd). This declined by approximately 20% in obtaining a volumetric concentration ratio of 2.44 over 4 hr of operation. However, operating temperature was low (40F) and the total bacterial plate count of the concentrated milk was fewer than 1000 per ml. Cheesemaking experiments are planned with skim milk concentrated in this manner. Compositional changes during ultrafiltration of skim milk are depicted in Figure 9 and, as expected, follow an essentially similar pattern to that of whey.

The 500 pounds of permeate collected in the above trial had a solids content of 4.5%, wet basis, and a pH of 6.71. Its solids composition was 9.3% incombustible matter, 3.6% nitrogenous compounds, and 87.1% lactose. It was combined with 250 pounds of cottage cheese whey (pH 4.5) and concentrated by a falling film vacuum pan evaporator to 54% total solids. This concentrated product was used to fortify beverage skim milk at four levels : 1/2, 1, 1 1/2, and 2% added solids. These milks were evaluated by a thirty-member preference flavor panel after one day and again after eight days' storage at 35F. Panelists were asked to give a preference rating to the samples using a good quality reference skim milk fortified with 1 1/2% added skim milk solids. The results obtained are summarized graphically in Figure 10. Most commercial levels range between 1/2 and 1% fortification and in this region, the skim milks fortified with the blend of sweet permeate and acid whey were not significantly different from the reference. At the higher levels of addition, panelists were able to detect a slightly sour - salty character. Overall, the results were considered encouraging from the standpoint of utilizing what may otherwise have been considered to be waste materials. Also, removing permeate from skim milk prior to cheesemaking will increase the yield of cheese per batch as well as reduce the volume of acid whey. Several experiments in this area are planned.

MONITORING OF RETENTATE COMPOSITION DURING MEMBRANE PROCESSING

An associated development at the University of Wisconsin has been the use of a Solution Analyzer to provide rapid estimates of total solids, lipid, solids-not-lipid, and protein in the concentrated fluid(17). The Solution Analyzer employs an acoustic principle in that it measures the so-called triggering frequency of a sound wave projected through the sample of interest. In the analyzer, a voltage pulse of short duration is produced by an oscillator to a ceramic element which acts as a transducer. Through piezoelectric action, the transducer is deformed and vibrates at high frequency. A short pulse of compressional energy is produced which travels through the sample, is reflected to a receiving transducer, and is returned to the oscillator. The returning signal activates another pulse and the number of activations per second is termed the triggering frequency, f_t, where t is the temperature of analysis. Any changes in f_t reflect changes in sample composition provided other conditions, such as temperature, are held constant.

To obtain measurements, a standard f_t value is

obtained for distilled water at two temperatures, 33 and 65C. New f_t values are obtained with the sample of interest in the sample cup instead of water and the changes in triggering frequencies from those of water, Δf_t, are obtained. From these values, approximations of total solids, lipid, solids-not-lipid, and protein can be made by solving the simultaneous equations

$$\Delta f_t = b_1 X + b_2 Y + C$$

$$\text{for } t = 33 \text{ and } 65C$$

where Δf_t is the change in triggering frequency at 33 or 65C; b_1 is the regression coefficient for extractable lipids at 33 or 65C; X is the extractable lipids content; b_2 is the regression coefficient for solids-not-lipid or protein at 33 or 65C; Y is the solids-not-lipid or protein content; and C is the ordinate at 33 or 65C.

Hence, Δf_t is related directly to sample composition. Statistical comparison of results obtained using the Solution Analyzer and standard methods of analysis indicated that the method was both accurate and precise. Table 2 summarizes statistical data for total solids and protein determinations.

SUMMARY

Membrane processing is a recent development and much investigation of the potential applications is still pilot scale. Commercial application in the dairy industry is now considered feasible where it is used to produce undenatured whey protein concentrates and to reduce the BOD of dairy waste streams. To meet 3A sanitary standards, capital costs are high but when compared to other concentrating methods, operating costs are quite low. It is hoped that advances in sanitary aspects and overall design improvements will open the way to broader food industry acceptance.

ACKNOWLEDGEMENT

The generous support lent by Abcor, Inc., Cambridge, Mass., and Havens Corp., San Diego, Calif., in freely providing pilot scale equipment for experimentation is gratefully acknowledged.

BIBLIOGRAPHY

(1) Blatt, W. F., A. Dravid, A. S. Michaels, and L. Nelsen. 1970. Solute polarization and cake formation in membrane ultrafiltration : causes, consequences, and control techniques, p. 47 -97. In Flinn, J. E. [ed.] Membrane science and technology. Plenum Press, New York.

(2) Chian, E. S. K. and J. T. Selldorff. 1969. Ultrafiltration of biological materials. Process Biochem. 4(9) : 47 - 51.

(3) deFilippi, R. P. and R. L. Goldsmith. 1970. Application and theory of membrane processes for biological and other macromolecular solutions, p. 33 - 46. In Flinn, J. E. [ed.] Membrane science and technology. Plenum Press, New York.

(4) Fenton-May, R. I., C. G. Hill, Jr., and C.H. Amundson. 1971. Use of ultrafiltration/ reverse osmosis systems for the concentration and fractionation of whey. J. Food Sci. 36 : 14 - 21.

(5) Fenton-May, R. I., C. G. Hill, Jr., C. H. Amundson, M. H. Lopez, and P. D. Auclair. 1972. Concentration and fractionation of skim milk by reverse osmosis and ultrafiltration. J. Dairy Sci. 55 : 1561 - 1566.

(6) Horton, B. S. et al. 1970. New method for economic control of pollution caused by cheese wheys and recovery of cheese whey proteins by ultrafiltration. SOS/70, Washington, D. C.

(7) Horton, B. S., R. L. Goldsmith, and R. R. Zall. 1972. Membrane processing of cheese whey reaches commercial scale. Food Technol. 26(2) : 30 - 35.

(8) Horton, B. S. 1973. Whey processing by sanitary CIP membrane separation equipment. Amer. Dairy Review 35(4) : 45 - 50.

(9) Klein, C. L. and Wong Chang-ming. 1969. Study reveals potential for profit in waste recovery through desalination. News release, Nov. 2, 1969, Office of Saline Water, U. S. Department of the Interior.

(10) Lowe, E., E. L. Durkee, R. L. Merson, K. Ijichi, and S. L. Cimino. 1969. Egg white concentrated by reverse osmosis. Food Technol. 23 : 79 - 82.

(11) McDonough, F. E. and W. A. Mattingly. 1970. Pilot-plant concentration of cheese whey by reverse osmosis. Food Technol. 24(2) : 88 - 91.

(12) Melachouris, N. 1969. Discontinuous gel electrophoresis of whey proteins, casein, and clotting enzymes. J. Dairy Sci. 52(4) : 456 - 459.

(13) Porter, W. L., J. Siciliano, S. Krulick, and E. G. Heisler. 1970. Reverse osmosis : application to potato-starch waste effluents, p. 220 - 230. In Flinn, J. E. [ed.] Membrane science and technology. Plenum Press, New York.

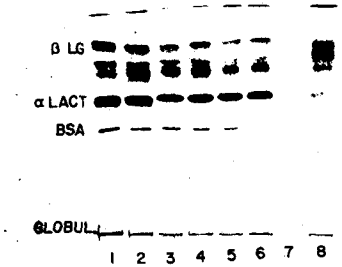

Figure 7. Discontinuous acrylamide gel electrophoresis of whey proteins from whey protein concentrates (WPC) produced by ultrafiltration.

Key
1. Raw milk
2. Pasteurized cheddar cheese whey (0 time)
3. Whey concentrated to 17.3x protein, wet basis, in a Havens unit
4. Spray dried WPC, 19.2x protein, Havens
5. Whey concentrated to 7.6x protein, wet basis, in an Abcor unit
6. Spray dried WPC, 11.1x protein, Abcor
7. Permeate solids from Havens
8. Conventionally produced spray dried whey powder

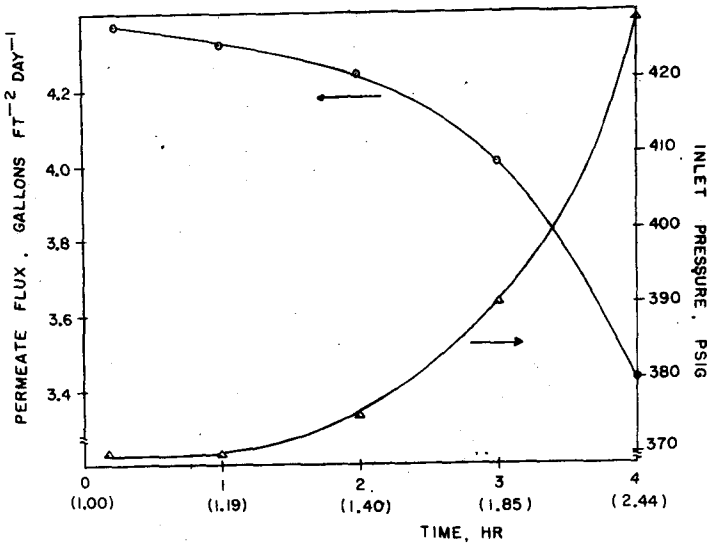

Figure 8. Ultrafiltration of skim milk. Changes in permeate flux rate (gallons per square foot per day) and inlet pressure (psig) versus time (hr). Volumetric concentration ratios corresponding to the times shown are in parentheses.

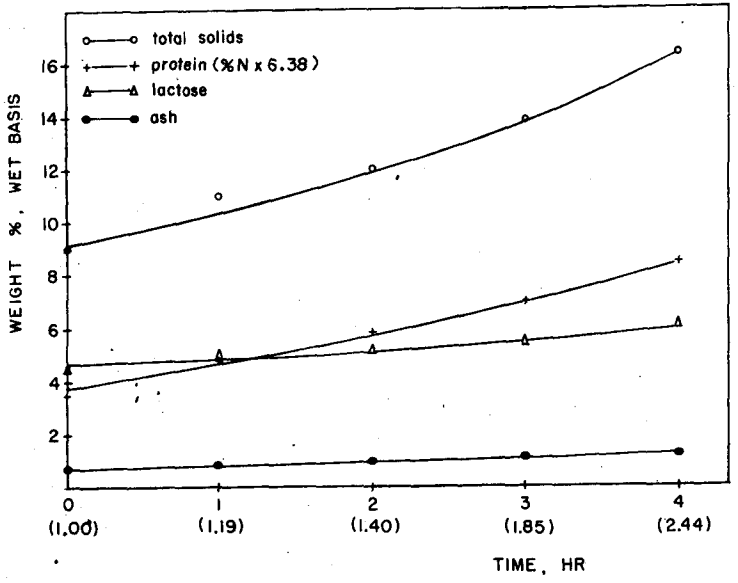

Figure 9. Ultrafiltration of skim milk. Changes in retentate composition with time (hr). Volumetric concentration ratios corresponding to the times shown are in parentheses.

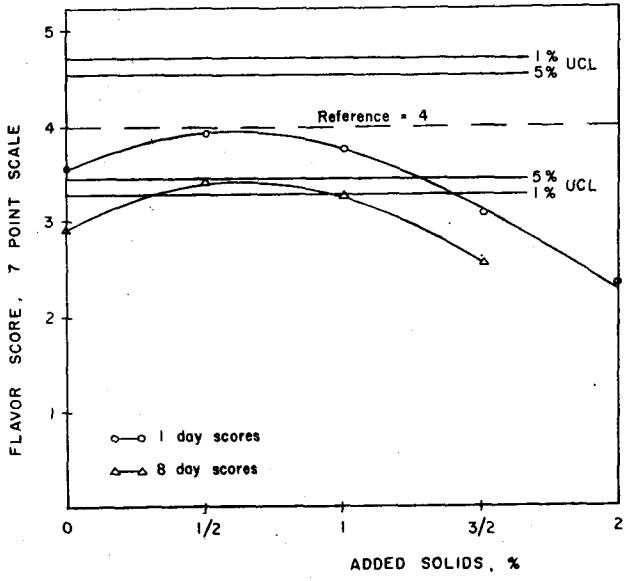

Figure 10. Flavor scores of beverage skim milk fortified with a 2:1 blend of permeate : acid whey solids versus percentage fortification. Evaluations were conducted with fresh milk and after eight days' storage at 35F.

(14) Sourirajan, S. 1970. General applicability of the technique, p. 1 - 54. In S. Sourirajan, Reverse osmosis. Logos Press, London.

(15) Sourirajan, S. 1970. Preparation and performance of porous cellulose acetate and other membranes for reverse osmosis application, p. 55 - 153. In S. Sourirajan, Reverse osmosis. Logos Press, London.

(16) Underwood, J. C. and C. O. Willits. 1969. Operation of a reverse plant for the partial concentration of maple sap. Food Technol. 23(6) : 79 - 82.

(17) Wesenti-Pulle, M. P., W. C. Winder, C. H. Amundson, and C. G. Hill, Jr. 1971. An ultrasonic method for determining composition of cheese wheys during concentration with ultrafiltration and reverse osmosis. Presented at the 1971 Winter conference of the American Society of Agricultural Engineers.

(18) Wyeth, T. C. 1972. Chemical analysis of whey powders. J. Soc. Dairy Technol. 25(3) : 136 - 139.

Table 1. Analyses of feed to, and first and second stage permeates from two-stage membrane processing of cottage cheese whey(3).

Component	Feed	First-stage permeate	Second-stage permeate
Total solids, %	7.8	6.8	1.8
Protein and amino acid nitrogen, %	0.6	0.15	0.002
Lactose, %	3.9	3.5	0.05
Lactic acid, %	0.52	0.52	0.11
Chemical oxygen demand, ppm	65,600	54,100	800

Table 2. Statistical summary of the analyses of various concentrated cheese wheys for the extractable lipids and protein content by the Solution Analyzer and standard procedures(17).

Concentrated Cheese Whey	Multiple Correlation Coefficients		Standard deviation of differences		Range of differences	
	33 C data	65 C data	Extractable lipids (%)	Protein (%)	Extractable lipids (%)	Protein (%)
Cottage	0.959	0.967	± 0.00	± 0.06	0.00	+ 0.08 to -0.09
Cheddar	0.973	0.999	± 0.03	± 0.06	+ 0.03 to -0.04	+ 0.10 to -0.06
Swiss	0.981	0.998	± 0.02	± 0.05	+ 0.04 to -0.03	+ 0.07 to -0.06

WASTE WATER TREATMENT INSTRUMENTATION

James A. Deakins, P.Eng.
Electrical Engineer
D. E. McGregor and Associates Limited
Toronto, Ontario

INTRODUCTION

With to-day's concern for the environment, domestic and industrial waste treatment plants are being constructed at a considerable rate. In order to produce a good clear plant effluent that is essentially free of contaminants, that will not pollute our receiving water, good control of the sewage and sludge flows at various stages through the plant is essential and instrumentation plays an important role in this process.

To-day, I would like to illustrate and discuss the various stages of sewage flow and sludge removal and flows through a typical sewage treatment plant. In addition, I shall try to indicate the instrumentation used at each stage in a sewage treatment plant to provide the required indication and control.

FLOWS IN A TYPICAL SEWAGE TREATMENT PLANT

To begin with, let us look at Fig.1 which is a flow diagram for a typical sewage treatment plant. The incoming sewage flows to the wet well of the plant pumping station from the incoming sewers. Using centrifugal pumping units, the sewage is pumped to a high enough level so that it will flow through the rest of the plant by gravity. The pump intakes are usually protected by large bar screens.

Next, the sewage flows through the Detritor where heavy inorganic particles, such as sand, are removed by settling. The effluent from the Detritor flows to the Primary Clarifier tanks where the retention time of the flow through the tanks is approximately two hours. The sewage entering the Primary Clarifiers contains mainly organic and chemical matter. In the Primary Clarifiers all heavy organic matter will settle on the bottom of the clarifier as raw primary sludge and any light organic matter, such as grease and oils, will rise to the surface as scum. Approximately 50% of the organic matter will be removed in the Primary Clarifiers. The remaining organic and chemical matter remains disbursed in colloidal form or dissolved in the water.

This colloidal organic and dissolved chemical matter in the sewage flows from the effluent of the Primary Clarifiers to the influent of the aeration tanks where the retention time at rated design flow is approximately four to six hours. If extended aeration is used, such as in the food industry, the retention time can be up to thirty hours. In the aeration tanks where there is a biological breakdown of the organic compounds in the sewage due to the aerobic bacteria which is produced in the aeration tanks. The aerobic bacteria breaks down the organic matter to produce basically carbon dioxide and water.

The effluent (mixed liquor) from the aeration tanks flows to the Secondary Clarifiers.

In recent years there has been a great deal of research done on nutrient removal, which is the removal of phosphates and nitrogen compounds which cause algae growth in the receiving waters. There are many sewage treatment plants in operation for phosphates removal.

The effluent from the secondary treatment will enter the Chlorine Contact Chamber. The effluent from the Chlorine Contact Chamber will enter the nearest receiving waters, such as rivers, lakes and oceans.

With the removal of solids from the sewage there is always grit, sludge and scum to contend with. In the Detritor there is grit removal, which is the result of the settling of inorganic compounds, such as sand. This grit is virtually free of the decomposible organic compounds so that it can be trucked to a land fill site. The sludge from the primary clarifiers is called raw primary sludge and it is usually pumped to the dewatering building or to the digesters. The scum from the primary clarifiers can be heated and pumped to the dewatering building. There is no sludge collected from the aeration tanks since the particles are kept in suspension by the agitation caused by the compressed air or by the mechanical aerators. The sludge from the secondary clarifiers is pumped to the aeration tanks and to the sludge thickening building or directly to the digesters. The digested sludge from the digesters is pumped to a sludge lagoon, liquid fertilizer tank truck or dewatering building.

The sludge thickening building is a means of producing a thicker sludge before it enters the digesters or dewatering building.

The dewatering process is a process to produce "sludge cake" by removing the water from the sludge by means of vacuum filters or centrifuges. This process reduces the water content in the sludge to approximately 70%.

The incineration building is a process for drying the sludge cake to about 10% moisture or for burning all combustible matter. The ash from the incinerators can then be pumped to an ash lagoon. The exhaust gases from the incinerator have to be scrubbed in order to produce a flume that will not pollute the air or cause undesirable odours.

INSTRUMENTATION

Pumping Station

The instrumentation in the pumping station as shown in Fig.2 could consist of the following:

a) Wet Well Level Indication and Control
b) Flow Indication-Recorder-Totalizer
c) pH Indicator Recorder
d) Temperature
e) Total Organic Carbon (TOC) Indicator Recorder.

The level control could consist of an air bubbler system, including a small compressor, ¼" bubbler line to the wet well, pneumatic to electrical transducer, indicator-recorder and transmitter. Basically air is blown down the bubbler line in the wet well and the back pressure of zero to 15 p.s.i. is an indication of the liquid level in the wet well with zero p.s.i. being the zero reference level and 15 p.s.i. the maximum reference level. If the transducer is a pneumatic to current transducer, the output could be 4 to 20 milli-amperes corresponding to 0 to 15 p.s.i. This 4 to 20 milli-ampere signal could be used to drive an indicator-recorder and at the same time provide a signal to a motor control system to control constant and variable speed pumping units to maintain a preset wet well level.

"pH" indicates the acid-base condition of the water which should be maintained at approximately 7.0 in order not to upset the aerobic reaction in the aeration tanks and the anaerobic reaction in the digesters. The pH can be continuously monitored with a pH analyser, which can be located in the influent channel to the detritors. High and low pH contacts should be provided on the analyser to send alarms to a control room to alert the operator so that corrective measures can be taken. For industrial waste, the pH analyser could be used in conjunction with a flow signal to add corrective chemicals to their waste to maintain a pH of approximately 7.0 prior to their effluent entering a municipality sewerage system or receiving waters.

The flow to a sewage treatment plant can be measured by a Parshall Flume. The depth of the water in the Parshall Flume is a measure of the flow. The depth of the water in the Parshall Flume can be measured by a bubbler system, similar to the one described for the wet well. This flow is usually calibrated in million gallons per day (MGD). A totalizer is used to give the total gallons passing through the Parshall Flume. This rate of flow signal is used to pace chemical feeders for chlorinators, pH and nutrient removal.

Biochemical Oxygen Demand, BOD, gives an indication of the oxygen requirements for the organic matter in the sewage. BOD measurements are used to estimate the organic content in the sewage. Therefore, BOD removal is a prime function and indication of how well the treatment is working. A BOD measurement for a sample of sewage requires a 5-day laboratory test. However, there are Total Organic Carbon (TOC) analysers that are designed for continuous monitoring of waste waters, which measure the organic concentration in the water. TOC concentrations provide the operator with real time measurements of performance information. A TOC analyser could be located at the influent works, primary effluent and secondary or final effluent so that the total carbon removal can be indicated for each stage of the plant.

A typical wet well level and pump control system is shown in Fig.3.

Detritor Building

In the Detritor Building there is very little instrumentation as shown in Fig.4, entitled "Detritor (Grit and Screenings Removal)". The sewage flows through the grit channels where the heavy sand and other inorganic solids settle. To prevent the organic matter from settling and to provide pre-aeration, air is blown through an air-line with air bubble diffusers near the bottom of the grit channels which keeps the organic matter in suspension while the sewage flows through the detritor. In addition, the flow through the Detritor Building has to be fast enough so that the inorganic matter will not settle, but slow enough to allow the grit to settle. The flow rate is usually one to two feet per second.

Near the effluent of the Detritor Building, the sewage passes through mechanically raked bar screens which remove large articles such as rags and pieces of wood which could clog up sludge pumps and pipelines later on. The bar screens are usually controlled by timers and/or a differential air bubbler level system, which senses the level on both sides of the screens. When the difference in the level approaches a pre-set point, the rake drive is started to rake and clean the screens. The air bubbler system is similar to the one in the pumping station. Comminutors or barminutors which disintegrate the large particles are used in many plants rather than screens.

Primary Clarifier

The major equipment and instrumentation for the Primary Clarifiers is shown in Fig.5.

A sludge collection mechanism scrapes the raw primary sludge to a hopper while a skimmer skims the oils and greases to a scum trough.

The raw primary sludge is pumped to the digesters or to the dewatering building. The pumping is