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**Journal of
FOOD PROCESS
ENGINEERING**

**Edited by
D. R. HELDMAN**

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QUARTERLY

JOURNAL OF FOOD PROCESS ENGINEERING

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MEETINGS

JUNE 1981

June 1—5: INTERNATIONAL INSTITUTE OF REFRIGERATION, COMMISSION C1 (FREEZE—DRYING, CRYOBIOLOGY, MEDICAL APPLICATIONS). Theme: Techniques in low temperature technology for medical and biological applications. Warsaw, Poland. Contact: S. D. Augustynowicc, Research Center of Medical Technology, Low Temperature Department A1. Rewolncja Pazdziernikowej, 62, PO Box 54, 00—961, Warsaw, Poland.

June 3—5: PROGRESS IN FOOD ENGINEERING TECHNOLOGIES OF SOLID EXTRACTION, PURIFICATION, TEXTURIZATION SYMPOSIUM. Collegio dell Stelline, Milano, Italy. Contact: Symposium on Process in Food Engineering, Istituto di Technologie Alimentari, Universita via Celoria 2, 1—20133, Milano, Italy.

June 7—10: 41ST ANNUAL MEETING AND FOOD EXPO, INSTITUTE OF FOOD TECHNOLOGISTS. Georgia World Congress Center, Atlanta, Georgia. Contact: C. L. Willey, Institute of Food Technologists, Suite 2120, 221 North LaSalle Street, Chicago, Illinois 60601.

June 14—18: ANNUAL MEETING OF SOCIETY FOR CRYOBIOLOGY. St. Louis, Missouri. Contact: Lesley Lenny, The New York Blook Center, 310 E. 67th Street, New York, New York 10021.

June 21—24: SUMMER MEETING OF AMERICAN SOCIETY OF AGRICULTURAL ENGINEERS. Agricultural Energy Challenge: More Production — Less Consumption. Sheraton Twin Towers, Orlando, Florida. Contact: ASAE Headquarters, PO Box 410, St. Joseph, Michigan 49085.

June 21—24: 24TH ANNUAL CANADIAN INSTITUTE OF FOOD SCIENCE AND TECHNOLOGY. Research: Whose Business? Winnipeg Convention Center — Holiday Inn, Winnipeg, Canada. Contact: Barry McConnell, Conference Chairman, Department of Food Science, University of Manatoba, Winnipeg, Canada R3T 2N2.

June 25: ADVANCED MICROCOMPUTER INTERFACING SHORT COURSE. Sheraton Twin Towers, Orlando, Florida. Contact: M. A. Purschwitz, ASAE Headquarters, PO Box 410, St. Joseph, Michigan 49085

June 22—July 3: PACKAGING SYSTEMS LABORATORY. School of Packaging, Michigan State University, East Lansing, Michigan. Contact: C. J. Mackson, Director, School of Packaging, Michigan State University, East Lansing, Michigan 48824.

MEETINGS

JULY 1981

July 5—10: 4TH INTERNATIONAL CONFERENCE ON SURFACE AND COLLOID SCIENCE. Jerusalem, Israel. Contact: N. Garti, The Hebrew University, Jerusalem, Israel.

July 13—16: SEMINAR ON THE ECONOMICS OF PACKAGING AND DISTRIBUTION SYSTEMS. School of Packaging, Michigan State University, East Lansing, Michigan. Contact: C. J. Mackson, Director, School of Packaging, Michigan State University, East Lansing, Michigan 48824.

July 20—31: DYNAMICS OF PACKAGING. School of Packaging, Michigan State University, East Lansing, Michigan. Contact: C. J. Mackson, Director, School of Packaging, Michigan State University, East Lansing, Michigan 48824.

July 27—31: ADVANCES IN FOOD SCIENCE AND TECHNOLOGY. Massachusetts Institute of Technology. Contact: Director of Summer Session, Room E19—356, Massachusetts Institute of Technology, Cambridge, Massachusetts 02139.

AUGUST 1981

August 3—6: INTERNATIONAL INSTITUTE OF REFRIGERATION, COMMISSION C2 (FOOD SCIENCE AND TECHNOLOGY), D1 (REFRIGERATION STORAGE), D2 (REFRIGERATED LAND TRANSPORT), D3 (REFRIGERATED SEA TRANSPORT) MEETING. Boston, Massachusetts. Theme: Advances in technology in the chilling, freezing, processing, storage and transport of fish, especially underutilized species. Contact: International Institute of Refrigeration, 177 Boulevard, Malesherbes, 75017 Paris, France.

August 3—7: FOOD RHEOLOGY: PRINCIPLES AND PRACTICE. Massachusetts Institute of Technology. Contact: Director of Summer Session, Room E19—356, Massachusetts Institute of Technology, Cambridge, Massachusetts 02139.

August 10—14: CRYOGENIC ENGINEERING CONFERENCE AND CRYOGENIC APPLICATIONS SHOWCASE. Town and Country Hotel, San Diego, California. Contact: S. Friedman, Cryogenic Society of America, 9911 W. Pico Blvd., Suite 630, Los Angeles, California 90035.

August 11—13: SEMINAR ON SOLVING CORROSION PROBLEMS AND AIR POLLUTION CONTROL EQUIPMENT. Brown Palace Hotel, Denver, Colorado. Contact: National Association of

MEETINGS

Corrosion Engineers, PO Box 218346, Houston, Texas 77218.

August 17—20: SEMINAR ON ADVANCES IN HOT MEAT PROCESSING — INTERNATIONAL IUFOST. Sponsored by the Polish Academy of Sciences and Committee of Food Technology and Chemistry. Rydzyna near Poznan, Poland. Contact Professor W. Pezacki, Institute of Food Technology, Wojska Polskiego, 31, 60—624, Poznan, Poland.

August 24—28: INTERNATIONAL FEDERATION OF AUTOMATIC CONTROL CONGRESS VIII. Kyoto, Japan. Contact: IFAC Secretariat, Schlossplatz 12, A2361, Laxenburg, Austria.

SEPTEMBER 1981

September 7—11: ALIM—EN—TEC 81. Theme: Technology for Food Processing and Packaging. Mexico City, Mexico. Contact: Mexican Ministry of Industrial Promotion. LANFI, Public Relations Department, Av. Industria Militar #261, Mexico 10DF.

September 14—17: AGRITECH 81. International Agricultural Mechanization and Technology Exhibition. Telaviv, Israel. Contact: Government of Israel Trade Center, Industrial Division, 350 Fifth Ave., New York, New York 10118.

September 25: FOCUS ON FOOD SCIENCE SYMPOSIUM III. Theme: Modern Meat Technology — Microbial Considerations. Kansas State Union, Kansas State University, Manhattan, Kansas. Contact D. Y. C. Fung, Food Science Graduate Program, Leland Hall, Kansas State University, Manhattan, Kansas 66506.

OCTOBER 1981

October 2—6: BAKERY EXPO 81. Las Vegas, Nevada. Sponsored by the American Baking Association and the Baking Equipment Manufacturers Association. Contact: 1981 International Baking Industry Exposition, PO Box 37270, Washington, D. C. 20013.

October 15—16: THIRD BI-ANNUAL SEMINAR AND PACKAGING SHOW. School of Packaging, Michigan State University, East Lansing, Michigan. Contact: C. J. Mackson, Director, School of Packaging, Michigan State University, East Lansing, Michigan 48824.

NOVEMBER 1981

November 8—11: THE NATIONAL FROZEN FOOD CONVENTION. San Francisco Hilton, San Francisco, California. Contact:

MEETINGS

National Frozen Food Convention, Administrative Office, 604 W. Derry Road, PO Box 398, Hersey, Pennsylvania 17033.

November 15—19: FOOD AND DAIRY EXPO 81. Georgia World Conference Center, Atlanta, Georgia. Contact: Dairy and Food Industry Supply Association, 5530 Wisconsin Ave., Suite 1050, Washington, D.C. 20015.

DECEMBER 1981

December 15—18: ASAE WINTER MEETING: LAND AND WATER RESOURCES — A CRITICAL RESPONSIBILITY. Palmer House, Chicago, Illinois. Contact: ASAE Headquarters, PO Box 410, St. Joseph, Michigan 49085.

RETORT POUCH: THE DEVELOPMENT OF A BASIC PACKAGING CONCEPT IN TODAY'S HIGH TECHNOLOGY ERA¹

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INTRODUCTION

One indication of progress and maturity for a technological concept is that, in any introduction to discussion on the subject, descriptive details, characteristics, and founding principles can be minimized. The retort pouch is a multi-layer poly mer-foil pouch used (Fig. 1) in lieu of the metal can or glass jar for shelf-stable foods or refrigerated or frozen foods. Although, from technological and commercial acceptance aspects, the retort pouch's applicability should be judged alone without comparison, the comparison with cans for preservation mode, frozen "boil-in-bag" items for end product utility, and to both for the cost-value relationship is inevitable.

As is true with the other process techniques and advances discussed in this review, the retort pouch concept has been developed in an era of high biological, physical, chemical, mechanical, and communication sophistication and sensitivity. This has meant that many known questions had to be delineated and answered before adoption or marketing of the end product could be considered.

After mention of some of the basic motivations toward considering the retort pouch and the ensuing questions and trade-offs that led to approaches taken, the more significant of these questions and their corresponding answers will be summarized. Next, a few representative technical areas will be covered in more detail to provide some sense of the range of problems and/or anticipated questions that required resolution. Then there will be assessments of the current technical and commercial status. Finally, there will be mention of current develop-

¹1980 Underwood-Prescott Symposium Lecture, Massachusetts Institute of Technology, Cambridge, Mass 02139 September 30, 1980.

ment needs, improvement opportunities, and a prognostication on the future of the retort pouch. Much of the discussion will cite Natick Research and Development Laboratories (NLABS) experiences. It must be strongly emphasized that independent, parallel, similar and vital contributions have been made, both domestically and internationally, by many academic institutions, food processors, equipment manufacturers and packaging material suppliers.

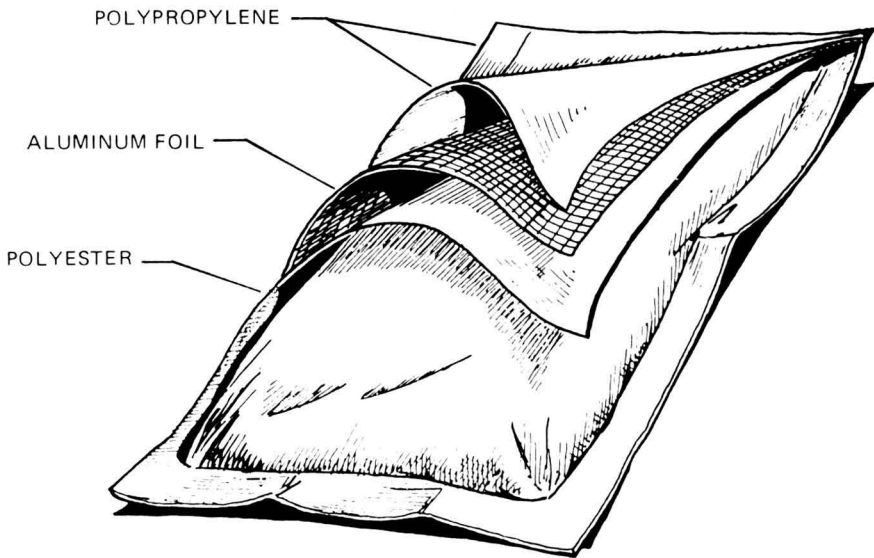


FIG. 1. RETORT POUCH

MOTIVATION

The very earliest motivation was probably simple curiosity. Once the early disclosures of successful laboratory results were noticed, however, potential advantages led to the consideration of the retort pouch for military and civilian use. For the U. S. military, the basic incentives for operational ration use were functional: a shape convenient to carry with negligible constraint to movement; softness to preclude injury during crawling, etc.; convenient opening; a ready-to-eat product; cube savings; and some weight savings. At this early time, the late 50's, quality equivalency, not improvement was the criterion. It was also accepted that the concept should have commercial appeal so that a reasonable production base and nominal costs could exist.

Commercial exploitation surfaced first in Europe with limited production of marketed items in Italy in 1960, Denmark in 1966, and in Japan in 1969. Motivation for these market entries was abetted by the lack of significant alternative preservation modes, such as frozen foods; or packaging, such as glass; and high costs for those alternative packaging materials. Tsutsumi (1972) credits the advent of pasteurized, flexibly-packaged products in Japan after World War II with facilitating the evolution to retort-sterilized items. Producibility existed; only a few technical barriers to manufacturing had to be overcome. The bottom line incentive, an acceptable cost-value relationship, existed for each of those specific geographical markets.

In North America, outside of the military, some early commercial test packs (Goldfarb 1970, 1971), and Canadian efforts (Anon. 1975), there has been less motivation to aggressively develop and market the pouch. Packers have not felt a great need to develop and establish a new package to compete with current products. During the early 1970's, interest, based upon potential quality, convenience, and established durability, was high but soon waned during the FDA review of packaging material acceptance. The energy crisis has been timely in motivating renewed commercial interest. A proper cost-value relationship for the pouch seemed easier to reach, and energy is another motivation for those who are active in this area.

FUNDAMENTAL TECHNICAL CONSIDERATIONS

Once the decision had been made to further investigate and develop the retort pouch, there was the necessary evaluation of alternate processing approaches, selection of research to perform, and systematic establishment of a responsive data base. This, based on the literature and discussions with the technical communities, is basically the approach followed in Japan, Europe, and North America.

Szczablowski (1971) discussed several of these as related to the U.S. Army Research and Development Laboratories' program. Examples of these and the respective decisions made by the military to meet their own perceived needs are:

- (1) Should aseptic sterilization techniques or post-packaging heat processing procedures be followed? The need to package solid foods precluded aseptic packaging at the time. At that time, the late 1950's, and still true to some extent, convincing food processors that the pouch concept was feasible and reliable was felt to be quite formidable without the burden of having to develop and

demonstrate new sterilization procedures and process equipment. Furthermore, many foods to be packaged in pouches were solids and the thin cross section of the pouch was viewed as a means to maintain quality through process.

- (2) Were the candidate three-ply films adequate performers or should emphasis be put on the development of a single film? The film and adhesive suppliers, independently and with cooperative support, proved that the three-ply approach was adequate and that films *per se* should not be a pacing problem.
- (3) What is the state-of-the-art of the technology required for implementation? Reviews, aided by industry seminars, revealed that all the basic production concepts existed but performance criteria had to be tightened and some modifications were necessary to accommodate the singular characteristics of the flexible pouch. This conclusion would help in communicating the producibility of retort pouches to industry. Elements of this assessment included:
 - a. The potential of using a steam-flush for package headspace air removal.
 - b. A decision to remain with a side seal package design rather than to explore a drawn pouch concept.
 - c. Recognition, that to achieve higher filling speeds, continuous equipment could be designed, and
 - d. Water cook retorting was feasible.
- (4) What initial data is required to support a decision for larger scale, serious development? The answer was product acceptability /shelf-life and package durability ; therefore, the first tests performed were storage tests of representative products and an assessment of resistance to abuse on drop, vibration, and personal field usage conditions (the last for the military , in field jacket pockets over a clothing-wear obstacle course).

DATA BASE

The next period of development, early 60's to mid 70's, resulted in quantitative answers to questions: related to the fundamental considerations and to data on which final decisions for military adoption, regulatory agency approvals, and commercialization could be made. The data was used to confirm the initial findings and capabilities of the concept. Several reviews of these extensive developments now exist (Lampi 1977, 1979; Tsutsumi 1972; Nughes 1974; Yamaguchi 1979). In brief, summary form, data obtained during this period established for:

Materials

- (1) A reliable retortability screening procedure using actual product and simulation of periodic retort pressure fluctuations.
- (2) A basis for initial, conditional approval of films relative to the quantity and safety of extractives.
- (3) Film structures that neither contributed off-flavors nor depleted desired product flavor over a period of time.
- (4) Limits of retortability.

Products

- (1) Basic formulations; comparative, hedonic acceptance levels and shelf-life estimates for a wide variety of basic and formulated foods.
- (2) Markets external to the United States, some favorable and some marginal.

Production

- (1) The initial definition of necessary unit operations, Fig. 2, applicable equipment and specifications for each function.
- (2) Initial production quality assurance tests and performance criteria.
- (3) Evolutionary improvements in the design and performance of equipment.
- (4) An affirmation that retort pouches can be prepared under commercial and commercial-type conditions at a high reliability level.
- (5) In some non U.S. environments, a favorable cost/value relationship.

Durability

- (1) Favorable controlled laboratory and shipping test measurements of abuse resistance in direct comparison to cans.
- (2) Shipping experience, both controlled and uncontrolled, that indicated adequate ruggedness.

In several geographical areas where commercialization occurred early, much of the data base was accumulated concurrently with actual manufacturing and marketing experience.

SELECTED TECHNICAL EFFORT

To typify the range of considerations and to reinforce and iterate that significant amounts of data and background are available, three research and development areas will be described in detail.

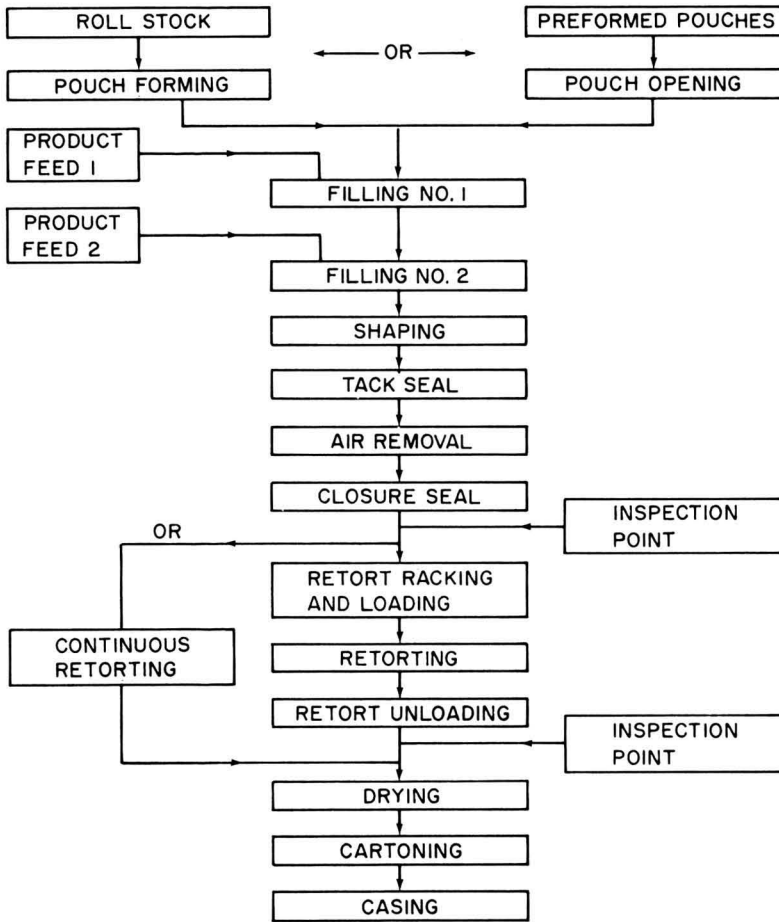


FIG. 2. FLOW CHART OF PRIMARY UNIT OPERATIONS

Definition of Defects

During early testing of pouches for durability and resistance to abuse, it was noted that food-filled pouches with easily visible holes in the pouch body did not show signs of spoilage (swelling) when immersed in a suspension of gas-producing bacteria. The biotester (Maunder *et al.* 1968) was developed to induce, through a kneading action, momentary pressure changes and penetration of bacteria through pouch defects; the biotester thereafter became a laboratory standard to determine pouch integrity. During that general time, when no experience on the incidence or frequency (and therefore, importance)

of pouch leakers existed, studies were initiated to investigate the smallest size hole in the three-ply materials that bacteria would pass through in the biotester, the utility of using a gas flow measurement in lieu of physical dimensions for sizing, and the nature, mainly size, of flex, puncture and abrasion caused defects.

The procedure for investigating hole size versus bacterial penetration was:

- (1) Using a laser to drill holes in pouch blanks; preliminary sizing by flat field microscope.
- (2) Forming a pouch from the blank; sterilizing the pouch with nitrous oxide (it was feared that absorption of gases, such as ethylene oxide, might alter the size of the hole).
- (3) Filling pouches aseptically with semi-solid nutrient agar containing dextrose; removing residual air, and sealing.
- (4) Biotesting with *Aerobacter aerogenes* followed by incubation at 37° C (100° F) for 14 days; checking for swelling.
- (5) Measuring the rate of gas flow through the holes.
- (6) Scanning electron microphotographing of defects; calculating hole diameter.

The results were too meager for reliable statistical interpretation. They did, however, present an indication, when rationalized by the use of physical principles, that penetration of bacteria was unlikely to occur with holes less than 10 μ in diameter.

Out of 500 laser-exposed pouch blanks, only 30 permitted gas flow. These were prepared as filled pouches and biotested. On final SEM examination, only twelve defects were clean enough and regular enough in shape to rely on for accurate analysis of results. Based on results with these twelve, as shown in Fig. 3, no penetration occurred in five of six pouches with holes 11 μ in diameter and smaller. A gas flow rate less than 0.1 atm. cc/s of helium at a ΔP of 192 mm Hg would also indicate holes where penetration would be unlikely. To present a perspective of the findings: the 10 μ hole diameter is 1/10 the thickness of the three-ply pouch material, the diameter of an *Aerobacter aerogenes* bacterium is 0.5 μ . The physics rationale was that approximately three pounds pressure is required to overcome capillarity in a 10 μ diameter hole and a 3-4 psi pressure differential is reasonable for the biotester action.

Next, experiments were run to determine the size defect that might be expected from flexing in the MIT Folding Endurance Tester; puncture with a fine, fire-hardened tungsten wire; or abrasion of a fold against a relatively smooth fibreboard surface. The results showed that the smallest defects caused by each type of abuse were:

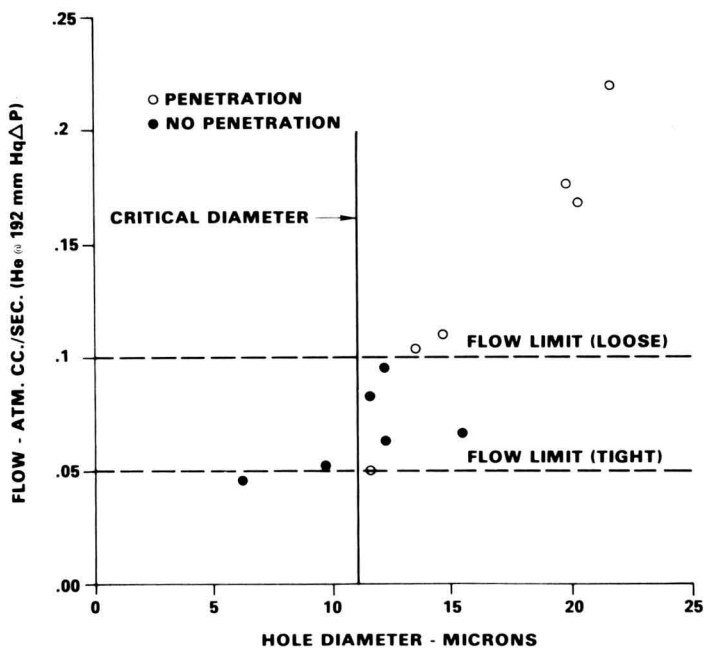


FIG. 3. LASER DRILLED DEFECTS MEASURED SIZE VS. FLOW RATE WITH NOTATIONS ON BACTERIAL PENETRATION

Flexing - 11 μ diameter hole
 Puncture - 100 μ diameter hole
 Abrasion - 24 μ diameter hole

These tests indicated that none of the three possible causes of abusive pinholing created leaks smaller than the 11 μ tentative critical diameter established by earlier tests. They also indicated that since the two films tested required from 1000 to 5000 flexes for failure, flexing is not a likely cause of failure and that abrasion and puncture caused relatively large pinholes.

These experiments formed a basis for evaluating the suitability of automatic leak testing procedures. No practical leak detection procedures were found, and since only one package in 100,000 is likely to have a leak below 100 μ in size (i.e., unlikely to be detected with the naked eye), justification to develop such instrumentation would be difficult. At this stage, these findings are rather academic; but at that time, mid 60's, pending real-life experience, they were essential as back-up.

Seal Area Cleanliness

Contamination of the pouch closure seal area during filling, and occasionally during air removal, has been the cause of most production-related pouch defects. Because of variations in the physical characteristics among products, and even among formulations for a single product, no detailed, universally applicable specifications to avoid seal area contamination can be given.

The approaches to elimination or drastic reduction of seal area contamination have been pragmatic, experience-conditioned engineering. Without detailing the various forms each step or device can take, steps to safeguard clean filling surfaces are:

- (1) Match the filler (auger, gear pump) and nozzle (wide - or narrow plug valve, etc.) to product characteristics by actual tests.
- (2) Specify bottom-to-top filling and no filling to seal area.
- (3) Eliminate nozzle drip by means of built-in suction features; external suction rings and/or sheet metal guards synchronized to physically prevent drip from contaminating seal surfaces.
- (4) Control pouch opening by conveyor clamps on both edges of the pouch, external suction cups and/or "spoonbill" forming devices.
- (5) Use hinged guards that swing down into the package opening to physically protect inner seal surfaces.
- (6) Control the handling of pouches between unit operations to preclude splash.
- (7) Avoid occluded air in the product and control rate of headspace air removal to prevent splash.
- (8) Consider the use of a tack or partial seal before air removal.
- (9) Control product consistency.

An alternate approach to get clean seal surfaces consists of a steam flush to remove contamination. Schulz and Mansur (1969) indicated that steam flushing not only cleaned seal surfaces, but removed residual air from the pouch. A curved sealing bar with transverse radius against a silicone-rubber anvil was then used to seal through residual moisture and grease. This technique has evolved to commercial use. Tsutsumi (1974) reported on an effective triple-bar sealing technique to cope with seal area liquid contamination: an initial hot bar makes the basic seal; a second, slightly lower temperature bar flattens any blister caused by vaporization of contamination during the initial sealing; a third ambient temperature bar performs a flattening action on the still heated polymer.

Significance of Inter-Lamina Bond Strength

The basic and obvious significance of inter-lamina bonding for retort pouch films is that the components must remain intact through retorting, handling, and storage. Beyond that, however, are the subtleties of procedures for evaluating resistance to external forces and of establishing realistic specifications.

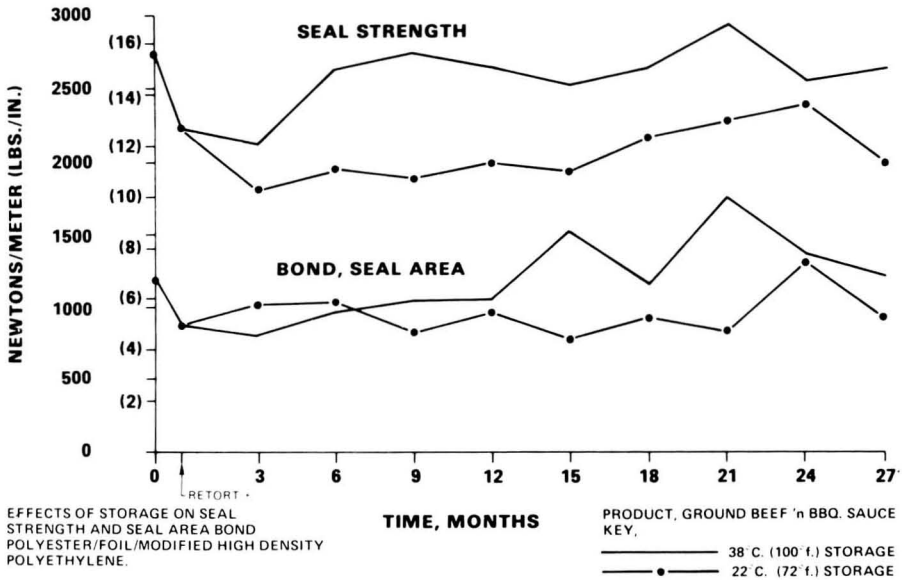


FIG. 4. EFFECTS OF STORAGE ON SEAL STRENGTH AND SEAL AREA BOND

At U.S. Army Natick Research and Development Laboratories (NLABS), the initial exposure to the vagaries of bond strength measurement came when pouches, filled with water, successfully passed laboratory retort tests, but failed dramatically in large scale production trials. Refinements in the screening procedure were made; the use of actual product (especially where there were lipids) was specified and the laboratory retort cycle included ± 2 psi fluctuations to simulate pressure variations that could be experienced in some commercial retorts.

Schulz (1973) showed the effect of retorting on the bond strength of one material - a drop from 600 gm/cm to 200 gm/cm. Other materials exhibit a reverse trend - an increase on retorting. Similar differences between two totally satisfactory films (in terms of handling performance) occur in bond strengths over storage periods (as long as 27

months are shown in Fig. 4 and 5. Since the retort pouch seals are total fusion, the performance of the pouch in terms of its capability to withstand vibration and drop is dependent on this seal area bond.

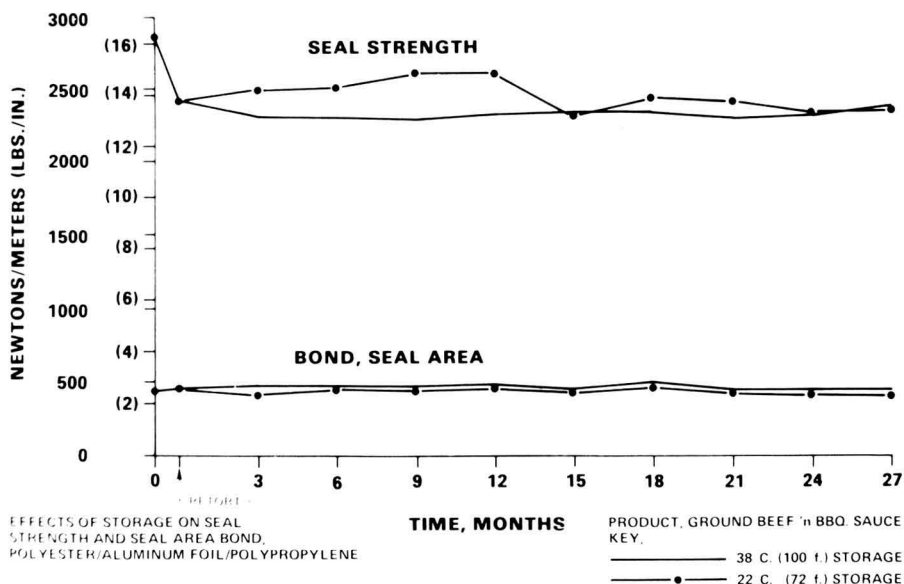


FIG. 5. EFFECTS OF STORAGE ON SEAL STRENGTH AND SEAL AREA BOND

CURRENT STATUS

Generally, the technology has been established. The basic questions on feasibility, durability, and applicability have been answered. There is trust that the pouch can be manufactured and will perform. The Japanese, primarily, are continuing to advance the art and make refinements to products, package and process. In the U.S., the energy crisis is one reason food processors remain interested, but involvement has been minimal and slow. The most commonly cited deterrent in the U.S. has been the speed of filling and sealing which, when compared to can lines running 150 to 400 or more per minute, is a slow 30 to 60 per minute. We apparently have a "chicken or egg" stalemate: processors are reluctant to invest in equipment apt to be rapidly obsolete, yet, equipment suppliers are reticent to invest in manufacturing higher speed equipment without some assurances of an adequate market.

Outside the U.S., commercialization has proceeded with multiple 60 per minute machines.

Technology

A summary of the current state of the technology is:

(1) Films —

For 116°C to 124°C processes:

9 to 25 micron polyester/9 to 25 micron foil/75 micron polyolefin (modified polyethylene or ethylene - propylene copolymers and blends). (It should be noted that U.S. FDA approval is currently limited to 121°C.)

For processes up to 138°C:

12 micron polyester/9 micron foil/15 micron oriented Nylon-6/50 micron polypropylene. (It should be noted that U.S. FDA approval is currently limited to 121°C.)

(2) Products —

Over 100, ranging from commodity vegetables to "ready meals".

(3) Package Design —

Flat four seal; ranges from 4-3/4 in. x 7-1/4 in. x 7/8 in. for 5 to 10 oz. contents to 12 in. x 18 in. x 1-1/2 in. for institutional packs, for 5-7 lb contents. With folding carton or polymer bag overwrap.

(4) Pouch Packaging Equipment —

From roll stock, intermittent motion packager for 25 to 60 pouches per minute; can incorporate steam flush and closure sealing or prepare pouches for transfer to separate vacuum sealer, or could incorporate in-line vacuum sealing without transfer to separate machine.

From roll stock, continuous motion packager for 250 pouches per minute, currently in fabrication and is in tests for commercial processing performance.

From preformed pouches, filler-sealers for 25-60 per minute, can incorporate squeezing action or steam flush for air removal.

(5) Retorts —

Horizontal batch, water or steam-air cook, modified to assure uniform distribution of heating media; use of retort racks.

Horizontal batch, water or steam-air with separate heating media accumulation tank; suitable for high temperature (135°C) cooks.

Continuous horizontal or vertical retorts for water or steam-air.

(6) Cartoning —

Standard folding carton equipment.

Yamaguchi (1979) reported that based upon ten years of experience

in Japan, a basic production system has been designed and twenty of these systems have been set up.

Commercialization

The current state of commercialization is highest in Japan where, according to Yamaguchi (1979), retort pouched foods in 1978 accounted for 8.7% (\$259 million U.S.) of the pouch, can, and frozen food market (excluding beverages). Thirty-three firms, including 13 major ones, pack pouches. In Europe, outside of STAR in Italy, commercialization has been inconsistent. Mencacci (1975) cites an overall production level of 70-80 million pouches for 1974. The recent pouch line for commodity vegetables at Lustucru, near Cambrai, France, is cited as Europe's first production-speed retort pouch line.

In the United States, the military has taken the lead in both development and adoption of the retort pouch. The Boy Scouts of America, for their wilderness programs, have also found the retort pouch acceptable. These users - 24 million meals (40 million thermostabilized retortable pouches) a year for the military and 300,000 plus per year for the Scouts - are currently procuring items from six packers. Kraft and ITT Continental Kitchens are test marketing products; Specialty Seafoods of the state of Washington is offering smoked salmon. In Canada, Freddy Chef of Quebec and Magic Pantry of Ontario are producing pouched foods commercially.

Economics

A detailed analysis of the current status of the economics cannot be done within the limits of this dissertation and would be of marginal value, since each market target, physical plant requirements, and area energy constraints would require a specific assessment. In Japan, the economics are obviously favorable, and to a lesser extent, the same seems to be true for some products in Europe. Relative to the latter area, a recent comparison of costs to prepare 1000 g portions of pea soup in Denmark (Poulsen and Raahauge 1979) showed that a retort pouch approach costs 80% of the sterilized can approach, 40% of freezing, and 20% of freeze-dehydration. The costs of capital, energy, labor, and packaging were considered.

In the United States, Kraft and ITT Continental Kitchens have apparently found that the economics of production and the perceived cost/value relationship of the end product favorable enough to explore the concept further with test marketing. Asp (1979) of Hormel reported difficulty in establishing a clear favorable cost/value relation-

ship in comparison with cans, based upon consumer acceptance data and the current relatively slow packaging speeds. Cooper (1979) on reviewing six alternate form, fill, seal and retorting systems, reported that a 40 to 60 pouch per minute line would require an initial investment of \$400,000 to \$800,000. For a 7 to 9 oz entree, his analysis estimated a 1-1/2 year pay back period on the equipment investment. He also pointed out that foil containing laminates are increasing in cost at a significantly lower rate than metal food cans are.

Institutional Pouches

The current scene includes significant interest and development effort (Beverly *et al.* 1980) on institutional size pouches - up to 12 in. x 18 in. x 1-1/2 in. in size, holding 5-7 lb of food. Filler-sealers capable of handling 12 in. x 15 in. pouches (5 lb contents) are now available. In addition to the anticipation that quality improvements are possible because of the flat shape, interest in the larger sizes arises from energy and packaging cost considerations, where, as shown by Badenhop and Milleville (1980), the pouch has advantages. Further savings can accrue from the fact that less brine or sugar syrup is required since the package conforms to the essential product volume. The intent of the institutional pouch is competition with #10 cans for commodity-type foods and with half-size steam table trays and frozen packs for formulated foods for food service.

RESEARCH AND DEVELOPMENT NEEDS

There is adequate technology for those knowledgeable to proceed to commercialization. However, as true with any technical entity, the more the subject is examined, the more questions arise. Many relate to evolutionary improvements in product and process; many pertain to safety and safeguards; all relate to the realities of today's production and distribution environments, conditions and hazards - the nitty-gritty of day-to-day operations with a product that, unless handled properly, can constitute a health hazard.

Beverly (1980), commenting on a worldwide seminar on retort pouches, felt that many food processors left that meeting confused because of the many diverse opinions. This is undoubtedly true, but the presentations also signal that much data and experience are being accumulated. Kopetz *et al.* (1979), based on several studies, enumerated specifics of pouch processing that must be quantified. Badenhop and Milleville (1980) voiced an urgent need to develop industry-wide

processing and packaging specifications for the retort pouch. Collectively, these are legitimate urgings to identify and collect more data on critical pouch elements. To the above authors' admonitions, I would add collection and organization of current data, since in reviewing several recent publications, I sense a lack of communication or awareness of what already has been done. Also, why not a worldwide approach?

Some organization of this additional work can be made:

(1) Formulations

Andres (1979) noted that the thrust of pouch research has now moved into the phase of ingredient definition and formulation improvement. I agree that this is currently the greatest need, for the military as well as for commercial applications. Asp (1979) reported an inability to establish a consumer acceptance for pouched items in comparison to like items in cans and voiced a need to establish product differences. This should be a challenge for formulators working with process engineers. Dacyshyn (Anon. 1979), discussing success with a cabbage roll product, mentioned that 20 to 50 recipe tests are required to get the proper combination of ingredients and retort time.

(2) Manufacturing Practices

Establishment of good manufacturing practices and specifications through repetitive investigations on thermoprocessing and all interfacing unit operations to establish trends and to enumerate the effects of process variables and deviations from these; in short to continue to build confidence through accumulated experience.

(3) Continuous Processing Systems

As stated earlier, a continuous motion pouch form, fill, seal machine is in its final stages of development. A continuous retort was used with early vegetable packs and is currently used for pouched intravenous solution. The marrying of these two, or like equipment, into a continuous processing system is a logical next step. Production speeds will be increased, manual handling of pouches can be eliminated, product exposure to process variables will be predictable and standardized, and energy use will be optimized.

(4) Film Conversion

A frequently reported production difficulty has been an inadequate supply of acceptable three-ply laminated film. Perhaps the problem is more economic than technical; i.e., the need for an assured market large enough to justify investment and improvements in converting lines. However, since film costs are a

critical element in the cost/value relationship formula of the pouch, there would appear to be room for investigations to improve and lower the cost of film lamination. Undoubtedly the rapid inception of pouch programs has stressed the ability of suppliers to react adequately.

FUTURE OF THE RETORT POUCH

The retort pouch is a firm commercial entity in Japan and still tenuous in Europe. In Canada, there are two commercial ventures - one in Quebec and one in Ontario. In the United States, test marketing by two major firms is in progress, while the military has adopted the pouch as its standard operational ration entree.

Where the pouch goes from here is difficult to assess, but many are trying. Judging from the number of phone inquiries from market research firms and published reviews and predictions (Bakos 1979; Office of Technology Assessment, U.S. congress, 1978; Katzenstein 1975), there is a good future for the retort pouch, most likely surfacing in the mid-to-late 1980's decade. The impetus is the energy crunch, abetted by changing lifestyles (which surfaces both retail and institutional applications).

The growth, except for some special instances, will be in comparison to many of today's market innovations, relatively slow. Perhaps over the long run, this is best. I think it is. As Dacyshyn (1979) relates on his relatively small market for pouched items in Ontario, he is building customer loyalty. Hopefully, such factors can offset calamities that frequently occur when after large scale introduction of a product and signs of initial and rapid success, promotional support is withdrawn. Also, slow growth should re-inforce a firm high quality level, that "excessive" competition hopefully cannot cheapen.

Conventional canned and frozen foods will remain as our preservation mainstays for some years to come. Inroads by the pouch are certain and many applications will be based upon the retort pouches singular characteristics. Some future applications will be:

- (1) Military - already there
- (2) Campers meals - reflecting dissatisfaction with the cost and quality of freeze-dried foods
- (3) Individual and possibly some dual servings - for young marrieds, singles, and the elderly ("ready-meals")
- (4) Delicate sauces and adjuncts
- (5) Institutional packs - if costs remain competitive
- (6) Special diet items - high preparation labor, special needs, such as low sodium general hospital use.

Overall, my prognosis is that the initial motivating factors - simplicity, convenience, known technology, a shape conducive to optimal thermal processing; supplemented by the need to conserve energy - are still valid and will result in commercial application and market feasibility for the retort pouch.

REFERENCES

- ANDRES, C. 1979. Retort pouch research efforts shift to food formulations food processing *40* (a), 44-46.
- ANON. 1979. Basically lady, it's a soft can. *Meat Industry* *25* (9), 18-20, 78.
- ANON. 1975. Prototype line for the flexible, retortable pouch. *Can Packag.* *28* (6), 30-33.
- ASP, R. 1979. The flexible pouch, Oct. '77, and today. *Activities report of research and development assoc.* *31* (2), 30-35.
- BADENHOP, A. F. and MILLEVILLE, H. P. 1980. Institutional size retort pouches. *Food Processing* *41* (1), 82-86.
- BAKOS, J. B. 1979. Pouch - consumers' view - wanted or not wanted? *Activities report. Research and Development Assoc.* *31* (2), 45-48.
- BEVERLY, R. G. 1980. Retort pouch in the 80's. *Food Engr.* *52* (3), 100-103.
- BEVERLY, R. G., STRASSER, J. and WRIGHT, B. 1980. Critical factors in filling and sterilizing of institutional pouches. *Food Technology* *34* (9), 44-48.
- COOPER, J. K. 1979. Pouch suppliers' view - production on or off? *Activities report. Research and Development Assoc.* *31* (2), 19-22.
- GOLDFARB, P. L. 1970. Pouch for low acid foods. I. *Mod Packag.* *43* (12), 70-76.
- GOLDFARB, P. L. 1971. Pouch for low acid foods II. *Mod Packag.* *44* (1), 70-76.
- KATZENSTEIN, A. W. 1975. The food update delphi survey. *Food Prod. Dev.* *9* (5), 20, 22, 26.
- KOPETZ, A. A., PRANGE, C. A., and FLESSNER, R. J. 1979. The future in our hands: critical factors in retort pouch thermal process assurance. *Activities report. Research and Development Assoc.* *31* (2), 49-53.
- LAMPI, R. A. 1977. Flexible packaging for thermoprocessed foods. *Advances in Food Research* *23*, 306-428.
- LAMPI, R. A. 1979. Overview and General Production Requirements for Retort Pouch. *Proceedings. Seminario Internacional Sobre Embalagens Flexiveis Esterilizaveis. Instituto de Tecnologia de Alimentos. Campinas, Brazil.*
- MAUNDER, D. T., FOLINAZZO, J. F. and KILLORAN, J. J. 1968. Bio-test method for determining integrity of flexible packages of shelf-stable foods. *Food Technology* *22* (5), 81-84.
- MENCACCI, S. A. 1975. Equipment systems for sterilizable pouches - USA, Europe, Japan. *Activities report. Research and Development Assoc. Mil Food Packag. Syst.* *27*, 161-175.
- NUGHES, F. 1974. *European Development in Retortable Pouch Packaging*, Presented at Annual Packaging Institute Forum, Chicago, IL.
- Office of Technology Assessment. Congress of the United States. 1978. *Priorities for Technology Assessment. Retortable Pouch. In Emerging Food Marketing Technologies. A Preliminary Analysis.* p. 14, 86 pp.
- POULSEN, K. P. and RAAHAUGE. 1979. Ready made meals made by freezing

- and other preservation methods - a cost comparison. *Scandinavian Refrigeration*. April 79, 325-328.
- SCHULZ, G. L. 1973. *Test Procedures and Performance Values Required to Assure Reliability*, Proceedings of the Symposium on Flexible Packaging for Heat-Processed Foods, pp. 71-82. National Academy of Science National Research Council, Washington, DC.
- SCHULZ, G. L. and MANSUR, R. T. 1969. Sealing through contaminated pouch surfaces, Tech. Rep. 69-76 -GP. US Army Natick Laboratories, Natick, MA.
- SZCZEBLOWSKI, J. W. 1971. *An assessment of the flexible packaging system for heat-processed foods*, Tech Rep. 71-57-GP. US Army Natick Laboratories, Natick, MA.
- TSUTSUMI, Y. 1972. "Retort Pouch" - Its development and application to food-stuffs in Japan. *J. Plast.* 6, 24-30.
- TSUTSUMI, Y. 1974. The Growth of Food Packed in Retortable Pouches in Japan, Presented at Annual Packaging Institute Forum, Chicago, IL.
- YAMAGUCHI, K. 1979. Production of retort pouch in Japan. Proceedings. Seminario Internacional Sobre Embalagens Flexíveis Esterilizáveis. Instituto de Tecnologia de Alimentos. Campinas, Brazil.

MATHEMATICAL MODELING OF ROUGH RICE DRYING IN A SPOUTED BED

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ABSTRACT

A theoretical model, based on the well-known mass and energy transport equations for a continuous media, was rigorously developed to describe the liquid diffusion controlled drying of cereal grain in a batch operated, well-mixed spouted-bed dryer. The equilibrium moisture content equation and the effective diffusion coefficient are required to solve the heat and mass transfer equations. Numerical techniques were used to solve these equations.

The model was tested using laboratory-scale experiments where rough rice was dried in a spouted-bed. A good agreement was obtained between the experimental and theoretical model.

INTRODUCTION

It is well known that the drying of most cereal grains takes place during the falling-rate drying period. Several mechanisms have been proposed to describe the migration of moisture in capillary porous media. Brooker *et al.* (1974) list several possible mechanisms of moisture movement.

Only liquid and vapor diffusion, or a combination of both, have proven to describe the drying process of cereal grains, Becker and Sallans (1955), Brooker *et al.* (1974), Newman (1931) and Sherwood (1931).

Drying studies of commercial grain dryers are usually performed by using single-layer (or fully-exposed) drying equations also known as thin-layer drying equations. In a recent publication, Sharaf-Eldeen *et al.* (1979) presents a review of all the mathematical models proposed to describe the falling rate drying of fully exposed biological materials. These models are classified in the following groups:

- (1) semitheoretical and empirical models,
- (2) diffusion models, and
- (3) simultaneous heat and moisture transfer models.

Empirical and semitheoretical models have been extensively used for their simplicity of computations.

The models of the second group are based on Fick's second law of diffusion. The governing equation is a second order partial differential equation.

The solutions of diffusion models are carried out by assuming isothermal conditions and neglecting the heat transfer equations as compared with the mass transfer equations. Young (1969) has shown that the heat transfer equations can be neglected if the Lewis number is greater than 60. The Lewis number is expressed as the ratio between thermal and mass diffusivities. Steffe (1979) has shown that for rice, the Lewis number is equal to 7500 which is two orders of magnitude greater than the limit of 60.

Simultaneous heat and mass transfer models have been presented by several investigators, Husain *et al.* (1972, 1973) Brooker *et al.* (1974). Thin-layer models are more rational and accurate in describing the entire drying period of agricultural products, but in general they require sophisticated and expensive computer methods of solutions. They also involve many transport coefficients.

After reviewing the different models proposed to describe the drying of porous materials, Whitaker (1977a) states, "In all of the previous theoretical studies of drying, the governing differential equations... were inferred in a purely intuitive manner from the well-known point equations of continuum physics." In this publication he presents a rigorous theory of drying in porous media based on the transport equations for a continuous media. The porous material is assumed to be composed of three distinctive phases: a rigid solid phase, a liquid phase made up of pure water and a gas phase composed of vapor and an inert component that is insoluble in the other two phases. The point equations are averaged over a localized "small" volume and the governing "volume averaged" equations for a drying process in a porous media are obtained. In the present study, we use the same rigorous approach and derive the general transport equations that describe an assumed liquid diffusion controlled drying of rough rice in a batch operated, well-mixed spouted-bed dryer.

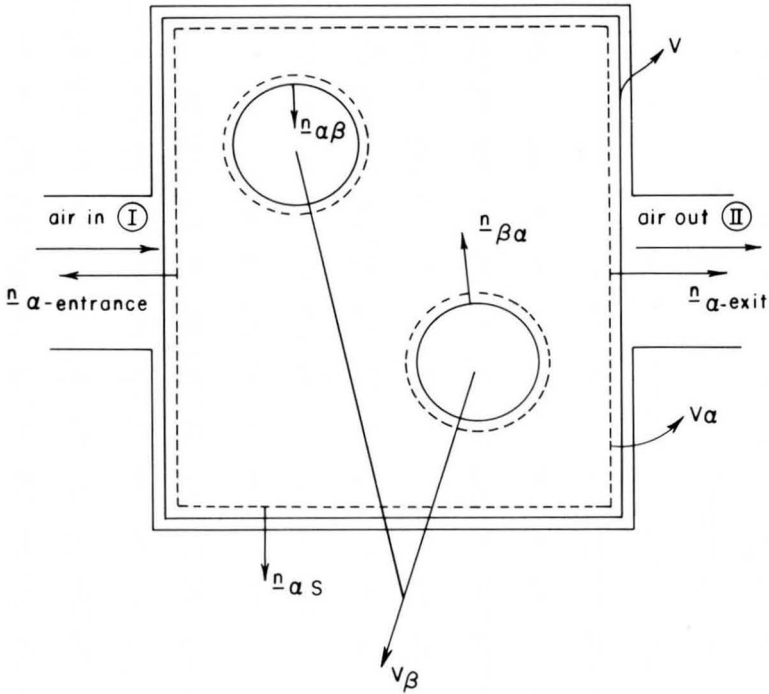


FIG. 1 BATCH OPERATED, WELL MIXED, SPOUTED-BED RICE DRYER

THEORETICAL DEVELOPMENT

Consider a well-mixed batch operated system as shown in Fig. 1.

Use a control volume v fixed in space composed by the whole bed and expressed as

$$v = V_{\alpha} + V_{\beta} \tag{1}$$

where both the volume of the gas-phase (V_{α}) and the volume of the grain-phase (V_{β}) are independent of time. The volume, V_{β} , comprises all grain phase (or rice). It is assumed that:

(a) The volume of every individual grain of rice remains constant throughout the drying process, thus any shrinkage of grain that could

occur is neglected.

(b) Although the volume of the drying-air gas-phase will penetrate within the assumed grain model, it is neglected and assumed that the volume of the α -phase remains constant.

(c) The drying-air leaves the system in thermal equilibrium with the grains of rice, this means that the temperature is not only uniform through the bed, but it is equal in both the α and β phases.

First, the governing point equations given elsewhere, (Bird *et al.* 1960) as:

Continuity Equation

$$\frac{\partial \rho}{\partial t} + \nabla \cdot (\rho \underline{v}) = 0 \quad (2)$$

Continuity Equation for i th species,

$$\frac{\partial \rho_i}{\partial t} + \nabla \cdot (\rho_i \underline{v}_i) = r_i \quad (3)$$

It is assumed that there is no chemical reaction ($r_i = 0$)

Thermal Energy Equation

$$\rho \frac{Dh}{Dt} = -\nabla \cdot \underline{q} + \frac{Dp}{Dt} + \nabla \underline{v} : \underline{\tau} + \bar{\phi} \quad (4)$$

Simplified Thermal Energy Equation for i th specie, Whitaker (1977a)

$$\frac{\partial}{\partial t} \left(\sum_{i=1}^{i=n} \rho_i \bar{h}_i \right)_{\phi} + \nabla \cdot \left(\sum_{i=1}^{i=n} \rho_i \underline{v}_i \bar{h}_i \right)_{\phi} = -\nabla \cdot \underline{q}_{\phi} + \bar{\phi}_{\phi} \quad (5)$$

The three definitions of averages that are used in the averaging process, Whitaker (1977a), of some function ψ defined every where in space can be written as

Spatial Average

$$\langle \psi \rangle = \frac{1}{V} \int_V \psi dV \quad (6)$$

Phase Average: in the α -phase for an α, β -phase system

$$\langle \psi_\alpha \rangle = \frac{1}{V} \int_V \psi_\alpha \, dV = \frac{1}{V} \left\{ \int_{V_\alpha} \psi_\alpha \, dV + \int_{V_\beta} \psi_\alpha \, dV \right\} \quad (7)$$

If we define $\psi_\alpha = 0$ in all other phases except in the α -phase, Eq. (7) reduces to

$$\langle \psi_\alpha \rangle = \frac{1}{V} \int_{V_\alpha} \psi_\alpha \, dV \quad (8)$$

Intrinsic Phase Average

$$\langle \psi_\alpha^\alpha \rangle = \frac{1}{V_\alpha} \int_{V_\alpha} \psi_\alpha \, dV = \frac{1}{V_\alpha} \left\{ \int_{V_\alpha} \psi_\alpha \, dV + \int_{V_\beta} \psi_\alpha \, dV \right\} \quad (9)$$

Applying the same restriction as in Eq. (9) is represented as

$$\langle \psi_\alpha^\alpha \rangle = \frac{1}{V_\alpha} \int_{V_\alpha} \psi_\alpha \, dV \quad (10)$$

The “averaging theorem” (Whitaker 1977a) used in the derivation of the volume averaged equations, which for a two phase system can be written as:

$$\langle \underline{\nabla} \psi_\alpha \rangle = \underline{\nabla} \langle \psi_\alpha \rangle + \frac{1}{V} \int_{A_{\alpha\beta}} \psi_\alpha \frac{n}{\alpha\beta} \, dA \quad (11)$$

In the following the same volume averaging process is used as given by Whitaker (1977a); within one grain of rice in order to obtain an intermediate form of the volume averaged equations that will represent the model. This averaging procedure is made to satisfy the following conditions (Gray 1975)

$$d \ll \ell \ll L$$

where

d is the distance over which the property under consideration varies significantly

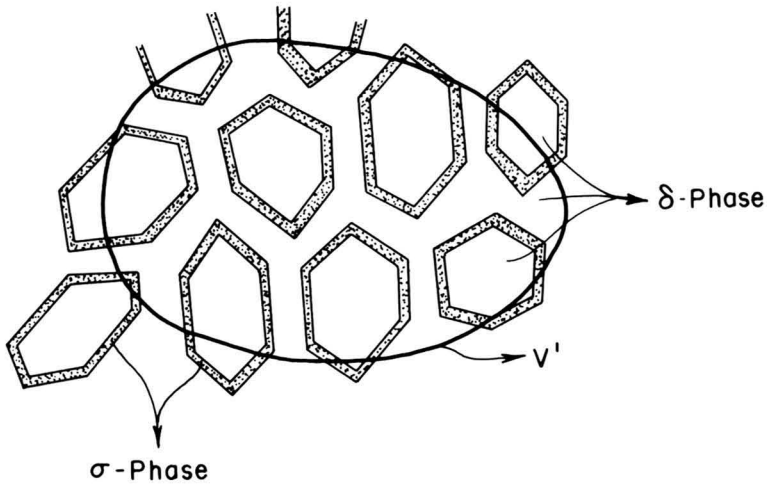


FIG. 2. MICROSCOPIC REPRESENTATION OF A GRAIN OF RICE $V' =$ AVERAGING VOLUME

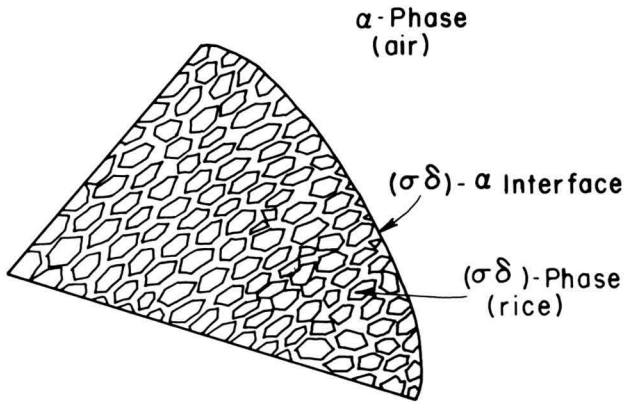


FIG. 3. PORTION OF AN INDIVIDUAL GRAIN OF RICE AND GRAIN-AIR INTERFACE

ϱ is the characteristic length of the averaging volume and
 L is the characteristic length of the global system

Grain-phase (β -phase)

Consider the β -phase as composed of rigid solid dry matter (i.e., insoluble solids, cell walls) and a liquid multiphase solution (bi-phase

composed by water and soluble solids). The rigid dry matter phase is designated as σ -phase and the liquid-phase as δ -phase.

Figure 2 shows a microscopic representation of a portion of rice grain with both the σ and δ phases and the averaging volume over which is performed the averaging process to obtain the intermediate form of the volume averaged equations. Figure 3 is a representation of a portion of the grain and the interface grain-air.

In this model it is assumed that:

(1) The intercellular void space (gas) is negligible as compared to the volume of the liquid phase, thus the intercellular spaces can be considered as completely filled with the liquid phase.

(2) The change with time of the volume of the liquid phase as the drying proceeds will be assumed small relative to its total volume within the range of moisture removal considered. In addition, it is assumed that the interface liquid-drying air remains fixed at the grain surface, thus neglecting the gas-phase formation of the interface as the water evaporates.

(3) The water migrates by the mechanisms of liquid diffusion inside the grain towards the surface where evaporation takes place.

(4) The grain of rice is considered a sphere.

(5) The water migration path is in the radial direction through the cell walls and the intercellular spaces.

(6) Temperature gradients within the grain are negligible as compared with moisture gradient (Lewis number much larger than 100), thus the grain temperature is uniform and a function of time only.

Intermediate Volume Averaged Equations. σ -phase (solid). Consider the σ -phase as a rigid matrix fixed in an inertial frame so the velocity of this phase is zero; then Eq. (2) is of no consequence and Eq. (4) is simplified to:

$$\rho_{\sigma} \left(\frac{\partial h_{\sigma}}{\partial t} \right) = - \underline{\nabla} \cdot \underline{q}_{\sigma} + \bar{\phi}_{\sigma} \quad (12)$$

In Eq. (12) both compressible work and viscous dissipation are neglected and the material derivative is expanded as

$$\frac{Dh_{\sigma}}{Dt} = \frac{\partial h_{\sigma}}{\partial t} + \underline{v}_{\sigma} \cdot \underline{\nabla} h_{\sigma} = \frac{\partial h_{\sigma}}{\partial t}$$

Notice that the density ρ_σ is constant, thus Eq. (12) can be written as

$$\frac{\partial}{\partial t} (\rho_\sigma h_\sigma) = -\nabla \cdot \mathbf{q}_\sigma + \bar{\phi}_\sigma \quad (13)$$

δ -phase (liquid). For the liquid phase Eq. (2) takes the form:

$$\frac{\partial \rho_\delta}{\partial t} + \nabla \cdot (\rho_\delta \mathbf{v}_\delta) = 0 \quad (14)$$

Eq. (3) becomes:

$$\frac{\partial (\rho_i)_\delta}{\partial t} + \nabla \cdot (\rho_i \mathbf{v}_i)_\delta = 0 \quad i=1,2 \quad (15)$$

and Eq. (5) is:

$$\frac{\partial}{\partial t} \left(\sum_{i=1}^n \rho_i \bar{h}_i \right) + \nabla \cdot \left(\sum_{i=1}^n \rho_i \mathbf{v}_i \bar{h}_i \right) = -\nabla \cdot \mathbf{q}_\delta + \bar{\phi}_\delta \quad (16)$$

using the definition:

$$\rho_\delta h_\delta = \sum_{i=1}^n \rho_i \bar{h}_i \quad (17)$$

and expressing the i th species velocity \mathbf{v}_i in terms of the mass average \mathbf{v}_δ and the diffusion velocity \mathbf{u}_i (Whitaker 1977a) as:

$$\mathbf{v}_i = \mathbf{v}_\delta + \mathbf{u}_i \quad (18)$$

Eq. (16) becomes:

$$\frac{\partial}{\partial t} (\rho_\delta h_\delta) + \nabla \cdot (\rho_\delta \mathbf{v}_\delta h_\delta) = -\nabla \cdot \mathbf{q}_\delta - \nabla \cdot \left(\sum_{i=1}^n \rho_i \bar{h}_i \mathbf{u}_i \right) + \bar{\phi}_\delta \quad (19)$$

where ρ_δ and \mathbf{v}_δ are given as:

$$\rho_\delta = \rho_1 + \rho_2 \quad (20)$$

$$\rho_\delta \mathbf{v}_\delta = \rho_1 \mathbf{v}_1 + \rho_2 \mathbf{v}_2 \quad (21)$$

where

ρ_1 = density of liquid water

ρ_2 = density of soluble solids

$$(22)$$

Now that we have the governing point equations for both the solid (σ) phase of the grain of rice we direct our attention to the boundary conditions that will apply on the control volume of Fig. 2.

Boundary Conditions. In Fig. 2 identify the following area within the limits of the control volume ν'

$$A_{\sigma\delta} = \text{solid-liquid interfacial area} \quad (23)$$

and note that

$$A_{\sigma\delta} = A_{\delta\sigma} \quad (24)$$

The boundary conditions at the solid-liquid interface are given as:

$$\text{BC 1:} \quad \mathbf{v}_\delta = \mathbf{u}_i \cdot \mathbf{n}_{\sigma\delta} = 0 \quad \text{at } A_{\sigma\delta} \quad (25)$$

$$\text{BC 2:} \quad \mathbf{q}_\sigma \cdot \mathbf{n}_{\sigma\delta} + \mathbf{q}_\delta \cdot \mathbf{n}_{\delta\sigma} = 0 \quad \text{at } A_{\sigma\delta} \quad (26)$$

$$\text{BC 3:} \quad T_\sigma = T_\delta \quad \text{on } A_{\sigma\delta} \quad (27)$$

where

$$\mathbf{n}_{\sigma\delta} = -\mathbf{n}_{\delta\sigma} \quad \text{on } A_{\sigma\delta} \quad (28)$$

In BC 1 it is required that the normal component of the diffusion velocity vector ($\mathbf{u}_i \cdot \mathbf{n}_{\sigma\delta}$) be zero at the fixed, impermeable surface while the tangential component ($\mathbf{u}_i \cdot \mathbf{\lambda}_{\sigma\delta}$) is non zero (Whitaker 1967).

Averaging Process. The analysis begins with the σ -phase by integrating the thermal energy Eq. (13) over V_σ and dividing the ν'' to obtain

$$\frac{1}{\nu'} \int_{V_\sigma} \frac{\partial}{\partial t} (\rho_\sigma h_\sigma) dV = -\frac{1}{\nu'} \int_{V_\sigma} \underline{\nabla} \cdot \mathbf{q}_\sigma dV + \frac{1}{\nu'} \int_{V_\sigma} \bar{\phi}_\sigma dV \quad (29)$$

The differentiation and integration can be interchanged in the first term of Eq. (29) using the Leibnitz formula for differentiating a triple integral, because $\underline{\nabla}_\sigma = 0$. Making use of the definition of phase average (Eq. 8):

$$\frac{\partial}{\partial t} \langle \rho_\sigma h_\sigma \rangle = -\langle \underline{\nabla} \cdot \mathbf{q}_\sigma \rangle + \langle \bar{\phi}_\sigma \rangle \quad (30)$$

Using the averaging theorem (Eq. 11) to represent $\langle \underline{\nabla} \cdot \mathbf{q}_\sigma \rangle$ as

$$\langle \underline{\nabla} \cdot \underline{q}_\sigma \rangle = \underline{\nabla} \cdot \langle \underline{q}_\sigma \rangle + \frac{1}{\nu'} \int_{A_{\sigma\delta}} \underline{q}_\sigma \cdot \underline{n}_{\sigma\delta} \, dA \quad (31)$$

Thus, Eq. (30) becomes

$$\frac{\partial}{\partial t} \langle \rho_\sigma h_\sigma \rangle = - \underline{\nabla} \cdot \langle \underline{q}_\sigma \rangle - \frac{1}{\nu'} \int_{A_{\sigma\delta}} \underline{q}_\sigma \cdot \underline{n}_{\sigma\delta} \, dA + \langle \underline{\phi}_\sigma \rangle \quad (32)$$

Equation (32) is the intermediate volume average equation for the solid σ -phase.

Now consider the liquid-phase (δ) and integrate the continuity Eq. (14) over V_δ and divide by ν' , following the same steps as above and imposing BC 1 the intermediate averaged continuity equation for the δ -phase as:

$$\frac{\partial}{\partial t} \langle \rho_\delta \rangle + \underline{\nabla} \cdot \langle \rho_\delta \underline{v}_\delta \rangle = 0 \quad (33)$$

Applying an identical procedure to Eq. (15) the intermediate averaged continuity equation for the i th species in the δ -phase can be obtained as

$$\frac{\partial}{\partial t} \langle \rho_i \rangle_\ell + \underline{\nabla} \cdot \langle \rho_i \underline{v}_i \rangle_\ell = 0 \quad (34)$$

In Eq. (34) the area integral is eliminated by making use of Eq. (18) along with BC 1.

The thermal energy equation for the i th species given as Eq. (16) upon integration over V_δ and dividing by ν' and performing the same steps as before gives:

$$\frac{\partial}{\partial t} \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i \right\rangle_\ell + \underline{\nabla} \cdot \left\langle \sum_{i=1}^{i=n} \rho_i \underline{v}_i \bar{h}_i \right\rangle_\ell + \frac{1}{\nu'} \int_{A_{\delta\sigma}} \quad (35)$$

$$\left(\sum_{i=1}^{i=n} \rho_i \bar{h}_i \underline{v}_i \right)_\ell \cdot \underline{n}_{\delta\sigma} \, dA = - \underline{\nabla} \cdot \langle \underline{q}_\delta \rangle - \frac{1}{\nu'} \int_{A_{\delta\sigma}} \underline{q}_\delta \cdot \underline{n}_{\delta\sigma} \, dA + \langle \underline{\phi}_\delta \rangle$$

It is recognized that the area integral on the left hand side of Eq. (35) is equal to zero. Thus, the intermediate volume averaged energy equation for the i th species and the δ -phase is:

$$\begin{aligned} & \frac{\partial}{\partial t} \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i \right\rangle + \nabla \cdot \left\langle \sum_{i=1}^{i=n} \rho_i \mathbf{v}_i \bar{h}_i \right\rangle \\ &= - \nabla \cdot \langle \mathbf{q}_\delta \rangle - \frac{1}{\nu'} \int_{A_{\delta\sigma}} \mathbf{q}_\delta \cdot \mathbf{n}_{\delta\sigma} \, dA + \langle \Phi_\delta \rangle \end{aligned} \quad (36)$$

The next step is to obtain the total intermediate averaged equations for the grain which can be done by adding the equations of both phases and imposing the assumption of "local equilibrium."

The total intermediate averaged continuity equation for the grain is given by Eq. (34) for the i th species while the total intermediate averaged thermal energy equation for the grain is obtained by adding Eq. (32) and (36). Making use of Eq. (24) and BC 2 (Eq. 26) and after grouping terms

$$\begin{aligned} & \frac{\partial}{\partial t} \left\{ \langle \rho_\sigma h_\sigma \rangle + \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i \right\rangle_\ell \right\} + \nabla \cdot \left\langle \sum_{i=1}^{i=n} \rho_i \mathbf{v}_i \bar{h}_i \right\rangle_\ell \\ &= - \nabla \cdot \left\{ \langle \mathbf{q}_\sigma \rangle + \langle \mathbf{q}_\delta \rangle \right\} + \langle \Phi_\sigma \rangle + \langle \Phi_\delta \rangle \end{aligned} \quad (37)$$

Equations (34) and (37) provide the necessary set of total intermediate averaged equations to perform the macroscopic mass and energy balances over all the grains of rice in the bed as shown in Fig. 1, with grain-air interface shown in Fig. 3.

Macroscopic Mass and Energy Balances. In Figure 3 we can identify two phases, one the surrounding drying-air or α -phase and the other the grain or $\sigma\delta$ -phase. The boundary conditions that apply at the $\sigma\delta$ - α interface in terms of mass diffusion fluxes for the i th species on both the α and $\sigma\delta$ phases and in terms of energy fluxes are listed below. The complete derivation of these boundary conditions have been done by Zuritz (1980), according to Whitaker's approach (1977 a,b). The i th species continuity equation boundary conditions are:

$$\text{BC 4:} \quad \langle \rho_i \mathbf{v}_i \rangle_\ell \cdot \mathbf{n}_{(\sigma\delta)\alpha} + (\rho_i \mathbf{v}_i)_v \cdot \mathbf{n}_{\alpha(\sigma\delta)} = 0 \quad i=1 \quad (38)$$

$$\text{BC 5: } \langle \rho_i \mathbf{v}_i \rangle_\ell \cdot \mathbf{n}_{(\sigma\delta)\alpha} + (\rho_i \mathbf{v}_i)_\nu \cdot \mathbf{n}_{\alpha(\sigma\delta)} = 0 \quad i=2 \quad (39)$$

$$\mathbf{n}_{(\sigma\delta)\alpha} = -\mathbf{n}_{\alpha(\sigma\delta)} \quad (40)$$

where the component $i = 1$ is the liquid water in the $\sigma\delta$ -phase and water vapor in the α -phase.

For the thermal energy boundary condition:

$$\begin{aligned} \text{BC 6: } & \left[\langle \mathbf{q}_\sigma \rangle + \langle \mathbf{q}_\delta \rangle + \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i \mathbf{v}_i \right\rangle_\ell \right] \cdot \mathbf{n}_{(\sigma\delta)\alpha} \\ & + \left[\mathbf{q}_\alpha + \sum_{i=1}^{i=n} (\rho_i \bar{h}_i \mathbf{v}_i)_\nu \right] \cdot \mathbf{n}_{\alpha(\sigma\delta)} = 0 \end{aligned} \quad (41)$$

Now that we have derived the necessary boundary conditions we start with the continuity equation for the i th species given as Eq. (34) in integrate over V_β and divide by ν to obtain

$$\frac{1}{\nu} \int_{V_\beta} \frac{\partial}{\partial t} \langle \rho_i \rangle_\ell dV + \frac{1}{\nu} \int_{V_\beta} \nabla \cdot \langle \rho_i \mathbf{v}_i \rangle_\ell dV = 0 \quad (42)$$

Interchanging differentiation and integration in the first term of Eq. (42) and applying the definition of volume average

$$\frac{1}{\nu} \int_{V_\beta} \frac{\partial}{\partial t} \langle \rho_i \rangle_\ell dV = \frac{\partial}{\partial t} \left[\frac{1}{\nu} \int_{V_\beta} \langle \rho_i \rangle_\ell dV \right] = \frac{\partial}{\partial t} \langle\langle \rho_i \rangle_\ell \rangle \quad (43)$$

In Eq. (43) there is an average of an average which is equal to the average (Whitaker 1968)

$$\frac{1}{\nu} \int_{V_\beta} \frac{\partial}{\partial t} \langle \rho_i \rangle_\ell dV = \frac{\partial}{\partial t} \langle \rho_i \rangle_\ell \quad (44)$$

Applying divergence theorem to the second term of Eq. (42) yields

$$\begin{aligned} \frac{1}{\nu} \int_{V_{\beta}} \nabla \cdot \langle \rho_i \mathbf{v}_i \rangle_{\ell} dV &= \frac{1}{\nu} \int_{A_{\beta\alpha}} \langle \rho_i \mathbf{v}_i \rangle_{\ell} \cdot \mathbf{n}_{\beta\alpha} dA \\ &+ \frac{1}{\nu} \int_{A_{\beta s}} \langle \rho_i \mathbf{v}_i \rangle_{\ell} \cdot \mathbf{n}_{\beta s} dA \end{aligned} \quad (45)$$

In Eq. (45) there are two different areas: the grain-air interface ($A_{\beta\alpha}$) and the grain-dryer walls interface ($A_{\beta s}$). At this last interface there is no mass flux taking place so the second area intergral is zero; furthermore, since the dryer walls are insulated it is assumed that there is no net heat exchange between the walls and the grains, therefore the remaining boundary conditions can be expressed as:

$$\text{BC 7:} \quad \langle \rho_i \mathbf{v}_i \rangle_{\ell} \cdot \mathbf{n}_{\beta s} = 0 \quad i=1,2 \quad \text{at } A_{\beta s} \quad (46)$$

$$\text{BC 8:} \quad \mathbf{q}_{\beta} \cdot \mathbf{n} = \mathbf{q}_s \cdot \mathbf{n}_{s\beta} = 0 \quad \text{at } A_{\beta s} \quad (47)$$

Before continuing it is convenient to notice that in Eq. (45) the interfacial area $A_{\beta\alpha}$ is used while in boundary conditions 4 to 7 $A_{(\sigma\delta)\alpha}$ is used and both areas represent the same interface thus relating each other by using the following expression:

$$A_{\beta\alpha} = A_{\alpha\beta} = N \times A_{(\sigma\delta)\alpha} \quad (48)$$

where N is the total number of grains in the bed.

Introducing Eq. (44) to (46) into (42)

$$\frac{\partial}{\partial t} \langle \rho_i \rangle_{\ell} + \frac{1}{\nu} \int_{A_{\beta\alpha}} \langle \rho_i \mathbf{v}_i \rangle_{\ell} \cdot \mathbf{n}_{\beta\alpha} dA = 0 \quad (49)$$

Defining the volume average density in terms of the intrinsic phase average for the bed as:

$$\langle \rho_i \rangle = \epsilon_{\beta} \langle \rho_i \rangle_{\ell}^{\beta} \quad i=1,2 \quad (50)$$

where

$$\langle \rho_1 \rangle_{\ell}^{\beta} \quad \text{water intrinsic phase average} \quad \frac{\text{g water}}{\text{m}^3 \text{ grains}} \quad (51)$$

and

$$\langle \rho_2 \rangle_{\ell}^{\beta} \text{ soluble solids intrinsic phase average } \frac{\text{g s.s}}{\text{m}^3 \text{ grains}} \quad (52)$$

and using

$$\epsilon_{\beta} (\rho_{\sigma}^{\beta} + \langle \rho_2 \rangle_{\ell}^{\beta}) \equiv \frac{\text{g dry matter}}{\text{m}^3 \text{ of bed}} \quad (53)$$

along with Eq. (50) to (52) Eq. (49) can be modified as

$$\begin{aligned} & \epsilon_{\beta} (\rho_{\sigma}^{\beta} + \langle \rho_2 \rangle_{\ell}^{\beta}) \frac{\partial}{\partial t} \left[\frac{\langle \rho_1 \rangle_{\ell}^{\beta}}{(\rho_{\sigma}^{\beta} + \langle \rho_2 \rangle_{\ell}^{\beta})} \right] \\ & + \frac{1}{\nu} \int_{A_{\beta\alpha}} \langle \rho_i \mathbf{v}_i \rangle_{\ell} \cdot \mathbf{n}_{\beta\alpha} \, dA = 0 \quad i=1 \end{aligned} \quad (54)$$

Define the mass rate of evaporation per unit volume of bed as:

$$\langle \dot{m} \rangle = \frac{1}{\nu} \int_{A_{\beta\alpha}} \langle \rho_i \mathbf{v}_i \rangle_{\ell} \cdot \mathbf{n}_{\beta\alpha} \, dA \quad i=1 \quad (55)$$

and from boundary condition 4 and Eq. (48), Eq. (55) can also be expressed as:

$$\langle \dot{m} \rangle = - \frac{1}{\nu} \int_{A_{\alpha\beta}} (\rho_i \mathbf{v}_i)_{\nu} \cdot \mathbf{n}_{\alpha\beta} \, dA \quad i=1 \quad (56)$$

It will be worthwhile to rewrite Eq. (49) as:

$$\frac{\partial}{\partial t} \langle \rho_i \rangle_{\ell} + \langle \dot{m} \rangle = 0 \quad (57)$$

The thermal energy equation given as Eq. (37) and integrated over V_{β} and divided by ν and following the above steps:

$$\begin{aligned} & \frac{1}{\nu} \int \frac{\partial}{\partial t} \left\{ \langle \rho_{\sigma} h_{\sigma} \rangle + \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i \right\rangle_{\ell} \right\} dV + \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i v_i \right\rangle \cdot \underline{n}_{\beta\alpha} dA \\ & = -\frac{1}{\nu} \int_{A_{\beta\alpha}} \left\{ \langle \underline{q} \rangle + \langle \underline{q}_{\delta} \rangle \right\} \cdot \underline{n}_{\beta\alpha} dA + \frac{1}{\nu} \int_{V_{\beta}} \left\{ \langle \bar{\Phi}_{\sigma} \rangle + \langle \bar{\Phi}_{\delta} \rangle \right\} dV \end{aligned} \quad (58)$$

Before solving Eq. (58) it is necessary to define some quantities (Whitaker 1977a).

Treating the enthalpy as independent of pressure express it as

$$h = h(T) \text{ only,} \quad \text{in all the phases} \quad (59)$$

also consider the heat capacities as constant so Eq. (59) takes the form of

$$h = C_p T + \text{constant} \quad \text{in all the phases} \quad (60)$$

therefore:

$$h_{\sigma} = (C_p)_{\sigma} T_{\sigma} + \text{constant} \quad (61)$$

Defining the phase and intrinsic phase averages of T_{σ} as:

$$\langle T_{\sigma} \rangle = \epsilon_{\beta} \langle T_{\sigma} \rangle^{\beta} \quad (62)$$

Express the partial mass enthalpy \bar{h}_i as

$$\bar{h}_i = h_i^{\circ} + (C_p)_i (T_{\delta} - T_{\delta}^{\circ}) \quad (63)$$

where h_i° and $(C_p)_i$ and T_{δ}° are constant quantities; and use the definitions of:

$$T_{\delta} = \langle T_{\delta} \rangle^{\beta} + \tilde{T}_{\delta} \quad \text{in the } \beta\text{-phase} \quad (64)$$

$$T_{\delta} = \tilde{T}_{\delta} = 0 \quad \text{in the } \alpha\text{-phase} \quad (65)$$

Starting with the first term in Eq. (58) and separating it into the two components and using Eq. (61) and (62) since ρ_σ is constant we obtain:

$$\frac{\partial}{\partial t} \langle \rho_\sigma h_\sigma \rangle = \epsilon_\beta \rho_\sigma^\beta (C_p)_\sigma \frac{\partial \langle T_\sigma \rangle^\beta}{\partial t} \quad (66)$$

and applying Eq. (63) to (65) to the second component of the first term of Eq. (58) which after slight rearrangement becomes:

$$\begin{aligned} \frac{\partial}{\partial t} \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i \right\rangle_\ell &= \sum_{i=1}^{i=n} \left[h_i^o - (C_p)_i T_\delta^o \right]_\ell \frac{\partial \langle \rho_i \rangle_\ell}{\partial t} \\ &+ \sum_{i=1}^{i=n} (C_p)_i \frac{\partial}{\partial t} \langle \rho_i T_\delta \rangle_\ell \end{aligned} \quad (67)$$

Introducing Eq. (57) into (67) and noticing that

$$\langle \rho_i T_\delta \rangle_\ell = \langle \rho_i \rangle_\ell \langle T_\delta \rangle^\beta + \langle \tilde{\rho}_i \tilde{T}_\delta \rangle_\ell \quad (68)$$

$$\begin{aligned} \frac{\partial}{\partial t} \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i \right\rangle_\ell &= - \sum_{i=1}^{i=n} \left[h_i^o + (C_p)_i (\langle T_\delta \rangle^\beta - T_\delta^o) \right]_\ell \langle \dot{m} \rangle \\ &+ \sum_{i=1}^{i=n} \langle \rho_i \rangle_\ell (C_p)_i \frac{\partial}{\partial t} \langle T_\delta \rangle^\beta + \sum_{i=1}^{i=n} (C_p)_i \frac{\partial}{\partial t} \langle \tilde{\rho}_i \tilde{T}_\delta \rangle_\ell \end{aligned} \quad (69)$$

The second term of Eq. (58) using Eq. (63) to (65) and the following relationship is obtained:

$$\langle \rho_i \underline{v}_i \langle T_\delta \rangle^\beta \rangle_\ell = \langle \rho_i \underline{v}_i \rangle_\ell \langle T_\delta \rangle^\beta \quad (70)$$

we obtain

$$\begin{aligned}
 & \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i \underline{v}_i \right\rangle_{\ell} \cdot \underline{n}_{\beta\alpha} \, dA \\
 = & \frac{1}{\nu} \int_{A_{\beta\alpha}} \sum_{i=1}^{i=n} \left[h_i^{\circ} + (C_p)_i (\langle T_{\delta} \rangle^{\beta} - T_{\delta}^{\circ}) \right] \langle \rho_i \underline{v}_i \rangle_{\ell} \cdot \underline{n}_{\beta\alpha} \, dA \\
 & + \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\langle \sum_{i=1}^{i=n} \rho_i (C_p)_i \tilde{T}_{\delta} \underline{v}_i \right\rangle_{\ell} \cdot \underline{n}_{\beta\alpha} \, dA \quad (71)
 \end{aligned}$$

Notice that the enthalpy in the second term of Eq. (71) is expressed in terms of the intrinsic phase average temperature $\langle T_{\delta} \rangle^{\beta}$ which will be constant over the area $A_{\beta\alpha}$. This allows one to move the enthalpy term outside the area integral and express the area integral according with Eq. (55), then Eq. (71) can be written as:

$$\begin{aligned}
 \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i \underline{v}_i \right\rangle_{\ell} \cdot \underline{n}_{\beta\alpha} \, dA &= \sum_{i=1}^{i=n} \left[h_i^{\circ} + (C_p)_i (\langle T_{\delta} \rangle^{\beta} - T_{\delta}^{\circ}) \right]_{\ell} \\
 \langle \dot{m} \rangle + \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\langle \sum_{i=1}^{i=n} \rho_i (C_p)_i \tilde{T}_{\delta} \underline{v}_i \right\rangle_{\ell} \cdot \underline{n}_{\beta\alpha} \, dA & \quad (72)
 \end{aligned}$$

Treat the third term of Eq. (58) and after imposing boundary condition 6

$$\begin{aligned}
 \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\{ \langle \underline{q}_{\sigma} \rangle + \langle \underline{q}_{\delta} \rangle \right\} \cdot \underline{n}_{\beta\alpha} \, dA &= -\frac{1}{\nu} \int_{A_{\beta\alpha}} \underline{q}_{\alpha} \cdot \underline{n}_{\alpha\beta} \, dA \\
 -\frac{1}{\nu} \int_{A_{\beta\alpha}} \left\langle \sum_{i=1}^{i=n} \rho_i \bar{h}_i \underline{v}_i \right\rangle_{\ell} \cdot \underline{n}_{\beta\alpha} \, dA &- \frac{1}{\nu} \int_{A_{\beta\alpha}} \left(\sum_{i=1}^{i=n} \rho_i \bar{h}_i \underline{v}_i \right)_v \cdot \underline{n}_{\alpha\beta} \, dA \quad (73)
 \end{aligned}$$

In Eq. (73) we have two enthalpy terms at the interface $A_{\beta\alpha}$, one for the liquid and the other for the vapor. Using the definitions given by Eq. (63) to (65) for the enthalpy terms and considering:

$$\langle T_\delta \rangle^\beta = \langle T_\alpha \rangle^\alpha = \langle T \rangle \quad \text{at the interface } A_{\beta\alpha} \quad (74)$$

(well mixed model)

Eq. (73) (with $i = 1$) becomes:

$$\begin{aligned} \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\{ \langle \mathbf{q}_\sigma \rangle + \langle \mathbf{q}_\delta \rangle \right\} \cdot \mathbf{n}_{\beta\alpha} \, dA &= -\frac{1}{\nu} \int_{A_{\beta\alpha}} \mathbf{q}_\alpha \cdot \mathbf{n}_{\alpha\beta} \, dA \\ &- \left[h_1^\circ + (C_p)_1 (\langle T \rangle - T_\delta^\circ) \right]_\ell \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\{ \langle \rho_i \mathbf{v}_i \rangle_\ell \cdot \mathbf{n}_{\beta\alpha} \, dA \right\} \\ &- \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\langle \sum_{i=1}^{i=n} \rho_i (C_p)_i \tilde{T}_\delta \mathbf{v}_i \right\rangle_\ell \cdot \mathbf{n}_{\beta\alpha} \, dA \\ &- \left[h_1^\circ + (C_p)_1 (\langle T \rangle - T_\alpha^\circ) \right]_\nu \left\{ \frac{1}{\nu} \int_{A_{\beta\alpha}} (\rho_i \mathbf{v}_i)_\nu \cdot \mathbf{n}_{\alpha\beta} \, dA \right\} \\ &- \frac{1}{\nu} \int_{A_{\beta\alpha}} \left(\sum_{i=1}^{i=n} \rho_i (C_p)_i \tilde{T}_\alpha \mathbf{v}_i \right)_\nu \cdot \mathbf{n}_{\alpha\beta} \, dA \quad (75) \end{aligned}$$

Use Eq. (55) and (56), the definition of

$$\left[h_{1\nu}^\circ - h_{1\ell}^\circ + (C_p)_{1\nu} (\langle T \rangle - T^\circ) - (C_p)_{1\ell} (\langle T \rangle - T^\circ) \right] = \Delta H_{\text{evap}} \quad (76)$$

to express Eq. (74) as

$$\begin{aligned} \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\{ \langle \mathbf{q}_\sigma \rangle + \langle \mathbf{q}_\delta \rangle \right\} \cdot \mathbf{n}_{\beta\alpha} \, dA &= -\frac{1}{\nu} \int_{A_{\beta\alpha}} \mathbf{q}_\alpha \cdot \mathbf{n}_{\alpha\beta} \, dA + \Delta H_{\text{evap}} \langle \dot{m} \rangle \\ &- \frac{1}{\nu} \int_{A_{\beta\alpha}} \left\langle \sum_{i=1}^{i=n} \rho_i (C_p)_i \tilde{T}_\delta \mathbf{v}_i \right\rangle_\ell \cdot \mathbf{n}_{\beta\alpha} \, dA - \frac{1}{\nu} \int_{A_{\beta\alpha}} \left(\sum_{i=1}^{i=n} \rho_i (C_p)_i \tilde{T}_\alpha \mathbf{v}_i \right)_\nu \cdot \mathbf{n}_{\alpha\beta} \, dA \quad (77) \end{aligned}$$

Now define the heat flux term at the interface $\beta - \alpha$ as:

$$\dot{Q}_{\text{SG}} = \frac{1}{\nu} \int_{A_{\beta\alpha}} \mathbf{q}_\alpha \cdot \mathbf{n}_{\alpha\beta} \, dA + \frac{1}{\nu} \int_{A_{\beta\alpha}} \left(\sum_{i=1}^{i=n} \rho_i (C_p)_i \tilde{T}_\alpha \mathbf{v}_i \right)_\nu \cdot \mathbf{n}_{\alpha\beta} \, dA \quad (78)$$

Finally, treat the last term of Eq. (58) to which we apply the definition of volume average to obtain:

$$\begin{aligned} \frac{1}{\nu} \int_{V_\beta} \left\{ \langle \bar{\Phi}_\sigma \rangle + \langle \bar{\Phi}_\delta \rangle \right\} dV &= \langle \left\{ \langle \bar{\Phi}_\sigma \rangle + \langle \bar{\Phi}_\delta \rangle \right\} \rangle \\ &= \left\{ \langle \bar{\Phi}_\sigma \rangle + \langle \bar{\Phi}_\delta \rangle \right\} \end{aligned} \quad (79)$$

Then using Eq. (66), (69), (72) and (77) to (79) in (58) and noticing that the last term of Eq. (72) cancels with the second right-hand term of Eq. (77) and that the first term of Eq. (69) cancels with the remaining term of Eq. (72), we have

$$\begin{aligned} \epsilon_\beta \rho_\sigma^\beta (C_p)_\sigma \frac{\partial \langle T_\sigma \rangle^\beta}{\partial t} + \sum_{i=1}^{i=n} \langle \rho_i \rangle_\ell (C_p)_i \frac{\partial \langle T_\delta \rangle}{\partial t} \\ + \sum_{i=1}^{i=n} (C_p)_i \frac{\partial}{\partial t} \langle \tilde{\rho}_i \bar{T}_\delta \rangle_\ell = -\Delta H_{\text{evap}} \langle \dot{m} \rangle + \dot{Q}_{\text{SG}} + \left\{ \langle \bar{\Phi}_\sigma \rangle + \langle \bar{\Phi}_\delta \rangle \right\} \end{aligned} \quad (80)$$

Equation (80) can still be transformed by using the definitions of phase and intrinsic phase average:

$$\langle \rho_i \rangle_\ell = \epsilon_\beta \langle \rho_i \rangle_\ell^\beta \quad (81)$$

and imposing the restriction of local thermal equilibrium, uniform grain temperature and well-mixed bed allows us to write:

$$\langle T_\sigma \rangle^\beta = \langle T_\delta \rangle^\beta = \langle T \rangle^\beta \quad (82)$$

and after rearranging

$$\begin{aligned} \epsilon_\beta (\rho_\sigma^\beta + \langle \rho_2 \rangle_\ell^\beta) \left[(C_p)_{dP} + \frac{\langle \rho_1 \rangle_\ell^\beta}{\rho_\sigma^\beta + \langle \rho_2 \rangle_\ell^\beta} (C_p)_{1\ell} \right] \frac{\partial \langle T \rangle^\beta}{\partial t} \\ + \Delta H_{\text{evap}} \langle \dot{m} \rangle = \dot{Q}_{\text{SG}} + \langle \bar{\Phi}_\beta \rangle \end{aligned} \quad (83)$$

where we have defined $\langle \bar{\phi}_\beta \rangle$ as

$$\langle \bar{\phi}_\beta \rangle = \left\{ \langle \bar{\phi}_\delta \rangle + \langle \bar{\psi}_\delta \rangle \right\} - \sum_{i=1}^{i=n} (C_p)_i \frac{\partial}{\partial t} \langle \bar{\rho}_i \bar{T}_\delta \rangle_\ell \quad (84)$$

and used

$$\left[\rho_\sigma^\beta (C_p)_\sigma + \sum_{i=1}^{i=n} \langle \rho_i \rangle_\ell^\beta (C_p)_i \right] = \rho_\sigma^\beta (C_p)_\sigma \quad (85)$$

$$+ \langle \rho_2 \rangle_\ell^\beta (C_p)_{2_\ell} + \langle \rho_1 \rangle_\ell^\beta (C_p)_{1_\ell}$$

and assumed that

$$(C_p)_\sigma \approx (C_p)_{2_\ell} \equiv (C_p)_{dp} \quad (86)$$

Equation (83) is the macroscopic thermal energy balance for the β -phase per unit mass of bed while Eq. (54) with (55) is the macroscopic mass balance.

Drying Air-phase (α -phase)

Governing Point Equations. Consider the drying air to be made up of the water vapor ($i = 1$) and inert gas or dry air ($i = 2$) and we write the counterpart of Eq. (20) and (21) as:

$$\rho_\alpha = \rho_{1_v} + \rho_{2_v} \quad (87)$$

$$\rho_{\alpha v} = \rho_{1_v v} + \rho_{2_v v} \quad (88)$$

where

$$\rho_{1_v} = \text{density of water vapor}$$

$$\rho_{2_v} = \text{density of dry air, constant} \quad (89)$$

The continuity equations for both the α -phase and the i th species in the α -phase are the same as Eq. (14) and (15) with the subscripts α instead of δ and v instead of ℓ , respectively. While the thermal energy equations are equal to Eq. (16) and (19) after changing the subscripts mentioned above.

Macroscopic Mass and Energy Balances. The assumption of well-mixed model allows us to perform the macroscopic balances over the entire α -phase within the bed as we did with the β -phase.

Analyzing Fig. 1 we see that A_α can be divided into the following areas:

$$\begin{aligned} A_{\alpha_e} &= \text{area at entrances and exits} \\ A_{\alpha\beta} &= \text{area at gas-grain interface} \\ A_{\alpha_s} &= \text{area at gas-walls interface} \end{aligned} \quad (90)$$

The boundary conditions at the gas-grain interface has been given as Eq. (38) to (41), and we now write the boundary conditions at the gas-walls interface as

$$\text{BC 9:} \quad \underline{v}_\alpha = \underline{v}_{i_v} = 0 \quad \text{at } A_{\alpha_s} \quad (91)$$

$$\text{BC 10:} \quad \underline{q}_\alpha \cdot \underline{n}_{\alpha_s} = \underline{q}_s \cdot \underline{n}_{s\alpha} = 0 \quad \text{at } A_{\alpha_s} \quad (92)$$

Following the averaging analysis made to the continuity equation for the i th species of the grain liquid phase given by Eq. (42) to (45) using Eq. (49), (56) and (90) to (92) we obtain the macroscopic mass balance for the water vapor in the α -phase as:

$$\begin{aligned} \rho_{2_v} \epsilon_\alpha \frac{\partial \langle \rho_1 / \rho_2 \rangle_v^\alpha}{\partial t} + \frac{\rho_{2_v} \dot{G}}{\nu} \left[\langle \rho_1 / \rho_2 \rangle_{\text{exit}_v}^\alpha \right. \\ \left. - \langle \rho_1 / \rho_2 \rangle_{\text{entr}_v}^\alpha \right] - \langle \dot{m} \rangle = 0 \end{aligned} \quad (93)$$

where

$$\langle \rho_i \rangle_v = \epsilon_\alpha \langle \rho_i \rangle_v^\alpha \quad (94)$$

$$\langle \rho_i \rangle_v^\alpha = \langle \rho_1 \rangle_v^\alpha, \quad i=1 \quad (95a)$$

$$\langle \rho_i \rangle_v^\alpha = \langle \rho_2 \rangle_v^\alpha = \rho_{2_v} = \text{constant}, \quad i=2 \quad (95b)$$

$$\dot{G} = A_{\text{exit}} \langle \underline{v}_1 \rangle_{\text{exit}_v} = A_{\text{entr}} \langle \underline{v}_1 \rangle_{\text{entr}_v} \quad (96)$$

Performing a similar analysis with the thermal energy equation as we did from equation (58) to (83) and using equations (90) to (92) yields the final form of the thermal energy equation for the α -phase in the bed as:

$$\begin{aligned} \epsilon_{\alpha} \rho_{2_v} \left[(C_p)_2 + \langle \rho_1 / \rho_2 \rangle^{\alpha} (C_p)_1 \right]_{\text{bed}_v} \frac{\partial \langle T_{\alpha} \rangle^{\alpha}}{\partial t} + \frac{\rho_{2_v} \dot{G}}{\nu} \left[(C_p)_2 \right. \\ \left. + \langle \rho_1 / \rho_2 \rangle^{\alpha} (C_p)_1 \right]_{\text{entr}_v} \left(\langle T_{\alpha} \rangle^{\alpha}_{\text{exit}} - \langle T_{\alpha} \rangle^{\alpha}_{\text{entr}} \right) = -\dot{Q}_{SG} + \langle \dot{\Phi}_{\alpha} \rangle \end{aligned} \quad (97)$$

Final Form of the Macroscopic Balances

We assume that the terms involving the time rate of change of both mass and temperature of the α -phase within the bed are negligible as compared with the corresponding convective terms. Besides assumption (c) allows us to define:

$$\langle T \rangle^{\beta} = \langle T_{\alpha} \rangle^{\alpha}_{\text{bed}} = \langle T_{\alpha} \rangle^{\alpha}_{\text{exit}} = \langle T \rangle \quad (98)$$

and we also define the following quantities:

$$\langle T_{\alpha} \rangle^{\alpha}_{\text{entr}} = \langle T \rangle_I \quad (99)$$

$$\langle \rho_1 / \rho_2 \rangle^{\alpha}_{\text{entr}_v} = Y_I \quad (100)$$

$$\langle \rho_1 / \rho_2 \rangle^{\alpha}_{\text{exit}_v} = \langle \rho_1 / \rho_2 \rangle^{\alpha}_{\text{bed}_v} = Y \quad (101)$$

$$\frac{\langle \rho_1 \rangle^{\beta}_{\ell}}{(\rho_{\sigma}^{\beta} + \langle \rho_2 \rangle^{\beta}_{\ell})} = \bar{M} \quad (102)$$

$$\rho_{\sigma}^{\beta} + \langle \rho_2 \rangle^{\beta}_{\ell} = \rho_{dp}^{\beta} \quad (103)$$

In view of Eq. (98) to (103) we write the final forms of the macroscopic balances as:

β - Phase

Macroscopic Mass Balance

$$\epsilon_{\beta} \rho_{dp}^{\beta} \frac{\partial \bar{M}}{\partial t} = - \langle \dot{m} \rangle \quad (104)$$

Macroscopic Thermal Energy Balance

$$\epsilon_{\beta} \rho_{dp}^{\beta} \left[(C_p)_{dp} + \bar{M} (C_p)_{1\ell} \right] \frac{\partial \langle T \rangle}{\partial t} + \Delta H_{\text{evap}} \langle \dot{m} \rangle = \dot{Q}_{SG} \quad (105)$$

where we have neglected the source term
 α - Phase

Macroscopic Mass Balance

$$\frac{\rho_{2v} \dot{G}}{\nu} (Y - Y_I) = \langle \dot{m} \rangle \quad (106)$$

Macroscopic Thermal Energy Balance

$$\frac{\rho_{2v} \dot{G}}{\nu} \left[(C_p)_{2v} + Y_I (C_p)_{1v} \right]_I (\langle T \rangle_I - \langle T \rangle) = \dot{Q}_{SG} \quad (107)$$

where we have also neglected the source term $\langle \Phi_{\alpha} \rangle$

Final Governing Equations

Combining equations (105), (106) and (107) we obtain:

$$\begin{aligned} \epsilon_{\beta} \rho_{dp}^{\beta} \left[(C_p)_{dp} + \bar{M} (C_p)_{1\ell} \right] \frac{\partial \langle T \rangle}{\partial t} + \Delta H_{\text{evap}} \frac{\rho_{2v} \dot{G}}{\nu} (Y - Y_I) \\ = \frac{\rho_{2v} \dot{G}}{\nu} \left[(C_p)_{2v} + Y_I (C_p)_{1v} \right]_I (\langle T \rangle_I - \langle T \rangle) \end{aligned} \quad (108)$$

Equation (108) gives us the time rate of change of the temperature within the bed as a function of the known quantities:

$$\epsilon_{\beta}, \rho_{dp}^{\beta}, (C_p)_{dp}, (C_p)_{1\ell}, \frac{\rho_{2v} \dot{G}}{\nu}, (C_p)_{2v}, (C_p)_{1v}, Y_I, \langle T \rangle_I$$

and the unknown quantities:

\bar{M} and Y

In order to develop an equation for \bar{M} we go back to the intermediate averaged continuity equation for the i th species in the δ -phase given as Eq. (34) and express \underline{v}_i according with the definition given as Eq. (18) for $i = 1$ to obtain:

$$\frac{\partial}{\partial t} \langle \rho_i \rangle_\ell + \underline{\nabla} \cdot \langle \rho_i \underline{v}_\delta \rangle_\ell = - \underline{\nabla} \cdot \langle \rho_i \underline{u}_i \rangle_\ell \quad (109)$$

The diffusive flux $\langle \rho_i \underline{u}_i \rangle_\ell$ can be expressed as, Bird *et al.* (1960)

$$\langle \rho_i \underline{u}_i \rangle_\ell = - \langle \rho_\delta D \underline{\nabla} (\rho_i / \rho_\delta) \rangle_\ell \quad (110)$$

Neglecting variations of D within the averaging volume allows us to move the molecular diffusivity outside the averaged term of Eq. (110), and we express Eq. (109) as:

$$\frac{\partial}{\partial t} \langle \rho_i \rangle_\ell + \underline{\nabla} \langle \rho_i \underline{v}_\delta \rangle_\ell = \underline{\nabla} \cdot \left[D \langle \rho_\delta \underline{\nabla} (\rho_i / \rho_\delta) \rangle \right]_\ell \quad (111)$$

If we assume that flux of component 1 is much larger than the flux of component 2; or $\rho_2 \underline{v}_2 \ll \rho_1 \underline{v}_1$, and using Eq. (18) for $i = 1$ in Eq. (21) we obtain

$$\rho_\delta \underline{v}_\delta = \rho_i (\underline{v}_\delta + \underline{u}_i)_\ell, \quad i = 1 \quad (112)$$

after a simple algebraic manipulation we can express the mass average velocity \underline{v}_δ as (Whitaker 1967):

$$\underline{v}_\delta = \frac{\underline{u}_i}{(\rho_\delta / \rho_i - 1)_\ell} \quad i = 1 \quad (113)$$

Taking into consideration assumption 2 that mathematically expresses

$$\Delta V_\delta \ll V_\delta \quad (113(a))$$

Consider the change of the density of the liquid phase ρ_δ as:

$$\underline{\nabla} \rho_\delta = \underline{\nabla} (\rho_1 + \rho_2) \quad (113(b))$$

So

$$\begin{aligned} \underline{\nabla} \rho_{\delta} &= \underline{\nabla} \left(\frac{m_1}{V_{\delta}} \right) + \underline{\nabla} \left(\frac{m_2}{V_{\delta}} \right) \\ &= \frac{1}{V_{\delta}} \underline{\nabla} (m_1) - \frac{m_1}{V_{\delta}^2} \underline{\nabla} (V_{\delta}) \\ &\quad + \frac{1}{V_{\delta}} \underline{\nabla} (m_2) - \frac{m_2}{V_{\delta}^2} \underline{\nabla} (V_{\delta}) \end{aligned} \tag{113(c)}$$

We see that Eq. (113b) can be expressed as

$$\underline{\nabla} (\rho_{\delta}) = \underline{\nabla} (\rho_1) \tag{113(d)}$$

$$\text{if } \frac{\underline{\nabla} V_{\delta}}{V_{\delta}} \ll \frac{\underline{\nabla} (m_1)}{m_1} \text{ and } \frac{\underline{\nabla} (V_{\delta})}{V_{\delta}} \ll \frac{\underline{\nabla} (m_1)}{m_2} \tag{113(e)}$$

where we have already considered $\frac{1}{V_{\delta}} \underline{\nabla} (m_2) \approx 0$ within the averaging volume.

The restriction imposed by Eq. (113e) should be a subject of further study and verification.

Then, assuming that Eq. (113e) is valid, Eq. (111) can be transformed into the following form (Zuritz 1980).

$$\frac{\rho \langle \rho_i \rangle_{\ell}}{\partial t} = \underline{\nabla} \cdot \left[D \langle \underline{\nabla} (\rho_i)_{\ell} \rangle \right] \tag{114}$$

The assumption of Fickian diffusion is tested later when the experimental data are compared to the theoretical model.

Using the averaging theorem to represent $\langle \underline{\nabla} (\rho_i) \rangle_{\ell}$ and defining a tortuosity vector that accounts for the internal geometry of the grain of rice (Gray 1979):

$$\underline{\tau}_{\delta} = \frac{1}{V} \int_{A_{\delta \sigma}} \rho_{i\ell} \underline{n}_{\delta \sigma} dA \tag{115}$$

and defining an effective diffusivity tensor as

$$\underline{\underline{D}}_{\text{eff}} \cdot \underline{\nabla} \langle \rho_i \rangle_{\ell} = D (\underline{\nabla} \langle \rho_i \rangle + \tau_{\delta}) \quad (116)$$

the intermediate averaged continuity Eq. (114) for $i = 1$ takes the final form as:

$$\frac{\partial \langle \rho_i \rangle_{\ell}}{\partial t} = \underline{\nabla} \cdot \left[\underline{\underline{D}}_{\text{eff}} \cdot \underline{\nabla} \langle \rho_1 \rangle_{\ell} \right] \quad (117)$$

which can be further transformed into:

$$\frac{\partial M}{\partial t} = \underline{\nabla} \cdot \left[\underline{\underline{D}}_{\text{eff}} \cdot \underline{\nabla} (M) \right] \quad (118)$$

where

$$M = \frac{\epsilon_{\delta} \langle \rho_1 \rangle_{\ell}^{\delta}}{\epsilon_{\sigma} \rho_{\sigma} + \epsilon_{\delta} \langle \rho_2 \rangle_{\ell}^{\delta}} \quad (119)$$

Equation (119) is related to Eq. (104) by the definition of phase average as:

$$\frac{\partial \bar{M}}{\partial t} = \frac{1}{\nu'} \int_{V_{\delta\sigma}} \left(\frac{\partial M}{\partial t} \right) dV \quad (120)$$

If we can solve Eq. (118) we will be able to calculate the unknown quantity \bar{M} by means of Eq. (120). Equation (120) is the most important to us since it predicts the rate of drying.

Equation (118) is subject to the following initial and boundary conditions:

$$\text{IC :} \quad t = 0, \text{ on } \nu_{\text{grain}}, M = M_i = \bar{M}_i \quad (121)$$

$$\text{BC 1 :} \quad t \geq 0, \text{ at the center of the spherical grain, } \underline{\nabla} (M) = 0 \quad (122)$$

$$\begin{aligned} \text{BC 2 :} \quad t > 0, \text{ at } A_{(\sigma\delta)\alpha}, \langle \rho_i \underline{\nu}_i \rangle_{\ell} \cdot \underline{n}_{(\sigma\delta)\alpha} \\ = - (\rho_i \underline{\nu}_i)_{\nu} \cdot \underline{n}_{\alpha(\sigma\delta)}, \quad i = 1 \end{aligned} \quad (123)$$

We can modify Eq. (123) by performing a rather lengthy analysis (Zuritz 1980) into:

$$\begin{aligned} \text{BC 2':} \quad t > 0; \text{ at } A_{(\sigma\delta)\alpha}; - \left[\underline{D}_{\text{eff}} \cdot \underline{\nabla} (M) \right] \cdot \underline{n}_{(\sigma\delta)\alpha} \\ = \frac{\nu \langle \dot{m} \rangle}{(\epsilon\rho)_{\text{dp}} A_{\alpha\beta}} ; i = 1 \end{aligned} \quad (124)$$

If we assume, referring to Fig. 3, that at the surface $A_{(\sigma\delta)\alpha}$ the grain moisture $M_{(\sigma\delta)\alpha}$ reaches instantaneous equilibrium with the bulk absolute humidity of the air within the bed Y the $\langle \dot{m} \rangle$ is given by Eq. (106) and BC 2' becomes

$$\begin{aligned} \text{BC 2':} \quad t > 0; \text{ at } A_{(\sigma\delta)\alpha}; - \left[\underline{D}_{\text{eff}} \cdot \underline{\nabla} (M) \right] \cdot \underline{n}_{(\sigma\delta)\alpha} \\ = \frac{\rho_{2v} \dot{G}}{(\epsilon\rho)_{\text{dp}} A_{\alpha\beta}} (Y - Y_I) \end{aligned} \quad (125)$$

Next we need to develop an equation that relates the absolute humidity of the air in equilibrium with the grain moisture at the interface $(\sigma\delta) - \alpha$. From the EMC empirical equation developed by Zuritz and Singh (1979) we have

$$M_{(\sigma\delta)\alpha} \equiv M_e = f_1 (RH, \langle T \rangle) \quad (126)$$

where

$$RH = f_2 (Y_{\alpha(\sigma\delta)}, \langle T \rangle) \quad (127)$$

In Eq. (127) we express $Y_{\alpha(\sigma\delta)} = Y$ then we should be able to express:

$$Y = f_3 (M_e (\langle T \rangle), \langle T \rangle) \quad (128)$$

We use the definition of absolute humidity Y (Brooker 1974),

$$Y_{\alpha(\sigma\delta)} = 0.622 \left[\frac{RH \times P_{vs}}{P - RH \times P_{vs}} \right] \quad (129)$$

where RH is given by the empirical EMC (Zuritz and Singh 1979) as

$$RH = 1 - \exp \left[- \left(1 - \frac{T_A}{T_C} \right)^{m'} \frac{K''}{T_A} M_e^a T_A^b \right] \quad (130)$$

and the saturation vapor pressure P_{vs} by the empirical relationship (Kiefer 1941) modified as:

$$P_{vs} = P_{vs}^o \exp \left[6887 \left(\frac{1}{T_A} - \frac{1}{T_A^o} \right) - 5.31 \ln \left(\frac{T_A}{T_A^o} \right) \right] \quad (131)$$

Thus the complete set of equations necessary to predict both drying rate and time rate of change of temperature of the grains within the well-mixed bed, is given by Eq. (108), (118), (121), (122), (125) and (129) to (131).

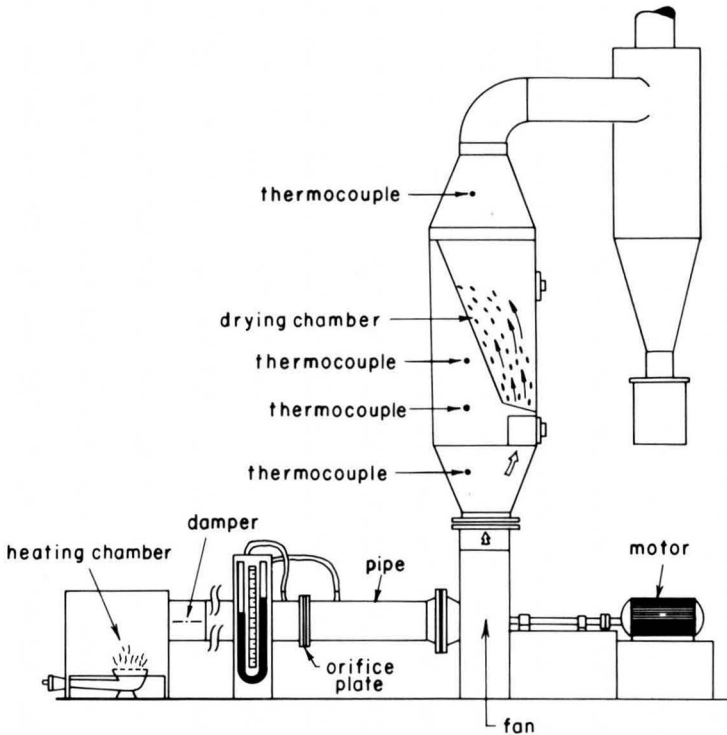


FIG. 4. SPOUTED-BED DRYER

(a) Heating chamber; (b) orifice plate; (c) pipe; (d) fan; (e) motor; (f) drying chamber; (g) damper; (h), (i), and (j) thermocouples.

MATERIALS AND METHODS

A medium grain rough rice, variety CSM5, was used in this investigation.

A spouted-bed dryer (provided by Sivetz Coffee Enterprises, Inc.) was used to dry the rice in separate batch operations (Fig. 4).

Four Copper-Constantan thermocouples, 24 gauge, were used to measure air and grain temperatures. Two more thermocouples were used to measure both wet and dry bulb ambient air temperatures.

Experiments were performed at three different air velocities and air temperatures. Each batch was dried in two passes with tempering between passes.

RESULTS AND DISCUSSION

Equilibrium moisture content data (EMC) used in this investigation were experimentally determined by Zuritz and Singh (1979) from 10°C to 40°C and by Zuritz and Singh (1980) from 40°C to 60°C.

The different parameters used to predict moisture content and temperature of rice during drying were those presented by Wang (1978) and are listed in the Nomenclature section.

The initial value of $D_{rr(\text{eff})}$ used with each trial was selected within the range of those determined by Wang (1978) and Steffe (1979). Then, the value was varied two or three times until both experimental and predicted values gave a reasonable correlation.

It should be mentioned that no optimization procedure was used to find the best curve to fit the experimental data, since the experiments were performed to check the validity of the proposed theoretical model rather than to determine the governing transport parameter $D_{rr(\text{eff})}$.

Predicted and experimental data for all the trials are given by Zuritz (1980) along with initial weight of rice and inlet air conditions.

Figure 5 shows both, predicted and experimental average moisture contents and temperature as a function of time for one of the trials. The same good agreement between experimental and predicted moisture content data as shown in Fig. 5 was found with all the sixteen experiments. It should be noted that the effective diffusivity varied from 6×10^{-6} to 1.9×10^{-5} cm²/min for all sixteen experiments (Zuritz 1980).

The differences between experimental and predicted temperatures during the first half of the drying period could be due to the fact that the air did not leave the bed in complete thermal equilibrium with the grain. In addition the thermocouple readings were affected by air

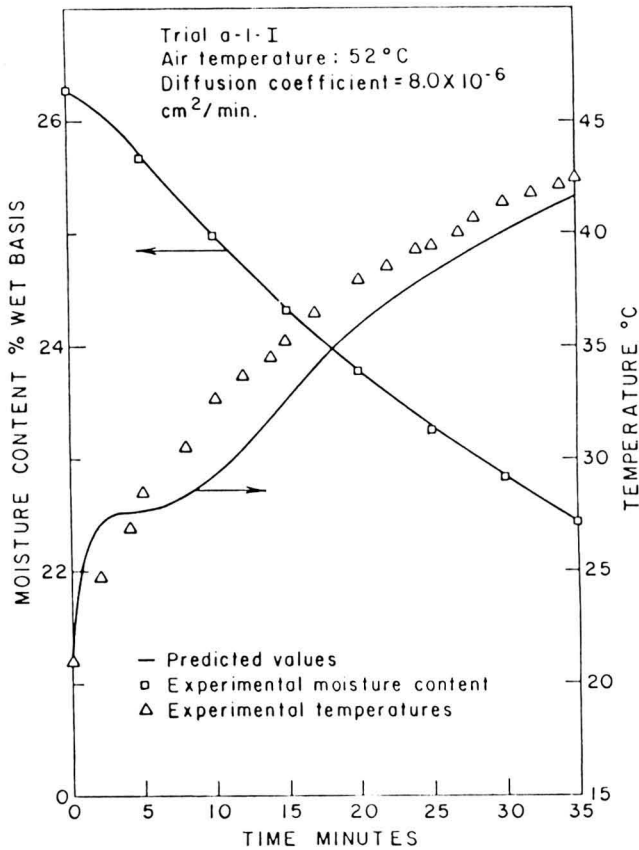


FIG. 5. PREDICTED AND EXPERIMENTAL AVERAGE TEMPERATURES AND MOISTURE CONTENTS OF RICE

flowing through the grain; therefore, those values represent air temperature rather than grain surface temperature.

It was observed that the difference between temperatures was less at lower inlet air temperature and during the second half of the drying period. This indicates that thermal equilibrium must have been approached toward the end of the drying period. In general these differences are not significant and the model seems to represent the physical phenomenon satisfactorily.

CONCLUSIONS

1. The proposed well-mixed model, derived from fundamental equa-

tions, is shown to be suitable to simulate drying of rough rice in a batch operated spouted bed.

2. There was an agreement between experimental and predicted values when the assumption of equilibrium between the bulk absolute humidity of the air in the bed and the surface-moisture content of the rice was used to express the flux of moisture at the grain-drying air interface.

3. The effective diffusion coefficient $D_{\text{err}(\text{eff})}$ was found to be a function of temperature and initial moisture content of the grain.

REFERENCES

- BECKER, H. A. and SALLANS, H. R. 1955. A study of internal moisture movement in the drying of wheat kernels. *Cereal Chem.* 32 (3), 212–216.
- BIRD, R. B., STEWART, W. E. and LIGHTFOOT, E. N. 1960. *Transport Phenomena*, John Wiley & Sons, Inc., New York.
- BROOKER, D. B., BAKKER-ARKEMA, F. W. and HALL, C. W. 1974. *Drying cereal grains*, The AVI Pub. Co., Inc., Westport, Conn.
- GRAY, W. G. 1975. A derivation of the equations for multi-phase transport. *Chem. Eng. Sc.* 30, 229–233.
- HUSAIN, A., CHEN, C. S. CLAYTON, J. T. and WHITNEY, L. F. 1972. Mathematical simulation of mass and heat transfer in high moisture foods. *TRANSACTIONS OF THE ASAE* 15 (4), 732–736.
- KIEFER, P. J. 1941. The thermodynamic properties of water and water vapor. *Monthly Weather Review* 69 (11), 329–331.
- NEWMAN, A. B. 1931. The drying of porous solids. *Trans. of the AIChE* 27, 203–220, 310–333.
- SHARAF-ELDEEN, Y. I., HAMDY, M. Y. and BLAISDELL, J. L. 1979. Falling rate drying of fully exposed biological materials: a review of mathematical models. *American Society of Agricultural Engineers Paper No. 79-6522*. ASAE, P. O. Box 229, St. Joseph, MI 49085.
- SHERWOOD, T. K. 1931. Application of theoretical diffusion equations to the drying of solids. *Trans. of the AIChE* 27, 190–202.
- STEFFE, J. F. 1979. Moisture diffusion and tempering in the drying of rough rice. Unpublished Ph.D. thesis, University of California, Davis, CA 95616.
- WHITAKER, S. 1967. Velocity profile in the Stefan diffusion tube. *I&EC Fundamentals* 6, 476.
- WHITAKER, S. 1968. *Introduction to Fluid Mechanics*, Prentice-Hall, Englewood Cliffs, N.J.
- WHITAKER, S. 1977a. Simultaneous heat, mass and momentum transfer in porous media: a theory of drying. *Advances in Heat Transfer* 13, 119–202.
- WHITAKER, S. 1977b. Toward a diffusion theory of drying. *I&EC Fundamentals* 16 (4), 408–414.
- YOUNG, J. H. 1969. Simultaneous heat and mass transfer in a porous hygroscopic solid. *TRANSACTIONS OF THE ASAE* 12 (2), 720–725.
- ZURITZ, C. A. SINGH, R. P. MOINI, S. and HENDERSON, S. M. 1979. Desorption isotherms of rough rice from 10°C to 40°C. *TRANSACTIONS OF THE*

ASAE 22 (2), 433–436, 440.

ZURITZ, C. A. and SINGH, R. P. 1980. Desorption isotherms of rough rice from 40°C to 60°C. Unpublished report, Department of Agricultural Engineering, University of California, Davis, CA 95616.

ZURITZ, C. A. 1980. Simulation of rough rice drying in a spouted-bed. Unpublished Ph.D. thesis, University of California, Davis, CA 95616.

NOMENCLATURE

Roman Letters

a	Constant defined in Eq. (130)
A	Surface area (cm^2)
b	Constant defined in Eq. (130)
C_p	Constant pressure heat capacity [$\text{J/kg}^\circ\text{K}$]
$\underline{\underline{D}}_{\text{eff}}$	Liquid phase effective diffusivity tensor [m^2/s]
$\underline{\underline{D}}_{\text{rr}(\text{eff})}$	Radial component of the liquid phase effective diffusivity tensor [m^2/s]
D	Liquid phase molecular diffusivity [m^2/s]
\dot{G}	Air flow rate [m^3/s]
\bar{h}	Enthalpy per unit mass [J/kg]
\bar{h}	Partial mass enthalpy [J/kg]
h°	Reference enthalpy
ΔH_{evap}	Enthalpy of evaporation per unit mass [J/kg]
K'	Constant defined in Eq. (130)
m'	Constant defined in Eq. (130)
$\langle \dot{m} \rangle$	Mass rate of evaporation of water per unit volume of bed
M	Grain moisture content dry basis, decimal
\bar{M}	Volume averaged grain moisture content, dry basis, decimal
M_s	Surface grain moisture content, dry basis, decimal
M_e	Equilibrium moisture content of rice, dry basis, percent
\underline{n}	Unit outwardly directed normal
N	Total number of grains of rice in the bed
p	Pressure
P_v	Vapor pressure
P_{vs}	Saturation vapor pressure
P_{vs}°	Saturation vapor pressure at $273.16^\circ\text{K} = 6.105 \text{ mb}$
P	Atmospheric pressure
\underline{q}	Heat flux vector

\dot{Q}_{SG}	Heat flux per unit volume of bed defined by Eq. (78)
r	Radius distance from the center of a sphere
R_s	Equivalent radius of rice
RH	Relative humidity of the air in equilibrium with the rice surface moisture content
t	Time (s)
T	Temperature ($^{\circ}C$)
T°	Reference temperature ($^{\circ}C$)
T_A	Absolute temperature ($^{\circ}K$)
T°_A	Reference absolute temperature = 273.16 $^{\circ}K$
T_C	Critical temperature of water ($^{\circ}K$)
\underline{u}	Mass diffusion velocity
\underline{v}	Fluid velocity
\underline{V}	Volume (m^3)
ν	Volume of the bed
ν'	Control volume within the grain of rice
$V_m(t)$	Material volume
\underline{w}	Velocity of an interface
\underline{Y}	Absolute humidity of the air

Greek Letters

ϵ	Volume fraction of a phase
ρ	Density
λ_o	Enthalpy of vaporization of water at reference temperature
$\underline{\tau}\delta$	Tortuosity vector defined by Eq. (115)
$\underline{\tau}$	Viscous stress tensor
$\underline{\Phi}$	Rate of heat generation

Subscripts

α	Drying air-phase in the bed
β	Rice-phase in the bed
dp	Dry product
δ	Liquid-phase within the grain of rice
σ	Solid-phase within the grain of rice
$\sigma\delta$	One grain-phase
e	Entrances and exits
$grain$	Individual grain of rice
i	Ith species property
I	Inlet air properties
ℓ	Refers to the liquid state

s	Solid fixed surfaces
v	Refers to the vapor state
w	Water property
$\alpha\beta, (\sigma\delta)\alpha$	Interface between respective phases
1	Water vapor in the gas-phase or liquid-water in the grain-phase
2	Dry air in the gas-phase or soluble solids in the grain-phase

Numerical Values of Parameters Used in the Predictions of Moisture and Temperature of Rice

$A_{(\sigma\delta)\alpha}$	Surface area of one grain of rice = 4.75×10^{-5} (m ²)
$(C_p)_a$	Constant pressure heat capacity of dry air = 1.0132×10^3 [J/kg °K]
$(C_p)_{dp}$	Constant pressure heat capacity of rice dry matter = 1.1221×10^3 [J/kg °K]
$(C_p)_v$	Constant pressure heat capacity of water vapor = 1.8841×10^3 [J/kg °K]
$(C_p)_w$	Constant pressure heat capacity of liquid water = 4.1868×10^3 [J/kg °K]
R'	Universal gas constant = 4.619×10^2 [J/kg]
R_s	Equivalent radius of one grain of rice = 1.84×10^{-3} (m)
V_{grain}	Volume of one grain of rice = 1.99×10^{-8} (m ³)
$(\epsilon\rho)_{dp}$	True density of one grain of dry rice = 1,230 [kg/m ³]

GAS PARTICLE HEAT TRANSFER COEFFICIENT IN FLUIDIZED PEA BEDS

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ABSTRACT

Individually Quick Frozen food production has led to the development of fluidized bed freezing equipment, where the previously cooled fluidizing air freezes the food by direct heat transfer. The operating conditions of these beds, as well as their design characteristics, are consequently related to the gas-particle heat exchange including the fluid dynamic complexity of aggregative fluidization.

The purpose of this paper is the experimental determination of gas-particle heat transfer coefficients for fluidized pea beds and their extension to conditions prevailing in industrial processing.

Heat transfer coefficients were determined for both, peas and porous alumina pellets, in a system where particles were dried by the fluidizing air. Transfer coefficients were evaluated for various gas velocities and bed depths.

The results obtained for alumina pellets and those for peas can be satisfactorily correlated by using the Archimedes number, and show good agreement with correlations obtained by other authors.

INTRODUCTION

The processing of granular frozen foods has led to the development of IQF (individually quick frozen) products which, by maintaining the granular nature of the foodstuff facilitate the partial use of packages or the marketing of frozen mixtures processed in different seasons.

The processing of such products, which include a wide variety (peas, potato or carrot cubes, french fried potatoes, artichokes, meat balls, shrimps, etc.) has been made possible by the use of fluidized bed continuous freezers where the previously cooled fluidizing air freezes the product by direct heat transfer.

The operating conditions of such beds, as well as the design charac-

teristics, are therefore related to gas-particle heat transfer with the addition of the fluid dynamic complexity of aggregative fluidization (bubbling beds).

Consequently, for purposes of calculation, information is required on gas-particle heat transfer in this type of bed, as well as on its dependence on the operating variables (gas velocity, depth of bed, size of particle, etc.).

The bibliography on gas-particle heat or mass transfer in fluidized beds is profuse and has been successfully summarized by various authors (Davidson and Harrison 1963; Barker 1965; Kunni and Levenspiel 1969; Davidson and Harrison 1971; Gutfinger and Abuaf 1974).

In general, the transfer coefficients have been evaluated under different conditions, alternating steady state techniques with transitory measurements and adopting different methods to obtain the corresponding temperatures. At the same time, determinations have been made in fluidized beds of several dimensions and covering a broad range of particle sizes. In general, it can be concluded that there is a large amount of existing information, although there are important discrepancies in the predictions of each author.

Nevertheless, for the working conditions prevailing in continuous fluidized beds used to freeze foods (shallow beds and Reynolds numbers $Re = G_o D_p / \mu$ above 1000) the number of contributions decreases drastically (Bradshaw and Myers 1963; Chang and Wen 1966; Lindauer 1967; Persson 1967; Mann and Feng 1968; Bhattacharyya and Pei 1974) and there are also discrepancies not only on the value of the coefficients but also with regard to their dependence on the parameters of the system.

The purpose of this paper is to determine the gas-particle heat transfer coefficients for fluidized pea beds and to extend predictions to the range of conditions prevailing in industrial processing (low temperatures).

However, in systems where only sensible heat is used, this determination involves difficulties in the evaluation of the temperature on the surface of the particles. Many of the discrepancies between authors are due, precisely, to the methods used to evaluate that temperature (Kettering *et al.* 1950; Wamsley and Johanson 1954; Heertjes and McKibbins 1956; Bradshaw and Myers 1963; Lindauer 1967; Mann and Feng 1968; Bhattacharyya and Pei 1974).

In order to work with a steady system that permits the exact determination of the condition existing in the gas-particle interphase, the heat transfer coefficient was obtained through a dehydration process which involved a simultaneous change of phase and the consequent transfer of matter between the particles and the gas. For this purpose, wet particles were placed in the bed and were dried by fluidization

with air. In this way, heat was transferred at the expense of the latent, rather than sensible heat of the solid, and a steady state could be achieved by maintaining the solid at the wet bulb temperature while the air temperature moves along the corresponding adiabatic line.

To comply with the conditions described, it is necessary to maintain the solid during the experiment in the so-called constant velocity drying period, wherein the surface attains a vapor pressure equal to that of water at the same temperature. For this reason, the length of that period was carefully determined in single particle experiments, so as to limit the measurement time.

Parallel to the determination of the heat transfer coefficient, the voidage of the bed was evaluated for different superficial gas velocities so as to find the minimum fluidization voidage. The pressure drop was also measured in terms of the air mass flow to determine the corresponding minimum fluidization velocity.

FLUID DYNAMICS OF THE BED

Materials and Methods

The experimental fluidized bed was of rectangular section (23.7 cm \times 31.8 cm) with glass side walls which made it possible to clearly see the behavior of the bed during the fluid dynamic experiments.

Air was injected by means of a 1.5 HP centrifugal blower with a maximum flow of 70 m³/min and a discharge pressure of 28 mm of water. Through a connecting sleeve, air was introduced into a chamber where three baffles deflected the flow towards the grid which constituted the base of the bed. Several types of grids were tested, good fluidization having been found with a perforated plate having 30% of free area, and 1.25 mm perforations in triangular arrangement.

The drop in pressure caused by this type of grid was sufficiently high to produce a uniform flow of air through the bed.

The air velocity was determined at the top of the bed by an Alnor velometer previously calibrated against a vane type anemometer.

The pressure drop both through the grid and through the bed was evaluated by static pressure inlets under the grid in each of the side walls jointly connected to a single pressure gauge in order to attenuate fluctuations due to lateral channeling of air.

Pressure drops were determined in terms of the mass flow for different heights of the fixed bed H_m . The height H to which the bed expanded was also measured for each superficial gas velocity v_o .

Table 1. Characteristics of the different beds analyzed

Material	$D_p \times 10^2$ (m)	$\rho_s \times 10^{-3}$ (Kg/m ³)	ϵ_{mf}	V_{mf} (m/s)		$v'_{mf} = v_{mf}/\epsilon_{mf}$ (m/s)
				Experimental from ΔP vs. Re	Calculated from ERGUN'S Equation	
Dry peas	0.56	1.23	0.36	1.19	1.15	3.30
Wet peas	0.70	1.03	0.40	1.50	1.46	3.75
Dry alumina	0.56	1.33	0.36	1.22	1.19	3.39
Wet alumina	0.56	1.59	0.39	1.69	1.68	4.33

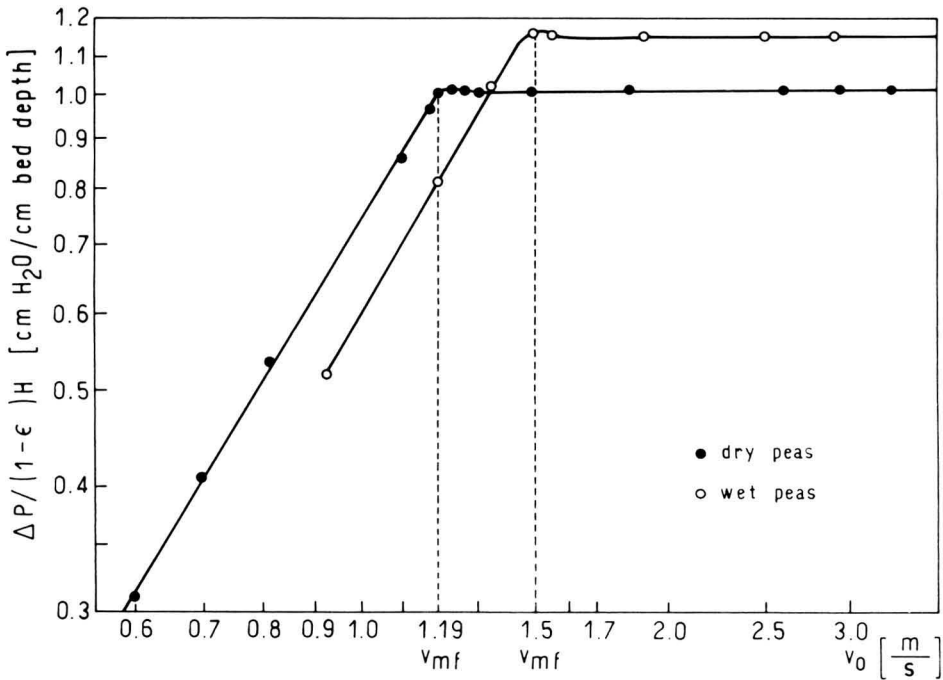


FIG. 1. PRESSURE DROP FOR DRY AND WET PEAS

Results and Discussion

Fluid dynamic behavior experiments were made with four materials: dry peas, wet peas, dry porous alumina pellets and wet porous alumina pellets. Table 1 shows the characteristics of each material.

Figure 1 shows the pressure drop for beds of dry and wet peas, respectively. The values were obtained with bed heights of 3, 4 and 5 cm.

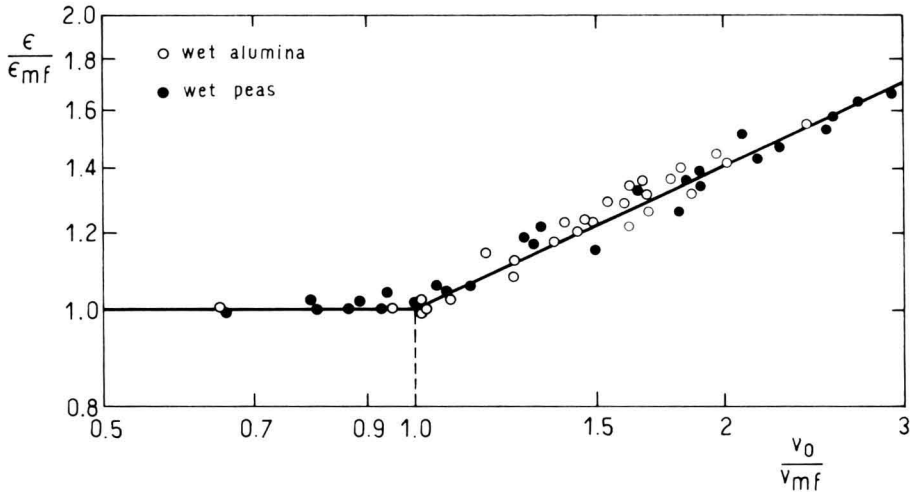


FIG. 2. VOIDAGE AS A FUNCTION OF SUPERFICIAL VELOCITY FOR WET ALUMINA AND PEAS

The figure makes it possible to determine the minimum fluidization velocity for each case, irrespective of H and ϵ . It will be observed that the minimum fluidization velocity for wet peas is higher than for dry peas.

This effect can be mainly attributed to a difference between the minimum fluidization voidage in wet and dry materials. As can be seen in Table 1 wet particles (peas or alumina spheres) are more loosely packed than the dry materials under minimum fluidization conditions.

The values of D_p , ρ_s and ϵ_{mf} in Table 1 were applied to Ergun's equation (Bird *et al.* 1964; Kunii and Levenspiel 1969) to calculate the minimum fluidization velocity v_{mf} as follows:

$$\frac{1.75}{\epsilon_{mf}^3} \left[\frac{D_p v_{mf} \rho}{\mu} \right]^2 + \frac{150(1-\epsilon_{mf})}{\epsilon_{mf}^3} \left[\frac{D_p v_{mf} \rho}{\mu} \right] = \frac{D_p^3 \rho (\rho_s - \rho) g}{\mu^2} \quad (1)$$

For calculation purposes, air density was taken as $\rho = 1.2131 \text{ kg/m}^3$ and viscosity as $\mu = 1.8 \times 10^{-5} \text{ kg/ms}$. The values of v_{mf} obtained are included in Table 1 and, as can be observed, show a satisfactory agreement with those evaluated experimentally by means of the pressure drop measurements.

Moreover, measurements were taken of the expansion of the bed in terms of the air superficial velocity v_0 and were used to determine the

bubble fraction therein. The results are presented in terms of ϵ/ϵ_{mf} in Fig. 2 for wet peas and wet alumina, respectively. The value of ϵ was obtained for each flow rate from the following equation:

$$\frac{1 - \epsilon}{1 - \epsilon_{mf}} = \frac{H_m}{H} \quad (2)$$

where H was evaluated by direct reading through the transparent walls of the bed.

Regression of data was done through the empirical equation:

$$\frac{\epsilon}{\epsilon_{mf}} = \left(\frac{v_o}{v_{mf}} \right)^n \quad (3)$$

Coefficient n , obtained by minimum squares of the logarithmic form of Eq. (3) was $n = 0.437$, which substantially agrees with that of the similar equation reported in the literature in terms of the terminal velocity of the particles (Davidson and Harrison 1971).

As can be seen in Fig. 2, the data for wet peas and alumina can be satisfactorily correlated with Eq. (3). This information is important for evaluating the bubble fraction in the bed and subsequent analysis of the factors that affect the particle-fluid transfer coefficient.

Thus, defining the fraction of the volume of the bed occupied by bubbles as δ , we have:

$$\delta = \frac{\epsilon - \epsilon_{mf}}{1 - \epsilon_{mf}} \quad (4)$$

where, according to Davidson's bubbling bed model, it has been assumed that the emulsion phase is under minimum fluidization conditions.

Substituting Eq. (3) in (4) we obtain:

$$\delta = \frac{\epsilon_{mf} \left[\left(\frac{v_o}{v_{mf}} \right)^n - 1 \right]}{1 - \epsilon_{mf}} \quad (5)$$

therefore, having determined the velocity (v_{mf}) and minimum fluidization voidage (ϵ_{mf}) by Eq. (5), the bubble fraction in the bed in terms of the superficial air velocity v_o can be established.

Figure 3 represents Eq. (5) for wet peas ($\epsilon_{mf}=0.40$ and $v_{mf}=1.50$ m/s)

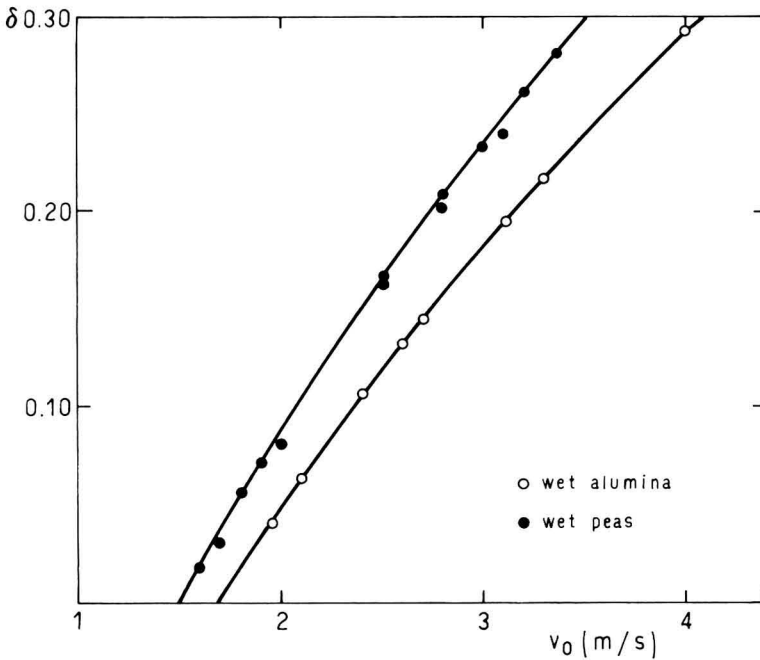


FIG. 3. BUBBLE BED FRACTION FOR WET ALUMINA AND PEAS

and alumina ($\epsilon_{mf}=0.39$ and $v_{mf}=1.69$ m/s). The values of δ calculated from Eq. (4) with the values of ϵ obtained experimentally have also been included.

It can be seen that the bubble fraction in peas is greater than in alumina, in accordance with its lower minimum fluidization velocity. It will also be observed that an important fraction of the flow circulates through the bubble phase. These effects undoubtedly influence the values of the heat transfer coefficients and their dependence on the governing parameters, as will be seen later.

FLUID-PARTICLE HEAT TRANSFER COEFFICIENT

As mentioned earlier, the method used to determine experimentally the gas-particle heat transfer coefficient, is a steady state method based on the heat transfer from air to particles and its elimination as latent heat through the consequent vaporization of water. This method makes it possible to know the exact temperature in the surface of the solids provided these are maintained in the constant rate period of drying

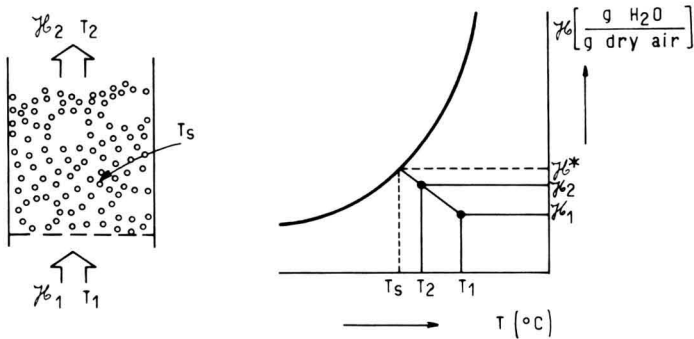


FIG. 4. SCHEME OF THE OPERATING CONDITIONS USED FOR THE HEAT TRANSFER COEFFICIENT MEASUREMENT

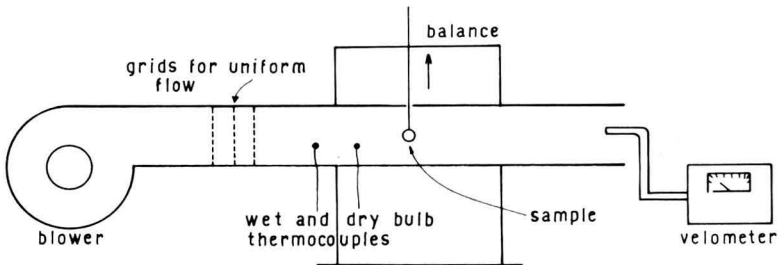


FIG. 5. EXPERIMENTAL DEVICE USED FOR MEASURING THE CRITICAL TIME

throughout the experiment and that the bed behaves adiabatically.

Consequently, once wet solids are introduced in the bed and discarding an initial period of fluidization during which unsteady transfer processes occur (temperature stabilization) or mechanical water loss processes take place (drops dragged along by the air current), a constant drying rate is reached in the bed characterized by the fact that the surface of the solid is maintained at the wet-bulb air temperature. If, furthermore, the equipment does not lose heat through its walls, the air will move along the corresponding adiabatic line of a psychrometric diagram keeping the particles interphase temperature constant throughout the bed. Figure 4 shows schematically this operating condition.

It is therefore necessary to check that the particles are kept throughout the experiment above the critical water content which marks the end of the period of constant drying rate. Calling critical time, the

period required to achieve this water content, its evaluation determines the measuring time of the heat transfer coefficient in the fluidized bed.

Calculation of Critical Time

Critical times were evaluated by means of experiments with individual particles whose weights were recorded continually and which were subjected to a current of air of known velocity, temperature and humidity (Fig. 5). The velocity of air coming from a blower was made uniform by means of grids placed before the weighing zone. The velocity was measured at the outlet by a velometer. Under these conditions it was possible to obtain a practically flat velocity profile (the Reynolds number in the duct was approximately 17,500 for the lowest velocity used).

The inlet air humidity was determined by measuring the dry and wet-bulb temperatures with thermocouples in series connected to a potentiometric recorder.

In order to simulate the drying conditions to which the particle would be subjected in the fluidized bed, and accepting the theory that in aggregative fluidization the emulsion phase is maintained under minimum fluidization conditions, circulating the excess air through the bubbles (Davidson and Harrison 1971), the velocity of the air impinging on the particle was the interstitial velocity calculated in the emulsion phase as $v'_{mf} = v_{mf}/\epsilon_{mf}$.

Experiments were made for various psychrometric conditions of the air, but within the range foreseen for operation in the bed. In all cases an unsteady initial period of approximately two minutes was found. The critical times were evaluated by making minimum squares with the data obtained which showed an undeniable linear behaviour, and then tracing the resulting straight line of statistical analysis.

Results show that the net measuring time for peas was about 3 min, which extends to 5 min, for porous alumina pellets.

Calculation Equations

Considering a fluidized bed having an S section and H height, where-in spherical particles are dried by means of a circulating fluid, assuming a perfect mixing in the solid and plug flow for the gas, a differential energy balance in a control volume formed by the space between particles which occupy a height dy , can be written in steady state as:

$$dQ = \rho v'_o S \epsilon \hat{C}_p dT = G_o S \hat{C}_p dT \quad (6)$$

where $G_o = \rho v_o = \rho v'_o \epsilon$ is the superficial mass air flux, \hat{C}_p its specific heat, T its temperature and S the bed section.

At the same time, a heat transfer coefficient h can be defined for the bounding surface of the control volume as follows:

$$dQ = h a S (T_s - T) dy \quad (7)$$

where a is the particle surface per unit of volume of the bed and T_s the surface temperature of the solids

From Eq. (6) and (7) it is obtained:

$$G_o \hat{C}_p dT = \frac{6(1-\epsilon)}{D_p} h (T_s - T) dy \quad (8)$$

where $a = 6(1-\epsilon)/D_p$ has been introduced into Eq. (8), D_p being the diameter of the spherical bed particles.

Integrating Eq. (8) between the inlet and outlet of the bed, assuming h and T_s as constant, it is obtained:

$$\ln \frac{(T_s - T_2)}{(T_s - T_1)} = - \frac{h 6(1-\epsilon) H}{G_o \hat{C}_p D_p} \quad (9)$$

or, introducing Eq. (2):

$$\ln \frac{(T_s - T_1)}{(T_s - T_2)} = \frac{6 h(1-\epsilon_{mf}) H m}{G_o \hat{C}_p D_p} \quad (10)$$

It should be noted that determination of the heat transfer coefficient h demands no energy losses through the walls of the bed, as well as knowledge of temperature T_s on the surface of the solids.

Working under adiabatic conditions and with the particles above the critical moisture, the T_s temperature can be taken as equal to the wet-bulb air temperature. As the air changes its temperature and humidity along the adiabatic line, (Fig. 4) the T_s temperature will be constant throughout the bed. Consequently, h was calculated from Eq. (10) measuring the temperatures as described hereunder.

Materials and Methods

The fluidized bed used for the heat transfer experiments was similar to that described in a previous section although with a smaller section (25.4 cm \times 17.7 cm). Its sides were isolated with expanded polysty-

rene for the purpose of ensuring an adiabatic exchange process.

Both the wet and dry-bulb temperatures were determined at the inlet and at the outlet by means of thermocouples of copper-constantan connected to a multichannel potentiometric recorder. The coincidence of the wet-bulb temperature in the inlet and outlet was an indication of the aforementioned adiabatic condition.

To shorten the time for attaining the steady state, previously to the run, wet particles were fluidized for a few minutes and then removed and replaced by the experimental particles. These particles were kept until used in chilled water at a temperature equal to the predicted wet bulb temperature for the run so as to minimize the transient period.

Porous alumina pellets with a particle diameter of $D_p = 0.56$ cm and peas with $D_p = 0.7$ cm were used. Rehydration of alumina pellets between runs was made applying vacuum over the particles and subsequently immersing them in distilled water for 24 h.

Once the particles were placed in the bed, the fluidization commenced, discarding the measurements during the first 2 min corresponding to the transient period in particles, thermocouples, walls, grid, etc., and the remosion of the excess water in the form of fine drops carried by the air flow.

According to the evaluation made for the critical time, measurements were performed from 2 to 7 min in the case of porous alumina and from 2 to 5 min for the case of peas.

During the first experiments, at the same time the change in air humidity was measured, particle samples were taken from the bed and the water content determined. In this way, the water balance between the air and the particles was ascertained.

Results and Discussion

The repeated values of individual runs were averaged, showing a maximum dispersion of 5.7%. Experiments were performed with superficial velocities v_o from 1.8 to 3.5 m/s and bed heights H_m from 2.5 to 6.0 cm.

Experimental values for some of the runs performed with pea beds are shown as example in Table II.

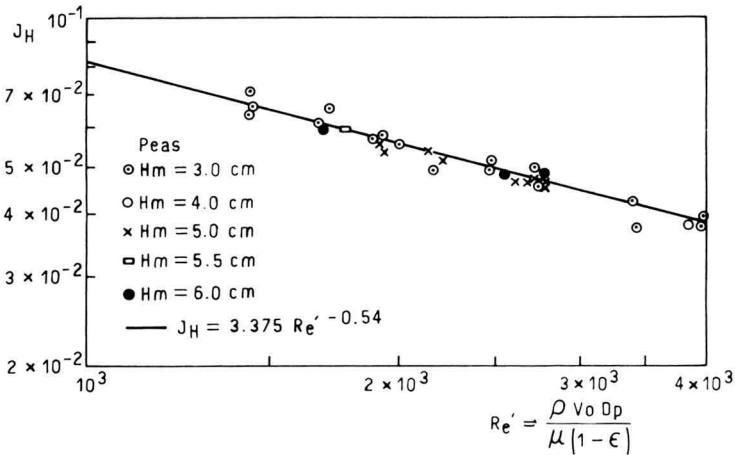
Complete data on peas are plotted in Fig. 6 in terms of the Colburn's factor $j_H = Nu Re^{-1} Pr^{-1/3}$ as a function of the modified Reynolds number $Re' = \rho v_o D_p / \mu(1-\epsilon)$.

As can be seen, the heat transfer coefficient is independent of the bed height.

Data from Fig. 6 have been correlated by using the equation:

Table 2. Pea beds heat transfer coefficient measurements

Peas $D_p = 0.7$ cm		$\epsilon_{mf} = 0.4$			$v_{mf} = 1.5$ (m/s)		
$H_m \times 10^2$ (m)	v_o (m/s)	T_1 (°C)	T_2 (°C)	T_s (°C)	\bar{h} (W/m ² °K)	j_H	Re'
3.0	1.80	30.80	23.30	20.40	180.00	0.0660	1434
3.0	2.20	26.50	21.93	19.53	187.00	0.0554	1914
3.0	2.60	25.00	18.45	14.33	198.00	0.0495	2446
3.0	3.20	20.95	18.83	17.20	217.00	0.0436	3384
3.0	3.60	29.9	24.87	19.83	196.22	0.0353	3970
4.0	3.50	20.32	17.20	15.45	221.00	0.0392	3838
4.5	2.80	19.95	14.32	15.70	216.00	0.0479	2722
5.0	2.20	28.23	24.13	23.32	189.00	0.0566	1904
5.0	2.40	29.15	22.43	20.95	195.00	0.0537	2147
5.0	2.80	28.18	24.33	23.33	210.00	0.0494	2700
5.5	2.10	28.18	20.45	19.33	187.00	0.0586	1785
6.0	2.00	18.32	14.58	14.20	196.90	0.0608	1689
6.0	2.60	20.95	15.95	15.25	213.00	0.0597	2478
6.0	2.80	21.8	16.7	15.75	210.00	0.0482	2757

FIG. 6. GAS PARTICLE HEAT TRANSFER COEFFICIENT IN TERMS OF COLBURN'S FACTOR j_H , FOR PEAS AS A FUNCTION OF THE MODIFIED REYNOLDS NUMBER

$$j_H = a Re'^b \quad (11)$$

where $a = 3.375$ and $b = -0.540$ are constants obtained by applying

the least square method to the logarithmic form of Eq. (11). The correlation coefficient was 0.96. The value of b is in good agreement with Reynolds dependence reported by other authors who worked with different materials (Bradshaw and Myers 1963; Chang and Wen 1966; Bhattacharyya and Pei 1974). The analysis of Table 2 also shows a slight dependence of the heat transfer coefficient on the superficial air velocity.

The approximate constancy of the heat transfer coefficient for increasing superficial air velocities can be interpreted in terms of Davidson's (1971) bubbling bed model which recognizes that the excess flow over the minimum fluidization conditions circulates in the form of bubbles through the bed, generating a high degree of by-pass and consequently the fluid-particle coefficient does not increase in the same proportion as v_o increases.

EXTENSION OF RESULTS TO FLUIDIZED BEDS FREEZERS OPERATING CONDITIONS

As shown in Table 2 room temperatures were used for the gas-particle heat transfer coefficient measurement. However, fluidized beds used for freezing of foods works at temperatures in the order of -40°C .

Under these conditions, important changes in properties of air take place leading to variations in the heat transfer coefficient which must be calculated from the existing data.

Moreover, it would be worthwhile to extend the present information on peas to other spherical foods and to different particle sizes.

In order to reach this objective a complete dimensionless correlation of the gas-particle heat transfer coefficient in fluidized beds must be supplied so as to make it independent of the scale and properties of the system.

The fluid dynamic complexity of aggregative fluidized beds, where the size and velocity of the bubbles depends, in a manner that has not yet been fully established, on the size of the particles, their density, etc., complicates the selection of the relevant dimensionless groups of the system. However, a first approximation can be based on a procedure similar to that used for fixed beds wherein the space between particles is considered as a tortuous and anfractuous duct resulting (Bird *et al.* 1964):

$$\text{Nu} = \text{Nu}(\text{Re}', \text{Pr}) \quad (12)$$

where $\text{Nu} = h D_p / k$; $\text{Pr} = \mu \hat{C}_p / k$ and $\text{Re}' = G_o D_p / \mu(1 - \epsilon)$.

Nevertheless, the fluidized bed involves considerable differences with respect to fixed beds due to the presence of bubbles whose fraction depends on the flow circulated and which, obviously alters the shape and size of the duct taken as a basis for the fixed bed correlations.

Thus, for $Re' > 1000$, Burke-Plummer's equation (Bird *et al.* 1964) predicts:

$$\frac{1.75}{\epsilon_{mf}^3} \left[\frac{D_p v_{mf} \rho}{\mu} \right]^2 = Ar \quad (13)$$

where $Ar = D_p^3 g \rho (\rho_s - \rho) / \mu^2$ is the Archimides number which relates flotation and viscous forces.

Substituting Eq. (13) in Eq. (5) it is obtained:

$$\delta = \frac{\epsilon_{mf}}{(1 - \epsilon_{mf})} \left[\left(\frac{1.323 Re' (1 - \delta) (1 - \epsilon_{mf})}{\epsilon_{mf}^{3/2} Ar^{0.5}} \right)^n - 1 \right] \quad (14)$$

It will be noted that the equality of Re' is not a sufficient condition for the existence of an equal bubble fraction (equal shape and size of the duct considered). For this to occur and make it possible to generalize the data for particles having different diameters and/or densities, it is essential to introduce the Archimides number in addition to those included in Eq. (12). (In fact, equality in δ also requires equality in ϵ_{mf} . However, for spherical pellets there is little variation of ϵ_{mf} , depending only on the arrangement of the bed).

Chang and Wen (1966) who worked with different particle diameters and solids of different densities, propose adding the Archimides number in the correlation of the transfer coefficient. With this additional group these authors satisfactorily correlated all their data as well as those of Bradshaw and Myers (1963) through the following equation:

$$j_H = 0.097 Re'^{-0.502} Ar^{0.198} \quad (15)$$

Note that the dependence of j_H on $Ar^{0.198}$ is similar to the dependence of δ predicted by Eq. (13) according to $Ar^{0.5n} = Ar^{0.218}$.

In order to check the j_H dependence with Archimides number the data obtained from the experiences performed in the present work with wet porous alumina as well as those obtained with peas are included in Fig. 7 together with Chang and Wen Eq. (15). Data of Bradshaw and Myers (1963) are also plotted.

It will be seen that experimental results agree satisfactorily, con-

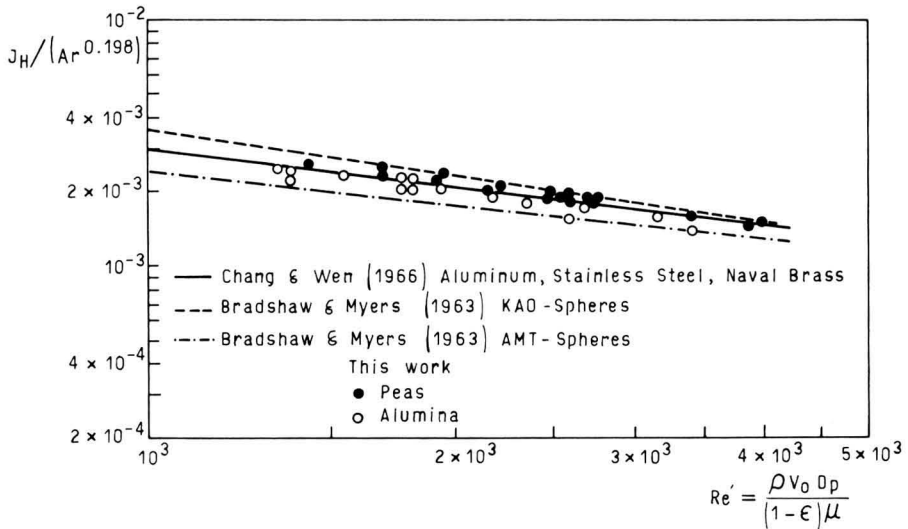


FIG. 7. COLBURN FACTOR CORRELATION FOR ALUMINA AND PEAS BEDS

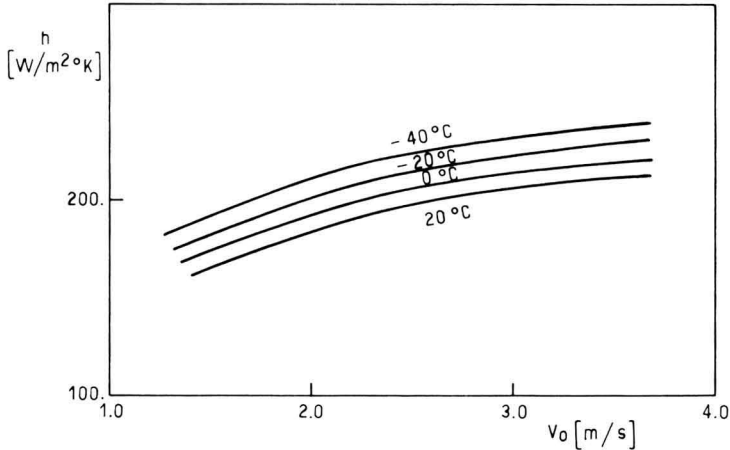


FIG. 8. GAS PARTICLE HEAT TRANSFER COEFFICIENT DEPENDENCE WITH AIR SUPERFICIAL VELOCITY v_o AND AIR TEMPERATURE FOR PEA BEDS

firming the usefulness of Eq. (14) in extending the results to spheres of different density and size of particle.

Figure 8 shows the gas-particle heat transfer coefficient dependence with the superficial velocity and the air temperature for the case of

peas as predicted by Eq. (15). Saturated air properties for each temperature were used for calculation and the bed voidage was supplied by Eq. (1) and (3). As can be seen the effect of air temperature on the heat transfer coefficient was small involving a change of 13% in the range from -40°C to 20°C for the case of peas. Anyway, Fig. 8 allows for a fast prediction of the heat transfer coefficients for pea beds at the temperatures normally used in fluidized freezers.

Moreover, by using Eq. (1), (3), and (15) the heat transfer coefficient for other spherical foods, as meat balls, grapes, cherries, etc., can be calculated.

CONCLUSIONS

- (1) The fluid dynamics of fluidized pea beds is analyzed, showing aggregative fluidization with volumetric bubble fractions of up to 30%.
- (2) The presence of bubbles justifies, through a bypass effect of circulated air, the relatively low values of the heat transfer coefficient in fluidized beds as well as the slight dependence on the superficial air velocity.
- (3) The minimum fluidization superficial velocity calculated from Ergun's equation shows good agreement with experimental values obtained for peas and alumina. An empirical equation for predicting bed voidage for different superficial air velocities is provided. This equation allows to account for the effect of air properties changes as well as for different solid densities and particle sizes.
- (4) The experimental method used for measuring the gas-particle heat transfer coefficient, based on an exchange of sensible and latent heat, generates reproducible results comparable with those of other authors. The heat transfer coefficient obtained was independent of the bed height.
- (5) In order to extend the obtained data to fluidized bed freezers operating conditions as well as to other spherical foods and particles sizes, a correlation with the Archimedes number is used and checked with experimental data obtained for peas and porous alumina pellets.
- (6) This correlation, proposed by Chang and Wen (1966), together with Eq. (1) and (3) allows for predictions of the heat transfer coefficient at low temperatures and for spherical foods of different particle sizes and densities.

NOMENCLATURE

a	=	surface area of particles per unit of volume of the bed, m^{-1} .
Ar	=	Archimides number. $Ar = D_p^3 g \rho (\rho_s - \rho) / \mu^2$
\hat{C}_p	=	specific heat of air $J/kg^\circ K$.
D_p	=	particle diameter, m.
g	=	gravity acceleration, m/s^2 .
G_o	=	superficial flux, $kg/m^2 s$.
h	=	heat transfer coefficient, $W/m^2^\circ K$.
j_H	=	Chilton-Colburn j-factor for heat transfer. $j_H = Nu Re^{-1} Pr^{-1/3}$
k	=	thermal conductivity of air, $W/m^\circ K$.
H	=	height of bed, m.
H_m	=	height of bed at minimum fluidization condition, m.
n	=	exponent from Eq. (4).
Nu	=	Nusselt number. $Nu = h D_p / k$.
Pr	=	Prandtl number. $Pr = \mu \hat{C}_p / k$.
Q	=	heat flow, W.
Re	=	Reynolds number. $Re = G_o D_p / \mu$
Re'	=	modified Reynolds number. $Re' = G_o D_p / \mu (1 - \epsilon)$.
S	=	bed section, m^2 .
T	=	air temperature, $^\circ C$.
T_1, T_2	=	air temperature at inlet and outlet, respectively, $^\circ C$.
T_s	=	temperature of solids, $^\circ C$.
t_c	=	critical time, s.
v_o	=	superficial air velocity, m/s.
v_{mf}	=	superficial air velocity at minimum fluidization conditions, m/s.
v'	=	interstitial air velocity, m/s.
v'_{mf}	=	interstitial air velocity at minimum fluidization conditions, m/s.
y	=	vertical coordinate, m.

Greek Symbols

δ	=	bubble fraction in the bed.
ϵ	=	bed voidage.
ϵ_{mf}	=	bed voidage at minimum fluidization conditions.
μ	=	air viscosity, kg/ms .
ρ	=	air density, kg/m^3
ρ_s	=	density of solids, kg/m^3 .

REFERENCES

- BARKER, J. J. 1965. Heat transfer in fluidized bed. *Ind. Eng. Chem.* 57 (5), 33.
- BHATTACHARYYA, D. and PEI, D.C.T. 1974. Fluid to particle heat transfer in fluidized bed. *Ind. Eng. Chem. Fund.* 13 (3), 199.
- BRADSHAW, R. D. and MYERS, J. E. 1963. Heat transfer and mass transfer in fixed and fluidized bed of large particles. *AIChE J.* 9 (5), 590.
- BIRD, B. R., STEWART, W. E. and LIGHTFOOT, E. N. 1964. *Fenómenos de Transporte*. John Wiley & Sons, Inc., New York.
- CHANG, T. M. and WEN, C. Y. 1966. Fluid to particle heat transfer in air fluidized beds. *Chem. Eng. Prog. Symp. Series.* 62 (67), 111.
- DAVIDSON, J. F. and HARRISON, D. 1963. *Fluidized Particles*. University Press. Cambridge.
- DAVIDSON, J. F. and HARRISON, D. 1971. *Fluidization*. University Press. Cambridge.
- GUTFINGER, C. and ABUAF, N. 1974. *Advances in Heat transfer*, pp. 10, 167, Academic Press, New York.
- HEERTJES, P. M. and MCKIBBINS, S. W. 1956. The partial coefficient of heat transfer in a drying fluidized bed. *Chem. Eng. Sci.* 5, 161.
- KETTERING, K. N.; MANDERFIELD, E. L. and SMITH, J. M. 1950. Heat and mass transfer in fluidized systems. *Chem. Eng. Progress.* 46 (3), 139.
- KUNII, D. and LEVENSPIEL, O. 1969. *Fluidization Engineering*. John Wiley & Sons, Inc., New York.
- LINDAUER, G. 1967. Heat transfer in packed and fluidized beds by the method of cyclic temperature variations, *AIChE J.* 13 (6), 1181.
- MANN, R. S. and FENG, L. C. L. 1968. Gas-solid heat transfer in fluidized beds. *Ind. Eng. Chem. Proc. & Development.* 7 (3), 327.
- PERSSON, P. 1967. Fluidizing technique in food freezing. *ASHRAE J.* 9, 42.
- WALTON, J. S., OLSON, R. L. and LEVENSPIEL, O. 1952. Gas-solid film coefficients of heat transfer in fluidized coal beds. *Ind. Eng. Chem.* 44 (6), 1974.
- WAMSLEY, W. and JOHANSON, L. N. 1954. Fluidized bed heat transfer. *Chemical Eng. Prog.* 50 (7), 347.

LITERATURE ABSTRACTS

ABSTRACTS FROM JOURNAL OF FOOD SCIENCE

AN EQUATION CORRELATING SHELF LIFE OF DEHYDRATED VEGETABLE PRODUCTS WITH STORAGE CONDITIONS. R. Villota, I. Saguy, M. Karel. *J. Food Sci.* 45, 398–399 + 401 (1980).

A correlation of time of failure and storage conditions was established for two classes of dehydrated products, namely vegetables and vegetable powders. Mathematical models developed were based on data reported in the literature for products stored in air or nitrogen with failure due to off-flavors or changes in color. A general mathematical model, in accord with theoretical considerations was found to be applicable in all cases studied.

MODELING AND SIMULATING A DISTRIBUTED PARAMETER TUNNEL-DRIER. R. Bertin, F. Pierronne, M. Combarous. *J. Food Sci.* 45, 122–125 + 137 (1980).

In this paper a distributed parameter model is proposed for the discontinuous drying operation of a tunnel-drier. This model is more complete than the previous ones in which continuous loading is currently assumed. The model is used for spherical fruits placed in a California tunnel of the parallel flow type. Numerical results are in good agreement with experimental information. The model can be used for optimum designing and automatic control.

LABORATORY FOOD EXTRUSION—THE DESIGN OF A HORIZONTALLY SPLIT BARREL. D. Paton, G. D. Robertson, G. E. Timbers, W. A. Spratt. *J. Food Sci.* 45, 224–227 (1980).

An extruder barrel has been designed and constructed of two identical half-barrels, joined along a horizontal axis. This arrangement allows a rapid disassembly to expose with minimum disruption the barrel's contents. Phase transition zones can be identified, measured and photographed and found to vary with the process conditions. This novel split-barrel greatly facilitates the study of internal events during food extrusion.

COMPUTER-AIDED PREDICTION OF BEET PIGMENT (BETANINE AND VULGAXANTHIN-1) RETENTION DURING AIR-DRYING. I. Saguy, I. J. Kopelman, S. Mizrahi. *J. Food Sci.* 45, 230–235 (1980).

A mathematical model describing the drying behavior of beet slices was developed based on data from an air-drying experiment which measured average moisture content and temperature. The model was based on application of Fick's law for unidimensional moisture flow in the falling drying rate period. The mathematical model and the data generated from the kinetic studies of temperature and moisture sensitive red beet pigments (betanine and vulgaxanthin-1) were combined in a computer program to simulate and predict beet pigment retention, as a function of the process variables. Predicted and actual experimental pigment retention agreed well, which indicates this approach's potential for simulating deterioration of beet pigments during air-drying.

EFFECT OF HEAT TREATMENT ON THE QUALITY OF COOKED CARROTS. K. Paulus, I. Saguy. *J. Food Sci.* 45, 239–241 + 245 (1980).

We investigated the effect of heat treatment on the quality of cooked carrots (Texture, chemical, and sensory evaluation). Slices of three carrot varieties (Rubika, Kundulus, and Rothild) were cooked at 90–120°C for 2–70 min. Compressive failure stress and rupture work were measured by a Zwick Universal Testing Machine. Textural softening could be expressed by an exponential equation of the type: $\delta = A \exp(-kt)$, where δ = rupture stress, t = cooking time, and A and k are constants. Plotting k versus $1/T$ (T = cooking temperature, °K) revealed an Arrhenius-type relationship with apparent activation energy of about 28, 27, and 22 Kcal/mole for Rubika, Rothild, and Kundulus varieties, respectively. Chemical analysis (dry matter and β -carotene) showed that the cooking process caused insignificant changes. Statistical analysis of the sensory assessments showed an overall significant preference for 3 mm slice thickness and 110°C cooking temperature.

ENERGY CONSUMPTION AND THE POTENTIAL FOR CONSERVATION IN A SPRAY-DRIED COFFEE PLANT. M. Okada, M. A. Rao, J. E. Lima, M. Torloni. *J. Food Sci.* 45, 685–688 (1980).

Consumption of direct energy sources in a plant producing spray-dried coffee amounted to 12,930 kcal/kg of instant coffee (kg 1C). Thermal energy sources accounted for 95% of the energy and electricity for the rest. Spray-drying was the most energy intensive operation, requiring 5,040 kcal/kg 1C. Thermal energy consumption (Kcal/kg 1C) for roasting, extraction, and concentration was 890, 2,030, and 1,780, respectively. Thermal energy losses in the unit operations amounted to 55% of that consumed. Heat recovery and insulation techniques can be used to conserve about 70% of the losses.

METHODOLOGY FOR ENERGY RECOVERY FROM LOWER TEMPERATURE EXHAUST AIRSTREAMS OF FOOD PLANTS. M. E. Lazar, D. F. Farkas. *J. Food Sci.* 45, 689–691 (1980).

A cross flow plate air to air heat exchanger module of simplified design was constructed to obtain cost and operating data. The fabrication of corrugated metal spacers was found to be a major cost. Therefore design criteria were developed for a simple "do-it-yourself" corrugation system. Extrapolation of prototype costs to full size installations shows that a payback of 6 months or less is possible for typical food plant installations.

A METHOD FOR ESTIMATING THERMAL DIFFUSIVITY OF HEAT CONDUCTION FOOD IN A CYLINDRICAL CAN. J. I. Uno, K. I. Hayakawa. *J. Food Sci.* 45, 692–695 (1980).

A series of theoretical formulae were derived to simulate heat conduction in a finite cylinder whose surface heat transfer conductances on top, bottom, and side surfaces were neither infinite nor identical to each other. These formulae were utilized to determine thermal diffusivity of heat conduction food tilled in a cylindrical can and surface conductances on its surfaces. Error analysis shows 1mm deviations in locations, where the temperatures are monitored, as well as

similar deviations in the height and radius of a cylinder given significant variations in the thermophysical property values. It is also shown that these variations may be significantly reduced if one uses a cylinder whose height and radius are 100 mm or greater. Commercially processed apple sauce and 8% bentonite suspension were filled into cylindrical cans of two different sizes. These cans were then subjected to heat treatments in a still retort. Temperature data collected during these treatments were then utilized to estimate the thermophysical properties of sample materials. There was fair agreement between those estimated and respective literature values.

HIGH VACUUM FLAME STERILIZATION: PROCESS CONCEPT AND ENERGY USE ANALYSIS. P. A. Carroad, S. J. Leonard, J. R. Heil, T. K. Wolcott, R. L. Merson. *J. Food Sci.* 45, 696–699 (1980).

A flame sterilization procedure which has the advantages of high vacuum, high temperature, and short time processing for commercially sterilizing canned foods is described. By virtually eliminating the covering liquid needed in commercial packs, the high vacuum flame sterilization process yields a high quality product without diluting natural flavor attributes, without needing to add sugars or salts. Examining the energy efficiency of the process, on a pilot scale, high vacuum flame sterilization is seen to be 30% less consumptive of thermal energy on an energy per mass of fruit processed basis than conventional retorting.

DYNAMICS OF ENHANCED PROTEIN ULTRAFILTRATION USING AN IMMOBILIZED PROTEASE. S. S. Wang, B. Davidson, C. Gillespie, L. R. Harris, D. S. Lent. *J. Food Sci.* 45, 700–702 (1980).

An industrial grade protease was immobilized on a prototype, non-cellulosic tubular ultrafiltration membrane using a purely physical technique based on vacuum absorption and pH adjustment. A supermarket grade of nonfat dry milk was used to prepare a 0.001 g/ml feed solution which was pumped steadily through the tubular membrane module at 114 L/min. The temperature of the retentate stream in the module was maintained at 50°C and 276 kPag (40 psig) respectively. The prototype system was operated continuously for 240 h under total recycle conditions and the control (i.e., system without the enzyme) was similarly operated for 50 h. The degree of flux enhancement of the prototype system relative to the control was impressive. It was estimated that over a period of 240 h, 93% more permeate was collected for the prototype than for the control. A gradient, lumped parameter mass balance model was formulated for the dynamics of the protein gel layer. The model was simulated on a digital computer, and the various mass transfer and enzyme kinetic parameters were estimated using a Marquardt nonlinear least squares routine. The resulting curve-fit was satisfactory, considering the complexities of the process and the simplifying assumptions used in the model development.

CONCENTRATION OF WATERMELON JUICE. S. S. Huor, E. M. Ahmed, R. D. Carter. *J. Food Sci.* 45, 718–719 (1980).

A high temperature-short time process was used to produce a 65° Brix concentrate from watermelon juice. Color of the reconstituted juice was similar to color of

the fresh juice. The concentrate could be stored at -21°C for 18 months without color loss. The dominant pigment in the concentrate is lycopene and it could be extracted in a relatively pure form. Watermelon concentrate showed a pseudoplastic behavior with apparent viscosities lower than orange and grapefruit juice concentrates within the shear rates of 2–100 sec^{-1} .

ENERGY USE QUANTIFICATION IN THE CANNING OF CLINGSTONE PEACHES. P. A. Carroad, R. P. Singh, M. S. Chhinnan, N. L. Jacob, W. W. Rose. *J. Food Sci.* 45, 723–725 (1980).

Thermal and electrical energy usages in a California clingstone peach cannery were monitored to establish energy consumption levels and to identify energy intensive unit operations. Representative processing lines were instrumented with orifice plates for steam flow measurements and with electric watt transducers for electrical power monitoring to yield actual consumption values. Steam use in the lye peelers and retorts accounted for 98% of all energy consumed, indicating that further examination of these operations for modification or equipment redesign should be a first priority for energy conservation. Pumping of liquids and dry conveying of fruit and waste material accounted for 61% of all electrical energy use. Electrical energy use amounted to only 2% of the total energy used, however, on an equivalent fossil fuel use basis.

PILOT SCALE ULTRAFILTRATION OF CLARIFIED ALFALFA JUICE. B. E. Kunkles, R. H. Edwards, R. E. Miller, G. O. Kohler. *J. Food Sci.* 45, 730–734 + 739 (1980).

Pilot scale ultrafilter units were tested for use in concentrating and purifying soluble alfalfa leaf protein solutions. Tubular and hollow fiber type units with non-cellulosic membranes were employed. Operating temperature was generally maintained at 10°C because of microbial growth and precipitation of the heat labile protein at higher temperatures. Higher permeation rates and lower protein losses were obtained with the tubular unit. Concentrating the clarified alfalfa juice to 1/10 of the original volume produced protein concentrates containing about 50% crude protein and 10% ash. Diafiltration of the clarified alfalfa juice until the permeate volume was 10 times the original sample volume resulted in products containing 70–76% protein. Although diafiltration removed more than 86% of the orthodihydroxy phenolic compounds, the dried protein products were tan colored.

ENERGY ACCOUNTING IN CANNING TOMATO PRODUCTS. R. P. Singh, P. A. Carroad, M. S. Chhinnan, W. W. Rose, N. L. Jacob. *J. Food Sci.* 45, 735–739 (1980).

An energy accounting method was used to determine energy consumption in various unit operations in canning tomato juice, whole-peeled tomatoes, and tomato paste. Data on steam and electric consumption were obtained from a canning plant with the use of steam flow meters and electric transducers. Unit operations associated with the following equipment were investigated: crushers, hot-break heaters, pulpers, finishers, lye-bath peelers, evaporators and retorts. The average thermal and electrical energy intensities of canning tomato products were 538 Btu

and 0.0126 kWh per pound of tomatoes received, respectively. Energy intensive operations were identified as those associated with hot-break heaters, lye-bath peelers, evaporators, and retorts.

INSTRUMENT FOR TEXTURE OF SMALL CURD COTTAGE CHEESE AND COMPARISON TO SENSORY EVALUATION. C. A. Perry, P. A. Carroad. *J. Food Sci.* 45, 798—801 (1980).

The texture of washed and unwashed small curd cottage cheese is evaporated using the Instron Universal Testing Machine and Ottawa Texture Measuring System with perforated plate insert. Results indicate that such an instrumental system quantifies cottage cheese texture with excellent correlation to sensory evaluation of firmness.

PREDICTION OF WATER ACTIVITY OF AQUEOUS SOLUTIONS IN CONNECTION WITH INTERMEDIATE MOISTURE FOODS: EXPERIMENTAL INVESTIGATION OF THE a_w LOWERING BEHAVIOR OF SODIUM LACTATE AND SOME RELATED COMPOUNDS. J. Chirife, C. Ferro Fontan. *J. Food Sci.* 45, 802—804 (1980).

The water activity lowering behavior of sodium lactate, lactic acid, and other related hydroxy acids as well as its sodium salts, has been experimentally determined. The results for the sodium salts were very well fitted to Bromley's (1973) model for a_w in aqueous solutions of strong electrolytes, while the a_w data for the acids were very well described by Nerrish's (1966 equation. It was found that sodium lactate does not present any exceptional negative deviation from Raoult's law as it has been reported in the literature. The B number for sodium lactate ($10.050 \text{ kg/mol}^{-1}$) was found to be somewhat lower than that corresponding to sodium chloride. Moreover, lactic acid was found to have a small positive deviation from Raoult's law.

TEMPERATURE DEPENDENCE OF SENSORY QUALITY CHANGES DURING THERMAL PROCESSING. T. Ohlsson. *J. Food Sci.* 45, 839—847 (1980).

For the heat inactivation of microorganisms the temperature dependence is markedly higher than for the changes of sensory or nutritional quality during thermal processing. With the aid of the cook-value concept, the temperature dependence of the changes of various sensory attributes has been determined. Fish, meat, and vegetable products were processed in thin cans to different quality levels at different temperatures. Values from sensory analysis and color measurements were used to determine the rates of quality changes, and their temperature dependence, which ranged from $13\text{--}34^\circ\text{C}$ (average 23°C) for a tenfold shift in the rate of quality changes. The results are compared with literature values. Applications of the results are discussed.

SIMULATION OF THE THERMAL PROPERTIES OF CREAM STYLE CORN. P. K. Hardt, H. W. Adams, J. M. Henderson. *J. Food Sci.* 45, 840—843 + 847 (1980).

This investigation has identified a synthetic mixture that will model the heating characteristics and viscosity of cream style corn.

OPTIMAL STERILIZATION TEMPERATURES FOR FLAT CONTAINERS.
T. Ohlsson. *J. Food Sci.* 45, 848-852 + 859 (1980).

To reduce deterioration of the sensory and nutritional quality during heat sterilization, container geometries favorable for a rapid heat transfer e.g. retort pouches, should be chosen. Also processing conditions for optimal quality retention should be used. For solid foods packed in flat containers, calculations were performed at different sterilization temperatures and processing conditions. To describe the changes in sensory and nutritional properties a simple term, the C-value, was used. Optimal sterilization temperatures were chosen where the minimal quality changes (minimal C-values) were found. The calculations show that larger thickness means lower optimal temperature range. Low initial temperature or long retort come-up-time has only minor influence. A generalization of the results of the calculations is presented.

CORRECTION FACTOR OF COME-UP HEATING BASED ON CRITICAL POINT IN A CYLINDRICAL CAN OF HEAT CONDUCTION FOODS. J. Uno, K. Hayakawa. *J. Food Sci.* 45, 853-859 (1980).

A computerized procedure was developed for estimating the critical point correction factor, Cf, of come-up heating in thermal processes, which are applied to cylindrical cans of heat conductive foods. Sterilizing values at the thermal center of the can were used as a criterion for this estimation. Fifteen parameters are required to uniquely determine Cf values. These parameters include z value, thermo-physical properties, and can dimensions, as well as operational conditions. Through a dimensional analysis, ten dimensionless parameters were selected to estimate Cf values. A fractional central composite experimental design was applied to evaluate statistically the influence of each parameter on Cf. Through the evaluation, several regression equations were developed for their practical uses. A series of heat conduction experiments were conducted by using several different sizes of cans filled with food simulant, 8% bentonite suspension. Fairly good agreement was found between predicted and observed correction factors.

COMPRESSIVE DEFORMATION PATTERNS OF SELECTED FOOD POWDERS. R. Moreyra, M. Peleg. *J. Food Sci.* 45, 864-868 (1980).

The mechanical behavior of selected food powders was studied by an Instron UTM through compression, decompression, and stress relaxation tests. Powders in moist or cohesive state were considerably more deformable than in dry or noncohesive state. Upon decompression, cohesive powders remained compacted while the noncohesive powders returned to a free flowing state. This phenomenon was manifested in the ratio between recoverable and nonrecoverable work. Moisture also modified the stress relaxation pattern of the compact in a way that could be explained in terms of reorientation of liquid bridges and the plasticity of the bed solid matrix.

HEATING CHARACTERISTICS OF CONDENSED CREAM OF CELERY SOUP IN A STERITORT: HEAT PENETRATION AND SPORE COUNT REDUCTION. M. R. Berry Jr., J. G. Bradshaw. *J. Food Sci.* 45, 869-874 + 879 (1980).

The sterilization value (F_o) and heat penetration factors (f_h and j) were determined from time/temperature data as a function of container headspace, reel speed, and product consistency for cans of condensed cream of celery soup heated in an FMC Steritort. The product was formulated in the laboratory, and water was added to control consistency. Two commercial instruments were used to measure product consistency. Fill weight (headspace) was the most critical processing parameter for the simulated Sterilmatic (individual-serving-size cans) and Orbitort (institutional-size cans) processes. As the headspace bubble was eliminated, the product was heated by conduction in a manner similar to a still process. The degree of agitation of the product was also directly affected by reel speed and consistency. The integrated sterilization value (IS) was determined from the spore count reduction technique of process determination for selected processes and normally exceeded F_o . The difference was as much as 6.9 min in the 603 × 700 can.

ROLE OF NITRITE IN CURED MEAT FLAVOR: ANTIOXIDANT ROLE OF NITRITE. B. MacDonald, J. I. Gray, L. N. Gibbins. *J. Food Sci.* 45, 893–897 (1980).

The antioxidant activity of nitrite was investigated in model systems containing linoleic acid, Tween 20, and phosphate buffer. Results indicated that nitrite by itself can act as a prooxidant especially at concentration greater than 25 mg/kg (parts per million). However, the addition of nitrite to model systems containing prooxidants such as Fe^{2+} or $Fe^{2+}EDTA$ substantially reduced the rates of oxidation. Lipid oxidation catalyzed by aqueous beef extract also showed a marked decrease upon the addition of nitrite. Nitrite also produced a significant effect ($p < 0.05$) on heme-catalyzed lipid oxidation although the nature of the effect was not established. Further studies with an aqueous extract of pork demonstrated that dialysis removes a fraction which is largely responsible for the catalytic effect of meat extracts on lipid oxidation. Trace metal analysis revealed the presence of iron in the dialyzates from these extracts. Nitrite may function as a metal chelator to tie up these trace metals present in meat.

BLADE TENDERIZATION EFFECTS ON BEEF LONGISSIMUS SENSORY AND INSTRON TEXTURAL MEASUREMENTS. L. H. Hayward, M. C. Hunt, C. L. Kastner, D. H. Kropf. *J. Food Sci.* 45, 925–930 + 935 (1980).

Blade tenderization effects on longissimus steaks from 112 steers fed rations differing in energy levels and feeding times were studied using Instron measurements and sensory evaluations for detecting textural changes. Ribeye steaks were removed before tenderization and after one pass through a blade tenderizer. Sensory panel scores for myofibrillar and overall tenderness significantly increased and detectable amounts of connective tissue decreased with blade tenderization. Juiciness and flavor scores were not significantly affected by tenderization. Cooking loss increased 2% for blade tenderized steaks. Blade tenderization had little effect on textural differences of steaks from cattle fed different nutritional regimens other than improving connective tissue and overall tenderness scores for steaks from the control and submaintenance cattle groups. Blade tenderization significantly lowered Instron adhesion peak force and numerous Instron Warner-Bratzler measurements (e.g. initial-yield force, peak force, initial-yield angle, total work done). Instron compression values were reduced by blade tenderization.

PREPARATION OF A QUICK-COOKING BROWN RICE PRODUCT USING A CENTRIFUGAL FLUIDIZED BED DRIER. R. L. Roberts, R. A. Carlson, D. F. Farkas. *J. Food Sci.* 45, 1080-1081 (1980).

A quick-cooking brown rice product was developed using the centrifugal fluidized bed drier ("CFB") concept. The high heat transfer rates in the CFB equipment provided a dry-quick-cooking brown rice in about 5 min drying time, after pre-cooking. Preparation time for the finished product was 10-15 min, about one-fourth that, required for raw brown rice. Protein, vitamin, and mineral contents were comparable to raw brown rice and higher than white rice. Selected processing conditions were successfully used on a continuous CFB unit, producing about 40 lb/h finished product. Yields were high as very few broken kernels were observed.

EVALUATION OF THE FLOWABILITY OF MELTED MOZZARELLA CHEESE BY CAPILLARY RHEOMETRY. C. E. Smith, J. R. Rosenau, M. Peleg. *J. Food Sci.* 45, 1142-1145 (1980).

Flow curves of mozzarella cheese at 30, 40, 55 and 70°C were determined by a capillary piston-driven rheometer. The shear stresses and rates were calculated using the Bagley and Rabinowitch correction formulae. The standard test and correction procedure was applicable to melted mozzarella cheese, but not to other types of cheese (notably cheddar) in which slippage and other effects played a dominant role.

THEORETICAL ANALYSIS OF THE RELATIONSHIP BETWEEN MECHANICAL HARDNESS AND ITS SENSORY ASSESSMENT. M. Peleg. *J. Food Sci.* 45, 1156-1160 (1980).

"Hardness" is mainly perceived through the forces that the fingers, jaws or tongue exert. Since the mechanical resistance of the tissues involved is often comparable in magnitude to that of the tested object, the mechanical sensitivity must decline at a level determined by the resistance of both tissue and object. Sensory response to mechanical stimulus starts at zero ends up in saturation or pain and ought to have certain mathematical properties. Superposition of the response intensity and the mechanical sensitivity reveals a theoretical relationship between mechanical hardness and its sensory assessment. This relationship implies that scatter in hardness sensory scores is theoretically inevitable and that there must be a certain hardness range for which each organ has maximum sensitivity.

CONVECTION OVEN FRYING, HEAT AND MASS TRANSPORT IN THE PRODUCT. C. Skjöldebrand, B. Hallström. *J. Food Sci.* 45, 1347-1353 (1980).

The different origins of water losses from frying meat loaves in a specially designed convection oven are discussed. A theory is presented for how water content inside the crust depends on the time temperature history of two different recipes. A model is developed for calculating water content profile in the crust and the thermal conductivity of the crust. Meat loaves of an industrial recipe were fried for 1 h at different temperatures, humidities, and velocity of the air. The water content of the crust was shown to be very low but increased steeply close to the

evaporating zone. The flow of water close to the evaporating zone could be charged when using different recipes of minced meat. For example, when adding potato starch the water-holding capacity (WHC) is increased and the main water losses occur as evaporation.

CONVECTION OVEN FRYING: HEAT AND MASS TRANSFER BETWEEN AIR AND PRODUCT. C. Skjöödebrand. *J. Food Sci.* 45, 1358—1362 (1980).

Heat and mass transfer between air and product, fried in a specially designed forced convection oven were analyzed. Minced meat loaves of an industrial recipe were fried for 1 h at different temperatures, humidities, and velocities of the air. It was shown that the moisture loss rate was larger in the beginning of the frying than at the end. The heat transfer coefficients varied between 20 and 90 W/m²C.

FLOW PROPERTIES OF POWDERED SHORTENINGS. G. W. Baker, L. A. Johnson, E. W. Lusas, V. E. Sweat. *J. Food Sci.* 45, 1370—1375 (1980).

Mass flow rate, angle of repose, and angle of slide were used to survey flow properties of powdered shortenings. Samples containing 75% fat were encapsulated using sodium caseinate or various vegetable protein isolates at levels of 2, 4, and 8% protein. Powdered shortenings prepared with soy, peanut, and cottonseed proteins were more flowable through an orifice. Powdered shortenings prepared with sodium caseinate were slightly more flowable down inclines. Protein with greater solubility produced powdered shortenings with better flow properties but capsule formation did not occur in the absence of protein. Scanning electron microscopy indicated powdered shortenings with smoother surface topographies had better flow down an incline but poorer flow through an orifice.

KINETICS OF WATER DIFFUSION AND STARCH GELATINIZATION DURING RICE PARBOILING. A. S. Bakshi, R. P. Singh. *J. Food Sci.* 45, 1387—1392 (1980).

A mathematical model describing simultaneous water diffusion and gelatinization of starch during rice parboiling is proposed. Soaking experiments were conducted on a short-grain rice variety. A computer-aided nonlinear optimization technique is used to find diffusion coefficients and reaction rate constants in the temperature range 50—120°C. It is concluded that the parboiling process is limited by the reaction of starch with water below 85°C. The activation energy was found to be about half in case of diffusion limited process than that of reaction limited process. The study showed that it is possible to reduce energy input if rice is dehusked before parboiling. The mathematical model shows good correlation between observed and predicted value of water uptake.

BULK DENSITIES OF COOKIES UNDERGOING COMMERCIAL BAKING PROCESSES. M. P. Hwang, K. Hayakawa. *J. Food Sci.* 45, 1400—1402 + 1407 (1980).

The densities of food products undergoing baking processes in commercial multi-zone band ovens of a local plant were measured with a volume replacement method along with a photographic procedure. Three kinds of cookies were selected as the

sample products. It was found that the densities are closely related to the temperature and moisture content of cookies during the baking processes. This relationship is statistically represented by polynomial regression equations. The correlation coefficients were better than 0.970. According to the examination of changes in volume of cookies undergoing baking processes expansion ratios were found to be 2.0—2.5.

YIELDS AND SOLIDS LOSS IN WATER AND STEAM BLANCHING, WATER AND AIR COOLING, FREEZING, AND COOKING OF BROCCOLI SPEARS. P. A. Carroad, J. B. Swartz, J. L. Bomben. *J. Food Sci* 45, 1408—1410 (1980).

Blanching and cooling process alternatives were compared for over-all product yield and solids content using broccoli spears. Blanching was in steam or water, the latter including fresh water and recycling water in a screw blancher system. Cooling was evaporative, with and without sprays or by flaming. Product was frozen in an air blast freezer and cooked. Solids content and yields were measured after each operation. Results quantify the lower solids loss of steam blanching compared to fresh water blanching, but indicate that recycling blanching water may result in even lower losses than steam blanching.

OXYGEN MASS TRANSFER EFFECTS ON THE DEGRADATION OF VITAMIN C IN FOODS. D. H. Mohr Jr. *J. Food Sci.* 45, 1432—1433 (1980).

In the aerobic oxidation of nutrients such as vitamin C the rate of oxygen mass transfer and the chemical reaction rate interact to determine the actual degradation rate. A mass transfer analysis has been employed to provide an explanation for a set of apparently anomalous experimental measurements of vitamin C degradation in an intermediate moisture food. This analysis also illustrates the importance of mass transfer effects in the measurement or prediction of aerobic degradation rates.

PREDICTION OF NUTRIENT RETENTION IN THERMALLY PROCESSED HEAT CONDUCTION FOOD PACKAGED IN RETORTABLE POUCHES. P. F. Castillo, J. A. Barreiro, G. R. Salas. *J. Food Sci.* 45, 1513—1528 (1980).

A model was developed to predict the retention of nutrients with first order kinetics of thermal degradation in foods packaged in retortable pouches and sterilized by thermal conduction. The validity of the model was verified experimentally on a simulated food. The model was effective in predicting the temperature at the center of the container at the end of the heating period and, to a lesser extent at the end of the cooling period. The prediction of the fraction of nutrient retained at the end of processing compared favorably to the fraction obtained experimentally, and they were within the 90% confidence intervals of the experimental fractions.

OPTIMAL STERILIZATION TEMPERATURES FOR SENSORY QUALITY IN CYLINDRICAL CONTAINERS. T. Ohlsson. *J. Food Sci.* 45, 1517—1521 (1980).

Optimization calculations are presented for sterilization of foods in cylindrical cans. The C-value is used to describe the sensory quality changes during processing.

The optimal temperatures for sensory quality were the same as for maximal thiamin retention. They were 116–119°C for conduction heated foods of commonly used can sizes. The calculations showed that variations in the retort temperature program or starting temperature have minor effects on the optimal temperatures. The results are generalized using the *t*-slope-value of heat penetration measurements. It is proposed that optimal sterilization temperatures may be determined for any product or can size using presented diagrams. Its practical application is exemplified.

FLUIDIZED BEDS AS TURBULENCE PROMOTERS IN THE CONCENTRATION OF FOOD LIQUIDS BY REVERSE OSMOSIS. R. De Boer, J. J. Zommerman, J. Hiddink, J. Aufderheyde. *J. Food Sci.* 45, 1522–1528 (1980).

Fluidized beds offer a potential improvement of reverse osmosis processes for food liquids less fouling of the membrane, and reduced energy consumption. Our experiments were concerned with tubular systems in which fluidized beds of glass, steel, and lead beads were used. Glass beads appeared to be preferable, since they caused little damage to the membrane. Only with the larger glass beads (3 min) did the membrane skin become corrugated, so that the rejection decreased. The permeate flux for Gouda cheese whey was almost equal to that of an empty tube for which the velocity was about thirty times higher. The erosive action of the glass beads on the fouling layer was partially responsible for this effect. For reverse osmosis of skim milk and potato fruit water the bed did not reduce the fouling layer to a sufficient extent and, therefore, had a lower permeate flux than an empty tube.

CONTINUOUS EXPLOSION—PUFFING OF APPLES. J. F. Sullivan, J. C. Craig Jr., R. P. Konstance, M. J. Egoille, N. C. Aceto. *J. Food Sci.* 45, 1550–1555 + 1558 (1980).

The explosion-puffing process produces excellent dehydrated apple products that can be used as crisp snacks, instant applesauce, and ingredients for pies, tarts, and cobblers. To make these products more competitive and facilitate commercialization of the process, a continuous explosion-puffing system (CEPS) has been developed. The CEPS process has been optimized for orchard run Rome Beauty apples. Leaching loss as well as SO₂ absorbance and processing losses were determined. Drying profiles from a continuous belt drier were made. Pressure, temperature, and feed moisture were used as control variables for CEPS optimization studies. Analytical tests for bulk density, rehydration, color differences, percent disintegration, and hydroxymethylfurfural were performed on each sample. From these responses a bias was discovered, necessitating a second study. The second optimization study of these responses produced an optimal region. Any point in this region gives a value of pressure, temperature, and moisture. When these conditions are used for CEPS an excellent apple product results. Orchard run Winesap apples were processed through CEPS at an optimal condition determined for Rome Beauty apples to evaluate varietal differences. While their response was significantly different, the product was acceptable.

ENERGY USE, COST, AND PRODUCT QUALITY IN PRESERVING VEGETABLES AT HOME BY CANNING, FREEZING, AND DEHYDRATION. F. Drew, K. S. Rhee. *J. Food Sci.* 45, 1561–1565 (1980).

Carrots, zucchini, and summer squash were processed by canning, freezing and dehydration according to procedures used in the home. Energy used was measured and the cost of preservation was calculated with the packaging and amortized equipment cost included. Reduced ascorbic acid and carotene contents were determined within a week after processing and after 6 months of storage. Sensory quality of the processed vegetables was evaluated in less than a month after processing. The vegetables were cooked with salt added for sensory evaluation. Freezing required the shortest processing time and resulted in products with the highest sensory quality, with ratings well above minimum acceptability, and the most ascorbic acid (assuming the canned liquor is not used) and carotene. More energy (electricity) was required to process and store vegetables by freezing than by canning or dehydrating with a commercially available electric dehydrator; however, the overall cost for freezing preservation was estimated to be as low as or lower than the cost for the other two preservation methods if the freezer could be kept full. With the freezer only half-full, freezing would cost more than canning or dehydration for a long term (>6 months) storage. Canning used much less electric energy and was slightly cheaper in overall preservation cost than dehydrating with the electric dehydrator, and the canned products retained much more ascorbic acid and carotene than the dehydrated products. Nearly all ascorbic acid was lost during dehydration. The canned and dehydrated products were rated low in sensory quality, with ratings mostly below minimum acceptability.

SOME FACTORS CONTROLLING THE KINETICS OF MOISTURE MOVEMENT DURING AVOCADO DEHYDRATION. S. M. Alzamora, J. Chirife. *J. Food Sci.* 45, 1649–1651 + 1657 (1980).

The effect of some factors such as initial oil content, blanching and freezing before drying, on the rate of moisture movement during avocado dehydration was studied. It was found that the diffusion coefficients of water in avocado were strongly affected by its oil content.

A STUDY OF THE WATER ACTIVITY LOWERING BEHAVIOR OF POLYETHYLENE GLYCOL IN THE INTERMEDIATE MOISTURE RANGE. J. Chirife, C. Ferro Fontan. *J. Food Sci.* 45, 1717–1719 (1980).

This work reports on the experimental determination of water activity (a_w) in aqueous solutions of polyethylene glycol 200, 400, and 600, respectively. The measurements were made in the a_w range of most interest to intermediate moisture foods formulation (about 0.80–0.95). The observed behavior was very well fitted by Norrish's (1966) equation and a simple relationship was found between Norrish's correlating constant and molecular weight of polyethylene glycols. It was found that this relationship may be explained through a theoretical analysis of the thermodynamic behavior of macromolecules in solution.

THERMAL DIFFUSIVITY OF MODEL MEAT ANALOG SYSTEMS. S.S.H. Rizvi, J. L. Blaisdell, W. J. Harper. *J. Food Sci.* 45, 1727–1731 (1980).

Model meat analogs composed of soy spun fiber, egg albumen, and wheat gluten were thermally processed at five different temperatures under transient heat transfer conditions. Thermal diffusivity values were obtained during this period using

Dickerson's method. The subject samples were held at the final temperature for 1 h and thermal diffusivity values were obtained using the method of Pflug *et al.* (1965); values obtained by the two methods matched favorably within the experimental variations. The thermal diffusivities of the system varied with water content in the temperature range 71.1–93.3°C. Above this range, water alone could not reliably predict the thermal diffusivity values. Increase in processing temperature lowered the thermal diffusivity values.

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

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EDITORIAL OFFICE: Prof. D. R. Heldman, Editor, Journal of Food Process Engineering, Michigan State University, Department of Food Science and Human Nutrition, East Lansing, Michigan 48824 USA.

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