

MODELLING AND SIMULATING COMMERCIAL HOT WATER BLANCHING OF POTATOES

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Abstract

A method of applying the leaching model for hot water blanching to predict component losses during commercial scale potato processing was developed. The method involves characterizing an individual blancher to adjust the model to account for nonuniform mixing flow. The leaching model is based on the assumption that the water in the blancher is completely back mixed and the temperature and concentration are uniform throughout. The concentration of the blanch water for back mix flow was calculated using the mean blancher temperature. The blanch water concentration for nonideal flow (e.g., plug flow) corresponding to the temperature distribution was calculated. Using some combination of back mix and plug flow the blanch water concentration was calculated. Comparing the calculated and experimental values, a factor to adjust the model-calculated concentration to agree with the real value was determined. This factor then is used to correct future calculations made on this particular blancher with the model. Potassium is used as a key component because it is stable for sampling and analysis. However, the method should work for other components.

Introduction

Hot water blanching is one of the most important unit operations in food processing. Blanching deactivates enzymes, retards browning, and leaches water soluble material from the food pieces. Over 7 billion kg of potatoes are processed each year (1) to French fries, potato chips, hash browns, flakes, etc., and most are hot water blanched. Leaching during blanching is important because it lowers the reducing sugar content allowing the production of a lighter, more attractive product. However, blanching also removes nutrients, uses energy, and water.

We have developed a computer program for simulating potato processing. Unfortunately, mathematical models for most food related unit operations do not exist. Kozempel, *et al.* (3, 5) developed a predictive mathematical model for hot water blanching which applies to these solutes: soluble

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solids, potassium, glucose, sugars, magnesium, phosphorous, vitamin C, thiamin, riboflavin, and niacin. This is an ideal model in that it assumes that the blancher can be considered a well-mixed tank with uniform temperature and solute concentration. The ideal model, as previously presented and verified, accurately predicts component losses when the blancher is fully back mixed. However, the blancher is normally fully back mixed only at the bench level as was the case when this model was developed. Channeling, plug flow, and dead water regions are to be expected in pilot plant and commercial scale systems and, because of these, it is unreasonable to expect an ideal model to predict accurately without correction. The application of the model through the pilot plant level to commercial type blanchers is presented.

Materials and Methods

Pilot Plant—The blancher is a Thermascrew, model TL-36-K2210, made by the Rietz Manufacturing Company. The blancher is nominally 3 m long and the screw 0.9 m in diameter. Water was heated by steam sparging at 205 kPa at 8 points on the bottom. The range of blanch temperature used was 74 C to 85 C. The height of the exit overflow from the blancher sets the volumetric hold-up of the blancher. The blancher was modified to vary the overflow height and independently set the water capacity of the blancher from 541 to 874 l. The water flow rate was controlled with a Foxboro control valve, model VISD-1A3A00A, at values from 143 to 966 Kg/hr. Control was difficult because of sudden step changes in pressure of the supply water. The potato residence time was 16 to 22.7 min at 183 to 325 Kg/hr.

Potatoes (cvs. Russet Burbank and BelRus) used in the experiments were grown in Maine. Potatoes to be blanched were peeled in either a steam peeler at 1310 kPa (175 psig) for 18 sec or in a lye (20% NaOH) peeler at 66 C for 10 min, sulfite rinsed ($\frac{1}{4}$ % NaHSO₃), and cut into 1 cm dice in an Urschel cutter model G-A. The exit water concentration reached a steady state in 2 to 3 hr at which time data (process parameters) were recorded.

Commercial—The single commercial blancher was custom built in the production plant. It is nominally 4.1 m long with a screw diameter of 0.9 m and held 7937 l of water. Steam was sparged at 448 kPa (50 psig) maintaining a blanch temperature at 88-91 C. The potato feed rate was 1814-4082 Kg/hr of either "round white" or Russet Burbank. Potato residence time was 8 to 14 min. The water feed rate ranged from 454 to 1270 Kg/hr.

The processing plant uses three blanchers in series for the French fry line. They are similar to the single blancher but 6.7 m long with a screw diameter of 1.75 m. Each holds 13210 l.

Analyses-Water—Blanch water samples were filtered through 2 V folded filter paper prior to analysis to remove insoluble solids. Potassium was determined directly on potato water samples acidified with HNO₃, in an atomic

absorption spectrophotometer, Perkin Elmer 306. Glucose was determined within 4 hr of sampling to minimize loss of glucose (2), with a YSI Model 27 Industrial Sugar Analyzer.

Solid content of samples was determined by evaporation of the bulk of the water on a steam bath, then finish drying to a constant weight in a forced air oven at 105 C.

Potatoes—For soluble solids and soluble potassium from the juice, potatoes were ground in a Waring Blendor, the juice pressed through cheese cloth, and filtered through Whatman 2 V folded filter paper. Before determining soluble glucose (2) the filtrate was clarified with PbAc to remove possible interfering reducing substances. Lead was removed from the clarified juice using sodium oxalate. The filtrate was diluted, and subsequently analyzed for glucose utilizing the YSI sugar analyzer. Results are reported as solubles. Moisture content of potato was determined by AOAC Method 7.003 (6).

Results and Discussion

As reported previously (3), the mathematical model describing the loss of solute from potatoes during hot water blanching at 77 C is;

$$\frac{dS}{d\theta} = \frac{PMC_1 \psi + PMS \psi + W(S_1 - S)}{V_e + \left(\frac{PM\tau}{2}\right)\psi} \quad [\text{eq. 1}]$$

where $dS/d\theta$ is the rate of change in concentration of the blanch water and

$$\psi = 1 - \frac{8}{\pi^2} \text{Exp}(-\pi^2 D\tau/L^2)$$

At steady state $dS/d\theta = 0$ and eq. 1 simplifies to eq. 2;

$$S = \frac{PMC_1\psi + WS_1}{PM\psi + W} \quad [\text{eq. 2}]$$

All parameters in the model are process parameters except C_1 and D . D is the diffusivity coefficient for each solute and is known (3). For potassium, D is 0.423 cm²/hr, for glucose 4.057 cm²/hr, and for soluble solids, 0.274. Coefficient C_1 is the equilibrium solute concentration used in the model which may be considered the effective solute concentration in the water phase of the potato. C_1 must be determined.

Both D and C_1 vary with temperature. However, in our recent study (5) we found it best to hold D constant and vary C_1 with temperature. C_1 is a function of the analytically determined concentration of the solute in the potato C_o ;

$$C_1 = k \times C_o \quad [\text{eq. 3}]$$

The constant, k , is specific for each solute and varies with temperature according to a hyperbolic function, the hyperbolic tangent (Tanh). Equations 4 to 7 are for potassium, glucose, and soluble solids (5), respectively.

$$k_K = 0.13 + 0.621 \left[\frac{1.0 - \text{Tanh}(X)}{2} \right] \quad [\text{eq. 4}]$$

$$k_{glu} = 0.047 + 0.362 \left[\frac{1.0 - \text{Tanh}(X)}{2} \right] \quad [\text{eq. 5}]$$

$$k_{ss} = 0.01 + 0.848 \left[\frac{1.0 - \text{Tanh}(X)}{2} \right] \quad [\text{eq. 6}]$$

where

$$X = 6 \left[\frac{T + 1.11}{122.22} \right] - 3 \quad [\text{eq. 7}]$$

This ideal model (eqs. 1-7) assumes the blancher is back mixed with no channeling. It was verified when the blancher was back mixed (3, 5). The flow in commercial blanchers is a combination of back mixed, plug flow, etc. The pilot plant unit was not back mixed. How can this model be used for nonideal blanchers?

It was possible to do a residence time distribution study on our pilot plant blancher; but, the commercial plant could not be shut down to do a residence time distribution study. Some nonintrusive measurable variable is needed to determine a suitable correction factor for applying the ideal model to the nonideal pilot plant and commercial plant.

It was assumed that the flow consists of two types, ideal and one nonideal such as plug flow, and some combination of the two could be used when applying the ideal model to nonideal blanchers.

The exit concentration would then be equal to:

$$S = \alpha S_B + (1 - \alpha) S_P \quad [\text{eq. 8}]$$

where α is the correction factor representing the fraction of the flow which is back mixed. S_B and S_P are the solute concentrations calculated assuming all back mixed and all nonideal or plug flow, respectively.

Temperature can be measured to correct for channeling or estimate the residence time distribution. The temperature distribution is a function of the residence time distribution, is easily measured, and must be known to determine the equilibrium constant, C_1 , in the ideal model.

The blancher was hypothetically sectioned into zones of equal length or volume (nine for the pilot plant blancher) for monitoring temperature. Figure 1 is a typical plot of the temperature distribution versus zone position in the pilot plant blancher. In this blancher, the temperature rises approximately linearly to a maximum, T_B , and levels off at zone Z_c . Extrapolating back to zone zero gives the effective inlet temperature, T_o .

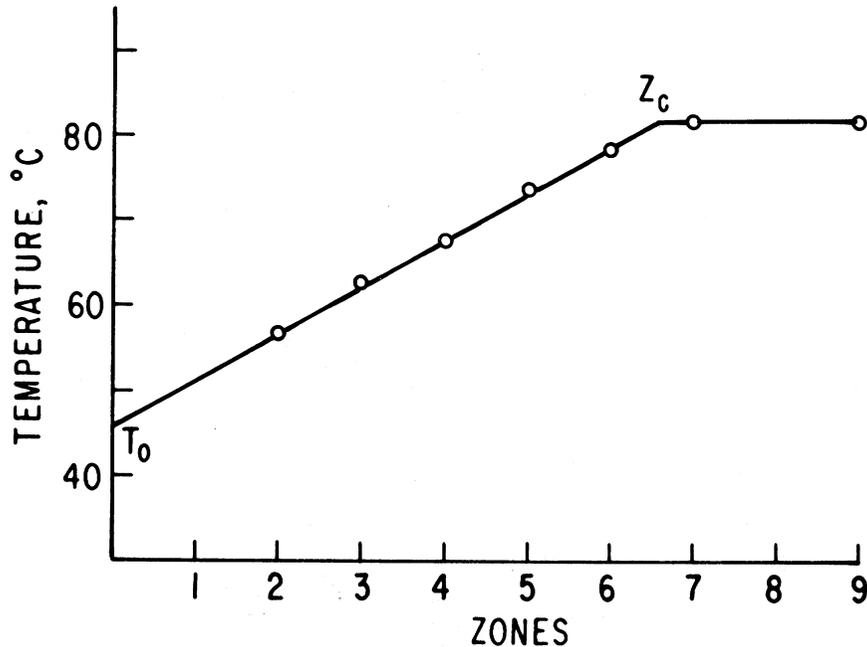


FIG. 1. Temperature distribution of pilot plant blancher as a function of hypothetical zones or artificial equi-volume sections of the blancher where zone 1 is the inlet and zone 9 is the exit.

By definition, a back mix blancher would have uniform concentration and temperature. Therefore, to determine the exit concentration assuming complete back mixing, S_B , a mean temperature for each run was calculated from similar plots. Equation 9 gives the calculation of the mean temperature, \bar{T} ;

$$\bar{T} = T_o + \frac{[0.5 (T_B - T_o) Z_c + (T_B - T_o) (Z - Z_c)]}{Z} \quad [\text{eq. 9}]$$

Applying the ideal model at the mean temperature, \bar{T} , gave the blanch water concentration assuming complete back mixing, S_B . Sufficient steam (enthalpy) was mathematically added in these calculations to raise the blanch temperature to T_B . This slightly increased the water flow rate in each zone.

Solute concentration in the exit blanch water (S_p) was then calculated using this temperature distribution. Since plug flow can be considered as an infinite number of back mixed units in series, S_p could be determined by applying the ideal model over an infinite number of differential back mixed units in series. Since temperature was measured over a finite number of zones, plug flow was approximated by applying the ideal model over this finite number of back mixed units. The model was applied to each zone se-

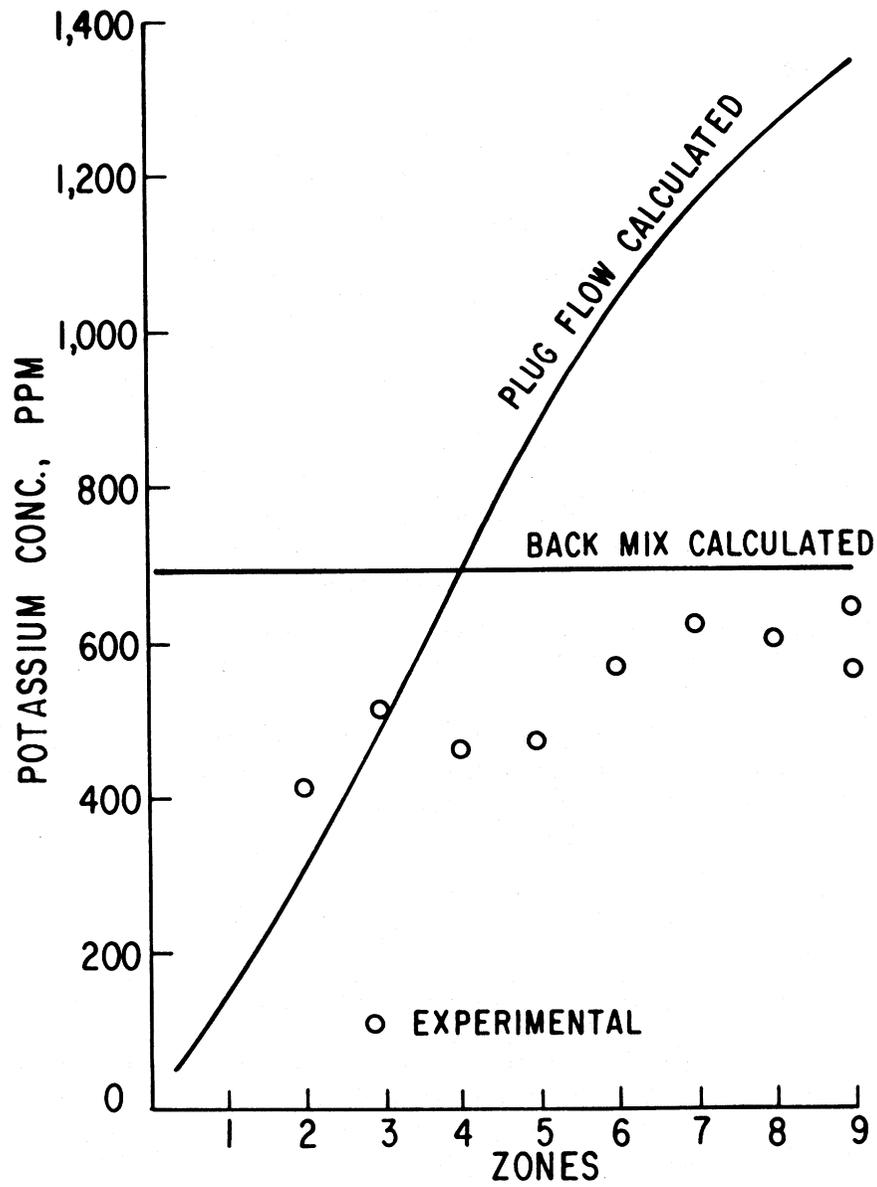


FIG. 2. Comparison of experimental potassium concentration with concentration calculated assuming back mix or plug flow in pilot plant. The zones are artificial equi-volume sections of the blancher where zone 1 is the inlet and zone 9 is the exit.

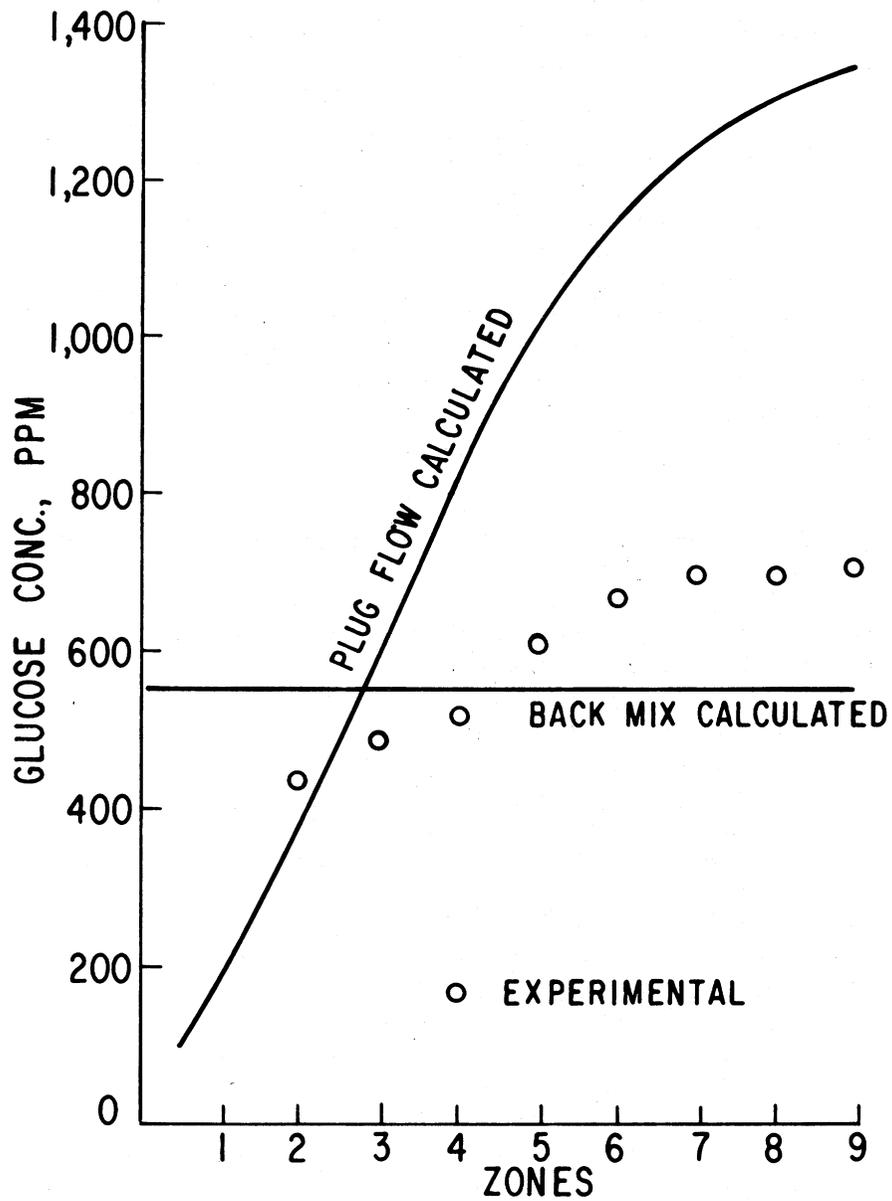


FIG. 3. Comparison of experimental glucose concentration with concentration calculated assuming back mix or plug flow in pilot plant. The zones are artificial equi-volume sections of the blancher where zone 1 is the inlet and zone 9 is the exit.

quentially, at the temperature of that zone, using the exit concentration of the previous zone, S_p , as the inlet concentration S_i .

Since the exit temperature is the same for either back mixed or plug flow, both model applications used the same amount of steam. Water flow rate was increased by the amount of added steam condensate.

Figures 2 and 3 give typical pilot plant results for potassium and glucose with curves calculated using the ideal model for back mixed and approximated for plug flow, applying the ideal model sequentially for a finite number of zones. It would not be expected for either calculated curve alone to agree with the experimental data, but rather some combination of the two (eq. 8). However, in all runs the exit concentration calculated using the plug flow model, S_p , was much greater than experimental, while the back mixed value, S_B , was approximately equal to or greater than experimental. When solving eq. 8 for α , the value of α in the pilot plant blancher was usually about 1, and $(1 - \alpha)$ was approximately zero. Dropping the second term of eq. 8 gives eq. 10.

$$S = \alpha S_B \quad [\text{eq. 10}]$$

Eq. 10 is a simple correction factor equation and applies when S_B is equal to or greater than experimental.

Table 1 lists the process variables and correction factors for potassium and glucose for the pilot plant blanch experiments. The average correction factor and standard deviation, respectively, was 1.12 and 0.14 for potassium and 1.34 and 0.08 for glucose. These factors apply specifically and only to the pilot plant blancher.

Organic compounds such as starches and sugars (e.g., glucose, fructose, and sucrose) are biologically unstable. Potassium, however, is stable. Therefore, potassium is an excellent key component with which to study and optimize a blanch process. It should indicate the effect of leaching on other components.

However, as seen in Figures 2 and 3, experimental glucose values were higher relative to the model than potassium. Since these are pilot plant results, there was little delay before analysis with minimal or nil glucose decomposition. The potatoes were in storage for several months at approximately 3 C so the glucose concentration in the potatoes had increased. Unfortunately, the storage conditions were not uniform so it was difficult to obtain a truly representative potato sample to evaluate C_o .

How does this approach work when simulating a commercial blancher where it is more difficult to obtain accurate data and analyses? Data were collected on four different occasions in 1981 and 1982 (Table 2) from a blancher in a Maine potato processing plant.

The primary product at the processing plant studied was frozen French fries. Nonuniform potato slivers screened from the French fry line were fed

TABLE 1. — *Pilot plant blancher data and model results.*

Temp. °C	Potato feed rate ¹	Residence time min	Water feed rate ¹	Potassium				Glucose			
				C _o ppm	S _{EXP} ppm	S _B ppm	$\alpha = \frac{S_{EXP}}{S_B}$	C _o ppm	S _{EXP} ppm	S _B ppm	$\alpha = \frac{S_{EXP}}{S_B}$
82	183	16	265-295	5156	610-644	655	0.96	6350	710	527	1.35
81	346	18.5	143	5140	1832-2057	1400	1.39	8800	1850	1438	1.28
82	226	18.5	234	5242	946-990	934	1.04	6300	885-910	705	1.27
74	325	18.5	143	5242	1667	1377	1.20	8900	1775	1390	1.28
85	227	22.7	871	4880	410-460	404	1.08	6100	400-460	307	1.37
75	234	22.6	996	4300	320	308	1.04	—	363	248	1.47
79	303	18.3	531	4130	631	560	1.12	5467			

¹Kg/h.S_{EXP} Exit water concentration determined experimentally.S_B Exit water concentration calculated assuming back mixed.C_o Analytically determined solute concentration in the juice in the potato. α Fraction of total flow which is back mixed.

TABLE 2. — Commercial blancher data and model results.

Run date	Temp. °C	Potato feed rate ¹	Resid. time min	Water feed rate ¹	Potassium				
					C _v ppm	S _{EXP} ppm	S _B ppm	$\alpha = \frac{S_{EXP}}{S_B}$	S _{pre}
Sept. 1981	91	3856	8.0	454	4759	1325	2330	0.57	—
Dec. 1981	88	1814	8.67	757	3849	766	1370	0.56	781
April 1982	89	1814	14.0	1277	4492	1060	1390	0.76	792
Sept. 1982	88	4082	8.0	902	1930	1095	1930	0.57	1100

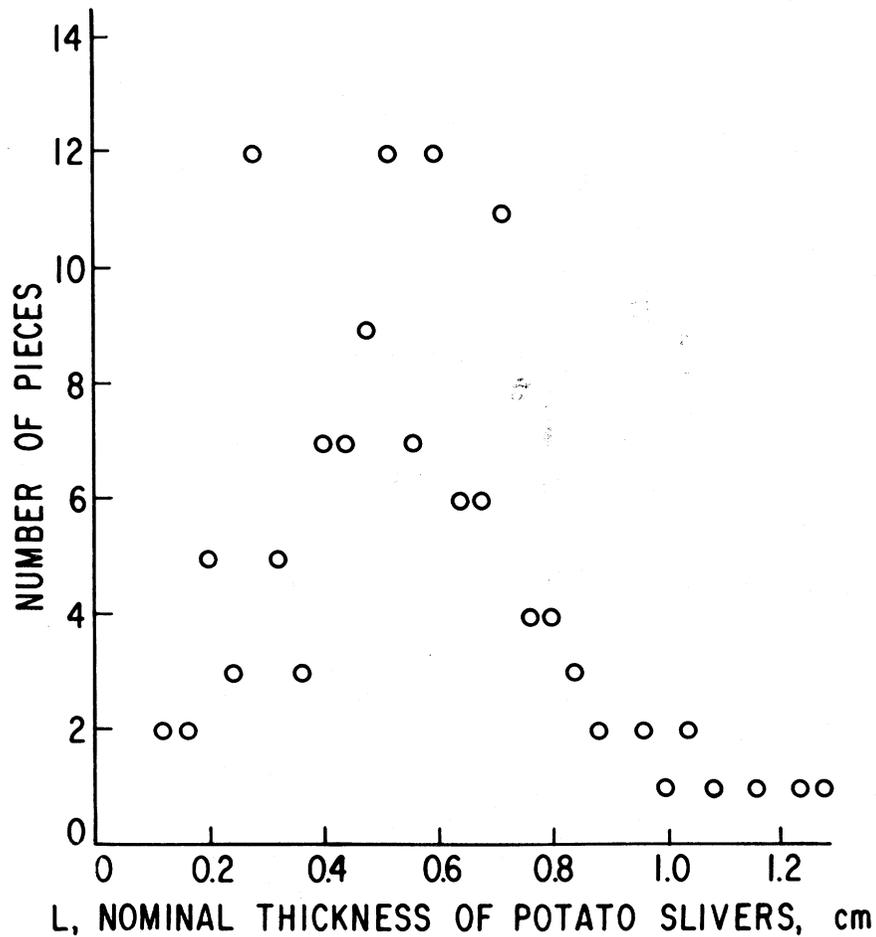
¹Kg/h.S_{EXP} Exit water concentration determined experimentally.S_B Exit water concentration calculated assuming back mixed.S_{pre} Exit water concentration predicted by $\alpha = 0.57$ (except base case of September 1981). α Fraction of total flow which is back mixed.

FIG. 4. Distribution of smallest dimension of potato slivers used as feed for the commercial blancher.

to the blancher. Because of the slivers, the feed was not a uniform nominal size as in the pilot plant. Figure 4 shows the particle size distribution. The piece size was approximated as 0.51 cm. Water flow rate was measured with a bucket and stop watch (typical values like 4.2 l/17 sec of 88 C water).

The first step in determining a correction factor for the blancher is to ascertain the temperature distribution to establish Z_c and T_o . T_b is a process parameter. Temperature distribution in September 1981 and in December 1981 (Figure 5) were in good agreement with Z_c of 8.8 and T_o between 55 C and 59 C.

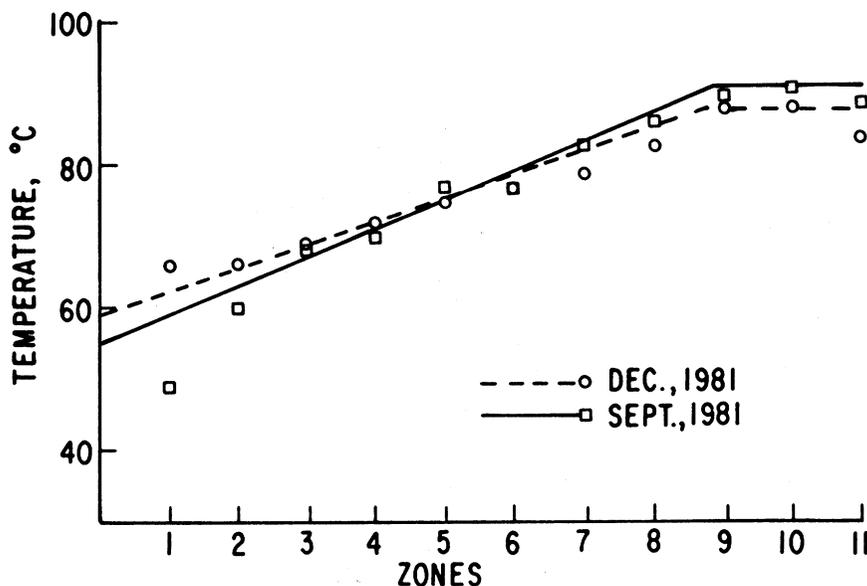


FIG. 5. Temperature distribution of commercial blancher as a function of hypothetical zones or artificial equi-volume sections of the blancher where zone 1 is the inlet and zone 11 is the exit.

Table 2 shows the process parameters and model results for four trials. The model results, S_b , are higher than experimental and showed there was significant by-passing in the blancher. Therefore, the nonideal term in eq. 8 ($1 - \alpha$) represented by-pass flow instead of plug flow. Since S_b was greater than experimental, eq. 10 was used instead of eq. 8. The correction factor, α , is remarkably consistent, with three of the four trials almost identical. Using the value of α from the September 1981 data (Table 2), the concentrations for the other three were predicted. Two of the three are almost perfect and the third is in error by 25%—fairly reliable predictions considering the accuracy of the data.

Other blanchers in this plant were studied—those in the French fry line. This line contains three blanchers in series, A, B, and C. French fry cut

potatoes enter blancher A, then B, and C. Water can be recycled from C, to B, to A with fresh water makeup in each. Figure 6 gives the temperature distributions for these units. These, too, were approximated as before, but with different values for Z_c and T_o for each. Because of recycle, the food process simulator program with the blanch subroutine was used to solve the trial and error calculation of this system. Table 3 lists experimental results for trials in April 1982 and September 1982, calculated results for April 1982, and predicted results in September 1982. Eq. 8 was applied to each blancher giving us three equations of the generic form;

$$S_n = \alpha_n S_{B_n} + (1 - \alpha_n) S_{P_n} \quad [\text{eq. 11}]$$

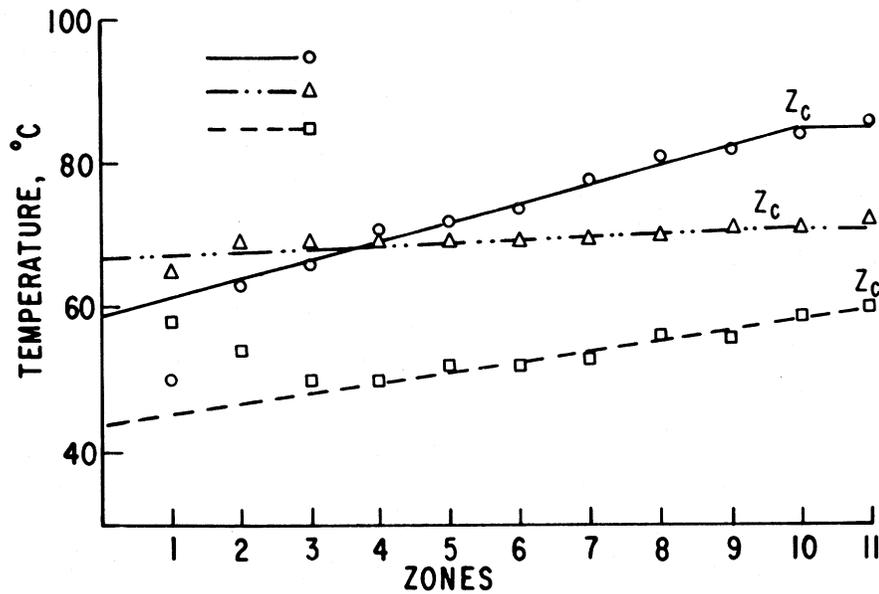


FIG. 6. Temperature distribution of commercial French fry blanchers A, B, and C as a function of hypothetical zones or artificial equi-volume sections of the blancher where zone 1 is the inlet and zone 11 is the exit. Z_c is the zone in the blancher where the temperature reaches a maximum.

where n is blancher A, B, or C. Eq. 8 was used because S_B was less than experimental. Solving the three equations for the values of α specific to each blancher, the concentrations for April 1982 were calculated as shown in Table 3. Using the values of α determined from the data of April 1982, the results for September 1982 were predicted. The agreement with experimental is remarkably accurate, considering the complexity in the commercial system and uncertainty in measuring the process parameters.

TABLE 3. — *Commercial French fry blanchers data and predicted results.*

Blancher	April, 1982				September, 1982			
	°C	K conc., ppm exp.	K conc., ppm calc.	α	°C	K conc., ppm exp.	K conc., ppm predicted	% Error
A	54	957	1180	0.629	56	614	673	9.1
		944				623		
		982				614		
B	76	1559	1470	0.521	69	842	758	10.0
		1540				842		
		1502						
C	83	1083	1080	0.700	77	860	861	0.5
		1131				844		
		1131						

calc. Calculated.

exp. Experimental.

α Fraction of total flow which is back mixed.

Conclusion

A method for simulating losses due to leaching in hot water blanching of potatoes on a commercial scale has been presented. Improved accuracy in simulation will require more accurate data (process parameters) collecting equipment in commercial plants. The combination of this model with the food process simulator should permit optimization of commercial blanching processes.

Nomenclature

- C_o = solute concentration in the juice within the potato, wt/wt, analytically determined
- C_1 = initial solute concentration in juice within the potato calculated from C_o wt/wt
- D = diffusivity, cm^2/hr
- k = ratio of C_1/C_o
- L = nominal thickness of cut pieces, cm
- M = potato moisture content, wt/wt
- P = potato flow rate, wt/hr
- S = solute concentration in the blanch water; also, solute concentration in the exit blanch water, wt/wt
- S_b = solute concentration in the exit blanch water assuming all back mixed, wt/wt
- S_p = solute concentration in the exit blanch water assuming plug flow, wt/wt

- S_1 = solute concentration in the inlet water to the branch, wt/wt
 T = temperature, °C
 T_B = maximum blanch water temperature, °C
 T_o = effective inlet temperature to blanch, °C
 \bar{T} = mean blanch temperature, °C
 V = volume of the blanch water
 W = inlet water flow rate including added steam condensate, wt/hr
 Z_c = zone in the blancher where the temperature reaches a maximum
 Z = number of hypothetical zones in the blancher
 α = fraction of total flow which is back mixed
 ρ = density of the blanch water, wt/vol
 θ = time, hr
 τ = extraction residence time, hr

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