TWIN SCREW EXTRUSION OF PEACH PUREE: RHEOLOGICAL PROPERTIES AND PRODUCT CHARACTERISTICS[1]

HULYA AKDOGAN[2] and TARA HABIG McHUGH

USDA, Agricultural Research Service,
Western Regional Research Center
800 Buchanan Street
Albany, CA 94710

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ABSTRACT

A semi-empirical, nonlinear model was developed to incorporate the effects of extruder operating conditions (temperature, moisture, and shear rate) on viscosity of peach puree in a lab-size twin screw extruder. A power law dependency for shear rate, an exponential dependency for moisture, and a modified first order transform function for temperature effects were employed. The proposed model fit the experimental data well. Motor torque and Specific Mechanical Energy (SME) were significantly affected by all extruder operating conditions. Experimental apparent viscosity varied from 40 to 130 Pa·s, depending on the operating conditions. Color indicatives (L, a, b) and extrudate density were mainly influenced by the moisture content. Water activity was only influenced by moisture.

INTRODUCTION

There is an ever-widening demand by the consumer for snack products with improved nutritive value and quality, novel shapes, tastes, colors, and textures. This demand can be met by restructured foods. Besides being a healthy food, restructured fruit products may also increase the utilization and consumption of fruits as value-added products, and as a new outlet for imperfect fruit (McHugh 1996). Restructured fruit pieces can also be used in baked goods and frozen

[1] Names are necessary to report factually on available data; however, the USDA neither guarantees nor warrants the standard of the product, and the use of the name by the USDA implies no approval of the product to the exclusion of others that may also be suitable.

[2] Author to whom the correspondence should be sent. Fax: 510-559-5851, Email: hulya@pw.usda.gov

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foods. By consuming 100% fruit products, consumers may add fruit to their diets without the inconvenience and cost of processing raw whole fruit (Hannigan 1983). The snack food and candy markets include many fruit-based snacks associated with the wholesomeness of the fruit, however, most are not available as 100% fruit. Such fruit snacks would be healthy alternatives to conventional fruit-containing confectionery snacks by being free of preservatives, stabilizers, added sweeteners, gelling agents, thickeners, salt, and lipids.

Extrusion makes formation of 100% fruit products possible, while, other manufacturing techniques must limit the concentration of fruit in the product. In conventional techniques such as starch molding, fruit content of the product is limited by the viscosity of the slurry. The increased viscosity at high fruit content cause flow problems when depositing fruit puree into the molding starch. However, extruders possess the capability to overcome this problem since they can handle highly viscous fruit melts at varying processing conditions (Moore 1989; McHugh 1996).

The literature on food extrusion mainly involves cereals. Since extrusion of 100% fruit products is quite novel, a literature search did not uncover many relevant information. The closest study found involved the inclusion of dried fruits such as prunes, figs, raisins, cranberries, and fruit juice concentrates, up to 7%, to extruded rice flour-based snack products (Maga and Kim 1989). The fruit and fruit juice added extrudates were more dense and slightly darker in color than the control (100% rice) and radial expansion was significantly lower upon fruit addition. A recent study by McHugh and Huxsoll (1999) investigated the effect of ingredients and extrusion conditions on product properties of peach and peach/starch gels. They found that moisture affected color and texture the most.

Extruder process parameters affect the viscosity of the melt and have a direct impact on the final product quality. Therefore, it is important to monitor the viscosity of the melt in the extruder. Extruder dies can be equipped with slit or capillary rheometers to measure the rheological properties on-line. The data obtained can be further used for rheological calculations. Several rheological models exist in the published literature; however, most of them are for starch-based or protein-based doughs (Cervone and Harper 1978; Harper 1981; Remsen and Clark 1978; Bhattacharya and Hanna 1986; Senouci and Smith 1988; Dolan *et al.* 1989; Wang *et al.* 1990; Altomare *et al.* 1992; Morgan *et al.* 1989; Mackey 1989; Lai and Kokini 1990). Rheological properties of fruit-based products are not expected to be the same as starch-based or protein-based products due to structural and compositional differences between fruits and starch or protein. Nevertheless, basic principles of rheology should still apply to fruit-based systems.

The objectives of this study were: (1) To analyze the rheological properties of peach puree extrudates by studying the interaction between the extruder operating conditions (moisture, temperature, and feed rate) and melt viscosity, die pressure drop, torque, and specific mechanical energy (SME); (2) To develop a semi-empirical model to describe viscosity changes of the peach puree melt; (3) To analyze product properties such as density, color, and water activity of peach puree extrudates as related to extruder operating conditions.

MATERIALS AND METHODS

Raw Material

Peach puree at 32° Brix was obtained from Sabroso (Medford, OR). Water content of the puree was reduced to approximately 4% final moisture by using a double drum dryer (Buflovak, Buffalo, NY) at 134C, 0.6 rpm, and a gap of 0.381 mm. This process of drum drying as a precursor to extrusion is the subject of a recent patent application by McHugh and Huxsoll (1999). Prior to extrusion experiments, drum dried puree was ground in a food processor (Model DLC-8S, Cuisinart, East Windsor, NJ) to decrease the particle size. Puree flakes were sieved with a U.S. Standard No. 60 sieve (opening size 250 μm) (Fisher Scientific, Santa Clara, CA) to eliminate the fines. The prepared material was placed in a glass jar and stored in a cold room at 5C. Prior to the extrusion, material was allowed to equilibrate at room temperature.

Extrusion

A lab-size corotating twin screw extruder (Haake-Leistritz Micro-18, Sommerville, NJ) was used to perform the extrusion experiments. A slit die/torque rheometer combination (Haake-Buchler, Paramus, NJ) was attached to the discharge end of the extruder to measure the rheological properties of the peach melt. The slit dimensions were 2.0 mm high and 20 mm wide. The extruder barrel was of modular design with six independent heating and cooling zones, each with an L/D of 5:1. Electrical heating and air cooling systems were used to maintain the barrel temperatures constant in the $\pm$1C range. Temperatures of each zone of the barrel and the slit die were set from the software and recorded together with the melt temperatures. Melt temperatures were measured by four thermocouples (Type J, Omega Engineering, Westport, CT) imbedded in the barrel, one in the die adapter piece, and one in the slit die/torque rheometer. Melt temperatures were also recorded by two thermocouples flush with the inside surface, one in the slit die and in the adapter piece. Deionized water was injected through the extruder barrel two barrel sections downstream from the feed port. A Bran+Luebbe metering pump (Model N-P31, Bran+Luebbe Inc., Buffalo Grove, IL) was used to pump water into the barrel. Three pressure transducers (Type PT422A, Dynisco, Sharon, MA) were used to measure the pressure along the slit die/torque rheometer. The pressure transducers were calibrated at each operating temperature for accurate readings. Two different pressure ranges of the transducers were chosen to increase the accuracy: 0-500 psi (3,447 kPa) and 0-1000 psi (6,895 kPa). The pressure transducers were located 3.81 cm from each other. Another pressure transducer was placed on the adapter piece (range: 0-1000 psi (6895 Pa)).

A K-Tron gravimetric feeder (Model K2V-T20, K-Tron Corp., Pitman, NJ) was used to meter the drum dried puree into the extruder. The first (feed port) and second sections of the barrel were unheated to avoid the sticking of puree particles to the walls of the feed port. The set temperatures for the rest of the barrel sections were kept approximately 10C cooler than the desired melt temperature at the die. Die melt temperatures of 30 to 60C were studied. Total feed rates (drum dried puree and water) were varied from 28 to 57 g/min. A constant screw speed was employed at 200 rpm.

A computerized data acquisition system (Rheocord 9000, Haake-Buchler, Paramus, NJ) was used to record set and melt temperatures, rotor speed, torque, and melt pressures along the die and on the adapter piece placed between the barrel and die. The data acquisition rate was every 6 s. Data collection was initiated after reaching the steady state (i.e. constant pressure and torque readings).

Color Analysis

A Minolta Chroma Meter (Model CR-200, Ramsey, NJ) was used to measure and compare Judd-Hunter L,a,b color values of the extrudates. L represents total lightness (100) or darkness (0) of the sample. Redness is represented by $+a$ values and greenness by $-a$ values. Yellowness is indicated as $+b$ values and blueness by $-a$ values. A standard white calibration plate was used to calibrate the colorimeter. Three measurements were taken at different locations on each sample.

Water Activity

Water activity of extrudates were measured using an AquaLab (Model CX-2, Pullman, WA) water activity meter per manufacturer's directions. All water activity measurements were made at 25C.

Bulk Density

Density was calculated by dividing the weight of the peach puree extrudate sample by its volume which was calculated directly from the dimensions of the sample as measured with a micrometer. Triplicate samples were determined at each operating condition.

Screw Configuration

The same screw configuration (Table 1) was used throughout the study. This configuration was selected to provide sufficient residence time and thorough mixing of the ingredients.

TABLE 1.

SCREW CONFIGURATION

element	axial length (mm)	pitch (mm) or stagger (°)
feed screw*	180	15
feed screw	90	10
feed screw	30	10
kneading block**	20	0°,-60°,60°,0°
kneading block	20	0°,-60°,60°,0°
feed screw	30	5
kneading block	20	0°,90°,0°,90°
kneading block	20	0°,-60°,60°,0°
kneading block	20	0°,30°,60°,90°
feed screw	20	5
kneading block	20	0°,30°,60°,90°
feed screw	60	5

*All feed screws are forward pitched.

**The kneading blocks are composed of four mixing disks.

Viscosity of the Feed Material

A concentric cylinder rheometer (Model DV-III, Brookfield Engineering Labs, Inc., Stoughton, MA) with a small sample adapter was used to measure viscosity of the rehydrated peach puree to 60° Brix. Temperature was regulated from 30C to 60C using a TC-500 water bath with programmable controller. Spindle speed was 0.1 rpm. Shear rate was held constant at 0.02 s^{-1}.

Moisture Content of the Feed Material

The moisture content of the feed was determined on a representative sample using AOAC Method No. 22.013 (A.O.A.C. 1970).

Statistical Analysis

PROC REG and PROC NLIN procedures on SAS (1995) were used to analyze the experimental data.

RESULTS AND DISCUSSION

The influence of three different extruder operating conditions (moisture, temperature, and feed rate) on peach puree rheological properties and product color, water activity, and density were examined. Added moisture at three different levels, 22, 26, and 30% (wb), was studied. The extrudate had difficulty maintaining its shape upon exiting the die above 30% added moisture. Below 22% added moisture, extrudates appeared tough and light in color. Melt temperatures in the range of 30 to 60C were studied. At each moisture, and temperature at least five different feed rates were used in order to cover a reasonably wide shear rate range at the die wall. Below the lowest feed rate used (28 g/min) the degree of fill in the extruder was not high enough to develop a steady pressure drop at the die. Above 57 g/min backflow problems were experienced due to a high degree of fill in the barrel. Preliminary experiments showed that the effect of screw speed on die pressure drop and torque was negligible; therefore, screw speed was held constant at 200 rpm. Bhattacharya and Hanna (1987) also found the effect of screw speed insignificant upon extruding corn gluten meal.

Rheological Evaluations

For analysis of the slit die measurements, the flow was assumed isothermal, laminar, and fully developed, thereby implying that the effect of viscous shear energy dissipation on temperature and the end effects could be neglected. The rheological calculations were based on linear pressure profiles, so no end effect correction was made. The edge effect was also neglected due to the fact that width to height ratio of the slit die was 10 (Akdogan *et al.* 1997). Pressure measurements along the slit rheometer for three different feed rates are shown in Fig. 1 (30% moisture and 30C temperature). All other cases exhibited an equally good linear decrease of pressure with the distance along the slit die.

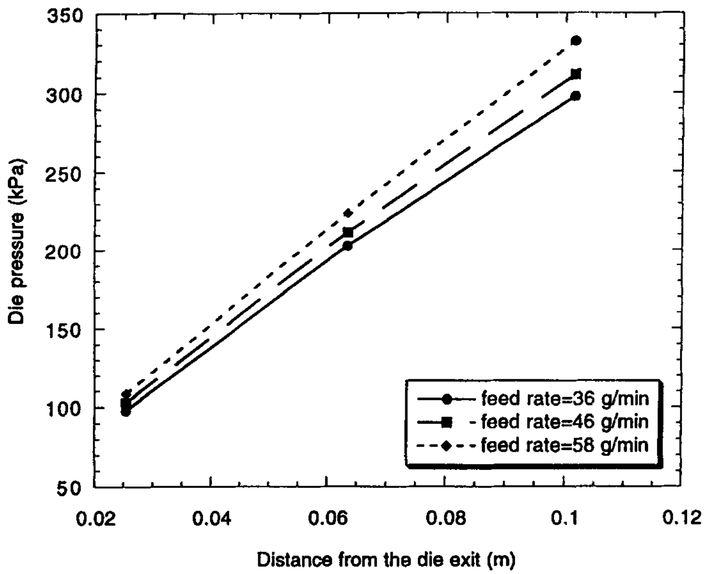

FIG. 1. PRESSURE GRADIENTS ALONG THE SLIT DIE AT DIFFERENT FEED RATES (Moisture = 30%, temperature = 30C)

Shear stress and shear rate were calculated using Rabinowitsch-Mooney equation (Steffe 1996):

$$\sigma_w = \frac{\Delta Ph}{2L} \tag{1}$$

$$\dot{\gamma}_w = \frac{\Gamma}{3}\left(2 + \frac{d(\ln\Gamma)}{d(\ln\sigma_w)}\right) \tag{2}$$

and

$$\Gamma=\frac{6\dot{Q}}{wh^2} \tag{3}$$

where σ_w is the wall shear stress (Pa), Γ is the wall shear rate (1/s), $\dot{\gamma}_w$ is the corrected wall shear rate (1/s), ΔP is pressure drop along the slit die (Pa), $\dot{Q}$ is the volumetric feed rate (m^3/s), w is the width (m), and h is the height (m) of the slit die.

Measurements of the pressure drop and feed rate were converted to shear stress at the wall and shear rate at the wall, respectively, using Eq. 1 and 2. The apparent viscosity of the melt is simply the ratio of wall shear stress to corrected wall shear rate (Steffe 1996):

$$\eta=\frac{\sigma_w}{\dot{\gamma}_w} \tag{4}$$

where η is in Pa·s.

Apparent viscosity is related to shear rate with a power law expression:

$$\eta_{\dot{\gamma}}=K_{\dot{\gamma}}\dot{\gamma}^{n-1} \tag{5}$$

for constant temperature and moisture content, n is the flow behavior index, and $K_{\dot{\gamma}}$ is the shear rate dependent constant.

The slope of the regression line of Ln (wall shear stress) vs Ln (wall shear rate) (Fig. 2) represents the flow behavior index, n. For all temperature and moisture cases, n was less than 1 indicating that the peach puree melt exhibited shear-thinning (pseudoplastic) flow properties. The R^2 range for the rheograms varied from 0.95 to 1. The slope, n, varied between 0.24 to 0.38. The intercept of the regression line (Fig. 2) gives the consistency coefficient, K, which varied from 704 to 3197 Pa·s^n depending on the temperature and moisture.

The effect of moisture on viscosity has been described using an exponential dependency (Harper 1981):

$$\eta_M=K_M\left(e^{mM}\right) \tag{6}$$

where, K_M is the moisture dependent constant, M is the moisture content, and m is moisture constant. Figure 3 illustrates the moisture effect on apparent viscosity at five different temperatures. As moisture increases, viscosity decreases. This is due to difference in the concentration of water.

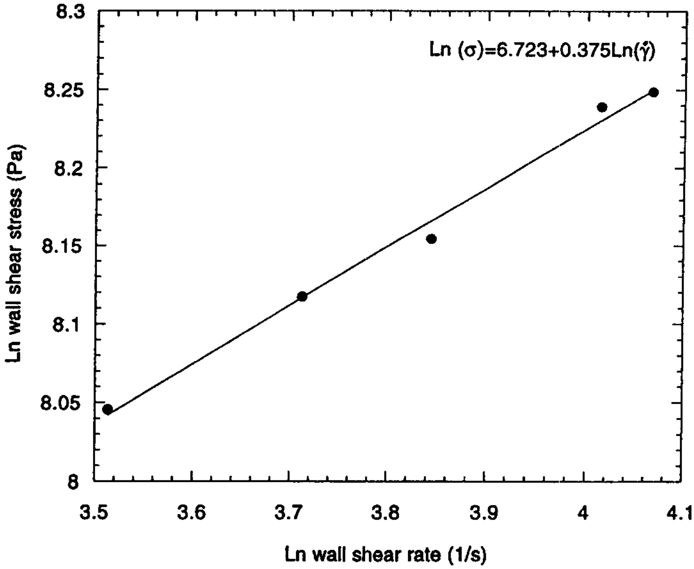

FIG. 2. RHEOGRAM FOR PEACH PUREE MELT
(Moisture = 30%, temperature = 60C)

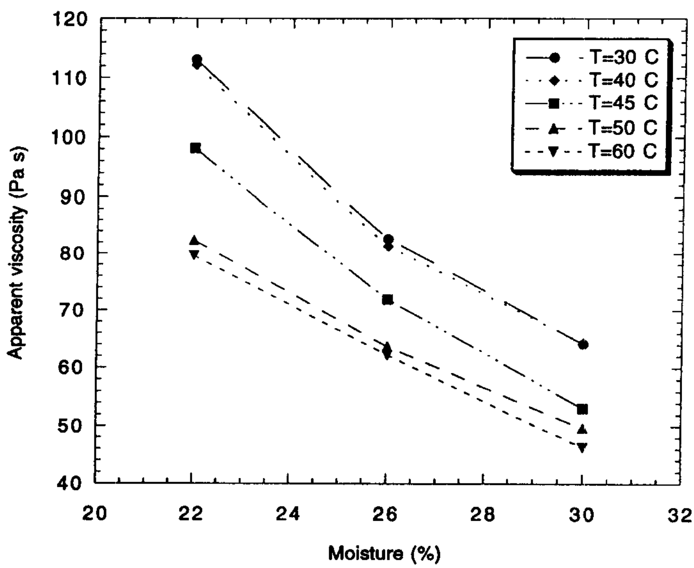

FIG. 3. EFFECT OF MOISTURE ON APPARENT VISCOSITY
(50 kg/h feed rate)

The temperature effect on viscosity showed an unexpected pattern. A considerable viscosity change was observed between 40 and 50C for all cases. The viscosity differences between 30 and 40C, and between 50 and 60C were much smaller. This pattern existed at all feed rates and moisture contents. Moreover, the lower the moisture in the system, the more pronounced the decline in melt viscosity between 40 and 50C. This is unlike many studies reported in the literature in which temperature data typically follows the Arrhenius type behavior (Saravacos 1970; Steffe 1996; Rao *et al.* 1981; Rao and Cooley 1983). However, their studies used fruit purees or concentrates lower than 65° Brix. In this study the effect of temperature on viscosity of peach puree is represented by a modified first order transfer function:

$$\eta_T = 1 - e^{-\frac{T_\Delta (T_\Delta - \beta)}{\alpha}} \tag{7}$$

where η_T is the normalized apparent viscosity (dimensionless), T is temperature (°C), T_{ref} is 30C, T_Δ is $T\text{-}T_{ref}$, α and β are constants. Apparent viscosity, η, is normalized accordingly:

$$\eta_T = \frac{\eta - \eta_i}{\eta_f - \eta_i} \tag{8}$$

where η_i is viscosity at T=30C, η_f is viscosity at T=60C, η is experimental viscosity at other temperatures (Pa·s). Figure 4 shows the experimental viscosity and viscosity predicted by Eq. 7 for all three moistures.

A separate experiment was conducted using a concentric cylinder rheometer to measure viscosity within the same temperature range and with 60° Brix material. More concentrated samples could not be handled due to the torque limitations of the viscometer. The temperature dependency of viscosity was found to be Arrhenius. An explanation for the discrepancy between the measurements of concentric cylinder rheometer and torque rheometer of the extruder might be due to the difference in mechanical energy imparted on the sample. A rather low shear rate (0.02 s^{-1}) was employed with the Brookfield rheometer. Whereas, higher shear rates and an aggressive screw configuration, with several kneading blocks and screw elements with narrow pitch, were used during extrusion. Mechanical energy input in the extruder barrel may have contributed to molecular breakdown of peach puree components (sugars and carbohydrate polymers) combined with relatively higher temperatures (>40C). These substances may lose their orderly structure and begin melting at elevated temperatures combined with high shear rates and mechanical energy input. More experimentation would be needed to verify this hypothesis.

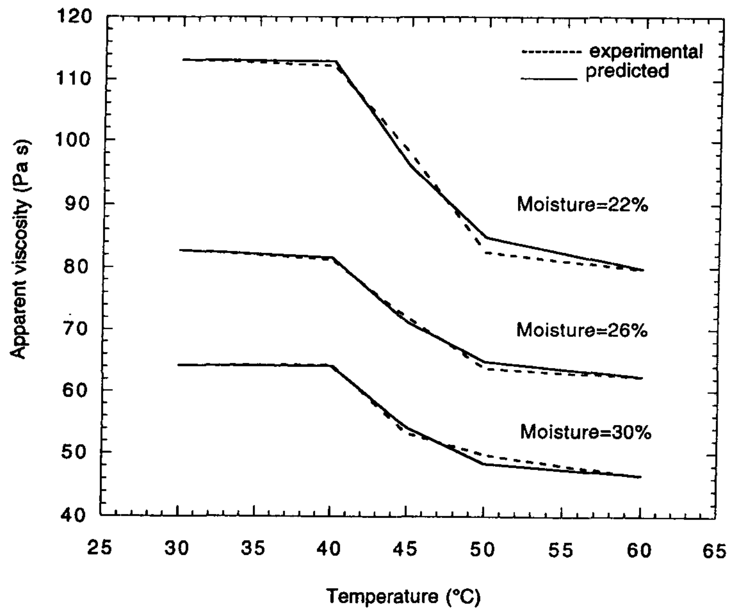

FIG. 4. PLOT OF APPARENT VISCOSITY AS A FUNCTION OF TEMPERATURE AT DIFFERENT ADDED MOISTURES
(50 kg/h feed rate)

Combining Eq. 5, 6, 7, and 8 yields the final viscosity model:

$$\eta = K(\dot{\gamma}^{n-1})(e^{mM})\left[(\eta_f - \eta_i)\left(1 - e^{-\frac{T_A(T_A-\beta)}{\alpha}}\right) + \eta_i\right] \tag{9}$$

The proposed model (Eq. 9) has an empirical basis; therefore, is valid only within the experimental range studied. A nonlinear regression analysis was performed on experimental data using the PROC NLIN procedure in SAS (1995) which uses least squares estimation. The Marquardt algorithm was chosen as the iterative method to determine the model parameter estimates. This required specifying the partial derivatives of the model with respect to each parameter but is a useful method when the parameter estimates are highly correlated, such as was the case with the rheological data. Parameter estimates of the model are reported in Table 2. Figure 5 shows the fit between the experimental apparent viscosity and viscosity predicted by the proposed model. The distribution along the 45° line indicates the degree of fit. The model's coefficient of multiple determination (R^2) was 0.91. The majority of the standardized residuals fell between $\pm$ 2 standard deviation range, therefore the error terms were considered normal. There was no systematic pattern in the residual distribution and the error variance appeared constant. No outliers were observed. Hence, it can be concluded that the model is appropriate for the experimental data.

TABLE 2.
PARAMETER ESTIMATES OF EQ. (9)

Parameter	Estimate
n	0.375
α, $(°C)^2$	152.31
β, °C	4.49
m	-0.48
K, s^{n-1}	18.35

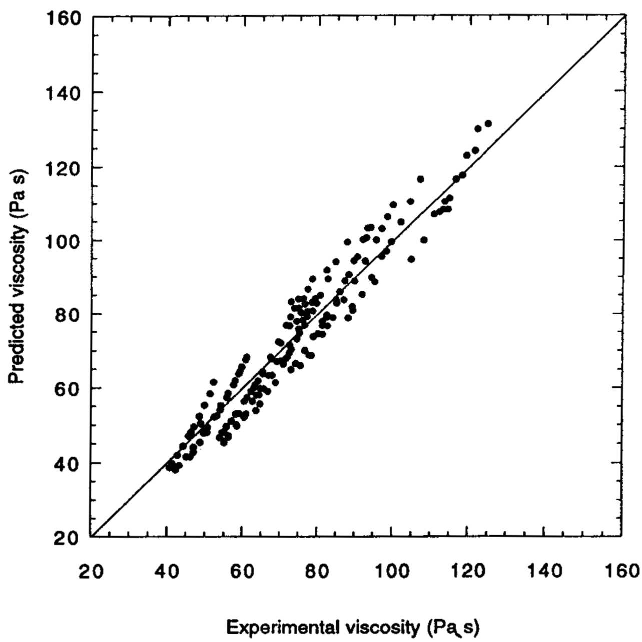

FIG. 5. PREDICTED VISCOSITY VS EXPERIMENTAL VISCOSITY OF PEACH PUREE MELT

The parameter estimates for n (0.375) and m (-0.48) were comparable to the literature values (Jao *et al.* 1978; Cervone and Harper 1978; Harper 1981; van Lengerich and Meuser 1987; Morgan *et al.* 1989; Lai and Kokini 1990; Altomare *et al.* 1992). However, their reported values were mostly for extruding cereal-based or protein doughs at higher temperatures. For peach puree, no values in the literature were available for temperature constants, a and b, because usually the temperature effect would fit an Arrhenius equation.

Power and Torque Levels

Power consumption should be significantly reduced upon extruding pure fruit products compared to cereals based on the results of Maga and Kim (1989) who reported torque values as high as 700 Nm for 100% extruded rice samples. For their 20% fruit added case, which was the highest amount of fruit added for their study, torque varied between 90 to 230 Nm, depending on the type of the fruit. Extrusion of 100% peach products did result in even lower torque values, 14 N·m being the maximum within the experimental range studied. Such low torque input to the system will result in less wear on the screw elements and barrel walls, cutting down on maintenance costs. This is especially important for a large scale production of similar fruit snacks. The nonabrasive nature of the fruit perhaps accounts for lower power consumption, hence lower motor torque. Also, low levels of power usage will lead to a lower manufacturing cost.

The measured torque or total torque, has two components: the torque required to work the product, net torque, and torque required to rotate the screws, the no-load torque. The no-load torque was determined by running the extruder with water only and consisted of 11 to 28% of the total torque depending on the operating conditions. The higher the moisture and temperature the lower the net torque. An increase in the feed rate while keeping the other process variables constant increased the torque. This can be attributed to an increase in the specific feed loading (SFL) in the barrel. SFL, which is a measure of the degree of fill in the extruder barrel, is defined as (Lu 1992):

$$SFL = \frac{\dot{m}}{60 \cdot N} \tag{10}$$

where SFL is the specific feed loading (kg/rev), $\dot{m}$ is the feed rate (kg/h), and N is the screw speed (rpm). Since the screw speed was kept constant at 200 rpm, feed rate increase resulted in an increase in SFL, hence an increase in torque.

A stepwise regression analysis was carried out with moisture, temperature, and feed rate as independent variables to determine the contribution of each variable on the variance of net torque which was the dependent variable. The PROC REG procedure with the stepwise selection option available on the SAS computer package (1995) was used. All process variables were found to be significant. Moisture contributed to the majority of the model R^2 (67.2%), whereas the temperature and feed rate contribution were 22.8%, and 10%, respectively. The regression model for motor torque is:

$$\text{Motor torque} = 44.8 - 44.6 \cdot M - 0.0946 \cdot T + 0.09 \cdot \dot{Q} \tag{11}$$

Specific Mechanical Energy (SME) Input

Torque also provides information to calculate specific mechanical energy (SME). SME may be defined as the energy provided by the motor drive to the material in the extruder per unit mass (Fichtali and van de Voort 1989):

$$\psi = \frac{60 \cdot 2\Pi \cdot \Phi \cdot N}{\dot{m}} \tag{12}$$

where Ψ is the specific mechanical energy (J/kg), Φ is the net torque (N$\cdot$m), N is the screw speed (rpm), and $\dot{m}$ is the feed rate (kg/h). Unlike torque, increasing feed rate causes SME to decrease, while keeping moisture and temperature constant. This is due to fact that SME is an indication of net energy input to the material per unit mass. According to the results of stepwise regression, the contribution of temperature and feed rate to the model R^2 was 31.1%, and 4.5%, respectively. The regression model for SME is:

$$\text{SME} = 1336.9 - 1135.7 \cdot M - 2.48 \cdot T - 1.276 \cdot \dot{Q} \tag{13}$$

Similar to the findings of torque, most of the contribution to model R^2 was due to moisture (64.4%). Lower moistures and lower temperatures correspond to higher SMEs.

Knowledge of process results is important for scaling-up purposes. The most difficult problem encountered in extrusion scaling-up operations is in obtaining the same product qualities on the large scale extruder as achieved on the pilot-plant or lab-size extruder. It is important to find a process parameter yielding a result that is independent of scale. The process parameters for extrusion scale-up could be shear rate, shear stress, energy input, residence time, product temperature, and die pressure drop. Valentas *et al.* (1991) reported that SME is a widely used primary scale-up criterion for a variety of products. Scalability is not the main focus of this research. More thorough experimentation and evaluation is recommended before using SME as scale-up criterion.

Product Properties

Physical properties of product are especially important for product acceptability by the consumer. Processing conditions usually have a significant impact on product properties. Quantification of the effect of extruder operating parameters, such as moisture, temperature, and feed rate, on the variance of product properties will aid to achieve an optimum product quality. Stepwise regression analysis from SAS (1995) was used in order to determine the contribution of each operating variable (moisture, temperature, and feed rate) on the variance of each extrudate property measured. Contribution of each variable to the model R^2 was quantified and reported. In most cases, interaction terms were not included in the model since no marked improvement was detected on R^2 upon including them. The variance inflation factors (VIF) were used to detect the presence of multicollinearity. VIF in excess of 10 is an indication of multicollinearity affecting the least squares estimates (Neter *et al.* 1990). VIF of all parameter estimates was 1.0, indicating that multicollinearity is not a problem.

Color

Product color characteristics are vitally important to consumers. Optimization of extrudate appearance requires a clear understanding of the effects of extrusion conditions on color.

Extrudate Lightness (L)

Regression was carried out with the three main variables, moisture, temperature, and feed rate. The R^2 for the model was 0.94. Moisture makes the highest contribution to R^2 of the extrudate lightness, 60.6%, compared to a 39.4% influence by temperature. The stepwise procedure rejected feed rate as insignificant in the model. Exclusion of feed rate from the model is also supported by Duncan's multiple range test ($\alpha=0.05$), which determined that there is no significant differences among extrudate lightness at different levels of feed rates at a given moisture and temperature. For Duncan's test, moisture, temperature, and feed rate were considered as treatments.

The regression model for L is:

$$L = 123.71 - 68.66 \cdot \mathrm{M} - 0.17 \cdot \mathrm{T} \tag{14}$$

Figure 6 illustrates the marked decrease in *L* as moisture and temperature increases. The highest *L* was obtained at 22% moisture and 30C temperature, and the lowest *L* was obtained at 30% moisture and 60C. The effect of moisture might be attributed to dilution effects. The temperature effect is perhaps due to browning of peach puree sugars and carbohydrate polymers, such as pectin and cellulosic substances, as they melt or undergo structural changes as temperature increases.

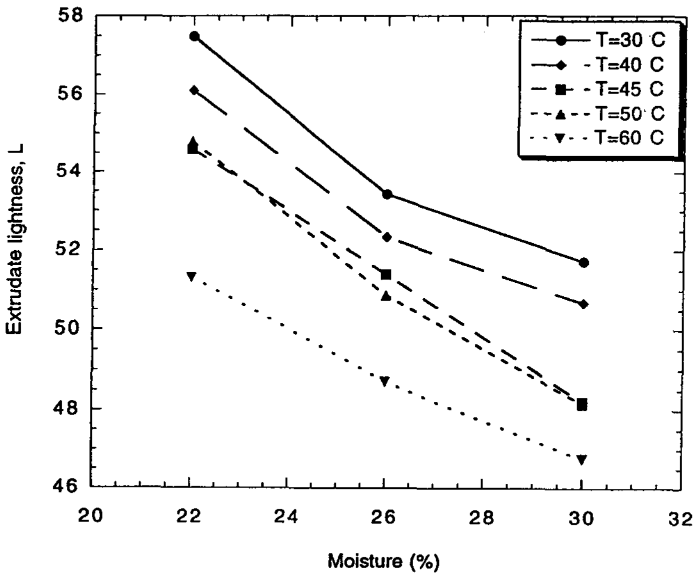

FIG. 6. EFFECT OF MOISTURE ON EXTRUDATE LIGHTNESS, *L*, AT DIFFERENT TEMPERATURES

Extrudate Redness (*a*)

The *a* values indicate redness of the sample with positive values. Values of *a* varied from +13.50 to +16.42 for the extruded samples, depending on the moisture and temperature. A stepwise regression analysis with main and first order interaction terms was performed on all experimental data. This revealed that, similar to lightness case, feed rate did not influence the variation in *a*. Moisture-temperature interaction influenced the model R^2 the most (95.2%). R^2 of the model was 0.86. The regression model for *a* is:

$$a = 2.562 + 20.53 \cdot M + 0.0228 \cdot T \qquad (15)$$

Extrudate redness was found to be directly proportional to moisture and temperature (Fig. 7). An increase in any of these variables resulted an increase in a. Possibly as temperature increased, a increased due to thermally induced reactions involving peach puree components.

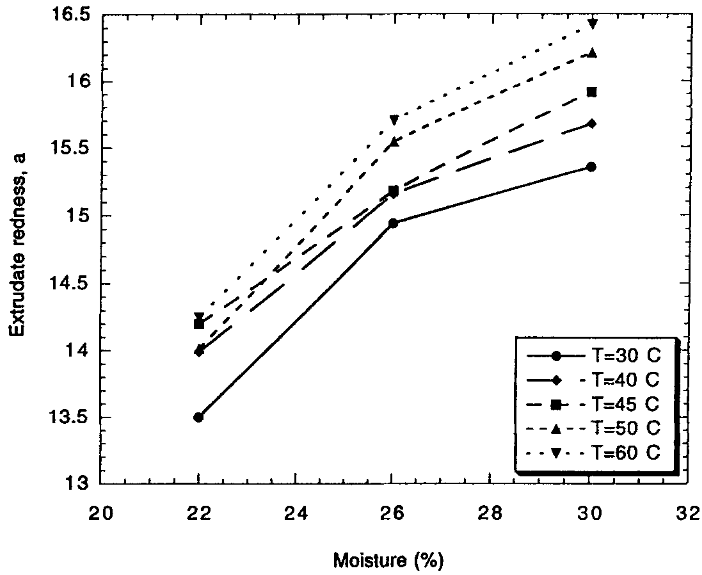

FIG. 7. EFFECT OF MOISTURE ON EXTRUDATE REDNESS, *a*, AT DIFFERENT TEMPERATURES

Extrudate Yellowness (*b*)

Values of b demonstrated a similar pattern to extrudate brightness, L. Stepwise regression analysis of b also did not include feed rate as another variable to the model. Moisture was found to be the most influential variable on the model R^2, having 60.7% of contribution. Increased moisture and temperature resulted in decreased b values (Fig. 8). R^2 of the model was 0.90. The regression model for b is:

$$b = 112.87 - 83.83 \cdot M - 0.205 \cdot T \tag{16}$$

The highest b was obtained at 22% moisture and 30C temperature (+33.15), and lowest b was obtained at 30% moisture and 60C temperature (+20.01). a and b was found to be inversely proportional throughout this study suggesting that thermal and moisture changes has opposite effects on these color indicatives.

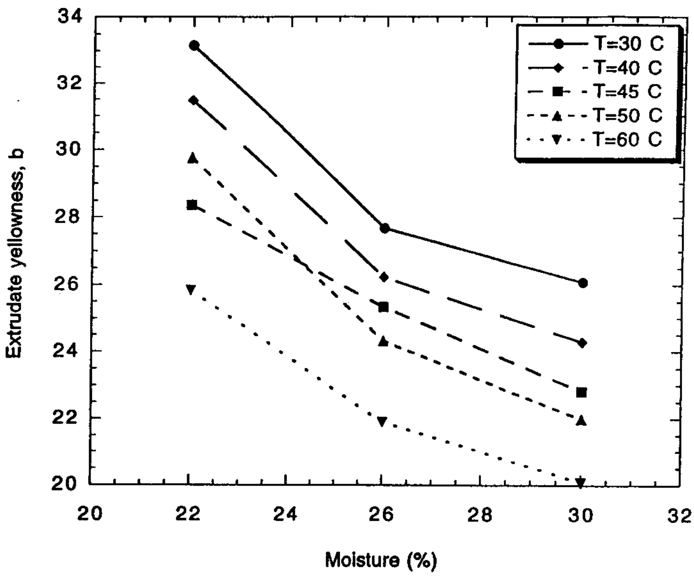

FIG. 8. EFFECT OF MOISTURE ON EXTRUDATE YELLOWNESS, *b*, AT DIFFERENT TEMPERATURES

Water Activity (a_w)

The lowest a_w was determined at 22% moisture and highest was at 30% moisture. Feed rate and temperature did not significantly affect a_w of extrudates. Moisture was the only variable influencing a_w of extrudates significantly. R^2 of the model was 0.96. The regression model for a_w is:

$$a_w = 0.554 + 1.092 \cdot M \tag{17}$$

Bulk Density

Bulk density of the extrudates was not influenced by temperature (1.4% contribution to model R^2) or feed rate (0.5% contribution to model R^2) markedly. Moisture was the only variable resulting in significant change in the density of the samples (98.1%). Expansion was negligible within the temperature range studied. R^2 of the model was 0.91. The regression model for bulk density is:

$$\text{Density} = 1.414 - 0.485 \cdot M - 1.88 \cdot 10^{-4} \cdot T + 2.31 \cdot 10^{-4} \cdot \dot{Q} \qquad (18)$$

CONCLUSIONS

Rheology of a novel fruit product, extruded 100% peach puree, was investigated to establish baseline models for food manufactured from this class of substrates. The apparent viscosity decreased as moisture and temperature were increased. A semi-empirical model was developed to predict melt viscosity of peach puree in a twin screw extruder. A power law type dependency for shear rate and exponential dependency for moisture was used. The temperature effect on viscosity was successfully described with a modified first order transfer function. A nonlinear regression analysis was performed to predict the viscosity under experimental conditions. The model fit the experimental data well.

Motor torque and SME were significantly influenced by moisture, temperature, and feed rate. An increase in moisture and temperature caused a decrease in torque and SME. An increase in feed rate increased torque and decreased SME.

Moisture was the dominant variable that influenced the color characteristics of the extrudates. Feed rate did not contribute to the variance of *L*, *a*, *b*, density, and a_w, revealing insignificant effects on these measured responses. Density and a_w were significantly influenced by moisture.

This work provides a basis for the extrusion of other fruit products which may exhibit similar characteristics. An extension of this study could be the correlation of extruder process parameters or system parameters to textural properties of various fruit extrudates, aiming to optimize the processing conditions of such products which would make them more acceptable by the consumer.

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