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**Journal of  
FOOD PROCESS ENGINEERING**

**Edited by D. R. Heldman, Michigan State University**

**FOOD & NUTRITION PRESS, INC.  
WESTPORT, CONNECTICUT 06880 USA**

**VOLUME 1, NUMBER 3**

**JULY 1977**

## JOURNAL OF FOOD PROCESS ENGINEERING

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All subscriptions and inquiries regarding subscriptions should be sent to Food & Nutrition Press, Inc., 265 Post Road West, Westport, Connecticut USA.

One volume of four issues will be published annually. The price for Volume 1 is \$45.00 which includes postage to U.S., Canada, and Mexico. Subscriptions to other countries are \$55.00 per year via surface mail, and \$62.00 per year via airmail.

Subscriptions for individuals for their own personal use are \$25.00 for Volume 1 which includes postage to U.S., Canada, and Mexico. Personal subscriptions to other countries are \$35.00 per year via surface mail, and \$42.00 per year via airmail. Subscriptions for individuals should be sent direct to the publisher and marked for personal use.

The *Journal of Food Process Engineering* is published quarterly by Food & Nutrition Press, Inc. — Office of Publication is 265 Post Road West, Westport, Connecticut 06880 USA.

Application to mail at second class postage rates is pending at Westport, Ct. 06880.

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Westport, Connecticut USA

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ISSN 0145-8876

Printed in the United States of America

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## EDITORIAL

It seems appropriate to pause for a moment and recognize the contributions of Dr. Karl F. Mattil. Dr. Mattil had served as a member of the Editorial Board for the *Journal of Food Process Engineering* for about one year before his death on October 4, 1977. Obviously he had made significant contributions to food engineering before his affiliation with this Journal, and these contributions should be recognized at this time and as a part of the record within the *Journal of Food Process Engineering*.

Dr. Mattil's educational background was in Chemistry, with B. S. and Ph.D. degrees from the Pennsylvania State University. In addition, he had two years of postdoctoral experience at the University of Pittsburgh. In 1944, Dr. Mattil joined Swift and Company Research Laboratories in Chicago and led the development of a process for crystallized modification of lard. In 1950 he was appointed head of Swift and Company's edible fats research division and pilot plant, and eventually became associate director of research in 1955. During these periods of time, he broadened his interest to processing of cottonseed, soybeans and peanuts as vegetable protein sources. It was during this period of time that Dr. Mattil developed an intense interest in the processing technology associated with vegetable protein sources.

In 1968, Dr. Mattil became Director of the Food Protein Research and Development Center at Texas A & M University. His research interests and leadership resulted in a nationally-known research program on processing and utilization of soybean, cottonseed, peanut, sunflower, and sesame crops as well as coconut protein, flour and fish protein concentrates. Under the leadership of Dr. Mattil, the Food Protein Research and Development Center has received worldwide recognition as a center of excellence for research on vegetable protein processing technology.

In addition to his membership on the Editorial Board for the *Journal of Food Process Engineering*, Dr. Mattil was a trustee of the Texas A & M Research Foundation and a member of the Soybean Utilization Advisory Panel of the American Soybean Association. He held membership and contributed significantly to activities in the American Oil Chemists' Society, the Institute of Food Technologists, the American

Association of Cereal Chemists, the American Peanut Research and Educational Association, as well as Phi Tau Sigma, Sigma Xi, and Alpha Chi Sigma.

Although the loss of Dr. Karl F. Mattil will be recognized in many circles, the contributions he has made during nearly forty years of research and leadership are available for others to use. These contributions will continue to be a source of inspiration for others in the scientific community to utilize and follow.

## MEETINGS

### MARCH 1978

Mar. 23-24: **14TH ANNUAL MEAT INDUSTRY RESEARCH CONFERENCE.** Center for Continuing Education, University of Chicago. J. Birdsall, Director of Scientific Activities, American Meat Institute, P.O. Box 3556, Washington, D.C., 20007.

Mar. 27-29: **PI/USA EASTERN REGIONAL PACKAGING FORUM AND PACKAGING PRODUCTS PANORAMA.** Sheridan, Philadelphia, PA. Packaging Institute, USDA, 342 Madison Ave., New York City, NY 10017.

Mar. 28-30: **WESTERN FOOD INDUSTRY CONFERENCE.** Freeborn Hall, University of California, Davis. J. C. Bruhn, Dept. of Food Science and Technology, Univ. of California-Davis, Davis, CA 95616.

### APRIL

Apr. 5-7: **DAIRY AND FOOD INDUSTRY SUPPLY ASSOCIATION. 59th ANNUAL MEETING.** Canyon Hotel Racquet and Gold Resort, Palm Springs, CA.

Apr. 11-13: **SIXTH ANNUAL INDUSTRIAL POLLUTION CONFERENCE.** Riverfront Convention Complex, St. Louis, MO. F. Harroun, Water and Wastewater Equipment Manufacturers' Assn., Suite 304, 7900 West Park Drive, McLean, VA 22101.

Apr. 11-13: **32D ANNUAL MEETING OF RESEARCH AND DEVELOPMENT ASSOCIATES FOR MILITARY FOOD AND PACKAGING SYSTEMS.** Drake Hotel, Chicago, IL. Col. Merton Singer, R & D Associates, 90 Church St., Room 1315, New York, NY 10007.

Apr. 17-20: **THIRD ANNUAL CONFERENCE ON TREATMENT AND DISPOSAL OF INDUSTRIAL WASTEWATERS AND RESIDUES.** Sheridan Houston Hotel, Houston, TX. Industrial Wastewater Residue Conference, Information Transfer Inc., Suite 202, 1160 Rockville Park, Rockville, MD 20852.

Apr. 25-28: **INTERNATIONAL INSTITUTE OF REFRIGERATION.** Commission C-2 (Food Science and Technology), Commission D-1 (Refrigeration Storage), and Commission D-2 (Refrigerated Land Transport), Budapest, Hungary. I. Ruttkay, Hungarian Scientific Society for Food Industry M E T E, V. Akadémia U. 1-3, 1361 Budapest, PF. 5, Hungary.

## **MAY**

May 8-10: **SEVENTH INTERNATIONAL CONFERENCE ON NOISE CONTROL ENGINEERING**, Inter-Noise 78. Jack Tarr Hotel, San Francisco, CA.

May 8-10: **32ND ANNUAL TECHNICAL CONFERENCE AND EXHIBIT**. American Society for Quality Control. Palmer House, Chicago, IL. D. C. Schmidt, Manager, Public Information Office, Dept. PI-132, Amer. Society for Quality Control, 161 W. Wisconsin Ave., Milwaukee, WI 53203.

May 9-10: **33rd ANNUAL PURDUE INDUSTRIAL WASTE CONFERENCE**. Stewart Center, Purdue University, W. Lafayette, IN. J. E. Etzel or J. D. Wolszon, Purdue Industrial Waste Conference, Purdue University, Civil Engineering Building, W. Lafayette, IN 47907.

May 16-18: **POWDER AND BULK SOLIDS 78 CONFERENCE AND EXHIBITION**. O'Hare International Trade and Exposition Center and Hyatt Regency O'Hare Hotel, Rosemont, IL. A. Kozlov, Industrial and Scientific Conference Management, Inc., 222 W. Adams St., Chicago, IL 60606.

## **JUNE**

June 4-7: **INSTITUTE OF FOOD TECHNOLOGISTS 38th ANNUAL MEETING AND TECHNICAL INDUSTRIAL EXHIBIT**. Dallas, Tex. C. L. Willey, Institute of Food Technologists, Suite 2120, 221 N. LaSalle St., Chicago, IL 60601.

June 27-30: **AMERICAN SOCIETY OF AGRICULTURAL ENGINEERS**. Annual Summer Meeting. Utah State University, Logan, Utah. J. L. Butt, Amer. Society of Agric. Engineers, 2950 Niles Rd., St. Joseph, MI 49085.

# ANALYSIS AND ECONOMIC EVALUATION OF CONCENTRATION ALTERNATIVES FOR LIQUID FOODS — QUALITY ASPECTS AND COSTS OF CONCENTRATION

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Received for Publication September 2, 1977

Accepted for Publication November 8, 1977

## ABSTRACT

*In the selection of concentration processes the net added value of the concentrate with respect to the capital invested should be maximized. The net added value expressed per unit weight of feed is the difference between the gross added value, being the increase in product value due to concentration, and the costs of concentration. The main factors affecting the gross added value and the concentration costs are treated systematically.*

*The main factors influencing the gross added value are: thermal quality degradation by chemical reactions; loss of volatile aromas; and the concentration that is obtained in the process. The first two factors generally affect the gross added value negatively. The thermal effects of the various processes are shortly summarized. In evaporative concentration and reverse osmosis, aromas can usually be recovered by steam stripping and distillation if the aromas in the food are at least a factor of three more volatile than water. For liquids containing aromas with a relative volatility close to or smaller than unity only freeze concentration can be used, if the quality is to be fully retained.*

*The approximate costs of concentration are given for dewatering capacities in the range of 1 to 50 tons per hour and 2000 to 7000 operational hr per year. As energy costs increase further, only concentration processes with a relatively low energy consumption will, at least for higher capacities, remain economically justified. The energy consumptions expressed in steam equivalents vary widely, viz. evaporation 0.16–1.3, freeze concentration 0.25–0.5, and membrane processes 0.01–0.48.*

## INTRODUCTION

Liquid foods like fruit and vegetable juices, milk, skim milk and

whey, beet and cane sugar juices, coffee and tea extract, beer and wines are very complex aqueous mixtures of organic and inorganic compounds. The water content is usually about 90% by weight, with extreme values generally between 75 and 97%. Partial removal of the water, reducing the water content to 10–25%, can be the last step in food processing prior to packaging, the concentrate being the end product, or can be a step prior to drying.

Objectives of producing concentrated end products are: reduction of packaging, transport and storage costs, induction of crystallization; reduction of water activity so as to enhance storage stability; improvement of convenience for the consumer, and in a few cases, improvement of sensory acceptability. Main objectives of concentration prior to drying are: reduction of the overall dewatering costs including capital and energy costs; improvement of retention of volatile aromas in the drying process; and reduction of the bulk volume of the dried product.

Unlike the concentration of most "chemicals," the concentration of liquid foods is a delicate affair. Even at moderate temperatures many of their constituents prove to be chemically unstable. At temperatures between 40 and 70°C enzymatically catalyzed reactions can alter food properties within a few minutes. To obtain good products, sanitary conditions must also meet strict standards. For many liquid foods the quality is moreover greatly dependent on the concentration and composition of odorous compounds. All aroma compounds are naturally volatile and can partly or even totally be lost by evaporation.

For the concentration of liquid foods three different main groups of processes can be used, viz. evaporation with or without aroma recovery by stripping and/or distillation, reverse osmosis or ultrafiltration with or without aroma recovery from the permeate by distillation, and freeze concentration.

For the evaporative concentration of foods, falling film evaporators, plate type evaporators, scraped surface film evaporators and centrifugal film evaporators, are most commonly used. Energy can be saved by carrying out the evaporation in multi-effect and by thermal compression. In a multi-effect arrangement the vapor generated by evaporation in one effect is used to provide the heat required to produce evaporation in a subsequent effect operated at a lower pressure. The temperature that is used in the effect with the highest pressure greatly affects the size of the heat exchanging surface and consequently the costs of the evaporator. Multi-effect systems may be operated in: reverse flow with the most concentrated product leaving the effect with the highest pressure; forward flow with the dilute feed entering the effect with the highest pressure; in parallel flow, where the feed is split among all

stages; and in combinations of the above. With forward flow quality degrading heat effects are minimal, whereas with counter-current flow the evaporation capacity is highest.

Membrane separations and freeze concentration can be carried out either in single stage or in multi-staged arrangements. Because the concentration rises from stage to stage, advantage is taken of the fact that the overall permeation flux and also the crystallization growth rate, which decreases with increasing concentration, are higher than in single stage operation.

This short survey may make it clear that the number of alternatives from which one has to choose is very great indeed. Yet, for any product to be used in a particular segment of the market in a particular country or part of a country and to be produced during a defined number of days at a defined quantity per hour there is only one process in one arrangement that performs the job optimally at given costs of utilities such as steam, cooling water and electricity. The selection is a purely economic problem.

The ratio of: (1) the product of the *net added value* resulting from concentration and the total tons of feed processed per year and (2) the capital to be invested in the concentration equipment (installed costs of equipment and cost of the building to be used for the process) has to be maximized. The net added value per ton of feed,  $S_{net}$ , is the difference between the gross added value per ton of feed,  $S_{gross}$ , and the costs,  $C_f$ , of the concentration process per ton of feed.

$$S_{net} = S_{gross} - C_f$$

The gross added value is the difference in sales value, expressed per ton of feed ex factory, between the concentrated product and the not concentrated liquid (feed). The concentration costs include capital costs, utility costs, labor costs etc.

In the following the factors affecting the gross added value and the factors affecting the costs to be made to obtain that desired gross added value will be treated in two separate chapters.

## THE GROSS ADDED VALUE

### Introduction

The gross added value is dependent on many changes which are brought about during processing. Certain changes may contribute posi-

tively to the product value for some products and negatively for other ones. The changes in functional properties can be classified as follows:

- (1) Selective physical losses; losses of a group or groups of constituents such as losses of volatiles (aroma losses) in evaporation and losses of constituents of small molecular size in membrane concentration. This in contrast with a-selective losses where part of the product is lost without a change in composition. In general a selective physical loss contributes negatively to the value, but the quality of milk containing mercaptans is improved by the loss of its volatiles. In the ultra-filtration of whey the loss of salts and sugars and more particularly of lactose also affects the value positively.
- (2) Physical changes; changes in rheological properties such as degree of gelling, viscosity, crystallization of constituents.
- (3) Changes by chemical reactions; enzymatic and non-enzymatic browning reactions, vitamin destruction, destruction of nutrients, development of flavors etc. In general the chemical reactions degrade the quality. Examples are the development of an undesirable cooked flavor, and discoloration. For maple juice, however, Maillard browning reactions are essential.
- (4) Concentration factor; the final concentration that can be obtained in the process affects the microbial stability of the products, the packaging, storage and transport costs and in case of drying the quality of the dried product and the integral dewatering costs.

The physical losses of constituents, physical changes and changes by chemical reactions have to be translated into increases or decreases in the value of the product. Quantitative valuing is the most difficult part in the selection of the optimum process for concentrating a particular product for a particular market segment.

The value that has to be assigned to the concentration that can be achieved with a particular concentration process depends again greatly on the market segment for which the product has to be used. Juices, for example, which have to be reconstituted to the original concentration before bottling, should be concentrated as highly as possible. For direct domestic use, however, the viscosity will be the limiting factor. If the functional properties of the single strength product and the concentrate are about equal, the gross added value can simply be calculated from the difference in the costs of transport, storage, and packaging of the single strength juice or extract and the concentrated product. It is evident that the added value increases the more remote the production

location is from the consumption location, and the shorter the season of production.

It is obvious that general rules for valuing the various changes in functional properties cannot be given; they are strongly product and market dependent. In this paper only the affects of process conditions on the various changes in product properties will be presented.

#### Effect of Process Conditions on Selective Physical Losses

**Physical Losses in Evaporation** — In evaporation, volatiles, including water and aromas, distribute themselves over liquid phase and vapor phase. The selective distribution with respect to water is expressed by the partition coefficient  $\beta_{i,w}$

$$\beta_{i,w} = \frac{C'_i/C_i}{C'_w/C_w} \quad (1)$$

in which  $C'$  is the concentration in the gas phase,  $C$  the concentration in the liquid phase and the subscripts  $i$  and  $w$  refer to an aroma component and water, respectively. When there is equilibrium between the phases, the partition coefficient is called the relative volatility,  $\alpha_{i,w}$ . When there is no phase equilibrium, the value of  $\beta_{i,w}$  is always closer to unity than  $\alpha_{i,w}$ .

When the distribution of volatiles is fully mass transfer controlled,  $\beta_{i,w}$  becomes about equal to unity, consequently

$$1 \leq \beta_{i,w} \leq \alpha_{i,w} \quad \text{if } \alpha_{i,w} > 1 \quad (2a)$$

and

$$\alpha_{i,w} \leq \beta_{i,w} \leq 1 \quad \text{if } \alpha_{i,w} < 1 \quad (2b)$$

The aroma of a food is composed of many different odorous components. The relative volatilities of most components in an aroma are larger or even significantly larger than unity. Exceptions are volatile acids and phenolic compounds, which can have relative volatilities smaller than unity. Because quality is controlled by the "bouquet" of the constituting aroma compounds, which may amount to 100 or more with widely differing relative volatilities, the potential for aroma loss of a liquid food can more conveniently be expressed by some mean of the relative volatilities of the components constituting the bouquet.

Thijssen (1970) has defined an effective relative volatility  $\alpha_{eff}$  which relates, according to equation 4, in a batch equilibrium evaporation process the sensorically observed retention of the strength of the total aroma with the fraction of water evaporated. The  $\alpha_{eff}$  values of, for example, apple juice, blackberry juice, blueberry juice, red raspberry juice and strawberry juice are 6.5, 4.5, 3.3, 4.5 and 10.3, respectively.

In evaporators with a large specific surface area of the liquid phase such as in long tube falling-film evaporators and plate type evaporators, phase equilibrium is closely approached. Equilibrium is generally not obtained in wiped film and centrifugal film evaporators.

*Equilibrium evaporation* — Aroma losses with the water vapor are very dependent on the number of effects used. If vapor and concentrate leaving an evaporator are about in equilibrium it can be derived for the fraction of the aroma,  $R_i$ , retained in the concentrate:

$$R_i = \frac{1}{\left(1 + \frac{\alpha_{eff} \cdot V_1}{F - V_1}\right) * \left(1 + \frac{\alpha_{eff} \cdot V_2}{F - (V_1 + V_2)}\right) * \dots * \left(1 + \frac{\alpha_{eff} \cdot V_n}{F - (V_1 + V_2 + \dots + V_n)}\right)} \quad Cont. \quad (3)$$

in which  $F$  is the feed rate and  $V_1, V_2, \dots, V_n$  denote the kilograms of vapor leaving the first, second and nth effect per unit time.

For a large number of effects, say 4 or more, equation (3) approaches the Rayleigh equation which describes the aroma retention in a batch evaporator

$$R_i = \left(\frac{F - V}{F}\right)^{\alpha_{eff}} \quad (4)$$

in which  $V = V_1 + V_2 + \dots + V_n$ , which is the total evaporation capacity.

From equations 3 and 4 it can be derived that in equilibrium evaporation at a constant concentration factor  $F/(F-V)$  the aroma retention decreases with a growing number of effects. For a value of  $\alpha_{eff} = 3$  and a concentration factor of 3 the aroma retention in the concentrate

leaving a single effect evaporator amounts to 14.3% whereas for a large number of effects the retention falls to 3.7%. It is evident from equations 3 and 4 that the aroma retention sharply decreases with increasing concentration factor  $F/(F-V)$ . In case of non equilibrium processes equations (3) and (4) can still be used by replacing the relative volatility by the partition coefficient  $\beta$ .

*Non-equilibrium evaporation* — In non equilibrium evaporation, the transfer of aromas from the liquid phase to the vapor phase is limited by the mass transfer in the liquid phase. In centrifugal film evaporators the value of  $\beta_{i,w}$  is about equal to  $\alpha_{i,w}$  at very low evaporation fluxes of about  $100 \text{ kg/m}^2 \text{ h}$  but approaches unity at high fluxes exceeding  $600 \text{ kg/m}^2 \text{ h}$  (Thijssen 1975). Consequently the aroma retention becomes independent of the value of the relative volatility. At a concentration factor of 3 the retention will be 33.3%. It is obvious that aromas with relative volatilities smaller than unity are lost to a greater extent in a non-equilibrium evaporation process than in equilibrium evaporation.

*Selective Physical Losses in RO and UF* — In reverse osmosis and ultrafiltration the water diffuses from a concentrated solution through a more or less selectively permeable membrane to an aqueous water absorbing phase. The driving force for water transport is a pressure difference over the membrane exceeding the difference in osmotic pressure. The process is called ultrafiltration if the membrane allows the passage of water and other small molecules such as salts but impedes the passage of molecules with a molecular weight of 500 or higher. The term reverse osmosis is used for tighter membranes which are still permeable for water but allow only a limited passage of small molecules including salts, monosaccharides and aromas. RO requires a much higher pressure, 25–100 atm, than UF where pressures of between 1 and 5 atm are used. Cellulose acetate is less and less used for membranes; its successors are polysulfone and other proprietary copolymer materials (Michaels 1974). For a known value of the partition coefficient  $\beta_{i,w}$  the retention of smaller molecules can be calculated with equations 3 and 4.  $\beta_{i,w}$  here refers to the relative partition of a component between the liquid food and the permeate with respect to that of water. The first term in the denominator of equation 3 is used if the concentration is carried out in only one step with a strong recirculation of the liquid to be concentrated. Equation 4 is applied if a large number of stages is used. With the membranes available on the market the lower molecular weight aromas such as esters, aldehydes, ketones and alcohols up to about four C atoms are lost with the permeate.

*Selective Physical Losses in Freeze Concentration* — In freeze concentration the water is removed from the concentrate in the form of ice

crystals. Down to the eutectic temperature the ice crystals are of a very high purity. Consequently if the ice and concentrate are sharply separated, freeze concentration is in principle a highly selective dewatering process (Bomben *et al.* 1973).

**Recovery of Volatiles** — Volatiles which would otherwise be lost almost completely in evaporative concentration or partly in RO, depending on the molecular size of the aroma constituents, can be recovered at least partly, by either separating them from the feed before the concentration step or by recovering them from the vapor produced in the evaporator(s) or the permeate leaving the RO apparatus. The dilute aromas in the distillate or permeate can be concentrated in a distillation column. For evaporation the most effective way is steam stripping of the aromas from the feed prior to evaporation in a counter-current vapor-liquid contacting device, followed by concentration of the aromas by distillation. For RO the aromas are preferably recovered from the permeate. For some foods aroma recovery by stripping and/or distillation does not give satisfactory results because they contain aroma components that are essential to the quality but which have  $\alpha$  values close to unity. An example is coffee extract.

The recovery of aroma in single effect and triple effect evaporation by stripping and distillation at minimal energy consumption is presented schematically in Fig. 1 and 2. In Fig. 2 vapor and liquid flows are in counter-currents. The recovery of aroma from the permeate in RO is presented schematically in Fig. 3. The heat exchangers are not shown in these figures. A *partial* aroma recovery can be obtained by concentrating the aromas in the permeate in a second RO treatment at low temperatures, 7°C, with tight membranes (Matsuura *et al.* 1974).

#### Effect of Process Conditions on Physical Changes

Owing to the increase in dissolved solids concentration, part of the constituents become super-saturated and start to flocculate or to crystallize. The lower the temperature is, the higher is this effect in general. The viscosity also starts to rise sharply with increasing concentration. To prevent excessive thickening some liquid foods such as fruit juices have to be depectinized by means of enzymes prior to concentration (Mannheim and Passy 1974).

#### Effect of Process Conditions on Chemical Changes

The rates of physical processes such as evaporation and crystallization and the rates of chemical processes are all dependent on concentration and temperature. The process rate can be defined as the change

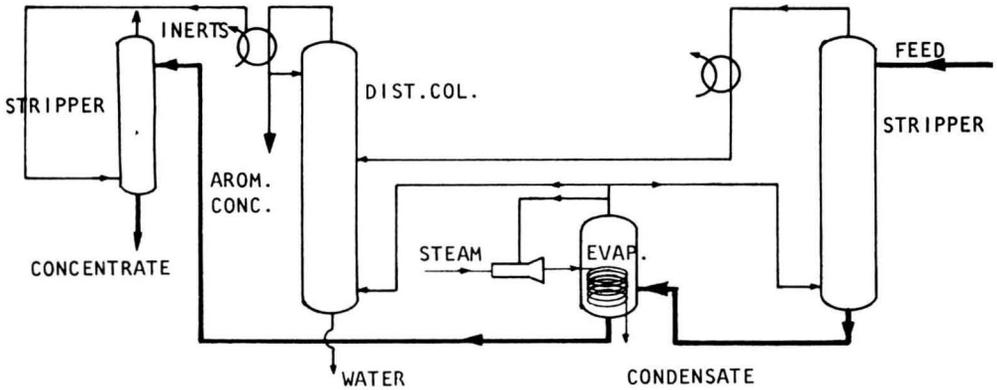


FIG. 1. SINGLE EFFECT EVAPORATION WITH AROMA RECOVERY

Aromas with relative volatility  $\geq 3.5$  are recovered for 90%.

with time of the concentration of the component (water or recovery component) under consideration.

**Effect of Concentration** — The dependence of the process or reaction rate,  $r_i$ , on the concentration of a constituent  $C_i$  is indicated by

$$\frac{-dC_i}{dt} \equiv r_i = k_i C_i^n \quad (5)$$

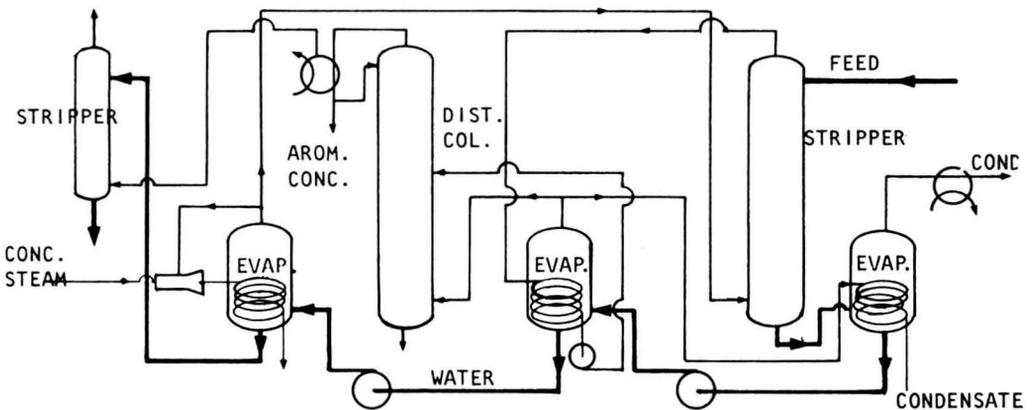


FIG. 2. TRIPLE EFFECT EVAPORATION WITH AROMA RECOVERY

Countercurrent flows. Aromas with relative volatility  $\geq 3.5$  are recovered for 90%

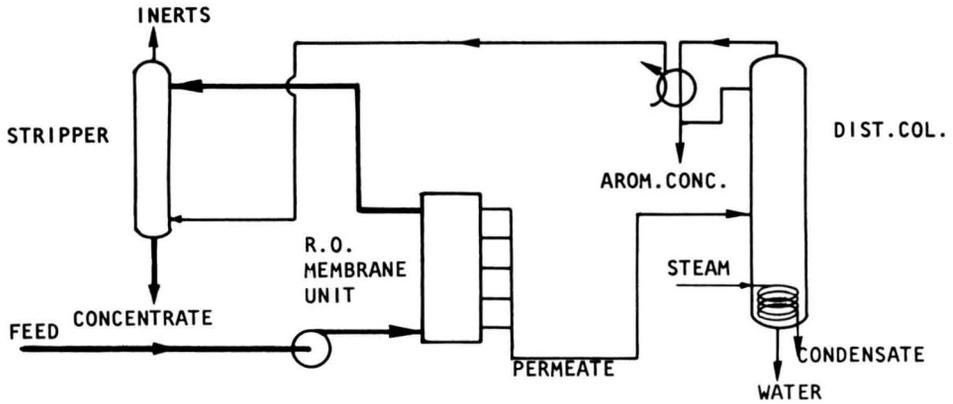


FIG. 3. REVERSE OSMOSIS WITH AROMA RECOVERY FROM THE PERMEATE

Aromas with relative volatility  $\geq 3.5$  are recovered for 90%.

in which  $k_i$  is the reaction rate coefficient.  $C_i$  can be the water, concentration, or concentration of a chemical constituent such as a flavor or aroma component.

For physical processes, the value of  $n$ , which indicates the order of the process in the concentration of constituent  $i$ , generally is between 0 and 1. For chemical reactions in foods, such as degrading reactions, a first order dependence in the concentration of the chemically unstable constituent is assumed in general.

In the case of a first order chemical reaction process, it follows from equation (5)

$$\frac{-dC_i}{dt} = k_i C_i \quad (6)$$

and after integration

$$\frac{C_{i,t}}{C_{i,o}} = \exp(-k_i t) \text{ at } T = \text{constant} \quad (7)$$

In the case of a concentration process, the concentration of the heat labile constituent decreases because of the chemical conversion but increases due to the water removal. Because of this combined effect the concentration of the heat labile constituent can more conveniently be expressed on dry matter basis.

The relation between the concentration  $C_i$  ( $\text{kg}/\text{m}^3$ ) and the concentration on dry matter basis  $X_i$  (kg of constituent per kg dry solids) reads

$$C_i = \frac{X_i}{X_w} C_w \quad (8)$$

in which in this case the subscripts  $i$  and  $w$  refer to the chemically instable constituent  $i$  and water, respectively. To represent the change of  $C_i$  with time due to the combined effect, mentioned above, equation (6) has to be extended to

$$\frac{dC_i}{dt} = -k_i C_i + X_i \frac{dC_w}{dt} \quad (9)$$

which after substitution of equation (8) and integration results in

$$\frac{X_{i,t}}{X_{i,o}} = \exp(-k_i t), \text{ at } T = \text{constant} \quad (10)$$

Because the water concentration does not appear in equation (10), it is clear that at constant residence time and constant process temperature  $T$ , quality loss by chemical conversion is not influenced by the concentration factor.

For a constant value of  $k_i$  and a small change in  $C_i$  with time, the value of  $C_i$  is almost proportional with time.

Of course, undesirable flavors can be caused by reaction products of consecutive reactions. In that case not only the mean residence time but also the residence time distribution has to be kept as small as possible; single pass plug-flow of the liquid food should be aimed at.

**Effect of Temperature** — The effect of temperature on the process and reaction rate and consequently also on  $k_i$  can usually be represented by the Arrhenius equation:

$$k_i = k_{i,T_r} \exp \left[ - \left( \frac{E}{R} \frac{1}{T} - \frac{1}{T_r} \right) \right] \quad (11)$$

in which  $E$  is the activation energy,  $R$  the gas-law constant,  $T$  is the absolute temperature and  $k_{i,T_r}$  is the value of  $k_i$  at reference temperature  $T_r$ .

The temperature dependences of physical rate processes, enzymatically catalysed reactions and non-enzymatic chemical reactions are listed in Table 1 along with  $k$  values at  $100^\circ\text{C}$ .

Table 1. Activation energies and  $k$  values at  $100^\circ\text{C}$  and high water activities of physical process rates, enzymatic reactions and nonenzymatic chemical reactions (Harris and Karmas 1975; Fennema 1975)

Group	$k_{100^\circ\text{C}}$ ( $\text{s}^{-1}$ )	$E$ (KJ/mole)
Physical rate processes		
evaporation rate and crystallization rate	strongly variable	8-40
Enzymatic reactions	—	17-60
Chemical reactions	—	60-210
thiamine destruction	$2 \times 10^{-5} - 6 \times 10^{-5}$	80-210
chlorophyll destruction	$10^{-4} - 10^{-3}$	30-80
Maillard reactions	$10^{-6} - 10^{-4}$	100-200
development of flavor artifacts (threshold quality changes)	?	60-130

Because the activation energies of chemical degrading reactions are always significantly higher than those of physical rate processes such as evaporation rate or permeation rate, the change of quality due to chemical reactions will be minimal at the lowest possible temperature and shortest possible residence time. The increase in residence time at lower process temperature, due to a reduced water flux, is fully offset by much lower chemical reaction rates.

**Classification of Concentration Equipment with Respect to Chemical Quality Degradation** — From equations 6 through 11 it can be derived that a change in quality due to thermally degrading chemical reactions will remain constant if the following condition is satisfied

$$\ln t = b_i + \frac{E_i}{RT} \quad (12)$$

in which  $t$  is the residence time at process temperature  $T$  and  $b_i$  is a constant of the product. Consequently when the product residence

time in the process equipment is plotted against the reciprocal of the absolute process temperature, straight lines with slope  $E$  indicate the conditions yielding equal thermal degradations for those values of  $E$ . In this way Fig. 4 presents the residence times and temperatures which are normally used for single pass operations in single up to four effect plate evaporators, single up to four effect long tube falling film evaporators, single and double effect centrifugal film evaporators, single effect scraped film evaporators, multi-staged RO and UF and multi-staged freeze concentration. The ranges of residence times and temperatures are only partly based on information and leaflets provided by manufacturers. Consequently the figures are for the responsibility of the authors. For some products the time-temperature combinations may fall outside the ranges indicated. For data on a specific product one should apply to equipment manufacturers.

Because the average of the activation energies of thermal degradation reactions causing off-taste, off-flavor, discoloration and nutrient loss is 83.6 KJ/mole (20 Kcal/mole) (Harris and Karmas 1975; Fennema 1975), this value has been used as an example in Fig. 4. The full lines in the figure are conditions for constant quality loss due to degrading reactions at the indicated values of  $k_{i,100}$  and  $E=83.6$  KJ/mole. For this value of  $E$  we see that for heat labile foods the quality is best preserved in the following sequence.

- |                                                            |                                                       |
|------------------------------------------------------------|-------------------------------------------------------|
| increasing potential<br>for thermal quality<br>degradation | 1. freeze concentration                               |
|                                                            | 2. single effect centrifugal film evaporator          |
|                                                            | 3. double effect centrifugal evaporator               |
|                                                            | 4. single effect scraped film evaporator + RO and UF  |
|                                                            | 5. single effect falling film and plate evaporator    |
|                                                            | 6. double effect falling film and plate evaporator    |
|                                                            | 7. triple effect falling film and plate evaporator    |
|                                                            | 8. quadruple effect falling film and plate evaporator |

In the figure the dotted lines represent the threshold conditions for freshly squeezed orange juice and a table wine (both estimated by own laboratory experiments) above which the quality is perceptibly affected.

For proper valuing of the effects of the various concentration processes on nutrient loss by chemical conversion the  $k_{100}^{\circ\text{C}}$  and  $E$  values of the product under consideration have to be known. Many kinetic data have been collected for microorganisms, chlorophyll destruction, and vitamins. However, studies of sensory quality degradation in foods are very scarce. For an accurate evaluation of sensory quality effects

the values of  $b_i$  and  $E_i$  in equation 11 should be experimentally obtained by determining at two temperature levels the times required to obtain a perceptible (threshold) change in quality in closed tubes without gas-headspace (no oxidation or aroma loss).

When using Fig. 4 for sensory quality effects, it has to be taken into account that above the threshold conditions the relationship between  $k_i t$  and quality obeys a power function

$$\text{sensory quality} = \text{constant} \times (k_i t)^n$$

Because the value of  $n$  may be expected to be higher than unity (Moskowitz 1974), a doubling of the reaction rate coefficient  $k_i$  at constant time or a doubling of the residence time at constant  $k_i$  more than doubles the reduction in sensory quality.

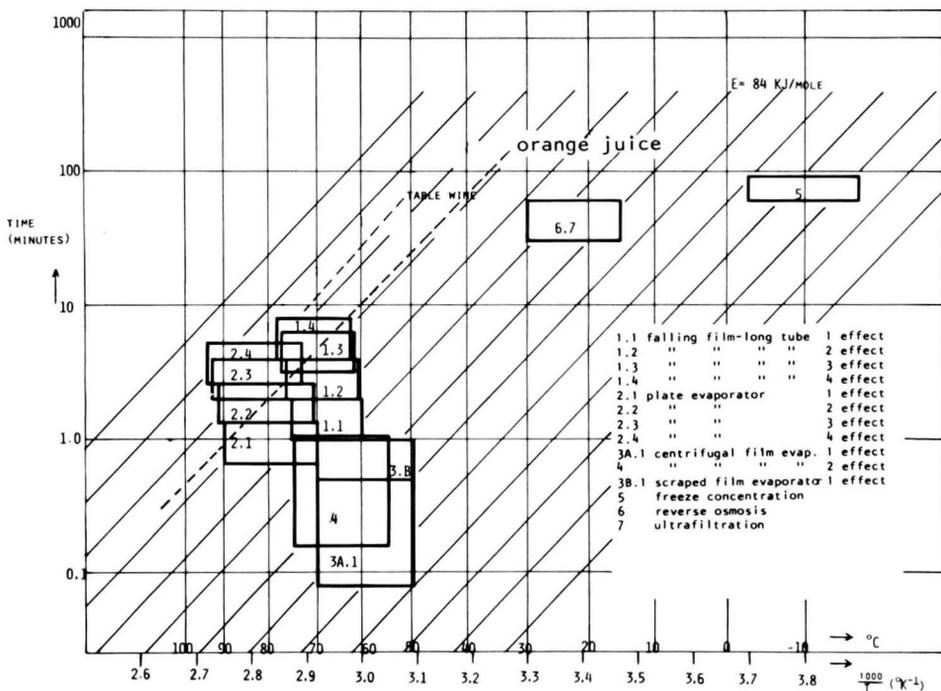


FIG. 4. RELATIONSHIP BETWEEN RANGE OF RESIDENCE TIMES AND RANGE OF TEMPERATURES APPLIED IN VARIOUS CATEGORIES AND VARIANTS OF CONCENTRATION PROCESSES

Each solid line represents conditions of equal thermal degradation for  $E=86.6$  KJ/mole and various values of  $k_i, 100^\circ C$ .

**Effect of Type of Process and Process Conditions on Concentration Factor**

The maximum concentration that can be obtained is primarily dependent on the viscosity at the maximum admissible process temperature. Liquid foods containing: (1) gels, such as pectins, and (2) polymers, such as proteins, exhibit much higher viscosities at equal concentration than solutions containing mainly low molecular dissolved solids such as sugars. The maximum viscosities which can be handled are given in Table 2 (approximations and/or average values taken from leaflets from equipment suppliers).

Table 2. Maximum concentrate viscosity

Concentration process	Viscosity (cP)
scraped film evaporator	40,000
centrifugal film evaporator	20,000
plate evaporator	400
falling film-long tube evaporator	200
freeze concentration	200
reverse osmosis	50

For a depectinized fruit juice the ranges of product concentration which can be obtained commercially in scraped film evaporators, centrifugal film evaporators, plate evaporators, falling film evaporators, freeze concentration and RO are 75–85, 70–80, 65–75, 60–70, 45–55 and 25–30 wt%, respectively.

**CONCENTRATION COSTS****Introduction**

The costs,  $C_f$ , of concentration expressed per ton of feed are related to the concentration costs per ton water removal,  $C_w$ , by the relation

$$C_f = \left(1 - \frac{X_p}{X_f}\right) \cdot C_w$$

in which  $X$  denotes kg of water per kg dissolved solids. The subscripts  $f$  and  $p$  refer to the feed and concentrated product, respectively. The concentration costs per ton of water removal are the sum of the labor

costs  $C_1$ , the capital costs  $C_c$ , the maintenance costs  $C_m$ , the utility costs including costs of electricity, steam, process water, cooling water and chemicals for cleaning  $C_u$ , the costs of a-selective product losses  $C_p$ , and other somewhat imponderable cost factors  $C_o$ .

$$C_w = C_1 + C_c + C_m + C_u + C_p + C_o$$

### Review of the Cost Components

**Labor Costs** — Labor costs are very dependent on the degree of automation of the process unit, the capacity per unit and the number of units being used and vary strongly from country to country.

**Capital Costs** — The capital costs are directly proportional to the installed costs of the concentration equipment, including costs of space and auxiliary equipment such as storage tanks etc.; they are directly proportional to the interest and depreciation and are inversely proportional to the number of productive operation hours per year and the capacity of the concentration unit. For most evaporation equipment the productive operation hours are, because of cleaning and maintenance, about 90% of the operational hours. For membrane processes even somewhat more than 10% might be required for cleaning and maintenance.

The capital costs of the equipment per ton of water removal decrease in general with increasing capacity of the unit. It has become common practice to present investment costs in the form of a power equation

$$\frac{I_{cap\ 2}}{I_{cap\ 1}} = \left[ \frac{Cap_2}{Cap_1} \right]^m$$

in which  $I_{cap}$  and  $Cap$  are the installed costs and the capacity of the unit, respectively. Power exponents of the main concentration equipment are collected in Table 3.

The capacities above which the capital costs per ton water removal do not further decrease are about 1.5, 5, 5, 15, 30 and 75 tons water removal per hour for scraped film evaporator, centrifugal film evaporator, freeze concentration, RO and UF, plate evaporator and long tube falling-film evaporators respectively.

**Maintenance Costs** — Maintenance costs vary between 2% and 5% annually of the investment costs. For falling film and plate evaporators 2% is usually taken whereas for centrifugal evaporators and membrane concentration about 5% is considered to be appropriate.

Table 3. Values of power exponents, m, of concentration equipment

Type	Capacity tons water removal per hour			
	1-5	5-10	10-50	> 50
Falling film, 3 effect		0.4		
Falling film, 5 effect			0.65	0.9
Plate evaporator, 2 effect		0.55	0.70	0.9-1.0
Centrifugal film evaporator	0.65-0.70	1	1	1
Scraped film evaporator	0.3 -0.5	1	1	1
Freeze concentration	0.65	1	1	1
Reverse osmosis (RO) and Ultra filtration (UF)	0.65	0.7-0.9	1	1

**Costs of Utilities** — The main utility costs are those of energy and water. In order to allow cost comparisons that are independent of rapidly changing values of currencies and of increasing costs of fuel, energy consumptions are conveniently expressed in steam equivalents. The steam equivalent of a specific concentration process is the ratio of the total utility costs per ton water removal of that process to the costs of one ton of steam. For most countries the ratio of the cost of electricity to that of steam and the ratio of the cost of cooling water and process water to that of steam, are practically constant and are only slightly dependent on fuel costs. In the present study we take the following constants

$$\frac{\text{cost of 100 kwh}}{\text{cost of 1 ton steam}} = 4.33$$

and consequently,

$$\frac{\text{cost of 1 KJ electrical energy}}{\text{cost of KJ steam}^1} = 2.84$$

$$\frac{\text{cost of 1 m}^3 \text{ cooling water}}{\text{cost of 1 ton steam}} = 3.33 \times 10^{-3}$$

The utility costs per ton of water removal are now simply calculated by multiplying the costs of one ton of steam with the number of

<sup>1</sup> Based upon heat generated by 1 ton of saturated steam of 10 atm at a condensation temperature of 100°C and 1 atm.

steam-equivalents of the process. The energy consumption per ton water removal expressed in steam equivalents are given for various concentration processes in Table 4.

Table 4. Energy consumptions of various concentration processes expressed in steam equivalents

Process	Steam equivalents	
	Without aroma recovery	With aroma recovery
1. evaporation		
1 effect evaporators	1.20-1.28	1.25-1.32
1 effect falling film evap. with thermal compression	0.44	0.60
1 effect plate evap. with thermal compression	0.65	—
2 effect falling film evap. with thermal compression	0.36	0.53
2 effect plate evap. with thermal compression	0.42	0.57
3 effect falling film evap. with thermal compression	0.29	0.50
3 effect plate evap. with thermal compression	—	—
4 effect falling film evap. with thermal compression	0.18	0.45
4 effect plate evap. with thermal compression	0.29	—
5 effect falling film evap. with thermal compression	0.16	—
2. freeze concentration		0.25-0.50
3. reverse osmosis	0.01-0.02	0.48
4. ultra filtration	0.01-0.02	—

**Costs of Non-selective Product Losses** — Product losses during concentration can be quite substantial. In evaporation about 0.1% of the feed can be lost with the condensate. In membrane processes a-selective product losses can occur because of pin-holes in the membranes. In freeze concentration substantial amounts of the product can be lost with the ice. Product losses increase the concentration costs very sharply. The authorities may levy a charge for the BOD of disposed condensate or melted ice.

**Other Cost Aspects** — There are also cost factors which vary a great deal with company, site and product. One of these factors for example

may be the cost to be attributed to a low "turn down" ratio in concentration capacity.

#### Capital Costs and Costs of Maintenance and Utilities for the Various Processes

**Basis for the Cost Calculations** — Because labor costs vary widely from factory to factory and the costs of product losses are dependent on the value of the product, only general figures for the sum of capital costs, maintenance costs and costs of utilities will be given. Again it has to be emphasized that these figures are only the best possible approximations. For tailor-made cost data application should be made to equipment manufacturers, stating product specifications and expected number of production hours per year and capacity per hour. The dewatering costs can then be based upon actual installed costs, utility requirements, and correct values for utilities etc. and an amortization procedure which might deviate from the one used in this study. In the calculations the following figures have been used:

(1) Annual capital recovery factor of installed costs based upon an amortization in 7 years and an interest of 10%	20%
(2) Costs of steam/ton	\$ 12
(3) Costs of electricity per 1000 kwh	\$ 52
(4) Costs of cooling water per 1000m <sup>3</sup>	\$ 40

**Dewatering Costs of Processes Without Aroma Recovery** — Figures 5 and 6 show the calculated costs per ton water removal without aroma recovery for 2000 and 7000 operation hours per year corresponding to effective production hours of 1800 and 6300, respectively for long tube-falling film evaporators, plate evaporators, scraped film evaporators, centrifugal film evaporators and RO and UF. The capital costs, maintenance costs and consumption of utilities are based on information provided by manufacturers in 1977. It can be derived from Table 4 how much the costs of utilities contribute to the dewatering costs. From Table 4 and Fig. 5 and 6 the dewatering costs can be calculated for any number of working hours per year.

The long tube-falling film evaporators and the plate evaporators have been economically optimized with respect to the number of effects. The number of effects corresponding to the minimum dewatering costs are given in the figure. For both falling film evaporators and plate evaporators thermo-compression is used. From the figure the following conclusions can be drawn:

By far the highest reduction in dewatering costs with increasing capacity is obtained with long tube falling film evaporators and plate evaporators. For 2000 hours per year the concentration of whey is

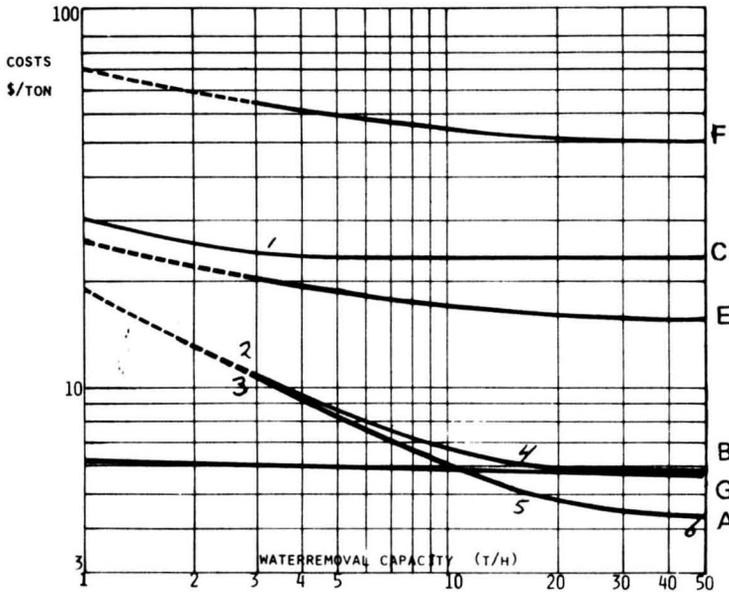


FIG. 5. DEWATERING COSTS IN DOLLARS PER TON WATER REMOVAL WITHOUT AROMA RETENTION OR RECOVERY FOR VARIOUS CONCENTRATION PROCESSES AT 2000 OPERATIONAL HOURS PER YEAR RESPECTIVELY, OF WHICH 90% ARE PRODUCTIVE HOURS

Labor costs and cost due to possible product losses are not included. The figures along the curves indicate the economic optimum number of effects

- A. multi-effect falling film evaporators with vapor compression for fruit juices 10–70%
- B. multi-effect plate type evaporators with vapor compression for fruit juices 10–70%.
- C. centrifugal + wiped film evaporators, one stage 10–70% for fruit juices.
- D. centrifugal + wiped film evaporators, two stage 10–70% for fruit juices.
- E. reverse osmosis flux  $13.5 \text{ l/m}^2/\text{h}$ , 10–30% for fruit juices.
- F. reverse osmosis flux  $3.5 \text{ l/m}^2/\text{h}$ , 10–30% for fruit juices
- G. ultrafiltration, whey, 6–18%, flux  $18 \text{ l/m}^2/\text{h}$

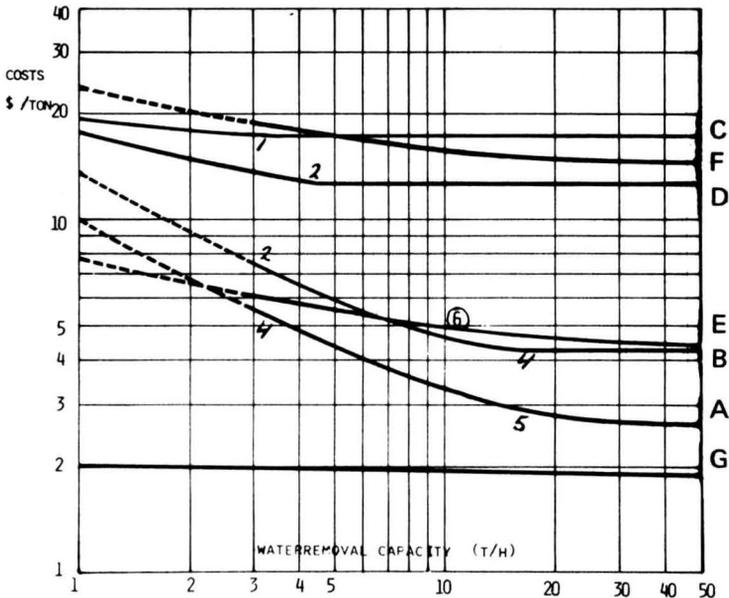


FIG. 6. DEWATERING COSTS IN DOLLARS PER TON WATER REMOVAL WITHOUT AROMA RETENTION OR RECOVERY FOR VARIOUS CONCENTRATION PROCESSES AT 7000 OPERATIONAL HOURS PER YEAR RESPECTIVELY, OF WHICH 90% ARE PRODUCTIVE HOURS.

Labor costs and cost due to possible product losses are not included. The figures along the curves indicate the economic optimum number of effects.

- A. multi-effect falling film evaporators with vapor compression for fruit juices 10–70%.
- B. multi-effect plate type evaporators with vapor compression for fruit juices 10–70%.
- C. centrifugal + wiped film evaporators, one stage 10–70% for fruit juices.
- D. centrifugal + wiped film evaporators, two stage 10–70% for fruit juices.
- E. reverse osmosis flux 13.5 l/m<sup>2</sup> /h, 10–30% for fruit juices
- F. reverse osmosis flux 3.5 l/m<sup>2</sup> /h, 10–30% for fruit juices
- G. ultrafiltration, whey, 6–18%, flux 18 l/m<sup>2</sup> /h.

cheapest by ultrafiltration up to a capacity of about 10 ton water removal per hour. Above this capacity multi-effect falling film evaporators become more economical. Falling film and plate evaporators show about equal costs up to about 10 tons per hour at 2000 hours. Plate evaporators have the advantage of a high turn down ratio in capacity and smaller space requirements. Because of their better steam economy, long tube falling film evaporators are clearly the most economical evaporators at 7000 hr. Reverse osmosis is the most expensive process, at 2000 hr, if the flux is limited to  $3.5 \text{ l/m}^2 \text{ h}$  due to concentration polarization. At a flux of  $13.5 \text{ l/m}^2 \text{ h}$ , which is reported to be feasible, reverse osmosis is more economical than single effect and double effect centrifugal film and scraped film evaporators at both 2000 and 7000 hr. However, the maximum concentration for fruit juices that can be obtained by RO is only 30%, whereas the maximum concentration in evaporators amounts to at least 70%.

**Dewatering Costs of Processes with Aroma Retention or Recovery —** The dewatering costs of processes with aroma recovery or with aroma retention are given for 2000 and 7000 hr per year in Fig. 7 and 8, respectively. Because both in evaporative concentration and in reverse osmosis aromas are, at least partially, lost with the vapor and permeate, aroma recovery by means of stripping and distillation has to be applied. For cost calculation an effective relative volatility of 3.5 at 90% recovery of these aromas has been taken. For lower  $\alpha_{\text{eff}}$  values the dewatering costs are significantly higher because of higher steam consumption. Because of the open pore structure of membranes applied in UF and the corresponding loss of low molecular compounds, no UF data are given in these figures. Freeze concentration does not require aroma recovery by stripping and distillation; because of the high selectivity, no aromas are withdrawn with the ice. Freeze concentration is in effect the only process that withdraws water without any quality loss. It should, however, be mentioned that unlike the dewatering costs of the membranes and evaporation processes which are based on actual price quotations by manufacturers, the dewatering costs of freeze concentration are extra-polated values. Freeze concentration units with capacities of 1 ton water withdrawal per hour or more are not yet commercially available.

The dewatering costs with aroma recovery appear to be the lowest for long tube falling film and plate evaporators, both for 2000 and 7000 operation hours per year. At capacities smaller than 20 tons/h the differences in costs between these two processes are not significant. More costly is reverse osmosis at a flux of  $13.5 \text{ l/m}^2 \text{ h}$ . At a flux of  $3.5 \text{ l/m}^2 \text{ h}$  reverse osmosis is most expensive of all. At 2000 hr/year freeze

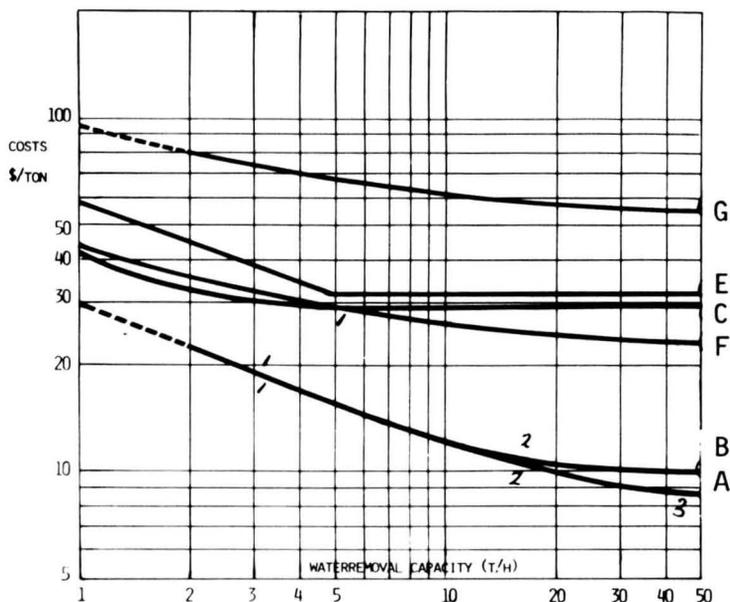


FIG. 7. DEWATERING COSTS IN DOLLARS PER TON WATER REMOVAL WITH AROMA RETENTION OR RECOVERY ( $\alpha_{i,w}$  3.5) FOR VARIOUS CONCENTRATION PROCESSES AT 2000 OPERATIONAL HOURS PER YEAR RESPECTIVELY OF WHICH 90% ARE PRODUCTIVE HOURS.

Labor costs and costs due to possible product losses are not included. The figures along the curves indicate the economic optimum number of effects.

- A. multi-effect falling film evaporators with vapor compression for fruit juices 10–70% + aroma recovery by stripping and distillation.
- B. multi-effect plate type evaporators with vapor compression for fruit juices 10–70% + aroma recovery by stripping and distillation.
- C. centrifugal + wiped film evaporators, one stage 10–70% + aroma recovery by stripping and distillation for fruit juices.
- D. centrifugal + wiped film evaporators, two stage 10–70% + aroma recovery by stripping and distillation for fruit juices.
- E. freeze concentration of fruit juices 10–50%, 0.25 steam equivalent.
- F. reverse osmosis flux of fruit juices 13.5 l/m<sup>2</sup>/h, 10–30% + aroma recovery by distillation
- G. reverse osmosis flux of fruit juices 3.5 l/m<sup>2</sup>/h, 10–30% + aroma recovery by distillation.

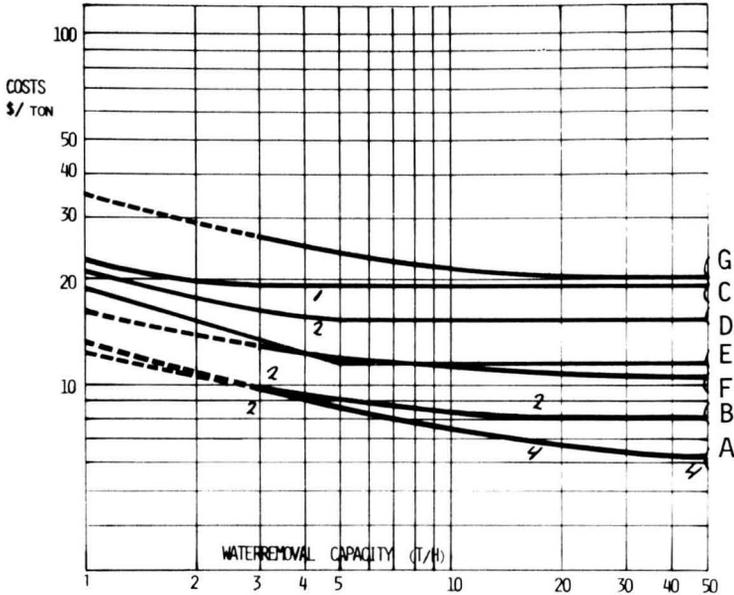


FIG. 8. DEWATERING COSTS IN DOLLARS PER TON WATER REMOVAL WITH AROMA RETENTION OR RECOVERY ( $\alpha_{i,w}$  3.5) FOR VARIOUS CONCENTRATION PROCESSES AT 7000 OPERATIONAL HOURS PER YEAR RESPECTIVELY OF WHICH 90% ARE PRODUCTIVE HOURS.

Labor costs and costs due to possible product losses are not included. The figures along the curves indicate the economic optimum number of effects

- A. multi-effect falling film evaporators with vapor compression for fruit juices 10–70% + aroma recovery by stripping and distillation.
- B. multi-effect plate type evaporators with vapor compression for fruit juices 10–70% + aroma recovery by stripping and distillation
- C. centrifugal + wiped film evaporators, one stage 10–70% + aroma recovery by stripping and distillation for fruit juices.
- D. centrifugal + wiped film evaporators, two stage 10–70% + aroma recovery by stripping and distillation for fruit juices.
- E. freeze concentration of fruit juices 10–50%, 0.25 steam equivalent.
- F. reverse osmosis flux of fruit juices 13.5 l/m<sup>2</sup> /h, 10–30% + aroma recovery by distillation.
- G. reverse osmosis flux of fruit juices 3.5 l/m<sup>2</sup> /h, 10–30% + aroma recovery by distillation.

concentration is slightly more expensive than centrifugal and scraped film evaporators with aroma recovery, the difference, however, is hardly significant. But, at 7000 hr per year freeze concentration is more economical than these evaporators. The costs of freeze concentration and those of RO at a flux of 13.5 l/h with aroma recovery by distillation are almost equal for this condition.

#### ACKNOWLEDGMENT

This study would not have been possible without the detailed information about capital costs, costs of installation, utility consumption and capacities which has kindly been supplied by the following equipment manufacturers:

Alfa-Laval, Lund, Sweden

Abcor Inc., Wilmington, Massachusetts, USA

A.P.V. Nederland B.V., Weesp, the Netherlands

Balfour Co. Ltd., Leven, Scotland

Danske Sukkerfabrikker, D.D.S., R.O.—Division, Nakskov, Denmark

Grenco B.V., 's-Hertogenbosch, the Netherlands

Pfaudler Werke A.G., Schwetzingen, W. Germany

Romicon Inc., Dordrecht, the Netherlands

Stork-Friesland B.V., Gorredijk, the Netherlands

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# LABORATORY EVALUATION OF PAPERBOARD CARTONS FOR THE NONREFRIGERATED STORAGE OF STERILIZED MILK<sup>1,2</sup>

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Received for Publication August 17, 1977

Accepted for Publication November 1, 1977

ABSTRACT

*The suitability of various carton materials for the nonrefrigerated storage of sterilized milk was investigated. One quart paperboard cartons were fabricated from the same base sheet of stock but varied in the type of sizing used to make them resistant to penetration by liquids and whether or not they were aluminum foil-lined. They were preformed and sterilized with ethylene oxide. The four types of paperboard were: (a) rosin (sizing) paperboard (R); (b) rosin paperboard with foil lining (RF); (c) cyanasize (sizing) juice paperboard (CJ); and (d) cyanasize-juice paperboard with foil lining (CJF). Each carton was aseptically filled and sealed, in a glovebox. Incubation was carried out at 20°C for up to nine weeks. Every week five cartons of each type were randomly selected and the milk tested for microbial stability and flavor. The candidate cartons were also tested for degradation of the physical characteristics of static bulge, wicking, tensile strength, and stiffness. Of these, it appears as if selection of carton type will be determined mostly by wicking resistance. The most acceptable carton type is CJF, which had minimal wicking, acceptable bulge, acceptable stiffness, and acceptable tensile strength during the testing period.*

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<sup>1</sup> Scientific Article No. A2309, Contribution No. 5311 of Maryland Agricultural Experiment Station Departments of Dairy Science and Agricultural Engineering.

<sup>2</sup> Supported in part by a grant from the International Paper Company.

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## INTRODUCTION

Heat treatments for milk employing the proper combination of time and temperature could result in a "commercially sterile" product. Historically milk so processed was not readily acceptable to the U.S. consumer. The major objection has been a "cooked" flavor resulting from the severe heat treatment required. The International Dairy Federation (IDF 1972) has defined these ultra high temperature (UHT) processes as pasteurization techniques with temperatures of at least 130°C in a continuous flow and holding times of approximately 1 second or more. Many UHT processes have the potential of producing a sterilized product.

Recently a falling-film pasteurizer-sterilizer (Slater 1973; Westhoff and Doores 1976) has been made commercially available. It offers the possibility of producing a "microbiologically stable" product with a minimal alteration of flavor (Westhoff and Doores 1976) and less protein denaturation (Turk and King 1977).

This investigation was undertaken to obtain preliminary data on possible flavor problems during long term nonrefrigerated storage of milk and to assess the suitability of various paperboard cartons during storage.

## MATERIALS AND METHODS

### Processing of Milk

Milk was processed by a DASI (Chevy Chase, MD) sterilizer at 143.3°C. The system has been described (Slater 1973) and the microbial stability of the milk established (Westhoff and Doores 1976). Briefly milk was preheated in a triple tube heat exchanger to 65.6°C and pumped to the heater (Fig. 1). The milk then flowed through the holding tube (1.79s) to the vacuum chamber for rapid cooling. Installation of the system included the placing of sampling ports (TRUTEST, Line and Tank Sampler Kit, Klenszade Pds., St. Paul, MN) in the stainless steel milk lines. The system was sterilized routinely prior to operation by recirculating water at 148.9°C for 20 min.

### Carton Preparation

Four types of one quart cartons were fabricated and preformed, with a 6.4 mm diameter hole punched in one side of the gable, by the International Paper Company. The types were: a) rosin paperboard (R),

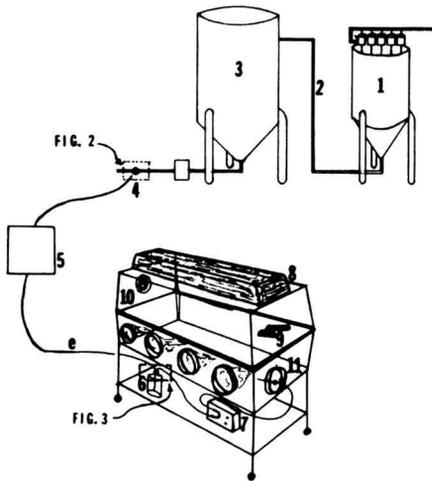


FIG. 1. THE HEATING SYSTEM AND THE ASEPTIC FILLING APPARATUS

(1) The DASI heater; (2) Holding tube; (3) Vacuum chamber; (4) TRUTEST Sampling Port with needle valve attached (For greater detail see Fig. 2); (5) Ice bath; (6) Glass reservoir bottle (For greater detail see Fig. 3); (7) Manostat pump; (8) Fluorescent and ultraviolet lights; (9) Heat extractor; (10) Air filter; and (11) Entry port.

the regular type of milk carton, b) rosin paperboard with foil lining (RF); c) cyanasize-juice paper stock (CJ), the type of carton used for orange juice, and d) CJ paper stock with foil lining (CJF). Rosin (R) and cyanasize (CJ) are two types of paperboard fabricated from the same base sheet of stock. They are different due to the sizing, which is the treatment of paper that gives it resistance to penetration by liquids. Rosin sizing is commonly used in milk packaging, while cyanasize is a more effective sizing agent which imparts increased resistance to fluids other than milk. The J in CJ refers to wet strength materials, which lend more strength to the paper when wet (Cantarella 1976). The cartons were placed in plastic bags (bags had a wick at closure to permit venting of gas), sterilized with ethylene oxide ( $\text{Et}_2\text{O}$ ), and shipped to the University.

#### Preparation of the Transfer System

A needle valve (VWR Scientific, No. 65667-004, Baltimore, MD), shown in Fig. 2, was screwed to a drilled and tapped solid end cap. Attached to the valve was a 61 m length of tubing (Tygon), with an inner diameter of 4.8 mm. A glass tube of 8 cm in length was affixed to the free end of the tubing, to be used as a connector. The glass connector and needle valve were wrapped in aluminum foil, and the entire apparatus sterilized as a unit by autoclaving at  $121^\circ\text{C}$  for 45 min. Before the sterilization of the pasteurizer-sterilizer, the needle valve was attached to one of the sampling ports in the milk line by the use of a hex nut (Fig. 2).

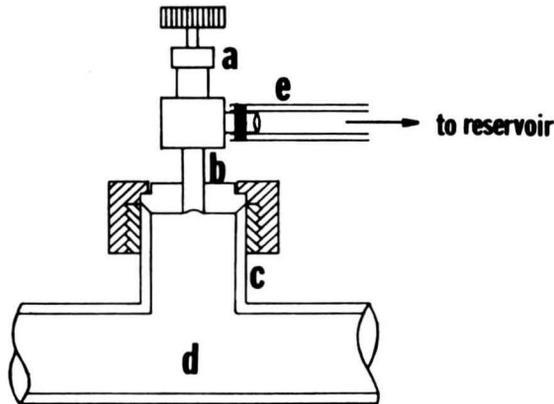


FIG. 2. SAMPLING PORT AND NEEDLE VALVE ASSEMBLY

(a) Needle valve; (b) Drilled and tapped end cap;  
 (c) TRUTEST Sampling Port; (d) Milk flow line;  
 and (e) Tygon tubing (61m).

#### Preparation of the Reservoir and Dispenser

Tubing sections f and h (Fig. 3) were attached to a 15 liter glass reservoir bottle, assembled as shown. Inserted in the free end of tubing section h was a short glass tube for filling the milk through the punched hole in the cartons. The free end of section f and the glass tube on the free end of section h were wrapped in aluminum foil, and the entire reservoir jug apparatus sterilized by autoclaving at 121°C for 45 min.

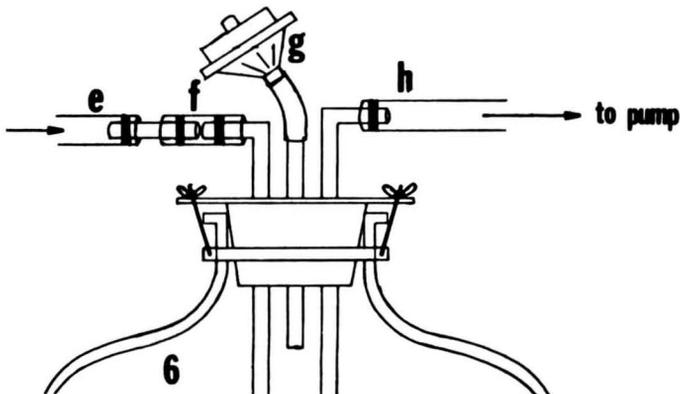


FIG. 3. RESERVOIR JUG

(e) Tygon tubing (61m); (f) Rubber tubing connector with clamps; (g) Gelman H4190 and air filter (Ann Arbor, MICH); and (h) Tygon tubing (3.66m).

### Preparation of the Glovebox

The preformed, plastic bagged, gas-sterilized cartons, and milk cases to support them, plus sealing tape, sterile glass bottles, and a sterile graduated cylinder, were placed inside a glovebox (Germ-Free Labs, six foot (1.8m) plexiglass glovebox, Miami, Fla.) (Fig. 1) and the lid shut. The filler tubing from the presterilized reservoir jug was "threaded" through a pump (MANOSTAT, No. 72-895-05, Silver Spring, MD), and the end placed inside the glovebox through an entry port. To sterilize or decontaminate the glovebox, all ports were sealed, and a stainless steel atomizer (Standard Safety Equipment Co., Palatine, Ill.) was used to spray 250 ml of 2% peracetic acid over all of the interior surfaces.

### Filling Procedure

The glass connector on the end of the 61 m length of tubing (Section e, Fig. 3) was dipped in alcohol and flamed, before attaching it to the flexible connector (Section f, Fig. 3) on the incoming milk flow line of the reservoir jug. Once this connection was made, the excess tubing length was immersed in an ice bath (Fig. 1, Number 5) to cool the milk. The plastic bags (containing cartons) were then opened inside the sterile glovebox and cartons were set upright in milk cases. During the filling operation the needle valve was opened and set to an appropriate flow rate. The output of the metering pump was adjusted to the desired volume with the sterile graduated cylinder. The cartons were then filled through the hole in the gable and sealed with foil tape. The tape was previously sterilized ( $\text{Et}_2\text{O}$ ) and placed into bags by the International Paper Company, and the bags were put in the glovebox before the peracetic acid treatment. In addition to the filling of the cartons, glass bottles, which were presterilized by autoclaving at  $121^\circ\text{C}$  for 30 min, were also filled.

Following the filling and sealing procedure, the glovebox was opened, and the cartons and bottles incubated at  $20^\circ\text{C}$  for up to nine weeks.

### Experimental Design

Sixty-one quart cartons, of each of two types, were aseptically filled and sealed in the glovebox in one run. Sixty cartons of each of the remaining two types were filled in a subsequent run. Glass bottles were filled during the course of both runs. Incubation was carried out at  $20^\circ\text{C}$  for up to nine weeks, with five samples of each type of carton, and glass bottles removed weekly for testing. The microbial stability of

the milk was tested by plating a 0.1 ml sample from each container with Standard Methods Agar (BBL). Those samples yielding no colonies after plate incubation at 32°C for 48 hr were presented to the taste panel. The official American Dairy Science Association score card was used (highest score a numerical value of 40) by a panel of six experts. A pasteurized, commercial milk sample (not stored) was also included each week.

#### Physical Characteristics of the Cartons

Static bulge, wicking, tensile strength, and stiffness were measured on all carton types. Static bulge was measured on one set of five cartons of each type for two replications; Bulge was measured in accordance with Technical Association of the Pulp and Paper Industry (TAPPI) to the nearest 1/32 inch (.79 mm) in the directions of panels 1-3 and 2-4. Bulge measurements were averaged and transformed by subtraction into bulge increases referenced to initial measurements. Values were then treated to remove time dependence in order that carton type and sample direction effects could be investigated by analysis of variance (ANOVA).

Wicking of cartons was determined following removal of the milk on a total of five cartons of each type per week for two replications per carton. An ink outline around the wicked area was transferred to wicking data sheets which were line drawings of an opened carton (Fig. 4).

The same cartons used for wicking were used to obtain samples for tensile strength testing (TAPPI procedure 404). Three ½" × 3" (1.27 × 7.62 cm) samples were cut in the machine direction from each of five cartons of each type for two replications of each carton type. TAPPI grips were used to attach these strips to the Instron universal tubing machine. Samples breaking at the clamps were not counted; they were replaced by other strips. Because our machine reads force in Newtons, a conversion factor of 0.2248 was multiplied to obtain pounds; doubling the result gives the force required to rupture a standard width 1" (2.54 cm) strip.

Stiffness was determined on three 1½" × 2¾" (3.81 × 6.99 cm) samples in both machine and cross directions were obtained from each of five cartons of each type each week. Testing was conducted according to the TAPPI procedure 489 on the Taber stiffness testing machine. Data for each carton and direction was the average of the three samples.

Analyses of variance were used to test statistical significance of treatment effects. This was not possible in one case (stiffness data), and analysis of covariance was used. In other cases, statistical assumptions were required which dictate that very conservative interpretations be

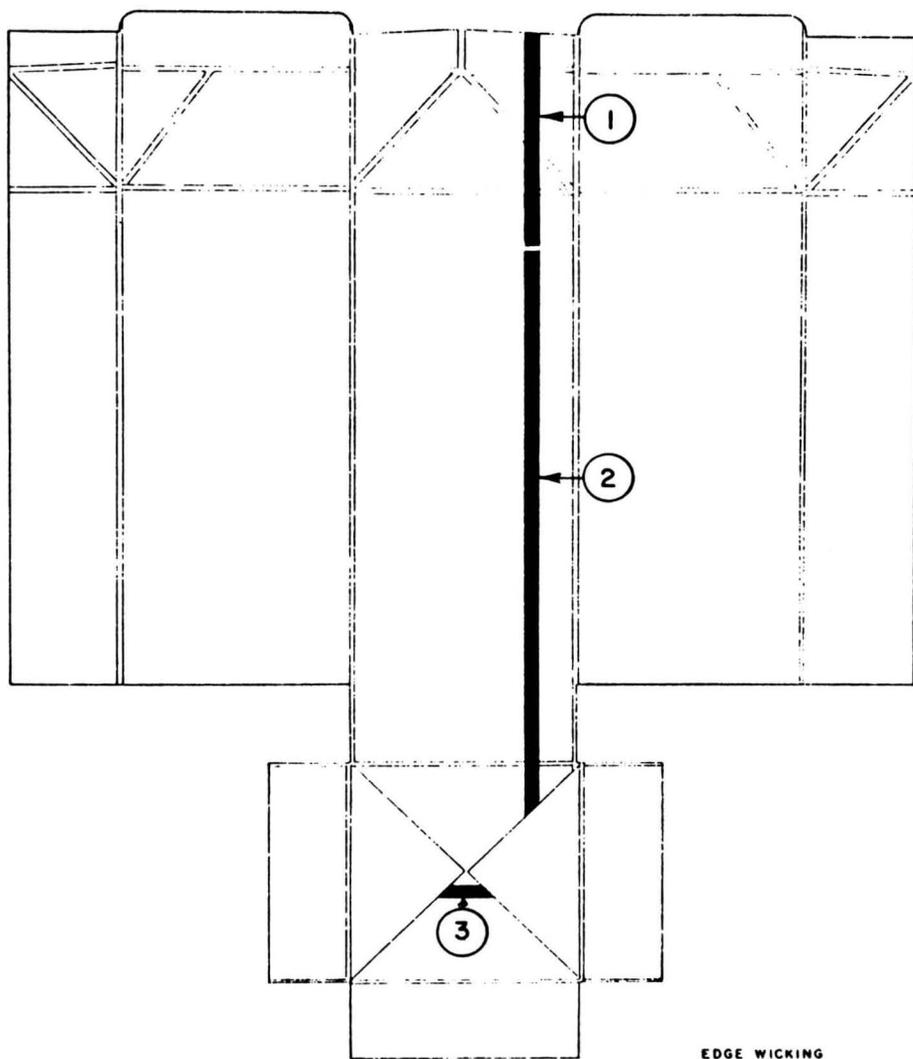


FIG. 4. LINE DRAWING OF OPEN CARTON SHOWING MAJOR WICKING AREAS

made of analyses of variance; therefore, the 1% level of type I error (interpreting a treatment effect as significant when, in reality, it is not) was used.

## RESULTS AND DISCUSSION

## Flavor Evaluation of the Milk

Table 1 contains the means and ranges of numerical flavor scores for sterilized milk stored in rosin paperboard cartons and glass bottles for up to nine weeks at 20°C. This numerical flavor evaluation ratings for pasteurized milk are also shown. This milk was commercially processed and purchased just prior to the taste panel each week, and was therefore not stored. Tables, 2, 3 and 4 contain similar data for the other carton types. In each case the scores for sterilized milk in glass bottles and nonstored pasteurized milk are compared. We have labelled the milk stored in glass bottles as "controls" since we were interested in determining if the paper cartons imparted any off flavors to the milk. Sterilized milk in glass bottles was not expected to score perfectly (a value of 40) but rather would provide a basis for comparison. Fifteen hundred total flavor evaluations of sterilized milk were made during this study. Only nine samples received a score of 32 (unacceptable). These are indicated as numbers in parentheses in Tables 1-4.

Determination of possible significant differences between scores was computed using the Student's t test to a 0.05 level of significance. At this level no significant differences were found between day 0 milk and milk stored for up to nine weeks at 20°C in any of the carton types or in glass bottles. It should be noted, however, that the wide ranges of scores would minimize the significant differences. Interestingly, scores for commercial pasteurized (nonstored) milk varied also (Tables 1-4). Differences between the pasteurized and the sterilized milk were minimal. When milk was stored at 20°C for up to eight weeks in any carton type and glass bottles, there was no significant difference (at the 0.05 level) between these scores and the pasteurized milk scores.

## Static Bulge

Fig. 5 illustrates carton type effect with time. Plotted values are averages of bulge increments in the 1-3 and 2-4 directions. The original data consistently shows the 1-3 direction to bulge more than the 2-4 direction. As noted, open symbols indicate half the data values missing during that week. The above procedure to remove time effects from the ANOVA minimizes the effect of missing data which is caused mainly by carton spoilage or inadequate original sample size.

In general, the addition of foil to the carton reduces its tendency to bulge with time. There appears to be a larger difference between CJ and CJF, than between R and RF. Also, the amount of bulge appears to be

Table 1. Flavor evaluations<sup>1</sup> of milk stored in rosin paperboard cartons

Weeks at 20° C	Rosin Cartons		Control <sup>2</sup>		Pasteurized Milk (not stored)	
	X	Range	X	Range	X	Range
0	38.1	36-40	37.2	32-40 (1) <sup>3</sup>	39.2	38-40
1	39.0	36-40	36.2	33-38	39.2	38-40
2	37.4	35.5-40	37.5	36-38	38.5	38.39.5
3	38.4	34-40	37.4	34-39	39.0	38-39
4	37.7	35-40	38.2	37-39	38.8	37-40
6	36.7	34.5-39	38.0	35-39	38.7	37-39.5
8	38.2	35-40	37.4	36-39	39.3	38-40
9	37.2	32-39 (1) <sup>3</sup>	37.8	36-40	38.8	37-40

<sup>1</sup> The official American Dairy Science Association scorecard was used, with the highest score a numerical value of 40

<sup>2</sup> Presterilized glass bottles, filled simultaneously with the cartons

<sup>3</sup> The number of samples receiving a score of 32 (unacceptable)

Table 2. Flavor evaluations<sup>1</sup> of milk stored in rosin paperboard — foil lined cartons

Weeks at 20° C	Rosin Foil Cartons		Control <sup>2</sup>		Pasteurized Milk (not stored)	
	$\bar{X}$	Range	$\bar{X}$	Range	$\bar{X}$	Range
0	38.3	35-40	37.2	32-40 (1) <sup>3</sup>	39.2	38-40
1	38.6	36-40	36.2	33-38	39.2	38-40
2	38.7	36-40	37.5	36-38	38.5	38-39.5
3	38.7	36.5-39	37.4	34-39	39.0	38-39
4	38.7	35-40	38.2	37-39	38.8	37-40
6	37.2	34.5-39	38.0	35-39	38.7	37-39.5
8	38.2	36.5-40	37.4	36-39	39.3	38-40
9	37.2	32-39.5 (2) <sup>3</sup>	37.8	36-40	38.8	37-40

<sup>1</sup> The official American Dairy Science Association scorecard was used, with the highest score a numerical value of 40

<sup>2</sup> Pasteurized glass bottles, filled simultaneously with the cartons

<sup>3</sup> The number of samples receiving a score of 32 (unacceptable)

Table 3. Flavor evaluations<sup>1</sup> of milk stored in cyanasize-juice paperboard cartons

Weeks at 20° C	CJ Cartons		Control <sup>2</sup>		Pasteurized Milk (not stored)	
	$\bar{X}$	Range	$\bar{X}$	Range	$\bar{X}$	Range
0	38.0	35-39.5	39.3	38-40	38.6	37-40
1	38.3	36-40	38.6	37-40	38.8	37-40
3	37.2	33-40	37.4	36-39	38.8	37-40
4	36.8	32-40 (1) <sup>3</sup>	36.5	36-38	37.9	37-38.5
5	37.8	35.5-39	36.0	32-40 (1) <sup>3</sup>	37.8	35-39
7	36.6	32-39 (1) <sup>3</sup>	37.3	35-39	38.3	38-39
8	37.1	34-39	37.4	36-39	38.2	37-39
9	36.8	32-40 (1) <sup>3</sup>	36.0	35-39...	38.6	37-39.5

<sup>1</sup> The official American Dairy Science Association scorecard was used, with the highest score a numerical value of 40

<sup>2</sup> Presterilized glass bottles, filled simultaneously with the cartons

<sup>3</sup> The number of samples receiving a score of 32 (unacceptable)

Table 4. Flavor evaluations<sup>1</sup> of milk stored in cyanasize-juice paperboard — foil lined cartons

Weeks	CJ Foil Cartons		Control <sup>2</sup>		Pasteurized Milk (not stored)	
	$\bar{X}$	Range	$\bar{X}$	Range	$\bar{X}$	Range
0	38.2	35-40	39.3	38-40	38.6	37-40
1	38.1	32-40 (1) <sup>3</sup>	38.6	37-40	38.8	37-40
3	37.5	33-40	37.4	36-39	38.8	37-40
4	36.4	30-40	36.5	36-38	37.9	37-38.5
5	37.5	36-40	36.0	32-40 (1) <sup>3</sup>	37.8	35-39
7	37.4	37-39	37.3	35-39	38.3	38-39
8	37.5	34-39	37.4	36-39	38.2	37-39
9	36.7	34-40	36.0	35-39	38.6	37-39.5

<sup>1</sup> The official American Dairy Science Association scorecard was used, with the highest score a numerical value of 40.

<sup>2</sup> Presterilized glass bottles, filled simultaneously with the cartons

<sup>3</sup> The number of samples receiving a score of 32 (unacceptable)

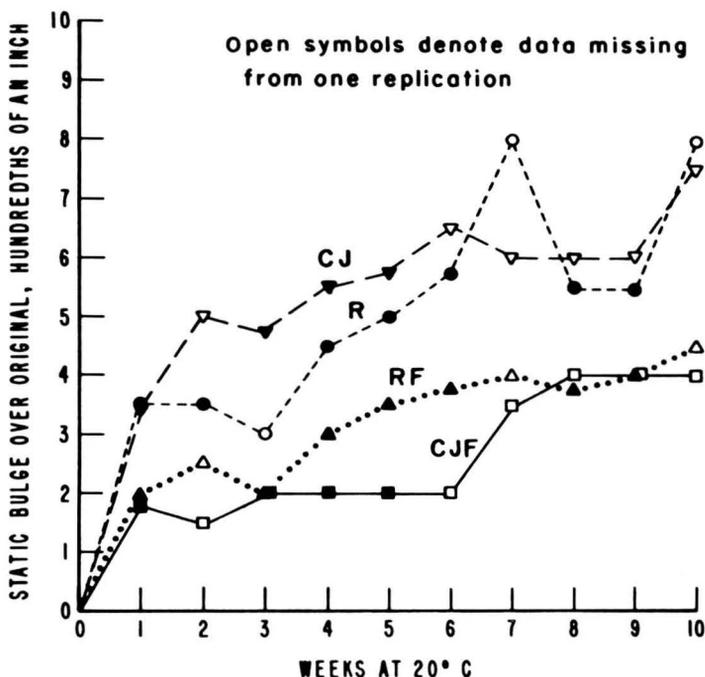


FIG. 5. STATIC BULGE INCREASES OF ONE QUART CARTONS CONTAINING STERILIZED MILK

R = rosin paperboard; RF = rosin paperboard with foil lining; CJ = cyanasize-juice paperboard; CJF = cyanasize juice paperboard with foil lining.

limited after a sufficiently long elapsed time. Approximately one-half of the long-range bulge increment appears in the first week. Both carton type and sample direction were important at the 0.01 probability level (Table 5).

Table 5. Static bulge difference ANOVA

Source	df	SS	MS	F
Total	8	0.00176	0.00022	
Mean	1	0.00001	0.00001	
Carton type	3	0.00119	0.00040	40 <sup>1</sup>
Direction	1	0.00053	0.00053	53 <sup>1</sup>
Error	3	0.00003	0.00001	

<sup>1</sup> Significant at  $\alpha = 0.01$

### Carton Wicking

Wicking, when present, occurred predominantly in three areas (Fig. 4): (1) along the side seam with the wicking front parallel to the seam, (2) at and above the level of the milk surface in the carton at the side seam with the wicking front more or less parallel to the seam, (3) at the bottom triangular area where the paperboard overlaps in the center with the wicking front roughly parallel to the cut edge of the board.

Average wicking area was the one measure picked to represent wicking tendencies of the cartons. Especially after several weeks, average wicking area was extremely variable and depended greatly on the influence of one or several cartons which appeared to wick much more than the others. Fig. 6 illustrates the average wicking area for all types of cartons over the ten week period. As noted, open symbols represent averages based on no more than half the maximum number of data values.

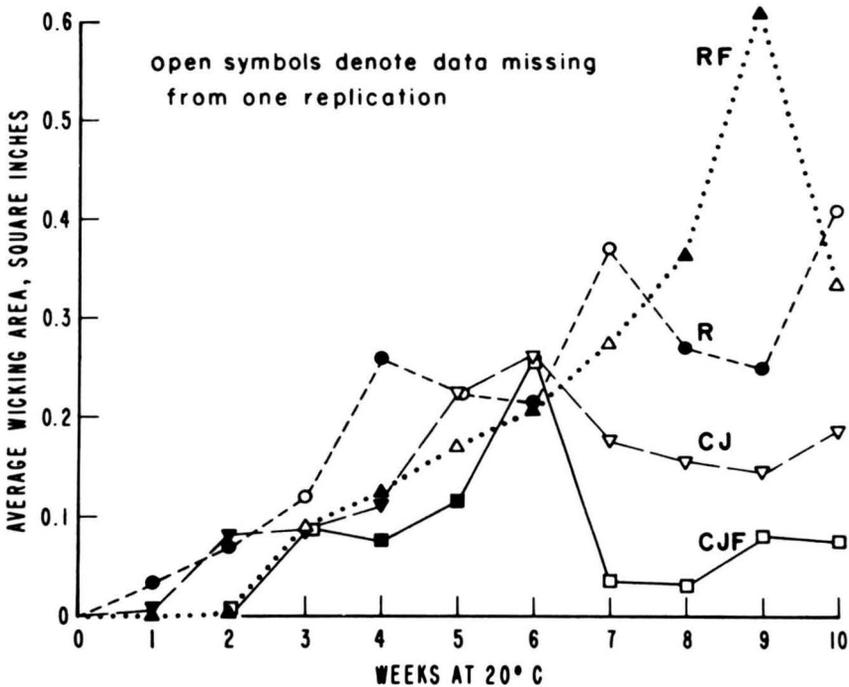


FIG. 6. AVERAGE WICKING AREA OF ONE QUART PURE-PAK CARTON CONTAINING STERILIZED MILK

R = rosin paper stock; RF = rosin paper stock with foil lining; CJ = cyanize juice paper stock; CJF = cyanize juice paper stock with foil lining.

Fig. 7 illustrates the results of counting the number of cartons above an arbitrary wicking area of  $0.150 \text{ in}^2$  ( $0.968 \text{ cm}^2$ ). If  $0.150 \text{ in}^2$  were considered unacceptable, well over 50% of R, RF, and CJ cartons would have failed by the fifth week. The CJF carton is clearly superior in wicking characteristics.

Data values were averaged for each week and carton type and considered as a randomized complete block experiment. The analysis of variance (ANOVA) is shown in Table 6. Both time and carton type significantly influence the data based on the 1% level of significance.

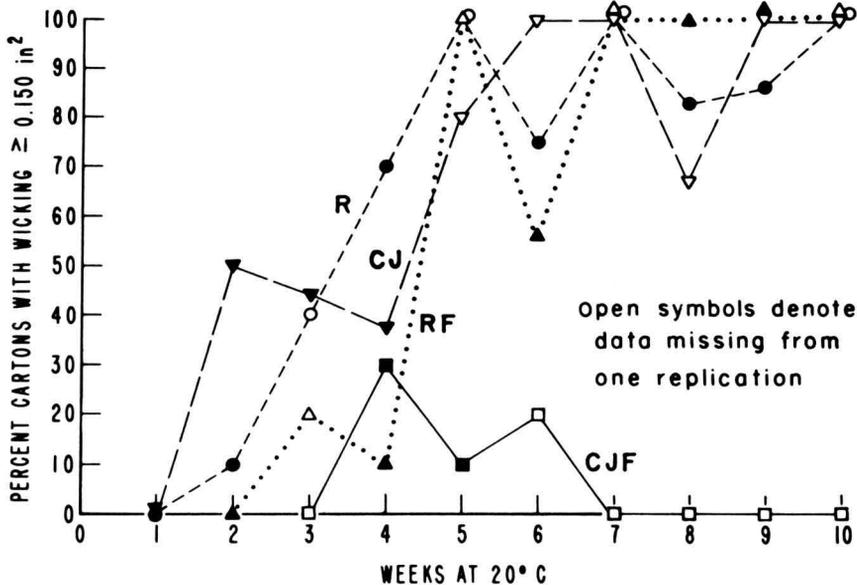


FIG. 7. PERCENT OF CARTONS EXCEEDING THE ARBITRARY WICKING AREA OF  $0.150 \text{ IN}^2$ .

R = rosin paperboard; RF = rosin paperboard with foil lining; CJ = cyanasize juice paperboard; CJF = cyanasize juice paperboard with foil lining.

Table 6. Average wicking area ANOVA

Source	df	SS	MS	F
Total	44	1.780		
Mean	1	1.001	1.001	
Carton type	3	0.128	0.0427	4.91 <sup>1</sup>
Time	10	0.389	0.0389	4.47 <sup>1</sup>
Error	30	0.262	0.0087	

<sup>1</sup> Significant at  $\alpha = 0.01$ .

### Carton Tensile Strength

Fig. 8 shows tensile strength with time for all four carton types. The reduction of tensile strength with time is not great, being in the range 7–20% over the full ten weeks. This factor does not appear to limit carton storage of sterilized milk for long periods of time. A randomized complete block analysis of variance (ANOVA) was computed, therefore, at the expense of some statistical sensitivity, determination of time  $\times$  carton interaction, and some statistical rigor. The ANOVA appears in Table 7. Both time and carton type have significant effects on tensile strength when tested at the 0.01 statistical level.

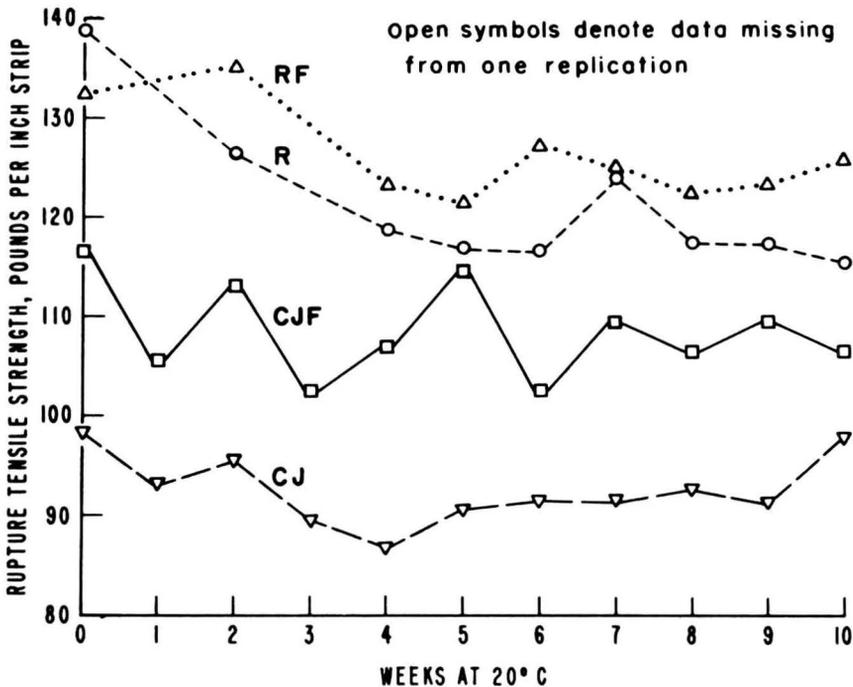


FIG. 8. REDUCTION OF TENSILE STRENGTH OF ONE QUART CARTONS CONTAINING STERILIZED MILK

R = rosin paperboard; RF = rosin paperboard with foil lining; CJ = cyanasize juice paperboard; CJF = cyanasize juice paperboard with foil lining.

### Carton Taber Stiffness

Fig. 9 shows Taber stiffness measurements averaged over both directions as they vary with time. Almost no decrease in stiffness is noted over the full nine weeks. Also, both RF and CJF curves are nearly the

Table 7. Tensile strength ANOVA

Source	df	SS	MS	F
Total	24	312,253.74		
Mean	1	307,844.42		
Time	5	529.02	105.80	7.18 <sup>1</sup>
Carton type	3	4,188.23	1,396.08	94.71 <sup>1</sup>
Error	15	221.09	14.74	

<sup>1</sup> Significant at  $\alpha = 0.01$

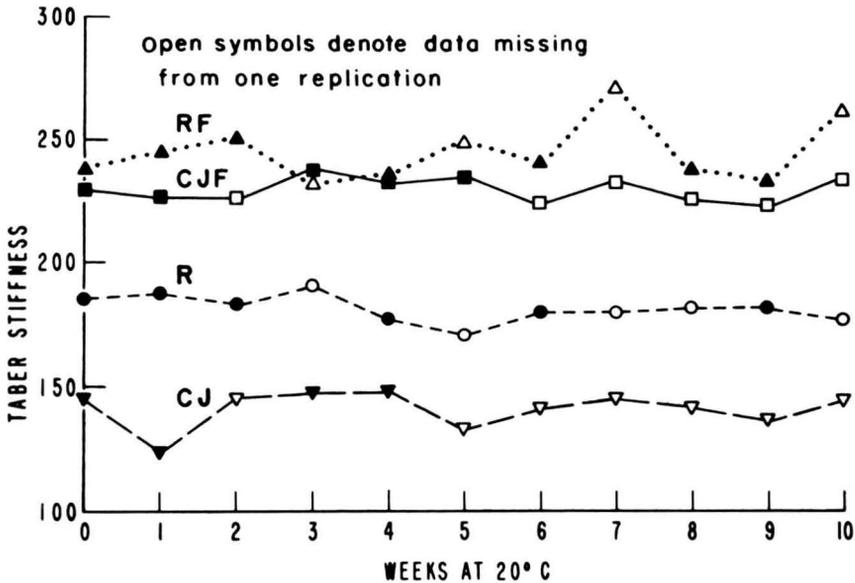


FIG. 9. EFFECT OF STORAGE ON TABER STIFFNESS OF ONE QUART CARTONS CONTAINING STERILIZED MILK

R = rosin paperboard; RF = rosin paperboard with foil lining; CJ = cyanasize juice paperboard; CJF = cyanasize juice paperboard with foil lining.

same, meaning that probably no practical distinction can be made between these two types of cartons. With little or no degradation in stiffness, this property will not limit usefulness of the cartons for long-term storage of milk.

#### ACKNOWLEDGMENTS

The authors thank Dr. L. W. Douglass for statistical assistance.

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# FIRMNESS AND THERMAL CONDUCTIVITY OF RED BEETS IN STORAGE

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Received for Publication August 24, 1977

Accepted for Publication December 5, 1977

## ABSTRACT

*Firmness and thermal conductivity of "Detroit Dark Red, Ferry's Strain" and "Ruby Queen" beets were determined after storage at: 0°C, 95% relative humidity (RH); 12.8°C, 24% RH, and 1.7°C, uncontrolled humidity. Firmness at fixed temperature and humidity conditions decreased with storage; it could be described by a first order rate equation. At a fixed temperature (1.7°C) and uncontrolled humidity, magnitudes of firmness varied as the humidity varied; in contrast, thermal conductivity was influenced very little. At 0°C and 95% rel. humidity, thermal conductivity remained essentially constant for four weeks. The magnitude of thermal conductivity, at room temperature and on the day of harvest, was about the same for four different varieties of beets: "King Red," "Detroit Dark Red, Ferry's Strain," "Ruby Queen," and "Detroit Dark Red, Morse's Strain;" it ranged between 0.579 and 0.600 W/m °K.*

## INTRODUCTION

Physical and thermal properties of foods play an important role in the engineering design of packaging and process equipment. While numerous studies have been conducted on the post-harvest physiology and storage life of fruits and vegetables, studies related to physical and thermal properties of fruits and vegetables in storage are not extensive.

Lutz and Hardenburg (1968) summarized the large number of studies on the storage life of fruits and vegetables. Thorne and Th.

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Meffert (1976) showed that storage temperature and life of fruits and vegetables can be related by a power type equation. Lentz *et al.* (1971) studied the factors affecting rate of moisture loss of stored fruits and vegetables, while Van den Berg and Lentz (1973) investigated decay and quality of onions and potatoes. Post-harvest texture change of asparagus was analyzed by Sharma *et al.* (1975); peak force measured in a shear cell was the studied texture parameter.

Beets are well adapted to storage; topped beets can be stored for 3 to 5 months at 0°C and 95% relative humidity (RH). However, the influence of storage conditions on texture and thermal conductivity has not been documented. Thermal conductivity and firmness data were obtained on beets in storage to ascertain: (1) the variation of these properties with beet variety, (2) whether the two properties could be described mathematically during storage at constant temperature and humidity, and (3) the influence of varying storage humidity, at a constant temperature, on the two properties. It is our hope that the study would provide insights on the importance of storage conditions for the determination of textural and thermal properties.

#### MATERIALS AND METHODS

“Ruby Queen” and “Detroit Dark Red, Ferry’s Strain” (DDR-FS) beets were grown on the experimental plots of the Geneva Agricultural Experiment Station. After harvest, the beets were topped, graded, and washed. Large beets (6 to 10 cm diameter) were selected for experimentation; the large sizes were used to facilitate measurement of thermal conductivity (Rao *et al.* 1975). The beets were stored in different chambers: 0°C, 95% RH; 12.8°C, 24% RH; and at 1.7°C. The 1.7°C chamber was a general purpose storage room and its humidity, which was not measured, varied with the quantity and nature of products (mainly vegetables) stored. The three storage conditions will be referred to as SC I, SC II, and SC III, respectively.

Deformation and puncture tests were conducted, at room temperature, on 30 beets after allowing about 4 hr for temperature equilibration. A 0.32 cm diameter punch was used in the puncture tests. For the deformation tests a force of 5 Newtons was used along with a cross head speed of 0.5 cm/min and a recorder chart speed of 50 cm/min. The tests were conducted with an Instron Universal Testing Machine; the machine and the test procedures have been reviewed and the original studies referenced by Bourne (1975).

Thermal conductivity of cylindrical samples from the beets was measured at room temperature by the line source method. The procedures were the same as those used for potatoes and squash (Rao *et al.* 1975). Five thermal conductivity determinations were made on samples taken from large beets, after they reached room temperature; the tests were performed after the deformation tests but before the puncture tests. The moisture content of the cylindrical samples was determined by freeze drying first followed by air drying (Rao *et al.* 1975). Because of rapid wilting of beets stored at SC II, it was not possible to perform thermal conductivity tests on them. Also, based on tests with SC III, little change in thermal conductivity was expected in beets stored at a high relative humidity (SC I). For this reason, with SC I beets thermal conductivity tests were conducted on the day of harvest, and three and four weeks after harvest.

Thermal conductivity tests were also conducted to determine the effect of variety and blanching on freshly harvested beets. For the effect of variety "King Red" and "Detroit Dark Red, Morse's Strain" (DDR-MS) were the additional varieties used, while for the effect of blanching "Ruby Queen" and "Detroit Dark Red, Ferry's Strain" beets were the only varieties used; the beets were blanched in boiling water for 20 min and cooled for 5 min in running cold water. The tests were completed within 4 to 6 hr after harvest.

## RESULTS AND DISCUSSION

Puncture force or its logarithm was not a satisfactory index of the physiological change in stored beets; this can be seen from the data for "DDR-FS" beets at SC II shown in Fig. 1. In contrast, the firmness data (expressed as reciprocal of the deformation) of beets at SC I and SC II were satisfactory as an index of physiological change; this can be seen from Fig. 2 and 3, where the log of firmness is plotted vs the storage time. In Fig. 3, the magnitude of firmness on the day of harvest is lower than that after 24 hr storage; this apparently reflects the loss of turgor pressure during the several hours between harvest and deformation test. As expected, the firmness at SC II decreases at a rate faster than at SC I.

It is also clear from Fig. 2 and 3, that the firmness of beets during storage can be expressed by a first order rate equation:

$$F = A e^{Bt} \quad (1)$$

where,  $F$  is the firmness,  $t$  is the storage time, and  $A$  and  $B$  are constants whose magnitudes depend on the storage conditions. Employing least square analysis, the magnitudes of  $A$  and  $B$  were estimated; they are listed in Table 1 along with the magnitudes of  $r^2$ .

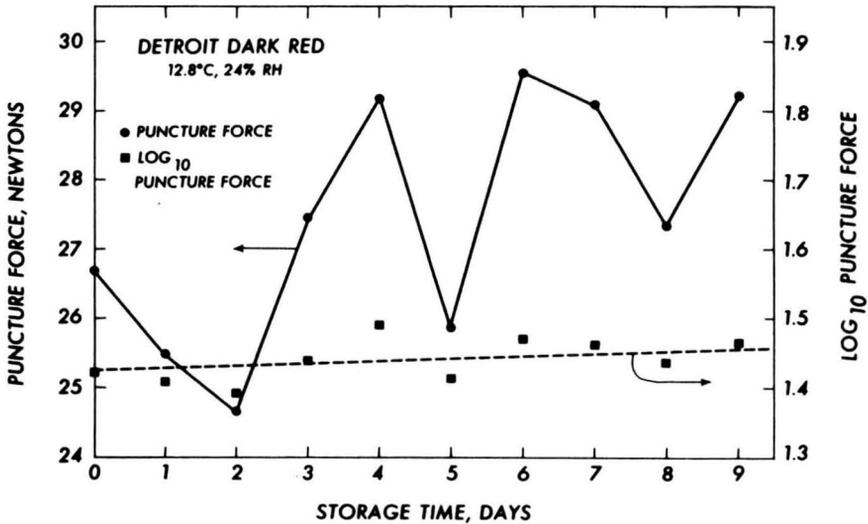


FIG. 1. PUNCTURE FORCE AND LOG<sub>10</sub> PUNCTURE FORCE OF DETROIT DARK RED (FERRY'S STRAIN) BEETS AS A FUNCTION OF STORAGE

The coefficient  $A$  represents the magnitude of firmness at time zero or at the time of harvest. The coefficient  $B$  can be used to obtain the fractional change,  $f$ , in firmness produced by one day storage:

$$f = \frac{Ae^{Bt} - Ae^{B(t-1)}}{Ae^{Bt}} \tag{2}$$

It can be readily shown that Eq. (2) reduces to:

$$f = 1 - e^{-B} \tag{3}$$

For small values of  $B$ , the approximation  $e^{-B} = 1 - B$  can be used; therefore,

$$f = B$$

For example, the error due to the approximation for  $B = -0.31$  is 6.3%.

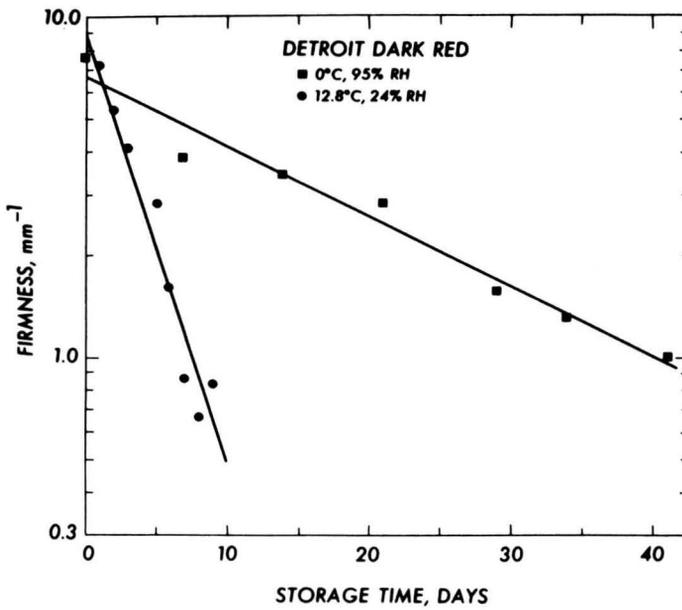


FIG. 2. SEMI-LOGARITHMIC PLOT OF STORAGE TIME VS FIRMNESS FOR DETROIT DARK RED (FERRY'S STRAIN) BEETS AT 0°C, 95% RH AND 12.8°C, 24% RH

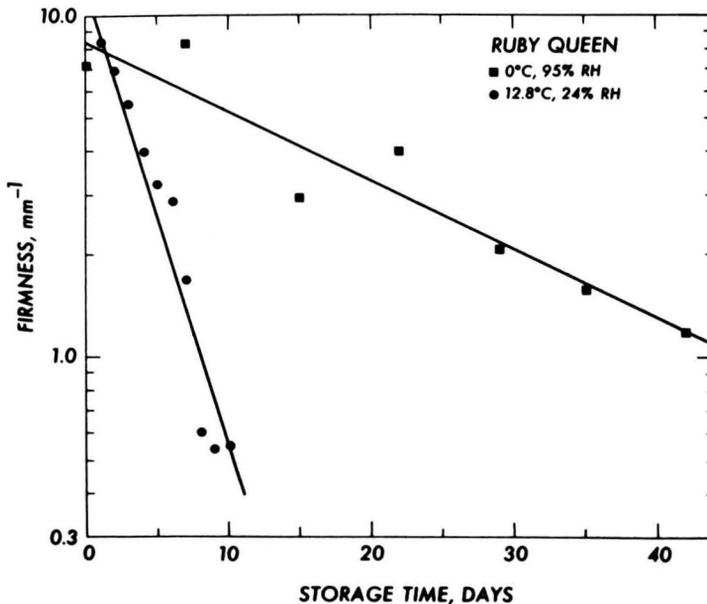


FIG. 3. SEMI-LOGARITHMIC PLOT OF STORAGE TIME VS FIRMNESS FOR RUBY QUEEN BEETS AT 32°F, 95% RH AND 12.8°C, 24% RH

The magnitudes of B listed in Table 1 reflect properly the different storage conditions; also, there is not a significant difference between the two varieties for a given storage condition. For a given variety, the magnitudes of A derived from the two storage conditions in Table 1 should have been equal; however, they are different by about 28%. The deviation is attributed to extrapolation of results based on a small amount of data, particularly at SC I.

Table 1. Magnitudes of regression coefficients for Equation 1.

Storage Condition	Detroit Dark Red			Ruby Queen		r <sup>2</sup>
	A	B	r <sup>2</sup>	A	B	
0°C, 95% Rel. Humidity	6.64	-0.047	0.967	8.33	-0.046	0.899
12.8°C, 24% Rel. Humidity	9.07	-0.293	0.953	11.71	-0.307	0.916

Magnitudes of firmness at SC III, shown in Fig. 4 and 5, clearly indicate the importance of humidity on firmness changes. For both varieties, the firmness remained approximately constant for about a day before the humidity variation was reflected in the magnitude of firmness. Moisture content of beet samples changed little during storage at SC III; for example, the moisture content of "DDR-FS" beets varied between 88.8% and 87.6%. For this reason, the variation in firmness is attributed to changes in turgor pressure. It is clear from these data that for determining textural properties of beets, temperature control alone is not sufficient. Further, it would be desirable to know the temperature and humidity histories from the time of harvest.

Thermal conductivity data at SC III are also shown in Fig. 4 and 5; as seen, thermal conductivity decreased at a slow rate. Variation in storage humidity did not cause large changes in thermal conductivity. Under SC I, the thermal conductivity remained essentially constant over a period of four weeks; the magnitudes of "DDR-FS" beets were: 0.579, 0.591, and 0.585 W/m °K after 0, 21, and 28 days from harvest, respectively, the corresponding figures for "Ruby Queen" beets were: 0.590, 0.582, and 0.584 W/m °K.

Sweat (1974) observed that the thermal conductivity of several fruits and vegetables (including beets) at ambient temperature and their moisture content were linearly related. Because the moisture content of the beets in this study did not change much during storage, the thermal conductivity also did not change significantly.

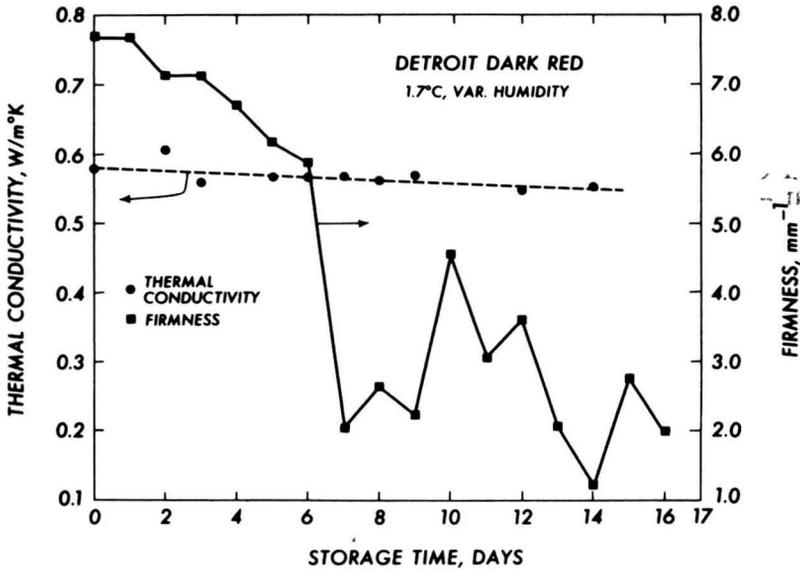


FIG. 4. FIRMNESS AND THERMAL CONDUCTIVITY OF DETROIT DARK RED (FERRY'S STRAIN) BEETS AT 1.7°C AND VARIABLE HUMIDITY

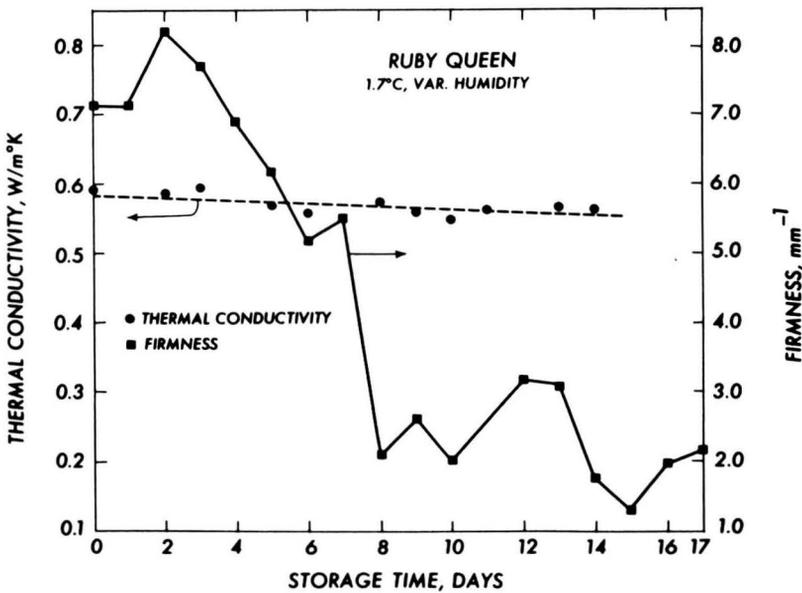


FIG. 5. FIRMNESS AND THERMAL CONDUCTIVITY OF RUBY QUEEN BEETS AT 1.7°C AND VARIABLE HUMIDITY

The thermal conductivity of "King Red" and "DDR-MS" beets on the day of harvest was 0.595 and 0.600 W/m °K, respectively. From these observations and the thermal conductivity values of "Ruby Queen" and "DDR-FS" beets, it can be concluded that there is little difference in the thermal conductivities of the various varieties. Blanching "Ruby Queen" and "DDR-FS" beets resulted in a negligible increase in the magnitude of thermal conductivity: from 0.553 to 0.586, and from 0.573 to 0.604, respectively. The observed increases are not only well within the reproducibility of thermal conductivity data with the line source technique, but could also be due to expulsion of air and the slight increase in moisture content resulting from blanching.

#### SUMMARY

The results of this study show that the firmness (as measured by deformation) of beets decreases with storage even at 0°C and 95% RH. At constant temperature and humidity storage, the decrease in firmness can be described by a first order rate equation. At 1.7°C, the firmness depended on the varying magnitude of the humidity in the storage chamber; in contrast, the thermal conductivity was not affected significantly. The thermal conductivity of four varieties of beets: "Ruby Queen," "DDR-FS," "King Red," and "DDR-MS" was of the same order of magnitude. Blanching caused a slight but not a significant increase in the magnitude of thermal conductivity of "Ruby Queen" and "DDR-FS" beets.

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# MASS TRANSFER CHARACTERISTICS OF HOLLOW FIBER ULTRAFILTRATION OF SOY PROTEIN SYSTEMS

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Received for Publication August 17, 1977  
Accepted for Publication December 5, 1977

## ABSTRACT

*Mass transfer characteristics and the applicability of theoretical transport models in ultrafiltration were evaluated in a pilot-scale hollow fiber system processing soybean water extracts. Permeate flux was found to increase linearly with transmembrane pressure and then reach an asymptotic value. Substantial deviations from the pure water permeation line and marked hysteresis effects were also observed, indicative of concentration polarization and additional hydrodynamic resistance at the membrane surface. Pressure-independence occurred at lower pressures with lower Reynolds Numbers ( $Re$ ) and higher Schmidt Numbers ( $Sc$ ). Flow rate and temperature significantly affected flux in the totally polarized region. The exponent in the flux- $Re^\alpha$  relationship varied from 0.6 at 20°C to 0.32 at 50°C. Theoretical laminar flow mass transfer models were found to predict Sherwood Numbers ( $Sh$ ) and flux lower than experimental values. The correlation that fit the data best was*

$$Sh = 0.181 (Re)^{.47} (Sc)^{.33}$$

## INTRODUCTION

Membrane separation methods are rapidly becoming an important industrial process for the concentration, purification and separation of colloidal or macro-molecular species in solution. Their main attractions in biological and food applications are mild operating conditions, simplicity, potentially high selectivity and, in the case of low pressure processes such as ultrafiltration, low energy requirements. Ultrafiltration is particularly attractive in protein systems which are quite

sensitive to extreme changes in their environment such as would occur during thermal evaporation and concentration. Of major interest in this area is the concentration and purification of enzyme solutions (Wang *et al.* 1969), recovery of protein from cheese whey (Horton 1974; Fenton-May *et al.* 1971) production of milk concentrates (Fenton-May *et al.* 1972) and soy protein concentrates low in undesirable components (Omosaiye *et al.* 1978; Omosaiye and Cheryan 1977; Cheryan and Schlessler 1977). The development of noncellulosic anisotropic membranes in a variety of configurations has given an impetus to studies of ultrafiltration applications in the food, pharmaceutical and biochemical areas (Porter and Nelson 1971; Porter 1972). Such high flux membranes, however, also lead to a rapid build-up of rejected solute on the membrane surface, a phenomenon known as "concentration polarization." This invariably reduces flux and, in the case of proteins that can gel easily, causes severe fouling of the membrane. Most studies of the flux behavior of ultrafiltration systems focus directly or indirectly on optimizing mass transfer characteristics, since the concentration polarization layer is the rate limiting step in the overall process (Blatt *et al.* 1970; Porter 1972; Goldsmith 1971). There is a paucity of data, however, obtained from relatively large-scale ultrafiltration units using systems of commercial interest. Most studies concerning the development of appropriate mass-transfer models have been done on small laboratory-scale apparatus using model protein systems. This pilot-scale study was undertaken to optimize operating parameters such as flow rate, temperature and feed concentration and to examine the applicability of existing boundary layer filtration mass transfer models in describing the hollow-fiber ultrafiltration of soy protein extracts. The overall commercial objective is to produce a soy protein isolate or concentrate low in undesirable components and with high-quality functional properties.

### Models for Predicting Flux

A number of theoretical models for various mechanisms of membrane transport are described in the literature. It is generally believed that the best description of macroporous membranes (i.e. membranes that retain only macromolecules such as proteins) assumes that pores are right-circular cylinders of uniform radius extending completely through the membrane, perpendicular to the surface (Lonsdale 1972; Baker *et al.* 1972). Solvent flux is then written in terms of Poiseuille's law

$$J = \frac{\epsilon r^2 g_c \Delta P_T}{8 \Delta x \mu}$$

where $J$	=	water flux
$\epsilon$	=	porosity
$r$	=	pore radius
$\Delta x$	=	membrane thickness
$\mu$	=	viscosity
$g_c$	=	gravitational constant
$\Delta P_T$	=	transmembrane pressure, the average of inlet and outlet pressures on the retentate side of membrane (assuming the permeate side is open to the atmosphere).

Real ultrafiltration membranes, however, do not have such perfect pores, and factors such as variations in pore size distribution, shape and tortuosity will affect the relationship expressed in equation 1. In addition, rejection of a solute by a membrane is not based on size alone and some interaction with the membrane probably occurs, especially if the membrane possesses a significant charge (Lonsdale 1972).

For a particular membrane-protein system, equation 1 is generally written as

$$J = A \frac{(\Delta P_T - \Delta \pi)}{\mu} \quad (2)$$

where  $A$  is defined as a membrane permeability coefficient, and  $\Delta \pi$  is the transmembrane osmotic pressure of the rejected solute against which the driving force  $\Delta P_T$  is applied.

Equation 2 has been found to be applicable over a limited range of operating conditions such as low protein concentrations and low pressures, particularly for the noncellulosic anisotropic membranes used in this study (Baker *et al.* 1972; Blatt *et al.* 1970; Cheryan and Schlessler 1977). At higher pressures and high protein concentrations, the flux typically becomes independent of pressure (see Fig. 3), and this has been shown to be the result of a build-up of rejected protein molecules on the membrane surface. This can reduce flux by one of two mechanisms: in one view, such as with reverse osmosis systems, the increased solute concentration on the membrane surface can result in a significantly higher osmotic pressure, resulting in a decrease in the driving force ( $\Delta P_T - \Delta \pi$ ) and flux. Kozinski and Lightfoot (1972) have suggested that the pressure difference between the pure water line and protein curve can be interpreted as a measure of osmotic pressure effects which may be the major contributor to lower flux. In protein ultrafiltration, however, osmotic pressures are small in view of the large molecular weights involved (7,000–600,000 in this case) although

Goldsmith (1971) has suggested that osmotic pressure may become significant during concentration due to the importance of the second virial coefficient term in the expression for calculating osmotic pressure.

The alternate view is preferred by many investigators (Blatt *et al.* 1970; deFelippi and Goldsmith 1970; Goldsmith 1971; Porter 1972), i.e. that the deviation represents a significant hydrodynamic resistance in the boundary layer. This model is schematically represented in Fig. 1 where concentration polarization is described to a first approximation by simple film theory. As the solution is ultrafiltered, solute is brought to the membrane surface by convective transport at a rate,  $J_s$ , where

$$J_s = J C_B \quad (3)$$

where  $J$  = permeate flux and  $C_B$  = concentration of solute in bulk of solution. The resulting concentration gradient causes the solute to be transported back into the bulk of the solution due to diffusional effects within the fluid above the membrane. Neglecting axial concentration gradients, the rate of back-transport of solute will be given by

$$J_s = D \frac{dC}{dx} \quad (4)$$

where  $D$  is diffusion coefficient and  $dC/dx$  is the concentration gradient over a differential element in the boundary layer.

Initially the rate of convective transport is greater than the back transport rate, but eventually a concentration is reached where no more solute can be accommodated in the surface layer due to restricted mobility of solute molecules; a "close-packed" configuration is obtained, denoted as  $C_G$  in Fig. 1 (Michaels *et al.* 1971). The two mechanisms balance each other and, at a steady state, equations 3 and 4 can be equated and integrated:

$$J = K \ln \frac{C_G}{C_B} \quad (5)$$

where  $K$  is the mass transfer coefficient.  $K$  has the same units as flux, and is equal to  $D/\delta$  where  $\delta$  is the thickness of the boundary layer over which the concentration gradient exists.

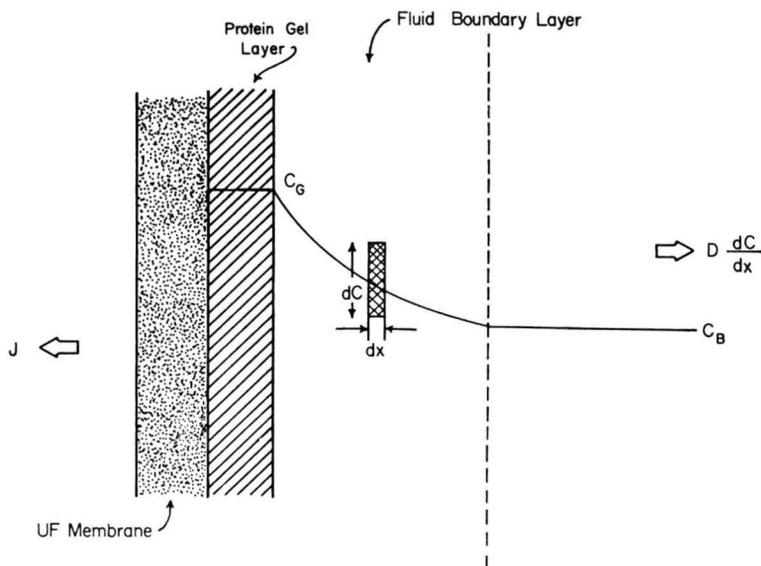


FIG. 1. SCHEMATIC ILLUSTRATION OF CONCENTRATION POLARIZATION

This model implies that the final ultrafiltration flux is controlled by the rate at which solute is transferred back from the membrane surface into the bulk fluid. Since, in most operations, the values of  $C_G$  and  $C_B$  are relatively fixed by physicochemical properties of the feed, efforts to improve permeate flux usually focus on enhancing  $K$  as much as possible by adjusting operating parameters. As mentioned frequently in the literature, any adjustment that increases flux without providing a compensating mechanism to increase rate of back-transport will be self-defeating. This explains why increasing transmembrane pressure has little effect on flux in the totally polarized region of operation, as seen in Fig. 3. The increased driving force merely resulted in a thicker or denser solute layer and flux rates are again governed by equation 5. It should also be remembered that equation 5 assumes a film model of mass transfer; other models would involve  $C_G$  and  $C_B$  in ways other than a simple logarithm of the ratio.

A number of qualitative relationships correlating the mass-transfer coefficient to physical properties and operating parameters have been developed. For all thin-channel configurations of interest, the basic equation is (Porter 1972; Goldsmith 1971):

$$\text{Sh} = A (\text{Re})^\alpha (\text{Sc})^\beta \quad (6)$$

$$\begin{aligned} \text{where } Sh &= \text{Sherwood Number} = \frac{K d_h}{D} \\ Re &= \text{Reynolds Number} = \frac{d_h V \rho}{\mu} \\ Sc &= \text{Schmidt Number} = \frac{\mu}{\rho D} \\ d_h &= \text{equivalent hydraulic diameter} \\ V &= \text{average velocity of fluid} \\ \mu &= \text{viscosity of fluid} \\ \rho &= \text{density of fluid} \end{aligned}$$

Values of  $\alpha$  and  $\beta$  will vary with system geometry and operating conditions.  $\beta$  is invariably equal to 0.33, while  $\alpha$  depends on flow conditions: 0.8–1.0 for turbulent flow and 0.33–0.50 for laminar flow (Porter 1972; Colton and Smith 1972; Henry 1971; Gröber *et al.* 1961; Kozinski and Lightfoot 1972; Goldsmith 1971).

This paper is restricted to a study of the totally polarized, mass transfer controlled region since the high pressure asymptote is the rate limiting region where prediction of flux is required. The low pressure region can readily be predicted by pure water permeability data and Poiseuille flow considerations, since membrane resistance is the controlling factor (Cheryan and Schlessler 1977; Kozinski and Lightfoot 1972). In the totally polarized region, however, flux behavior will be controlled by mass transfer characteristics of the system. Only laminar flow models have been considered since  $Re$  for hollow-fiber systems are typically 500–3000.

## MATERIALS AND METHODS

Ultrafiltration experiments were performed using a pilot-scale hollow fiber unit, Model HF1SS (Romicon, Inc., Woodburn, Mass.) The membrane was a noncellulosic anisotropic type, XM-50, with a nominal molecular weight cut-off of 50,000. The membrane was in the form of 660 hollow-fibers, 0.114 cm i.d. and 63.5 cm length. Total membrane surface area was 1.39 sq. m. Valves and pressure gauges at the inlet and outlet of the fiber cartridge enabled transmembrane pressure to be varied independent of pressure drop. Permeate flux was measured by a Gilmont flowmeter, Model 3205 (Cole-Palmer, Chicago, Illinois). Temperature was maintained to  $\pm 1^\circ\text{C}$  and both retentate and permeate were recycled to the steam jacketed feed tank. Feed volume ( $\sim 100$  l)

and flow rates (25–90 l/min) were much larger than the fiber cartridge hold-up ( $\sim 2$  l) and permeate flux, (0.3–2 l/min) that differential operation is assumed (Goldsmith 1971).

Proteins and other soy components were extracted from soybeans by soaking, blanching, grinding and filtration. The protein and total solids concentration could be adjusted by varying the bean:water ratio. Details of the extraction process are available elsewhere (Omosaiye *et al.* 1978; Cheryan and Schlessler 1977). Feed for each data set described in this paper was taken from the same lot of extracts. Bacterial growth was controlled by adding 100 ppm of thimerosal (Sigma Chemicals) to the feed. Total solids and nitrogen (N) contents were measured by standard methods (A.O.A.C. 1970) and protein expressed as Kjeldahl N  $\times 6.25$  after correcting for nonprotein-nitrogen. Analysis of retentate and permeate during runs revealed that of the 291 mg nitrogen per 100 ml feed, 10–20 mg nitrogen per 100 ml appeared in the permeate. However, typically about half the permeate nitrogen was soluble in 12% trichloroacetic acid indicating mostly nonprotein nitrogenous compounds, which may form 5–10% of the original soybean, was permeating through the membrane. Hence protein rejection was essentially greater than 97% (Cheryan and Schlessler 1977).

Viscosity was measured by Hoesppler falling-ball apparatus; the manufacturer's calibration of the balls was checked and standardized using deionized water and sucrose solutions of known viscosity.

## RESULTS AND DISCUSSION

### Physical Properties

Three physical properties are required for application of mass transport equations: density, viscosity and diffusivity. Density of soy protein extracts shown in Table 1 were obtained using pycnometers that had previously been calibrated at various temperatures using distilled water. Viscosity data at various concentrations and temperatures are shown in Fig. 2. To determine the possible presence of non-Newtonian behavior in the soy protein extracts, three different balls were used in the Hoesppler apparatus, with ball constants varying by a factor of 10 each, in effect resulting in shear rates varying by the same amount. No significant differences were found in viscosities at low concentrations, but consistent differences were noticeable at higher concentrations using different balls. Nonetheless, the data could be reasonably accurately expressed by the following equation

$$\mu = \mu' \exp(\kappa C^2) \quad (7)$$

where  $C$  = total solids concentration, and  $\mu'$  is the intercept and is the viscosity at  $C = 0$ . The value of  $\kappa$  was  $12.65 \pm 0.05 \times 10^{-3}$ . Although methods used here may not be sensitive enough to pick up non-Newtonian behavior in all concentration ranges, viscosity obtained by the Hoeppler falling ball method is at least a first order approximation of the true viscosity, even under the high shear rates common in hollow fiber systems. In addition, mass transfer effects reported in this paper were studied at a fixed low concentration at different temperatures, and hence the viscosity data should be applicable with no further correction. In any case, as will be seen later, errors in viscosity data will have a minor effect on the accuracy of flux prediction using laminar flow mass transfer models.

Table 1. Physical properties of soybean protein extracts<sup>1</sup>

Temp. (° K)	Density (gm/cc)	Viscosity (Cp)	Diffusivity (cm <sup>2</sup> /sec)	Schmidt Number
293	1.0050	1.412	$2.565 \times 10^{-7}$	$5.48 \times 10^4$
303	1.0029	1.098	$3.485 \times 10^{-7}$	$3.14 \times 10^4$
323	0.9973	0.813	$5.018 \times 10^{-7}$	$1.62 \times 10^4$

<sup>1</sup>Total solids = 3.6%, Protein = 1.65%, Fat = 0.94%, Soluble carbohydrate = 0.62%,  
ph = 6.7.

Diffusivity data for soy protein systems are scarce in the literature. Diffusion coefficients of solutes and water in liquid foods are of the order of  $10^{-7}$  to  $10^{-5}$  cm<sup>2</sup>/sec, tending to be lower at higher solute concentration or higher viscosities (Merson and Ginnette 1972). Very little exact data is available for food systems in general. Koshiyama (1968) has reported a value of  $3.85 \times 10^{-7}$  cm<sup>2</sup>/sec at 20°C for the 7S component of the soy protein system. This data was used as a reference point for calculating diffusivities at other temperatures using the Wilke modification of the Stokes-Einstein equation (Bretznajder 1971).

$$D = \frac{D_o \mu_o T}{T_o \mu} \quad (8)$$

Diffusivity data in Table 1 appear reasonable and check well with other reported  $D_{20,w}$  values of  $2.91 \times 10^{-7}$  cm<sup>2</sup>/sec (Wolf and Briggs 1959) and  $3.30 \times 10^{-7}$  cm<sup>2</sup>/sec (Koshiyama and Fukushima 1976) for the 11S soy protein. Diffusivity increases at an average rate of 3.2%/°C, which compares well with typical values of 3–3.4%/°C for globular proteins (Keller *et al.* 1971).

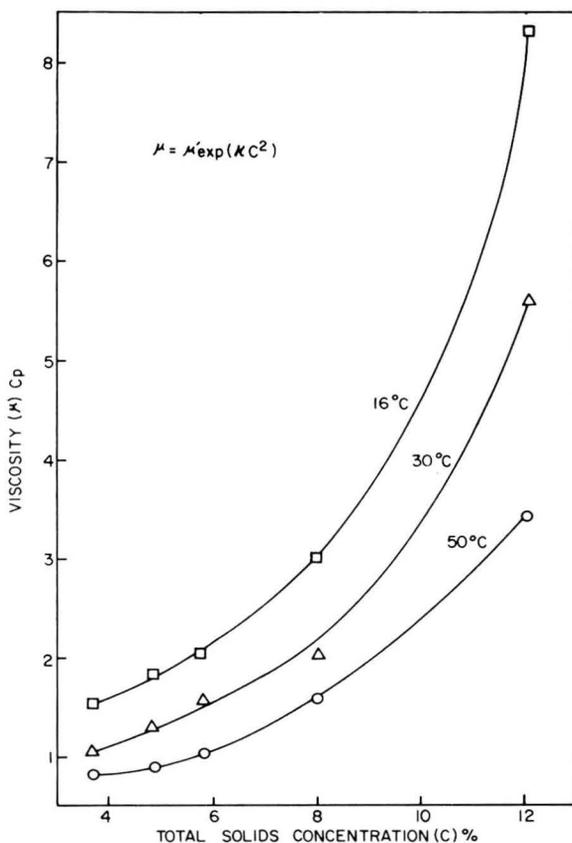


FIG. 2. VISCOSITY OF WATER EXTRACTS OF SOYBEANS

Initial concentration = 3.6%. Other concentrations obtained by ultrafiltration.

### Pressure-Flux Effects

Typical flux behavior obtained with this system is summarized in Fig. 3. The data points represent averages taken after the system had reached a steady state<sup>1</sup>. An indication of the extent of concentration polarization can be seen from: (a) the deviation of the curve from the pure water line, (b) the relatively rapid attainment of pressure-

<sup>1</sup> Actually a pseudo-steady state since flux does decline slowly over a longer period of days or weeks.

independent flux, (c) the marked hysteresis effect observed on lowering the pressure from the totally polarized region. Similar results, although of different magnitude, were obtained at other temperatures, flow rates and feed concentrations investigated (Cheryan and Schlessler 1977). Pressures at which curves deviate from equation 2 are determined by Reynolds and Schmidt numbers. For this system, deviation occurred at lower pressures with lower Reynolds numbers and higher Schmidt numbers, in general agreement with the observations of Kozinski and Lightfoot (1972) and Blatt *et al.* (1970).<sup>2</sup>

As mentioned earlier, it has been suggested that reduction in flux is due to a possible reduction in driving force caused by an increase in local osmotic pressure. The alternate hypothesis is presumed to dominate here, that the deviation is due to a significant hydrodynamic resistance in the boundary layer. It is difficult to differentiate between the two mechanisms. Goldsmith (1971) has pointed out that in both cases, flux is found to increase proportionally to  $K$  and decrease in proportion to  $\ln C_B$ , as defined in equation 5. However, although  $C_G$  varies very little with changes in operating parameters, its absolute value, obtained from graphs such as Fig. 4, will vary depending on the dominant mechanism. In general, highly structured macromolecules such as proteins will display gel-like behavior at concentrations of 10%–30% by weight (Blatt *et al.* 1970), and the  $C_G$  value obtained here is in good agreement with known gelation characteristics of soybean proteins (Smith and Circle 1972). This is in contrast to nongelling systems where  $J$  vs  $\ln C_B$  plots extrapolate to zero flux at concentrations below that of known gel formation, which can be explained reasonably well by an osmotic pressure difference across the concentration polarization layer exceeding the hydrostatic driving force. Second, the considerable amount of hysteresis observed when pressures were varied (Fig. 3) indicates that interactions had occurred between solute molecules, suggesting they were no longer completely mobile in solution, but probably approaching the “close-packed” configuration in the polarization layer as suggested by Michaels *et al.* (1971). It must be pointed out, however, that the concentration polarization layer formed was quite dynamic. At low concentrations and low transmembrane pressures, the flux could be almost completely restored, albeit rather

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<sup>2</sup> A reviewer has pointed out that equation 6 actually predicts higher values of  $K$  for higher Schmidt Numbers, i.e., deviation at lower pressures for lower values of  $K$ . This dichotomy could arise because  $D$  in  $Sh$  becomes less when conditions change so as to increase  $Sc$ , and that the change in  $D$  more than offsets the positive effect of higher  $Sc$  on  $Sh$ .

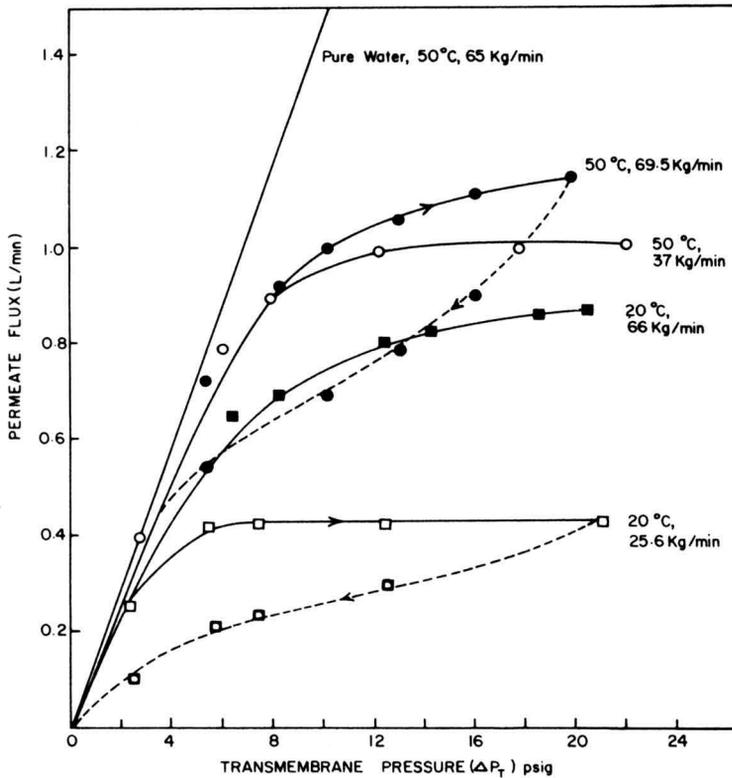


FIG. 3. EFFECT OF TRANSMEMBRANE PRESSURE, TEMPERATURE AND FLOW RATE ON PERMEATE FLUX

Arrows show direction of applied pressure.

slowly, to the original value when flow rates were changed from high to low values and vice versa.

**Mass Transfer Coefficient**

A previous study (Cheryan and Schlessler 1977) had shown that in the pressure-controlled region, activation energy for both flux and viscosity in the 20°C–50°C range were similar (3400 cal/mole), indicating that the model shown in equation 2 is a good expression for predicting flux behavior in that region. As expected, it was inadequate in the mass-transfer controlled region, and the slopes of the flux-temperature lines were found to vary with flow rates or fluid velocities. Dependency of flux with flow rate is more commonly expressed in terms of the Reynolds Number. Fig. 5 is a plot of log J vs log Re, and it shows that

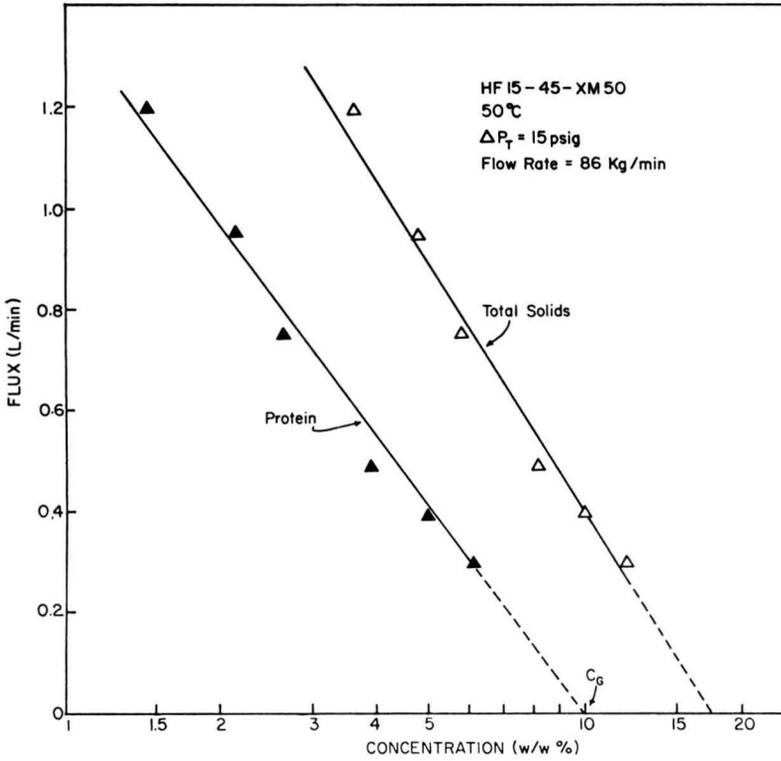


FIG. 4. EFFECT OF TOTAL SOLIDS AND PROTEIN CONCENTRATION ON FLUX

flux is proportional directly to  $Re^\alpha$ . However, the value of  $\alpha$  depended on the temperature, decreasing from 0.6 at 20°C to 0.32 at 50°C. Spriggs and Li (1976) present similar data, a decrease in the exponent of the turbulence factor (stirrer speed) with increase in temperature. Although these values of  $\alpha$  are in approximate agreement with a number of other studies with laminar flow systems, it is clear that even the combined effects of viscosity and flow rates, as expressed by the Reynolds number, cannot adequately account for changes in flux with temperature. Hensley *et al.* (1977) arrived at a similar conclusion with cottonseed protein extracts using different ultrafiltration equipment in the turbulent flow region.

The Arrhenius type model and the flux- $Re^\alpha$  model mentioned above fail to account for the fact that flux will also depend on other functions such as the Schmidt Number, which itself is a function of temperature (Pace *et al.* 1976). To evaluate these factors, values of a mass transfer coefficient  $K$  as defined in equation 5 were obtained at various flow

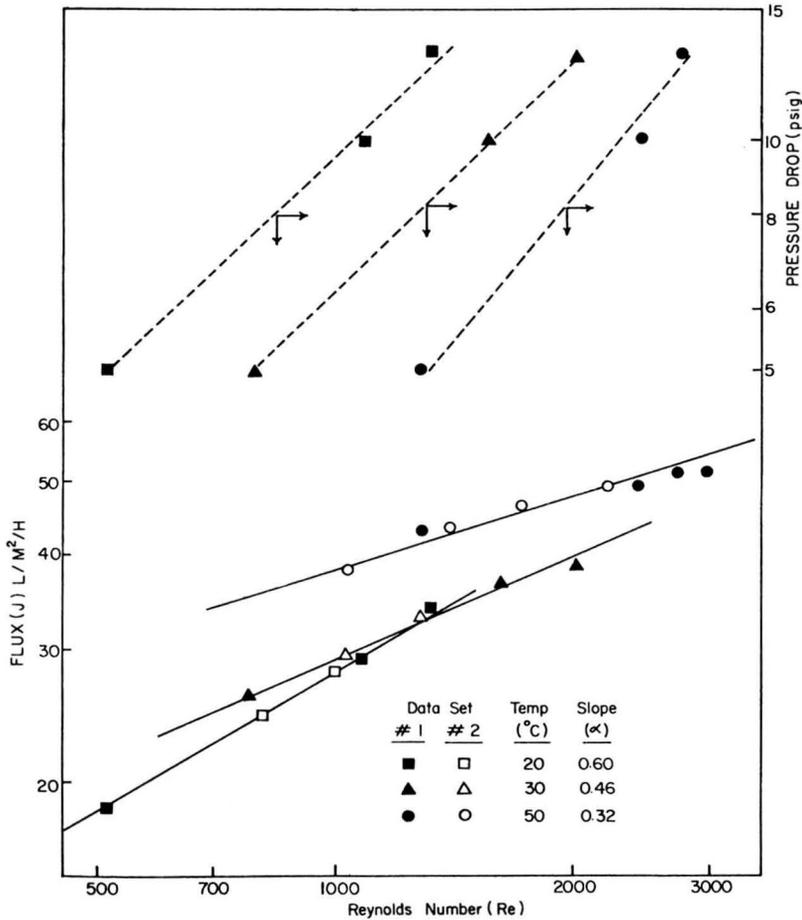


FIG. 5. RELATIONSHIP BETWEEN REYNOLDS NUMBER AND FLUX (FULL LINES) AND REYNOLDS NUMBER AND PRESSURE DROP (BROKEN LINES)

conditions from a series of plots similar to Fig. 4. Dimensionless quantities as expressed in equation 6 were calculated and plotted as shown in Fig. 6. The best fit to the data was obtained with a line giving a slope ( $\alpha$ ) of 0.47 with an intercept (A) of 0.181. This compares with values in the literature where  $\alpha$  ranges from 0.33 (Porter 1972; Henry 1971) to 0.50 (Goldsmith 1971; Colton and Smith 1972; Gröber *et al.* 1961), and values of A range from 0.2 to 0.33 when the authors' equations were applied to our system.

In laminar flow systems, the magnitude of  $\alpha$  and  $\beta$  depend primarily on the state of development of the velocity and concentration bound-

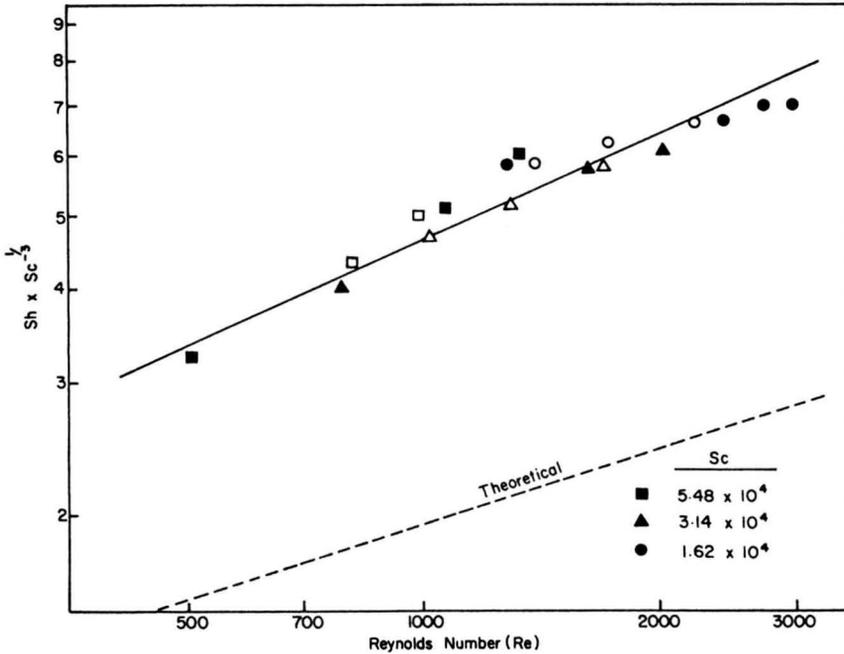


FIG. 6. MASS TRANSFER CORRELATION BETWEEN REYNOLDS NUMBER AND SHERWOOD AND SCHMIDT NUMBERS

Broken line indicates theoretical correlation expected using the Graetz or Levêque solutions.

ary layer profiles down the length of the channel. If both are fully developed,  $\alpha$  and  $\beta$  are zero (Blatt *et al.* 1970). If velocity profile is fully developed, but concentration boundary layer is developing down the entire length of the hollow fiber the Graetz or Levêque solutions can be used with  $\alpha = 1/3$  and  $\beta = 1/3$  (deFelippi and Goldsmith 1970). If both velocity and concentration profiles are developing,  $\alpha = 1/2$  and  $\beta = 1/3$  (Gröber *et al.* 1961; Porter 1972). During hollow-fiber ultrafiltration, a significant fraction of the solvent may be removed before the concentration gradient is fully developed (Lonsdale 1972). The length of the concentration profile entrance region ( $L$ ) can be calculated from equation 9 (Blatt *et al.* 1970).

$$L = \frac{0.1 \dot{\gamma}_w d_h^3}{D} \quad (9)$$

Shear rates ( $\dot{\gamma}_w$ ) for our system are of the order of 4000–14,000  $\text{sec}^{-1}$ , and D values about  $10^{-6}$   $\text{cm}^2/\text{sec}$ . This results in values of L much greater than the channel lengths used, indicating our use of the basic form of equation 6 with  $\beta = 1/3$  is valid.

The state of development of the velocity profile is not as clear, and this is what essentially determines  $\alpha$ . The channel length  $L^*$  required to attain fully developed laminar flow is given by

$$\frac{L^*}{d_h} = B \cdot \text{Re.} \quad (10)$$

where B varies from 0.029 (Gröber *et al.* 1961, Porter 1972) to 0.05 (McCabe and Smith 1956). In this study,  $L^*$  was found to range from a low of 1.66 cm to a high of 17.2 cm. This represents from 2% to 27% of the channel length over which the velocity profile is developing. This is probably not significant enough to justify usage of models where the velocity profile is assumed to be developing, although an Re power dependence of 0.47 was obtained here, close to the value of 0.50 suggested by Gröber *et al.* (1961). Pace *et al.* (1976) also obtained  $\alpha = 0.47$  and  $A = 0.15$  for the bovine serum albumin-sucrose system ( $6000 < \text{Re} < 20,000$ ). Porter (1972) and Blatt *et al.* (1970) have presented a number of cases where experimental slopes were higher than the 1/3 value predicted from theory.

The value of A will generally reflect physical property variations and conditions of the system. Kozinski and Lightfoot (1972) have presented an analysis of boundary layer filtration and expressed A in terms of parameters to correct for variations in diffusivity, viscosity and density at the wall and at bulk conditions. In addition, unlike in turbulent flow, laminar flow models where the concentration profile is developing down the entire length of the channel require a mass transfer dependency on channel length L. Equation 6 is frequently written such that

$$A = A' \left( \frac{d_h}{L} \right)^\alpha \quad (11)$$

with the value of  $A'$  ranging from 0.664 (Gröber *et al.* 1961) to 1.86 (Henry 1971).

The reasons for the discrepancy between experimental observations and theoretical expressions such as the Graetz or Levêque solutions as shown in Fig. 6 are not clear. Others (e.g., Porter 1972; Goldsmith 1971; Kozinski and Lightfoot 1972; Blatt *et al.* 1970) have also ob-

served that the theoretical models predict flux values much lower than those experimentally obtained. This implies that back diffusion from the membrane surface to the bulk stream is greater than expected, and is controlled by forces other than or in addition to concentration gradient, possibly by the tubular pinch effect discussed by Porter (1972). This could also merely be a reflection of the quality of the physical property data or the complexity of the system under study. The soy protein extract used as feed is a polyphasic system consisting of an emulsion (fat), suspension (proteins and insoluble carbohydrate), and solution (proteins, oligosaccharides, salts). The proteins themselves are a complex mixture ranging in molecular weights from 7000–600,000 (Smith and Circle 1972). Hence each component will contribute to a particular physical property in a different way and the data presented in this paper is a resultant of these effects. The greatest uncertainty is probably in diffusivity data, which was obtained indirectly by calculation. Viscosity variations and errors will have a small effect on flux prediction for reasons mentioned below.

#### Process Design of Ultrafiltration Systems

The real value of mass-transfer correlations such as those described above is in being able to predict flux, and hence to design an ultrafiltration process, for a specific application by choosing appropriate independent variables such as fiber diameter, length, temperature and flow rate to give as high a production rate as possible. Ideally, the first two should be such that entrance-region conditions are obtained during processing. In reality, they are practically fixed by availability of commercial equipment. Temperature should be as high as possible, but can be governed by membrane stability, protein denaturation and microbial growth. There may be a limit to flow rates also. Other factors being constant, an increase in flow rate in laminar flow systems does not have a very significant effect on flux due to the low value of  $\alpha$  in the flux- $Re^\alpha$  relationship. On the other hand, pressure drop is directly proportional to  $Re$  (McCabe and Smith 1956) as illustrated in Fig. 5, and pumping power varies as  $Re^2$ . In general of course, pumping requirements for such ultrafiltration systems are very low, typically of the order of 0.1–1.0 HP-HR/1000 gallons (deFelippi and Goldsmith 1970). Benefits of increasing flow rates on flux are more readily apparent in turbulent flow systems, since the Reynolds Number exponent is much higher.

The choice between laminar and turbulent flow systems will also depend to a large extent on the viscosity of the feed. All other factors being constant, equation 6 indicates that  $K$  or flux is proportional to  $\mu^m$ , where  $m$  can range from 0 to  $-1/6$  for laminar flow systems and  $-1/2$  to  $-2/3$  for turbulent flow systems. Since viscosity has a large effect on power consumption, flux:pumping power ratio would tend to favor the laminar flow approach for achieving high fluxes while processing concentrated protein solutions, such as is embodied in the hollow fiber configuration.

### CONCLUSIONS

An analysis of flux behavior during ultrafiltration of soybean water extracts shows that the effects of concentration polarization can be described using classic chemical engineering mass transfer models. These models can be used to predict performance characteristics of hollow fiber systems in the totally polarized, pressure-independent region where laminar flow conditions prevail. The precise form of the laminar flow model will depend on the state of development of the velocity and boundary concentration profiles. For highly viscous systems, it is preferred that entrance-region conditions be maintained throughout processing, thus taking advantage of the laminar flow-high shear approach to controlling concentration polarization. The exponent of the flux-Reynolds Number relationship varied with temperature; a better correlation could be obtained if diffusivity was taken into account by including the Schmidt Number in mass transfer expressions. However flux predicted by theoretical considerations is somewhat lower than experimental values. Theoretical models may be adequate for preliminary design purposes, but until further studies are made into causes for the discrepancy and solutions provided, mass transfer models will probably have to be supplemented by experimental data to enable proper design of ultrafiltration processes in complex food systems. The correlation that best fit the hollow fiber-soybean water extract system was

$$Sh = 0.181 (Re)^{.47} (Sc)^{.33} \quad (12)$$

### ACKNOWLEDGMENTS

The author thanks Mr. Joseph E. Schlessor for help in obtaining some

of the experimental data. Financial support was provided in part by the Agriculture Experiment Station, University of Illinois, Project No. 30-15-50-347.

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## LITERATURE ABSTRACTS

### ABSTRACTS FROM THE AICHE JOURNAL

Each of the following abstracts has been reprinted with permission from the *AICHE Journal*.

**A SOLID/LIQUID SEPARATION PROCESS BASED ON CROSS FLOW AND ELECTROFILTRATION.** J. D. Henry, Jr., Lee F. Lawler and C. H. Alex Kuo, *AICHE Journal* 23 (6), 851.

The cross flow/electrofiltration process combines migration of particles in the presence of electrical and shear fields to increase filtration rates. A mathematical model and the experimental data are presented for both a kaolin clay suspension and an oil in water chemically stabilized emulsion. When the filter is operated in a regime above the critical voltage, filtration rate-electric field strength dependence becomes linear, and both electroosmosis in the filtration medium and electrophoresis in the liquid film are the controlling mechanisms of transport. Both experimental data and the mathematical model indicate that the fluid circulation rate tangential to the filtration media does not necessarily increase filtration rate depending on the regime of operation. The model is used in conjunction with the experimental data to separate the liquid film resistance adjacent to the filter cake which is determined by fluid shear and electric field from the cake and filter medium resistances which are influenced by electroosmosis.

**MEAN TEMPERATURE DIFFERENCE: A REAPPRAISAL.** K. Gardner and J. Taborek, *AICHE Journal* 23 (6), 777 (1977).

The derivation of the mean temperature difference in heat exchangers is based on a number of assumptions or idealizations, the most important ones being that the heat transfer coefficient is constant throughout the exchanger; that the temperature of either fluid is constant over any cross section of its nominal path, that is, complete mixing, no stratification or bypassing; and that an additional assumption for shell and tube exchangers, which has not always been fully recognized, is that within one baffle crossing the shell fluid temperature change is small with respect to its overall change, (that is, number of baffles is large). In actual exchangers, any of the above assumptions are frequently subject to various degrees of invalidation. This paper examines the effects of deviating from the first two assumptions and presents a new solution to the third.

**DISPERSION IN LAYERED POROUS MEDIA.** M. B. Moranville, D. P. Kessler and R. A. Greenkorn, *AICHE Journal* 23 (6), 786 (1977).

Experimental data on directional dispersion coefficients in transversely isotropic media, a special case of anisotropic media, were determined in models consisting of

alternating layers of sintered glass and unconsolidated glass beads. Values of appropriate parameters to relate the velocity to the directional dispersion coefficient were determined for a variety of permeabilities and angles. Sensitivity of these parameters with respect to angle was determined, and results are presented which will permit estimate of when this effect is significant.

**SYNTHESIS OF HEAT EXCHANGE NETWORKS BY MIXED INTEGER OPTIMIZATION.** R. C. Kelahan and J. L. Gaddy, *AICHE Journal* 23 (6), 816 (1977).

Process synthesis involves manipulation of the process arrangement, while studying the variables of each arrangement, to arrive at the optimal process. If each process arrangement is treated as a discrete variable, process synthesis becomes a mixed integer optimization problem. This paper examines the synthesis of heat exchanger networks using the adaptive random search procedure, which can be used to search continuous and discrete independent variables simultaneously. The means of handling the heat exchanger arrangement as a discrete variable is discussed, and the incorporation of various synthesis heuristics is presented. The results of synthesis of  $2 \times 2$ ,  $2 \times 3$ , and  $3 \times 3$  networks are presented and compared with other methods of synthesis.

**THE EFFECT OF MASS TRANSFER ON APPARENT ASCORBIC ACID AUTO-OXIDATION KINETICS.** R. S. Davidson and R. A. Grieger-Block, *AICHE Journal* 23 (4), 529 (1977).

Previous studies of copper-ion catalyzed ascorbic acid autoxidation have indicated zero-order, first-order, or Michaelis-Menten dependence of rate on ascorbic acid or copper concentrations and first- or half-order dependence on oxygen concentration. A simple mathematical model of gas-liquid mass transfer with liquid phase reaction accounts for the various behaviors. Use of an experimental technique that eliminates the mass transfer step shows that the oxidation rate dependence is first order in copper, half order in oxygen, and Michaelis-Menten in ascorbic acid concentrations.

**STUDIES IN CHEMICAL PROCESS DESIGN AND SYNTHESIS.** N. Nishida, Y. A. Liu and L. Lapidus, *AICHE Journal* 23 (1), 77 (1977).

An algorithmic-evolutionary approach to the systematic synthesis of minimum cost networks of exchangers, heaters, and coolers is proposed. The new approach is easy to apply by hand calculations, requiring no special mathematical background and computational skill from the user. In addition to generating many cheaper networks for sample problems ranging in size from 4 to 10 streams as compared to previous studies, the proposed method provides an explicit theoretical guidance on the optimal exchange among hot and cold streams and on the optimal locations of heating and cooling utilities in the network. It also has a provision for the use of stream splitting and for generating a cyclic network. The new method is particularly effective in the synthesis of minimum cost networks for industrial crude unit preheat recovery.

**DIFFUSIONAL BOUNDARY-LAYER RESISTANCE FOR MEMBRANES WITH LOW POROSITY.** D. M. Malone and J. L. Anderson, *AICHE Journal* 23 (3), 177 (1977).

Diffusional mass transfer across membranes with uniform but low pore densities was studied experimentally as a function of stirring rate and pore area fraction. The results were analyzed in terms of a stagnant film boundary-layer model specially formulated for a heterogeneous membrane containing discrete pores. The overall membrane diffusional resistance is linearly related to the inverse of the pore area fraction of the membranes for a constant stirring rate. An equivalent boundary-layer thickness can be defined which is independent of membrane properties but a unique function of stirrer speed. These experimental boundary-layer thicknesses are greater by a factor of 3 than those predicted by a published Sherwood number correlation determined for homogeneous surfaces, but the stirring rate dependence is in excellent agreement with this same correlation.

**FLOW OF SLURRIES IN PIPELINES.** R. M. Turian and T. Yuan. *AICHE Journal* 23 (3), 232 (1977).

Pressure drop correlations for flow of slurries in pipelines were developed for each of the following four flow regimes; flow with a stationary bed, saltation flow, heterogeneous flow, and homogeneous flow. A total number of 2,848 data points, comprised of 1,912 collected from the published literature together with 936 taken using our own test pipelines and relating to ranges of the pertinent variables extensive enough to span all four flow regimes were used as the basis of these correlations. Also, these data were used in developing an associated quantitative regime delineation scheme. The correlations provide an improved predictive capability over previously available procedures and are also broader in scope. The delineation procedure developed here permits straightforward classification of the data according to the flow regime prevailing, and it is moreover inclusive of all the data and is self-consistent.

**GROWTH RATE OF AN ICE CRYSTAL IN SUBCOOLED PURE WATER.** J. P. Kallungal and A. J. Barduhn, *AICHE Journal* 23 (3), 294 (1977).

New and extensive experimental data on the rate of growth of ice crystals in the  $a$  axis direction in quiescent and slow flowing subcooled pure water show conclusively that thermal natural convection is an important heat transfer mechanism controlling the growth rate. At zero and low forced velocities, steady growth is observed only when the crystals grow horizontally or upward. Steady downward growth does not occur in quiescent water. This is consistent with the physical properties of water and the phenomenon of thermal natural convection.

Growth rates at high water flow rates vary as the square root of the forced velocity and the  $3/2$  power of the subcooling and follow the theory of Fernandez and Barduhn (1967) with the ice-water interfacial energy set at  $52 \text{ mJ/m}^2$  ( $52 \text{ erg/cm}^2$ ).

**PHASE DISTRIBUTION AND RESIDENCE TIME IN A SPOUTED BED.** G. C. Suci and M. H. Patrascu, *AICHE Journal* 23 (3), 394 (1977).

Particle movement in the core and in the annulus of a semicircular, vertical, spouted

bed was examined by high-speed tracer photography. Use of a functional analogy with a free jet establishes an equation correlating core air velocity and relative bed height. Dependence of bed void fraction and particle throughput vs. relative bed height shows good agreement with previous work. Particle mean retention time and air flow distribution in the annulus and the spout were obtained by calculational estimates.

A COMPREHENSIVE CORRELATING EQUATION FOR LAMINAR, ASSISTING, FORCED AND FREE CONVECTION. S. W. Churchill, *AICHE Journal* 23 (1), 10 (1977).

A correlating equation for assisting convection was developed by combining correlating equations for pure free and pure forced convection. These component equations are based on laminar boundary-layer theory for an isothermal, vertical plate. Theoretical values for assisting convection indicate that the third root of the sum of the third powers gives the best representation, as contrasted with the choice and rationalization of the second or fourth power by prior investigators.

This expression was modified by the addition of a limiting value  $Nu_0$  to obtain a better representation below the domain of boundary-layer theory and was generalized for uniform heating and for spheres and horizontal cylinders by the appropriate choice of the characteristic length.

TURBULENT MASS TRANSFER RATES TO A WALL FOR LARGE SCHMIDT NUMBERS. D. A. Shaw and T. J. Hanratty, *AICHE Journal* 23 (1), 28 (1977).

New measurements are presented on the influence of Schmidt number on the rate of mass transfer between a turbulent fluid and a pipe wall. It is found that for large Schmidt numbers the fully developed mass transfer coefficient is related to the friction velocity and the Schmidt number by the equation

$$\kappa_\gamma = 0.0889 \nu^0 \text{Sc}^{-0.704}$$

The experiments are accurate enough to rule out the  $\text{Sc}^{-2/3}$  or the  $\text{Sc}^{-3/4}$  relations commonly used, deduced from plausible limiting expressions for the eddy diffusivity close to a wall. It is argued that these expressions are valid only over a vanishingly small portion of the concentration field as  $\text{Sc} \rightarrow \infty$

DEGREES OF ANISOTROPY FOR FLUID FLOW AND DIFFUSION (ELECTRICAL CONDUCTION) THROUGH ANISOTROPIC POROUS MEDIA. G. Neale *AICHE Journal* 23 (1), 56 (1977).

This work involves a quantitative comparison between diffusion (or electrical conduction) and fluid flow occurring in the principal directions of certain simple anisotropic porous media, namely, clusters of parallel circular (or elliptic) cylindrical fibers. The degrees to which the observed macroscopic properties, permeability, and effective diffusivity (electrical conductivity) depend upon the direction of flow are determined analytically. Systems of parallel fibers are shown to be considerably more anisotropic with respect to fluid flow than with respect to diffusion (or electrical conduction), this effect being very pronounced with flattish fibers. It

would not appear that any general analogy exists between diffusion (electrical conduction) and fluid flow occurring within anisotropic porous media.

#### ABSTRACTS FROM TRANSACTIONS OF THE ASAE

Each of the following abstracts has been reprinted with permission from the *Trans. ASAE 19FE*.

**COOLING SHORT CYLINDERS IN AIR USING A WATER SPRAY.** L. D. Albright, *Trans. ASAE 19FE*, 762-770.

An experimental evaluation of forced air and water spray as a cooling means is described. Air velocities of 24.3 m/sec and spray rates of  $2.2 \times 10^{-5}$  m<sup>3</sup>/sec per square meter of sprayed area cool almost as rapidly as agitated water baths. Based on the results of the study, cooling processed cans with forced air and water spray is recommended to reduce water use in the food processing industry.

**HEAT TRANSFER DURING AIRCOOLING AND STORING OF MOIST FOOD PRODUCTS — II. SPHERICAL AND CYLINDRICAL SHAPES.** S. S. Murthy, M. V. Krishna Murthy and A. Ramachandran, *Trans. ASAE 19FE*, 577-583.

Heat transfer characteristics during aircooling of moist food products are analyzed taking into account the effects of the evaporation of the moisture film that usually exists on the product surface. The one-dimensional transient heat conduction equation in spherical and cylindrical coordinates is solved.

Charts useful in predicting the time-temperature histories and cooling loads are presented for the following values of dimensionless parameters:

Biot number,  $Bi$ : 0.1 to 10,  
Initial product temperature,  $t_i^+$ :  
2.5 ( $\approx 21C$ ) to  
4.0 ( $\approx 43C$ ) in increments of 0.5,  
Cooling air temperature,  $t_o^+$ : 1.25  
( $\approx 2C$ ) to 2.25 ( $\approx 15C$ ) in increments of 0.5.  
Relative humidity,  $\phi$ : 0.0 to 1.0 in increments of 0.2.

Relative humidity of the cooling air is an important parameter which decides the cooling rates. Surface evaporation effects are significant, particularly at low air humidities. The present analyses yield much better agreement with the experimental results compared to existing conventional methods.

**APPLICATION OF VIBRATIONAL SORTING TO BLUEBERRY FIRMNESS SEPARATION.** D. R. Bower and R. P. Rohrbach, *Trans. ASAE 19FE*, 185-191.

The increased use of mechanical harvesters in the blueberry industry has made it necessary to sort the fruit prior to sending it to the commercial fresh market if quality and shelf life are to be maintained. This paper explores an application of vibrational sorting to the separation of blueberries according to firmness.

The development of a vibrational sorter using an electromagnetically excited vibrating plexiglass plate is considered. Tests were run using the sorter to determine the

ability of the sorter to separate blueberries according to firmness and shelf life. The accuracy of sorting was determined through the use of quasi-static compression tests and shelf life storage tests of the sorted blueberries.

A theoretical consideration of the blueberry as a Kelvin model was developed and the equations of motion of the system determined. A dynamic testing procedure was also utilized to relate the spring constant and damping coefficient. The coefficient of restitution was then utilized to relate the model to the sorting phenomenon occurring with the vibrational sorter.

In addition to evaluations of the vibrational sorter for firmness separation, the conclusion presents a new dynamic test procedure to determine the actual values of the spring constant and damping coefficient for a Kelvin modeled blueberry under dynamic conditions more nearly like those found in a vibrational sorter.

**PERFORMANCE OF A ROTATING PERFORATED SLEEVE SPRAYER.** R. Bajracharya, J. R. Rosenau and P. M. Bluestein, *Trans. ASAE 19FE*, 754-757.

Milk droplets produced by a rotating perforated sleeve sprayer were dried in a laboratory spray drier. The examinations of the dried particle-size distributions revealed a high degree of uniformity and Sauter mean diameters in the range desirable for the practical purpose of spray drying of milk. The statistical analysis did not reveal a significant effect of the speed of rotation on the mean diameters; the Sauter mean diameter varied with feed rate to the  $-0.18$  power.

**LYSINE RETENTION IN FOOD DURING EXTRUSION-LIKE PROCESSING.** D. R. Thompson, J. C. Wolf and G. A. Reineccius, *Trans. ASAE 19FE*, 989-992.

Available lysine levels were measured as a function of process temperature time, product moisture content, pH, protein level, carbohydrate type and level. The results are reported in terms of activation energy (ca 34000 cal/mole) and reference reaction rate at 100° C (ca 0.24 mg/g sample/minute) for an initial zero order loss, and equations which represent the data in latter stages of the loss curve.

**A VACUUM BLENDING METHOD FOR PREPARING TOMATO SAMPLES FOR COLOR DETERMINATION.** D. E. Bilton and M. O'Brien. *Trans. ASAE 19FE*, 386-389.

An improved method for determining processing tomato color was tested against the standard method as used by the California tomato inspection service. The new procedure consists of blending the sample in a vacuum and producing a mascerate free of air bubbles by drawing a vacuum on the blender itself. The new vacuum Agtron color readings are not significantly different from the standard method, as concluded from one hundred paired comparisons. The effect of sample volume and screen mesh size on Agtron color reading was investigated. The new procedure is regarded as safer, more objective and requiring less time than the standard method.

**ENERGY ALLOCATION IN THE FOOD SYSTEM: A MICROSCALE VIEW.** S. J. Brown and J. C. Batty, *Trans. ASAE 19FE*, 758-761.

This study is based on the concept that while on-farm energy conservation efforts

are commendable, an even greater potential for energy conservation exists in other sectors of the food chain. Rather than examining the industry as a whole, the energy inputs to a one pound can of whole kernel corn was estimated at different stages of the food chain. These stages include agricultural production, processing, transport, marketing, shopping and home preparation. Analysis of results indicate that the agricultural sector has become heavily dependent upon fossil fuels and machinery to approximately the same extent as the other segments of the food chain. Thus, even effective on-farm energy conservation measures are not considered to increase the efficiency of the food chain unless they are coupled with significant changes in the energy costs of other sectors.

**SIMULATION OF LIQUID FOOD QUALITY DURING STORAGE.** R. P. Singh and D. R. Heldman, *Trans. ASAE 19FE*, 178-184.

This paper is concerned with the development of a computer-aided prediction model in selecting storage requirements best suited for liquid foods. Specific objectives included the development of a mathematical model of the oxygen diffusion coupled to a second-order chemical reaction and a simulation of the influence of light intensity on rates of quality degradation and oxygen uptake. Vitamin C concentration was selected as the quality index. Experimental and simulation results confirmed that the rate of vitamin degradation in liquid foods during storage is more accelerated in the presence of light than under dark conditions. The major portion of light-induced vitamin degradation in infant formula occurred within a 0-2 cm layer along the container wall. Agreement between simulation and shelf-life studies was within 1.43 to 5.10% standard deviation for initial ascorbic acid concentration and within 2.86 to 7.01% for initial dissolved oxygen concentration.

**COMPUTER SIMULATION OF POTATO PACKINGHOUSE OPERATIONS.** P. H. Orr and K. A. Ebeling, *Trans. ASAE 19FE*, 978-988.

This paper illustrates the use of a macroelemental, special purpose digital simulation language as a means for describing and quantifying the interactions that involve workers, machines, and microelemental units of products during packinghouse operations. The detail of the model required the use of a sequentially phased model rather than a single-phase one. The packinghouse operations were divided into eight sequential phases: (1) receiving; (2) presizing; (3) pregrading; (4) washing/drying; (5) grading; (6) sprout inhibition/transfer; (7) sizing; and (8) packing. The complete simulator can be used to conduct detailed investigations and obtain information about: (a) packinghouse equipment performance under abnormal operating conditions; (b) allocation of raw material and packaging costs under combinations of packinghouse operations; (c) attainable quality levels under various raw product input conditions; and (d) alternative plant operating policies.

**MATHEMATICAL MODELING OF THE POST-HARVEST TEXTURAL BEHAVIOR OF ASPARAGUS.** S. C. Sharma and R. R. Wolfe, *Trans. ASAE 19FE*, 984-987.

The post-harvest toughening of asparagus, which is caused mainly by lignification, was investigated. A mathematical model was developed in order to describe lignin

synthesis during post-harvest handling and storage. The model accounts for the influence of storage temperature, time, humidity, spear length, diameter, axial position and the height of the cut.

Experimentation yielded a prediction model which is in agreement with the theoretical formulation. Under the conditions of this study, humidity was found to have no significant effect and the effect of harvest height was only moderately significant. All other factors were found to have a major influence on toughness. The model was used to generate a graphical representation of the percentage of tough asparagus resulting from various storage conditions. These curves should be useful in defining post-harvest handling procedures conducive to maintaining asparagus quality.

**INFLUENCE OF HANDLING ON PICKLING CUCUMBER QUALITY.** D. R. Heldman, D. E. Marshall, L. R. Borton and L. J. Segerlind. *Trans. ASAE 19FE*, 1194-1200.

This investigation focuses (a) on the identification of handling steps between mechanical harvesting and brining where reduction of cucumber quality occurs; (b) on the estimation of the magnitudes of quality reduction, and (c) on the identification of the reasons for quality reduction due to severe damage.

Field experiments were carried out in order to collect samples at several stations between hand and mechanical harvesting locations and designated processing plants. The cucumbers were graded and the extent of visible damage determined.

Impact and tumbling experiments were performed in order to evaluate the effect of the handling on brine stock quality.

The extent of visible damage to greenstock cucumbers between harvesting and brining was closely related to handling steps where large-magnitude drops occur. It has been observed that carpel strength may increase with handling, although the relationship has not been sufficiently studied. Size-grading systems (vibratory) consistently increased the frequency of balloon bloaters in brine stock cucumbers. The frequency of internal fractures in brine stock was increased for drop distances greater than 0.8 m; nevertheless, repeated low-magnitude drops primarily influenced the frequency of balloon bloaters in brine stock cucumbers.

**ENGINEERING ANALYSIS OF PRAWN LARVA CULTURE.** J. Wang and J. S. Kuwabara, *Trans. ASAE 19FE*, 1197-1200.

A stepwise experimental investigation was designed in order to determine the optimum design criteria for a pilot post-larvae rearing equipment. The exploratory experiment revealed that temperature significantly influences prawn larvae culture. Next, the tolerance of prawn larvae to growth media temperature was investigated. The design criteria developed includes: (a) optimal range of physical variables (1) growth media exchange rate — at least once a day. Higher exchange rate and daily cleaning of the growth chamber highly desirable. (2) Growth media temperature — within the range of  $32.22^{\circ}\text{C} \pm 0.5^{\circ}\text{C}$ . (3) Larvae density —  $47.32-1 \pm 6.6/1$ . (b) Other design requirements include (1) close temperature control to within  $0.6^{\circ}\text{C}$  of desired level; (2) sufficient agitation within growth chamber is needed to maintain uniformity in growth media temperature; (3) automatic feeding and growth chamber cleaning.

**PREDICTION OF FREEZE-DRYING TIME FOR HYDROLYZED VEGETABLE PROTEIN MIXTURES.** R. P. Singh and D. R. Heldman. *Trans. ASAE 19FE*, 570-572.

The freeze-drying characteristics of concentrated liquids containing significant amounts of low molecular weight components has been studied. Experimental and mathematical procedures are discussed, which are used to measure relevant mass transport parameters describing the freeze-drying characteristics of a concentrated liquid product and to determine the influence of freeze rates on the transport parameters.

Moisture content history curves were obtained experimentally from freeze-drying trials. The procedure for developing empirical freeze-drying time correlations involves linearizing the moisture content curve in the temperature range where 85% water removal occurs.

Incorporating this linearization into a uniformly retreating ice front model to describe water vapor transport through the dry layer of the product, the diffusivity and external mass transfer coefficient were calculated. Thus, a freeze-drying time equation was developed to predict total freeze-drying times for products of any thickness under the experimental conditions used. It was confirmed that freezing rate and platen temperature significantly influence the product quality and freeze-drying time.

**COMPOSITE BIOMATERIALS AS STRUCTURAL MATERIALS FOR PACKAGING.** M. A. Hanna and N. N. Mohsenin. *Trans. ASAE 19FE*, 390-395.

This study was designed for the purpose of determining some mechanical properties of a composite biomaterial to be used as a structural packaging material. The materials investigated included mechanically fiberized alfalfa hay, ground corn, soybean meal and ammonium lignin sulfonate (binding agent).

The variables selected included the density of the fabricated sheet, the weight percent of binding agent, weight percent of fibrous material and percent protein of the mixture. The mechanical properties, including the modulus of rupture and apparent modulus of elasticity were determined from static bending tests. Maximum strength and the tensile modulus of elasticity were determined from the tension tests. Analysis of variance and multiple regression were used to analyze the experimental data and to determine the mechanical property relationships.

The mechanical properties of the biomaterial compared very well with the corresponding ones of the corrugated cardboard, except for the maximum tensile strength. The composite biomaterial did not have the same folding characteristics of cardboard.

The mathematical models can be used to predict the mechanical properties of composite biomaterials consisting of alfalfa fiber, binding agent, soybean meal and corn or to determine the set of composition and fabrication variables required to develop certain mechanical properties in a composite biomaterial.

**ENGINEERING ANALYSIS OF A MECHANIZED FRUIT GRADING TABLE.** R. B. Naugle and M. O'Brien, *Trans. ASAE 19FE*, 396-399.

This paper reports on research conducted during the 1973 and 1974 seasons to

increase the objectivity of and reduce the amount of labor in the inspection process by use of a mechanized grading table.

Testing was conducted in order to compare the reliability of the mechanized grading system and the manual one. Results indicate that the mechanized design decreases the time for grading by reducing by half the amount of fruit sample used in the "conventional" grading procedure. It eliminates the need for the O'Brien subsampler, the manual selection of the peelability sample and the problems of hardware associated with it by obtaining the comminuted sample from the half of the 91 kg sample of ungraded fruit. It also improves the efficiency of the manual grading process.

**AN ENGINEERING EVALUATION OF POTATO STORAGE VENTILATION SYSTEM PERFORMANCE.** L. A. Schapter, H. Cloud and D. Lundstrom. *Trans. ASAE 19FE*, 584-590.

An engineering analysis was conducted and field data was collected to determine design criteria of potato storage ventilation systems. A theoretical analysis was performed in order to develop relative distribution patterns for the static pressure and the amount of air discharged along the duct length. The computer simulation developed can be used to determine the effect of duct entrance air velocity and total slot discharge area on the air discharge distribution.

Data from measurements and simulated conditions substantiate storage operators' concerns about poor air distribution and reduced air flow in potato storage ventilation systems. Extremely variable air distribution patterns can exist in potato storage ducts that have large air discharge areas and thus operate at low static pressures. Satisfactory air distribution in a constant cross-sectional area duct, containing uniform slots can be obtained when the ratio of duct discharge area to duct cross-sectional area is about 1.0. This requires that the duct be constructed with a ratio of total slot area to duct cross-sectional area of about 4.

**PROGRAMMED TEMPERATURE CONTROL WITH A CASSETTE TAPE RECORDER.** J. L. Steele and C. N. Burkholder. *Trans. ASAE 19FE*, 573-583.

In order to carry out studies on the effect of various drying air temperature patterns on the ultimate quality of freshly dug peanuts, the need for developing a programmable temperature control equipment compatible with existing research facilities led to the design of a temperature control system which utilizes a conventional cassette tape recorder, a unit to generate or prepare the cassette tape program, and a second unit to decode the tape program upon playback and drive controller set point potentiometers in accordance with the program. The system has been used successfully for two peanut harvest seasons.

**DYNAMIC STRUCTURAL PROPERTIES OF SWEET POTATO.** V. N. Mohan Rao, D. D. Hamann and A. E. Purcell. *Trans. ASAE 19FE*, 771-774.

The present study was concerned with dynamic testing of dry and wet fleshed potatoes to determine the time dependent complex modulus and its variation along the longitudinal axis of the tubes. The relationship between fiber content of roots

and mechanical stiffness, and the use of the complex modulus as a means of detecting differences between cultivars were also investigated. The dynamic testing apparatus designed by Hamann with minor modification was used for sinusoidal uniaxial compression of cylindrical specimens. Results indicate that dynamic testing of sweet potatoes is a means of detecting differences between cultivars. The per cent fiber appears to be correlated to mechanical stiffness as measured by dynamic testing. Although results indicate that an exponential or logarithmic type relationship exists between fiber per cent and absolute modulus, the exact relationship deserves further investigation.

**ANALYSIS OF A SIMPLE PENDULUM IMPACTING DEVICE FOR DETERMINING DYNAMIC STRENGTH OF SELECTED FOOD MATERIALS.** V. K. Jindal and N. N. Mohsenin. *Trans. ASAE 19FE*, 766-770.

A pendulum type impactor offers a means of studying the impact behavior of certain agricultural materials. The measurements obtained with any such experimental setup need to be critically examined and verified in order to obtain some meaningful information from the impact test data. This study shows that the validity of the impact velocity and force relationships based on the physical configuration of a simple pendulum can be checked experimentally. The information recorded during impact tests can be used to determine several dynamic strength parameters of agricultural materials. Examples are presented to illustrate the use of the physical parameters obtained by this method.

#### **ABSTRACTS FROM THE INTERNATIONAL JOURNAL OF HEAT MASS TRANSFER**

Each of the following abstracts has been reprinted with permission from the *International Journal of Heat and Mass Transfer*.

**A PREDICTION OF THE BOUNDS ON THE EFFECTIVE THERMAL CONDUCTIVITY OF GRANULAR MATERIALS.** R. A. Crane and R. I. Vachon, *Int. J. Heat Mass Transfer 20*, 711-723 (1977).

Two models are developed to describe the effective thermal conductivity of randomly packed granular systems based on a one dimensional Ohm's Law method. These models are shown to represent upper and lower bounds on the effective conductivity of all normally distributed stochastic mixtures. An empirical factor has been obtained to account for three dimensional thermal effects. Comparisons to experimental data indicate that the modified correlation is generally accurate to within  $\pm 20\%$  over a wide range of constituent materials.

**LOCAL HEAT TRANSFER FROM A HORIZONTAL CYLINDER TO AIR IN CROSS FLOW: INFLUENCE OF FREE CONVECTION AND FREE STREAM TURBULENCE.** T. S. Sarma and S. P. Sukhatme, *Int. J. Heat Mass Transfer 20*, 51-56 (1977).

Local measurements have been made of the heat transfer from a horizontal circular cylinder to air in cross flow in forced convection and mixed convection. The studies have been conducted with the constant heat flux boundary condition at low

Reynolds numbers ranging from 500 to 4700 and at modified Grashof numbers ranging from  $0.8 \times 10^7$  to  $3.3 \times 10^7$ . The effect of varying the free stream turbulence intensity from about 0.5 to 20% has also been studied.

**MIXED CONVECTION FROM A HORIZONTAL CIRCULAR CYLINDER.** J. H. Merkin, *Int. J. Heat Mass Transfer* 20, 73-77 (1977).

The combined convection boundary layer on a horizontal circular cylinder in a stream flowing vertically upwards is studied in both the cases of a heated and cooled cylinder. It is found that heating the cylinder delays separation and can, if the cylinder is warm enough, suppress it completely. Cooling the cylinder brings the separation point nearer to the lower stagnation point and for a sufficiently cold cylinder there will not be a boundary layer on the cylinder.

**THE EFFECT OF SURFACE ROUGHNESS ON THE HEAT TRANSFER FROM A CIRCULAR CYLINDER TO THE CROSS FLOW OF AIR.** E. Achenbach, *Int. J. Heat Mass Transfer* 20, 359-369 (1977).

The influence of surface roughness on the heat transfer of a circular cylinder to the cross-flow of air has been studied. The Reynolds number was varied from  $2.2 \times 10^4$  to  $4 \times 10^6$ . The variation of the roughness parameter was  $0 < k_s/d < 900 \times 10^{-5}$ . The tests have been performed for the boundary condition  $T \approx \text{const}$ . For each of the rough cylinders, the static pressure distribution together with the local and total heat transfer have been measured. Particular attention has been paid to the transition from a laminar to a turbulent boundary layer as a function of Reynolds number and roughness parameter.

**MICROWAVE FREEZE-DRYING OF FOOD: A THEORETICAL INVESTIGATION.** T. K. Ang, J. D. Ford and D. C. T. Pei, *Int. J. Heat Mass Transfer* 20, 517-526 (1977).

An unsteady-state analysis of two dimensional freeze-drying with microwave internal energy generation is carried out, taking into account the differences of the transport parameters with respect to grain orientation, such as is found in food products. The anisotropic character of the material strongly influences the temperature profiles during drying. This importance is further amplified by a coupling effect between mass-transfer resistance, specimen temperature, and the absorption of microwave energy.

**LAMINAR FILM SURFACE EVAPORATION WITH UNIFORM HEAT FLUX IN A FAST ROTATING DRUM.** W. Roetzel, *Int. J. Heat Mass Transfer* 20, 549-553 (1977).

Effective cooling of a fast rotating drum can be obtained by evaporating a liquid flowing as a thin film on a conical inside surface. For special applications, a uniform heat flux and temperature over the outside cylindrical surface of the drum are desired. To obtain this, a wall profile is developed which yields a uniform overall resistance of wall and laminar liquid film.

NEW ANALYTICAL METHOD FOR A NONLINEAR DIFFUSION PROBLEM. M. Suzuki, S. Matsumoto and S. Maeda, *Int. J. Heat Mass Transfer* 20, 883-889 (1977).

Nonlinear parabolic partial differential equation with a concentration dependent diffusivity in semi-infinite region is solved by an analytical method under a constant flux boundary condition. The differential equation is transformed to a system of simultaneous linear ordinary differential equations. The solutions are represented by a series of products of the repeated integrals of the error function.

ANALYSIS OF MIXED FORCED AND FREE CONVECTION ABOUT A SPHERE. T. S. Chen and A. Mucoglu, *Int. J. Heat Mass Transfer* 20, 867-875 (1977).

An analysis is performed to study the flow- and heat-transfer characteristics of laminar mixed forced and free convection about a sphere. The transformed conservation equations of the nonsimilar boundary layers are solved by a finite difference method. Numerical results for gases having a Prandtl number of 0.7 are presented for buoyancy parameters which cover the entire regime of mixed convection, ranging from pure forced convection to pure free convection. In general, it is found that both the local-friction factor and the local Nusselt number increase with increasing buoyancy force for aiding flow and decrease with increasing buoyancy force for opposing flow. The effects of the variation of the local free stream velocities on the wall shear and surface heat-transfer results are also examined. With respect to the heat-transfer results, the buoyancy force effects on forced convection become significant for  $Gr/Re^2 > 1.67$  and  $< -1.33$ , respectively for aiding and opposing flows. The inertia force effects on free convection are found to be significant for  $Re^2/Gr > 0.01$ . The buoyancy-affected velocity profiles exhibit an overshoot beyond the local free stream velocity for aiding flow and an S-shape for opposing flow.

HEAT TRANSFER IN THE ENTRANCE LENGTH OF A HORIZONTAL ROTATING TUBE. J. E. McElhiney and G. W. Preckshot, *Int. J. Heat Mass Transfer* 20, 847-854 (1977).

An experimental study of heat transfer from steam condensing on the outside of a horizontal rotating tube through the tube wall to a laminar flow of cooling water on the inside of the tube is disclosed. A brief description of the apparatus and data gathering technique are given. The data for zero rotation are compared to literature values to establish the credibility of the experimental results. The overall heat-transfer coefficient and a previously developed model for the steam-side coefficient are used to calculate the coolant-film coefficient at a variety of rotational speeds. The data are reduced to dimensionless form, and a general correlation is presented. For rotations up to 40 rev/min the cooling-side coefficient is slightly improved; however, at rotations above 40 rev/min a severe deterioration of the cooling-side coefficient is observed. A phenomena of fluid mechanics is postulated to account for the loss of heat transfer and supporting evidence from the fluid mechanics literature is brought forth.

**THE LEADING EDGE EFFECT IN UNSTEADY NATURAL CONVECTION ON A VERTICAL PLATE WITH TIME-DEPENDENT SURFACE TEMPERATURE OR HEAT FLUX.** K. Mizukami, *Int. J. Heat Mass Transfer* 20, 981-989 (1977).

The propagation of the leading edge effect in unsteady natural convection on a semi-infinite vertical plate whose surface was subject to temperature or heat flux increasing exponentially with time or to a power-function change in temperature or in heat flux was analysed by means of the method of Goldstein and Briggs. The influence of the type of transient could be expressed in terms of several time-variables. Approximate expressions applicable also to other types of transient were derived.

**EFFECT OF APPROACH-FLOW VELOCITY AND TEMPERATURE NONUNIFORMITIES ON BOUNDARY-LAYER FLOW AND HEAT TRANSFER.** E. M. Sparrow, R. M. Abdel-Wahed and S. V. Patankar, *Int. J. Heat Mass Transfer* 20, 975-979 (1977).

The laminar flow and heat transfer on a flat plate subjected to nonuniform velocity and temperature profiles in the approaching free stream have been analyzed. The plate is situated in the laminar wake of an upstream plate. The extent of the approach-flow nonuniformities depends on the streamwise length of the wake relative to the length of the upstream plate. It was found that the effect of the nonuniformities is to reduce the wall shear and heat transfer on the downstream plate relative to their values for a uniform approach flow. Reductions of up to fifty percent were encountered. The extent of the reductions diminishes with increasing downstream distance, but non-negligible effects persist to a considerable length along the plate.

**THE THIRD KIND OF BOUNDARY CONDITION IN NUMERICAL FREEZING CALCULATIONS.** A. C. Cleland and R. L. Earle, *Int. J. Heat Mass Transfer* 20, 1029-1034 (1977).

Freezing of foodstuffs frequently involves boundary conditions of the third kind, which have proved to be difficult to handle and subject to error, mainly because the requirement to calculate surface temperatures implicitly. Experimental freezing studies on model and actual biological systems have been simulated by finite difference methods with a three time level scheme and boundary conditions of the third kind, for one dimensional heat transfer. For freezing times, from onset of cooling until the center reaches  $-10^{\circ}\text{C}$  experiment and prediction agreed to within  $\pm 14\%$  with 99% confidence, including experimental error estimated at  $\pm 4\%$ .

# GUIDE FOR AUTHORS

Typewritten manuscripts in triplicate should be submitted to the editorial office. The typing should be double-spaced throughout with one-inch margins on all sides.

Page one should contain: the title, which should be concise and informative; the complete name(s) of the author(s); affiliation of the author(s); a running title of 40 characters or less; and the name and mail address to whom correspondence should be sent.

Page two should contain an abstract of not more than 150 words. This abstract should be intelligible by itself.

The main text should begin on page three and will ordinarily have the following arrangement:

**Introduction:** This should be brief and state the reason for the work in relation to the field. It should indicate what new contribution is made by the work described.

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**Discussion:** The discussion section should be used for the interpretation of results. The results should not be repeated.

In some cases it might be desirable to combine results and discussion sections.

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DEWALD, B., DULANEY, J. T. and TOUSTER, O. 1974. Solubilization and polyacrylamide gel electrophoresis of membrane enzymes with detergents. In *Methods in Enzymology*, Vol. xxxii, (S. Fleischer and L. Packer, eds.) pp. 82-91, Academic Press, New York.

HASSON, E. P. and LATIES, G. G. 1976. Separation and characterization of potato lipid acylhydrolases. *Plant Physiol.* 57, 142-147.

ZABORSKY, O. 1973. *Immobilized Enzymes*, pp. 28-46, CRC Press, Cleveland, Ohio.

Journal abbreviations should follow those used in *Chemical Abstracts*. Responsibility for the accuracy of citations rests entirely with the author(s). References to papers in press should indicate the name of the journal and should only be used for papers that have been accepted for publication. Submitted papers should be referred to by such terms as "unpublished observations" or "private communication." However, these last should be used only when absolutely necessary.

Tables should be numbered consecutively with Arabic numerals. The title of the table should appear as below:

Table 1. Activity of potato acyl-hydrolases on neutral lipids, galactolipids, and phospholipids

Description of experimental work or explanation of symbols should go below the table proper.

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**EDITORIAL OFFICE:** Prof. D. R. Heldman, Editor, Journal of Food Process Engineering, Michigan State University, Department of Agricultural Engineering, East Lansing, Michigan 48824

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