

# The Backmixing-Diffusion Concept for Mass Transport Phenomena of Pineapple Juice Extraction Using a Reversing Continuous Countercurrent Extractor

Chairat Siripatana, Waigoon Rittirut, Thummarat Thummadetsak

**Abstract**— A backmixing-diffusion model was developed and proposed to describe the mass transport in a reversing continuous countercurrent extractor (RCCE) at steady-state condition. Three variables namely draft, retention time, and extraction temperature were studied according to various experimental trials. Equilibrium distribution coefficient and solute diffusivity were important parameters introduced to predict concentration profiles for both liquid and solid phases. It was found that backmixing-diffusion concept can be applied through the studied range of operation time. The differences between experimental data and the predictions for concentration profiles were due to the variations in Biot number and Peclet number along the length of extraction unit. Nevertheless, the simulation was successful for pineapple juice extraction process.

**Index Terms**— backmixing, countercurrent extraction, pineapple juice, reversing continuous countercurrent extractor, RCCE, solid-liquid diffusion

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## 1 SYNOPSIS

A study on mass transfer in a pineapple juice extractor was carried out. The extraction of pineapple juice from its peel was chosen as a case study using a reversing countercurrent extraction system (RCCE) based on Casimir [1]. A diffusion-backmixing model was used to describe the kinetics of mass transfer in the extraction unit. In estimating equilibrium distribution coefficient, solute diffusivity and time delayed due to plasmolysis, collection data from batch extraction experiments were used [2]. The first two parameters were introduced to predict concentration profiles of solute in both solid and liquid phases in the extractor using diffusion-backmixing model.

According to experimental work, various extraction trials were performed by varying three variables, namely : 1) draft, 2) retention time, and 3) extraction temperature. In the experimental range where temperature of 55-70 °C, draft of 1-2 and retention time of 50-90 min, it was observed that highest yield (70.9 %) was associated with highest draft and longest retention time. Furthermore, in all trials, after system reached a steady state condition, the solute concentration profiles along the extractor were of the same pattern. The diffusion-backmixing model was used to predict the concentration profiles satisfactorily. In general, the predictions agreed well with the experimental data in solid phase and can also approximate a closed trend in liquid phase.

Finally, the extraction system was analyzed critically. It was found that backmixing caused the reduction in driving force and the increasing in mass transfer of solute at solid-liquid interphases by convection. The combined effect of finite Biot number ( $Bi$ ) and Peclet number ( $R$ ) limited the maximum

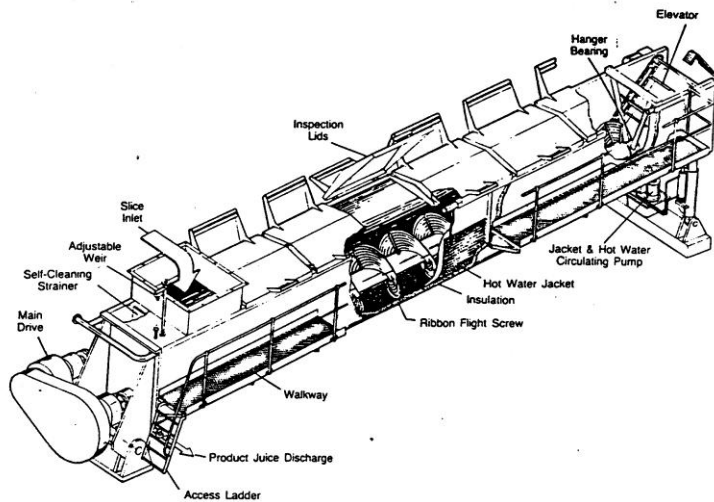
yield to about 70 % unless the design is modified. Furthermore, it can be concluded that the differences between experimental data and the predictions were due to the variations in  $Bi$  and  $R$  along the extractor which had been assumed constant for the model.

## 2 INTRODUCTION

Thailand is one of the biggest exporter of pineapple products. Although juice concentrate is not the main product of pineapple processing, the country earns over 100 million dollars each year since 2004 while the consumption need and export value got increasing annually [3]. Nevertheless, in order to be competitive it is necessary to improve the efficiency of the whole process by reducing energy costs and minimizing the environmental problems like reduction of the effluent load.

The juice is largely produced by mechanical expression. This method yield 50-60 % of the extractable solutes as compared to a yield of around 80-90 % obtained by diffusion extraction. However, very few big processing plants have adopted for the diffusion based process, thus the potential for juice concentrate production from pineapple wastes is still highly unexplored. In this study, a pilot-plant scale extractor described initially by Casimir [1] and subsequently used by others [4], [5] was considered to produce the juice from peels. The extractor details are given accordingly. The extractor illustration (Fig. 1) was also shown in the previous work [6], [7].

Siripatana [6] studied juice extraction processes from apples and pears using this extractor. More than five diffusion models were tested for describing mass transfer. It was found that the so-called "diffusion-backmixing model" gave the best prediction. Gunasekeran et al. [5] represented backmixing in a continuous diffusion processes by a series of well-mixed tanks. According to their experimental data and the predictions of their model, the extractor showed very low level in backmixing *i.e.* high Peclet numbers. This was not consistent with the earlier work of Siripatana [6].



**Fig. 1.** Reversing continuous countercurrent extractor (RCCE)  
**Source :** Siripatana [6]

Among comprehensive model based on diffusion and liquid phase backmixing theory was given by Lee and Schwartzberg [8]. The set of solution with any extent of backmixing was given. The solution is rather complex, yet it is not obvious how the solution can be made to reconcile for the asymptotic case (plug flow) with the well known solution given by Spanik [9].

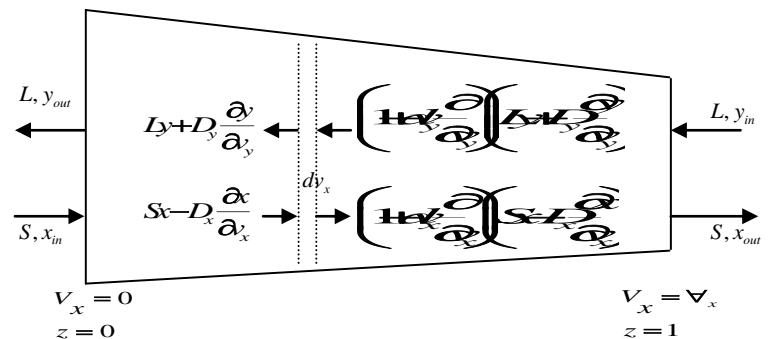
In addition, the earlier workers have ignored the time required for plasmolysis. The objectives of this investigation thus were to :- 1) explore the potential of continuous countercurrent diffusion extractor (CCDE), like RCCE unit, for the extraction of pineapple juice; 2) develop a mathematical model which can explain mass transport in necessity of fundamental parameters like in-solid diffusivity, draft of extraction (corres-

### 3 MODEL DEVELOPMENT

The solid and liquid phase in normal operation can vary greatly along the trough. The reason is due to inclination of the trough inside the extractor (slope) and the movement of solid and liquid phase relative to the screw conveyor. Nevertheless, the volumetric flow rate of both phases along the trough may be assumed constant at steady state conditions. For this reason, the model described here will be based on the volumetric flow rates of solid and liquid phases instead of their linear velocities. With this modification, the solid and liquid phase material balances will be equal to that conventional ones developed by Mecklenburgh and Hartland [10]. In this work, the solution can thus be rewritten to provide equivalent results as given by traditional backmixing [10] where average velocity is treated as constant.

Considering material balances around CCDE as shown in Fig. 2, the following assumptions have been made:-

- 1) The volumetric flow rate for both solid and liquid phases are constant.
- 2) The volumetric overall mass transfer coefficient ( $\kappa_x$ ), draft of extraction ( $\alpha$ ), and the equilibrium distribution coefficient ( $m$ ) are constant along the trough.
- 3) The volumetric dispersion coefficient ( $D_x$  and  $D_y$  for solid and liquid phases, respectively) are constant and independent of the position along the trough.



**Fig. 2.** Schematic view of material balance for differential volume,  $dv_x$ , along the trough of extraction unit.

Under the above assumptions, material balance around control volume  $dV$  is as follow:-

For solid phase:

$$(\text{solute flow rate at inlet}) - (\text{solute flow rate at outlet}) - (\text{rate of solute transferred to liquid phase}) = (\text{rate of solute accumulated in the volume})$$

- Chairat Siripatana is at School of Engineering and Resources, Walailak University, Thailand . E-mail: schairat@wu.ac.th
- Waigoon Rittirut is at School of Agricultural Technology, Walailak University, Thailand . E-mail: rwaigoon@wu.ac.th
- Thummarat Thummadetsak is currently working for Petroleum Authority of Thailand Company, Limited, Thailand E-mail: thummarat.t@ptplc.com

ponding to liquid to solid feed ratio) and time delayed due to plasmolysis (extraction time corrected for cell plasmolysis).

$$S \left[ \frac{\partial x}{\partial v_x} - \frac{\partial}{\partial v_x} \left( \frac{D_x}{S} \frac{\partial x}{\partial v_x} \right) \right] - K_x \frac{dA}{dv_x} (x-x^*) = \frac{\partial x}{\partial t} \tag{1}$$

At steady state condition, the above equation is thus reduced to:-

$$S \left[ \frac{dx}{dv_x} - \left( \frac{D_x}{S \nabla_x^2} \frac{d^2x}{dz^2} \right) \right] - K_x \left( x - \frac{y}{m} \right) = 0 \tag{2}$$

Where,

$$z = v_x / \nabla_x ; x^* = y / m ; \text{ and}$$

$$K_x = \frac{K_s dA}{dv_x}$$

Here

$K_x$  is volumetric overall mass transfer coefficient (s<sup>-1</sup>)

$S$  is volumetric flow rate of solid phase (m<sup>3</sup>/s)

$S_0$  is volumetric flow rate of solid phase at inlet of extraction unit (m<sup>3</sup>/s)

$S_f$  is volumetric flow rate of solid phase at outlet of extraction unit (m<sup>3</sup>/s)

$L$  is volumetric flow rate of liquid phase (m<sup>3</sup>/s)

$L_0$  is volumetric flow rate of liquid phase at inlet of extraction unit (m<sup>3</sup>/s)

$L_f$  is volumetric flow rate of liquid phase at outlet of extraction unit (m<sup>3</sup>/s)

$x$  is solute concentration in solid phase (kg/m<sup>3</sup>)

$y$  is solute concentration in liquid phase (kg/m<sup>3</sup>)

$v_x$  is accumulated volume of solid phase from its inlet point to position considered (m<sup>3</sup>)

$v_y$  is accumulated volume of liquid phase from its inlet point to position considered (m<sup>3</sup>)

$\nabla_x$  is total volume of solid phase in extraction unit

$\nabla_y$  is total volume of liquid phase in extraction unit

$D_x$  is volumetric dispersion coefficient in solid phase (m<sup>6</sup>/s)

$D_y$  is volumetric dispersion coefficient in liquid phase (m<sup>6</sup>/s)

$K_s$  is overall mass transfer coefficient (m/s), a parameter depending on solute diffusivity from diffusion theory

$dA^*$  is area of mass transfer in volume  $dv_x$  (m<sup>2</sup>)

$x$  is solute concentration in solid phase at equilibrium condition (kg/m<sup>3</sup>)

$y$  is solute concentration in liquid phase at equilibrium condition (kg/m<sup>3</sup>)

Here, subscript  $x$  is denoted for solid while subscript  $y$  is for liquid.

Converting equation (2) to dimensionless form,

$$\frac{1}{P} \frac{d^2x}{dz^2} - \frac{dx}{dz} = T_x (x - y/m) \tag{3}$$

Where

$$P = \nabla_x S / D_x = \text{Peclet number in solid phase,}$$

$$T_x = \nabla_x K_x / S = t_f K_x , t_f = t + t_l - t_p$$

Here,  $t_f$  is theoretical extraction time as described by equation (4). The equation was given in Siripatana's work (Siripatana, 1986) and then claimed by Thummadetsak (1996).

$$t_f = t + t_l - t_p \tag{4}$$

Where

$t$  is the actual extraction time.

$t_l$  is time lead due to initial solute distribution in solid. The parameter is introduced to correct for the time taken for the Sherwood number to allocate an asymptotic value, and

$t_p$  is time delayed due to plasmolysis.

It was noted that the concentration in solid phase ( $x$ ) in equation (3) is sometimes normally expressed as the average concentration ( $\bar{x}$ ) as shown in some works [6], [11].

(ii) For liquid phase:

Considering material balance of solute in liquid phase at inlet and outlet of differential volume,  $dv_y$ , the following equation is given

$$L \left[ \frac{\partial y}{\partial v_y} + \frac{\partial}{\partial v_y} \left( \frac{D_y}{L} \frac{\partial y}{\partial v_y} \right) \right] + K_s \frac{dA}{dv_y} (x - x^*) = \frac{\partial y}{\partial t} \tag{5}$$

Similarly, at steady state condition, the above equation can be rewritten in ordinary differential equations (ODEs) and then converted to dimensionless form,

$$\frac{1}{R} \frac{d^2y}{dz^2} + \frac{dy}{dz} = T_L (x - y/m) \tag{6}$$

where

$$v_y = \nabla_y v_x , \nu = L/S , y = mx^* , K_x = \frac{K_s dA}{dv_x} , z = v_x / \nabla_x$$

$$R = \nabla_y L / D_y = \text{Peclet number in liquid phase} = T_y = (\nabla_y / L) K_x = T_x (S/L)$$

Equation (1)-(6) are actually similar to what were shown in the previous work at steady state condition [12].

Considering characteristics flow of solid and liquid phase around the inlet and outlet of extraction unit in cooperation with the fact that there is no concentration change of  $x$  and  $y$  left out of extractor. The usual boundary conditions of equation (3) and (6) are as follow:-

$$x_{in} = x - \frac{1}{P} \frac{dx}{dz} \quad \text{at } z=0 \quad (7)$$

$$y_{out} = y, \quad \frac{1}{R} \frac{dy}{dz} = 0 \quad \text{at } z=0 \quad (8)$$

$$y_{in} = y + \frac{1}{R} \frac{dy}{dz} \quad \text{at } z=1 \quad (9)$$

$$x_{out} = x, \quad \frac{1}{P} \frac{dx}{dz} = 0 \quad \text{at } z=1 \quad (10)$$

Where

$x_{in}$  and  $y_{in}$  is the solute concentration in solid and liquid phase at inlet of the extractor, respectively.

$x_{out}$  and  $y_{out}$  is the solute concentration in solid and liquid phase at outlet of the extractor, respectively.

The solid phase concentration ( $x$ ) in equation (6) and the boundary conditions is normally expressed as the average concentration ( $\bar{x}$ ) as also mentioned in earlier works [6], [10], [11].

The solution of the above ODEs are also discussed by Mecklenburgh and Hartland [10] for the case where solid phase mixing is negligible. The solution is given as follow:

For  $Q=L/S \neq 1$ , let  $J=1/\alpha=Q/m$ ,  $m$  is distribution coefficient.

$$\frac{x_0 - x}{x_0 - y_0/m} = \frac{(1-w_1)H_1/q_1 - (1-w_1)H_2/q_2}{(1-FW_1)H_1/q_1 - (1-FW_2)H_2/q_2} \quad (11)$$

$$\left[ \frac{y - y_0}{mx_0 - y_0} \right] \alpha = \frac{(w_1 - H_1W_1)/q_1 - (w_1 - H_2W_2)/q_2}{(1-FW_1)H_1/q_1 - (1-FW_2)H_2/q_2} \quad (12)$$

Where  $q_1$  and  $q_2$  are roots of equations,

$$q^2 + q(T_x + R) = T_x R(1 - \alpha) / \alpha$$

$$H = 1 + q/R, \quad w = e^{qz}, \quad W = e^q$$

## 4 MATERIAL AND METHOD

### 4.1 Extractor

The configuration of the extractor is similar to what was mentioned in the previous work [1], [2], [6]. The relevant dimensions of the extractor are summarized as follow:- 2.2 m of overall length, 2.0 m of total screw length, 12 number of flights, 0.21 m of screw diameter, 21.8 m of trough width, and 30 m of trough depth. In this investigation, the operating conditions were in the following ranges.

- The forward and reverse rotational speeds of the screw were identical in all trials. Nevertheless, in each cycle, the screw moved in forward and reverse direction for 25 and 22 seconds, respectively, giving the percentage progression to 6.38.

- The trough was inclined at 4 degree to the horizontal.
- The solid feed rate was maintained at 12 kg/h by dropping 2 kg of raw material manually at 5-min. intervals. Before each run, the feed was preheated to about 70 °C for more than 5 min. to plasmolyze the cell structure so that time of cell plasmolysis ( $t_p$ ) was essentially zero.
- The average operating temperatures were controlled at 55, 62.5 and 70 °C for different runs.
- Retention times of solid were set as 50, 70 and 90 min.
- The following ratios of liquid to solid flow rate were investigated:- 1:1, 1.5:1 and 2.0:1. Drafts ( $\alpha=Lm/S$ ) were calculated by averaging the liquid to solid ratio at entering and leaving ends.
- The equilibrium distribution constant was found to be 1.14 for all experiments.

At steady state condition, both samples for solid and liquid phase were taken at different point along the length of extractor. The relative length ( $z$ ) for sampling position along extraction chamber was that:- 0, 0.1, 0.3, 0.5, 0.7, 0.9 and 1.0. Soluble solids ( $^{\circ}$ Brix) in solid and liquid phase were analyzed by refractive index using Abbe refractometer.

### 4.2 Parameter Determination

#### Solute diffusivity and time delayed due to plasmolysis

Batch extraction was carried out, the extraction curves and solid dimensions were used to calculate the diffusivities [2]. The time delayed due to plasmolysis was calculated by the method described in Siripatana [6] and then claimed in some work [13], [14].

#### Overall mass transfer coefficient

The overall mass transfer coefficient described here is actually "volumetric overall mass transfer coefficient,  $K_x$ " and is defined by the following basic equation.

$$-\frac{d\bar{x}}{dt} = K_x(\bar{x} - y/m) \quad (13)$$

Here  $K_x$  for different runs were evaluated directly from the actual concentration profiles. That is, firstly, the profiles were fitted to the fifth order polynomial with constant parameters. Then the average  $K_x$ s were determined by comparing with the solution of equation (13) such that the sum of square of error were minimized.

#### Peclet number in liquid phase

The liquid phase Peclet numbers ( $R$ ) were determined by treating it as parameters of the backmixing model (eqn. 1-12) which must be found by fitting experimental data to the model.

#### Solid characteristic dimensions

The raw material used in this work was the pineapple peels

which were regarded as slabs for the purpose of modeling. Since the external surface (skin) of the pineapple fruits is much less permeable to solute than the inner part, the diffusion through the skin was neglected. The mean rectangular dimensions were  $5.22 \times 14.9 \times 52.0 \text{ mm}^3$ .

**Biot number ( $Bi = K_s a / D$ )**

Following the approach developed by Siripatana [15] with some modifications for rectangular block, the following approximate relations can be obtained;

$$Bi = \left[ p / \left( \frac{1}{g(\alpha, Bi)} - 1 \right) \right]^{1/k}, \quad Sh = \frac{2K_x}{D} \left( \frac{1}{l_1^2} + \frac{1}{l_2^2} + \frac{1}{l_3^2} \right)^{-1} \quad g = \frac{Sh}{2h(\alpha, \eta)},$$

$$h(\alpha, \eta) = 2.460 + 0.4321\eta + (0.528 + 0.5435\eta) / \alpha + (0.03909 + 0.08694\eta) / \alpha^2$$

$$p = 2.1426 + 0.9314 / \alpha + (0.1575 + 0.9876 / \alpha), \quad k = 0.9781 - 0.01467\eta \quad (14)$$

Where

- $Bi$  is Biot number of mass transfer.
- $K_s$  is mass transfer coefficient (m/s)
- $K_x$  is volumetric overall mass transfer coefficient (1/s)
- $a$  is characteristic length (m)
- $\eta$  is shape factor ( $V$ ) - 1
- $D$  is diffusivities ( $\text{m}^2/\text{s}$ )
- $Sh$  is Sherwood number.
- $\alpha$  is draft.
- $l_1, l_2,$  and  $l_3$  is characteristic length (half length) of each dimension of rectangular block, as defined in Siripatana work [15].

**Time lead due to initial solute distribution in solid**

Following the approach used by Siripatana [6], the following approximate relation can be derived;

$$t_f = -\ln C_1^3 / q_1^2 D_s \left( \frac{1}{l_1^2} + \frac{1}{l_2^2} + \frac{1}{l_3^2} \right), \quad C_1 = \frac{2\alpha}{2.466\alpha - 0.525}$$

Here,  $q_1$  is the first root of  $\tan q_1 / q_1 = \alpha$ . The notation is given as follow:-  $t_f$  is time lead,  $D_s$  is solute diffusivity,  $\alpha$  is draft, and  $q_1$  is eigenvalue of differential solution problems as shown in Siripatana work [6].

**5 RESULTS AND DISCUSSION**

**5.1 Solute Diffusivity and Time of Plasmolysis**

The results from batch extraction trials of the raw material showed that, in the temperature range of 55-70 °C, the solute diffusivity ( $D_s$ ) and time of plasmolysis ( $t_p$ ) were linearly and exponentially dependent on temperature, respectively (Fig. 3). These results can be summarized as follows;

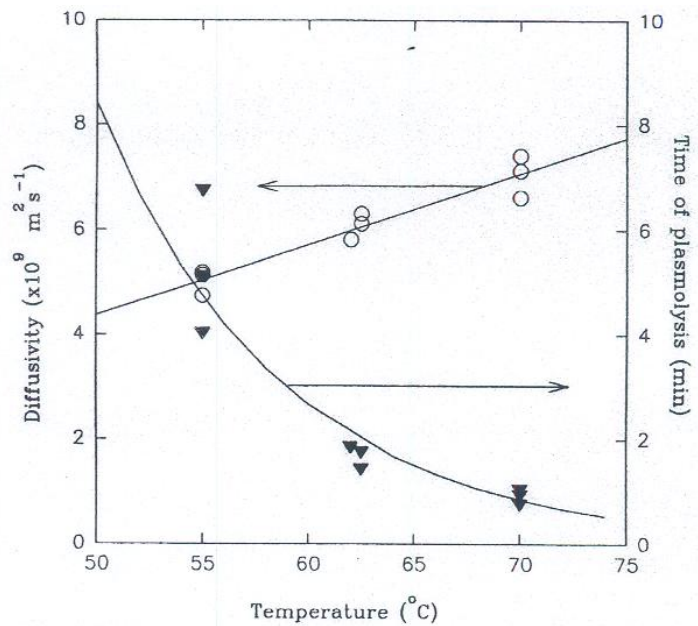
$$D_s = 1.347 \times 10^{-10} T(K) - 3.912 \times 10^{-8} \quad (\text{m}^2 \text{s}^{-1}) \quad (15)$$

$$t_p = 2664 \exp(-0.115[T(K) - 273.15]) \quad (16)$$

It should be note that the diffusivities of soluble solid,

which is mainly sucrose, were one order of the magnitude higher than that of sucrose in aqueous solution ( $0.62 \times 10^{-9} \text{ m}^2 \text{ s}^{-1}$  at 70 °C). This indicates that although the asymptotic extraction curves follow exponential pattern as predicted by diffusion theory, the main mechanism of solute transfer was not by molecular diffusion. Presumably, thermal contraction stress of the fiber matrix enhances solute transfer. This requires further investigation. Nevertheless, here it is sufficiency to claim that the observed value of diffusivities is 10 times higher than the molecular diffusion. This results agree with that given for pineapple slices by Beristain [16] and Sodchit [17].

In the investigated temperature range, time of plasmolysis ( $t_p$ ) approached zero exponentially as temperature increased from 50 to 70 °C. The similar trends were found by Siripatana [6] for diffusion of solute from pears and pineapples, and Sodchit [17] for solute infusion into pineapple slices.



**Fig. 3.** Solute diffusivity ( $\times 10^9 \text{ m}^2 \text{ s}^{-1}$ ) and time delayed due to plasmolysis (min) vs. temperature (°C).

**5.2 Overall Mass Transfer Coefficient**

The observed volumetric overall mass transfer coefficients ( $K_x$ ) were varied in the range of  $3.80 \times 10^{-4}$  to  $9.99 \times 10^{-4} \text{ s}^{-1}$  (Table 1). In spite of a considerable variation, it was quite obvious that an increase in total extraction time (a decrease in screw speed) was associated with the decrease in  $K_x$ . The effect was enhanced as temperature became higher (Fig. 4). It was also observed that  $K_x$  for the total extraction time of 50 min and temperatures of 62.5 and 70 C were slightly higher than that calculated by (14) using proper characteristics dimensions, infinite Biot number and diffusivities estimated from batch extraction. This was due to partial disintegration of solid particles due extra mechanical action of the screw which did not exist in batch extraction experiments. It should be emphasized that in fact the real cause of difference in  $K_x$  was the screw

action not the total extraction time. Thus, instead of reducing the total extraction time in order to increase  $K_x$  the extractor trough can be lengthened. This has an advantage of increasing  $K_x$  without getting side effect of lower residence time (which in turn lower yield).

### 5.3 Biot Number

Biot number ( $Bi=K_s a/D$ ) characterize the ratio of internal mass transfer resistance to resistance at the solid-liquid interface. Approximately it is related to  $\kappa_x$  by the following relation [15].

$$K_x = \frac{Sh}{2} D \left( \frac{1}{l_1^2} + \frac{1}{l_2^2} + \frac{1}{l_3^2} \right), \quad Sh = 2g(\alpha, Bi)h(\alpha, \eta),$$

$$g(\alpha, Bi) = \left( 1 + p / Bi^k \right)^{-1} \quad (17)$$

It is not surprising that the effect of total extraction time and temperature on  $\kappa_x$  is similar to that on  $Bi$  because, as shown in (17), if  $\alpha$  and  $D$  are constant  $\kappa_x$  will change in the same direction of the change in  $Bi$ . The relationship between Biot number vs. total extraction times and temperatures are shown in Fig. 5. It should be also stated that, in all experiments, Biot number was greater than 4, 6 and approach infinite for total extraction time of 90, 70 and 50 min. (screw speed of 2.09, 2.69 and 3.76 rpm), respectively. This indicated that the film resistance was relatively low and can be decreased easily by increasing the screw speed, nevertheless, with some sacrifice in the driving force due to higher backmixing.

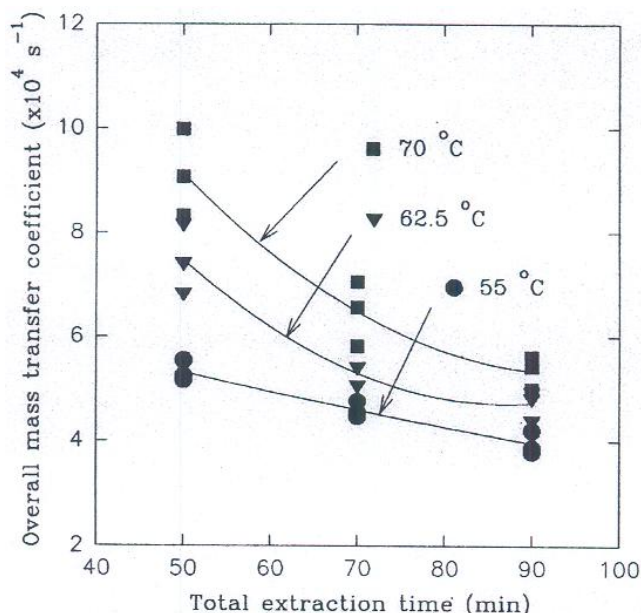


Fig. 4. Overall mass transfer coefficient vs. total extraction times and temperatures.

### 5.4 Peclet Number

It was confirmed that the raw material used in this work, the backmixing in solid phase was of minor importance [2]. In

contrast, the backmixing in liquid phase was quite considerable and the values of Peclet number ( $R$ ) are in the range of 1.5 to 3.5 (Fig. 6).

According to our experimental data, no clear effect of other factors except draft on the Peclet number could be identified. The effect of draft, nevertheless, had a certain trend, although the variations were quite large, and can be written in a following formula.

$$R = 4.12 - 1.04\alpha \quad (18)$$

If one notice that, by definition  $R = LV_L / D_y$ , one expects that, if  $D_y$  is constant, increasing  $\alpha$  (in effect increasing  $L$ ) will increase  $R$  proportionally which is exactly opposite of our experimental evidence. Thus clearly, as  $\alpha$  increased,  $D_y$  increased due to the stronger backmixing. Evidently, therefore the moving action of screw worked more efficiently when  $\alpha$  got higher.

Table 1. Summary of experimental trials and the parameters estimated in this work.

No	Draft ( $\alpha$ )	Temp. (°C)	Extraction* time (min)	$K_x$ ( $\times 10^4 s^{-1}$ )	Biot number ( $Bi$ )	Peclet number ( $R$ )	$K_x (Bi \rightarrow \infty)$ ( $\times 10^4 s^{-1}$ )	Yield (%)
1	1.41	55	50	5.17	19.4	3.2	5.97	59.6
2	1.18	55	70	4.76	10.7	3.4	6.13	61.9
3	1.16	55	90	4.20	6.6	3.2	6.15	63.7
4	1.76	55	50	5.55	66.0	3.5	5.79	63.7
5	1.69	55	70	4.55	10.1	2.5	5.82	65.0
6	1.67	55	90	3.87	5.5	2.3	5.83	67.5
7	2.17	55	50	5.25	35.0	1.5	5.67	65.0
8	2.21	55	70	4.47	10.1	1.9	5.66	67.7
9	1.98	55	90	3.80	5.36	2.3	5.72	70.2
10	1.28	62.5	50	6.83	43.6	2.7	7.31	61.1
11	1.15	62.5	70	5.42	8.26	2.4	7.45	62.1
12	1.10	62.5	90	4.95	6.01	2.6	7.51	62.7
13	1.58	62.5	50	8.17	$\infty$	3.3	7.01	64.9
14	1.45	62.5	70	5.42	8.99	2.4	7.18	65.8
15	1.44	62.5	90	4.83	5.94	2.6	7.19	67.7
16	2.12	62.5	50	7.42	$\infty$	1.9	6.87	66.3
17	1.99	62.5	70	5.07	7.5	2.4	6.91	68.4
18	1.84	62.5	90	4.40	4.7	1.9	6.97	69.4
19	1.24	70	50	9.08	$\infty$	3.6	8.53	64.0
20	1.10	70	70	5.82	6.27	3.3	8.71	63.6
21	1.11	70	90	5.00	4.16	3.2	8.69	64.9
22	1.62	70	50	9.99	$\infty$	2.7	8.21	67.2
23	1.52	70	70	7.05	16.9	2.1	8.27	68.3
24	1.52	70	90	5.43	5.46	1.8	8.27	67.2
25	2.10	70	50	8.33	$\infty$	2.2	7.97	70.0
26	1.75	70	70	6.56	11.8	1.8	8.13	70.6
27	1.51	70	90	5.62	6.06	2.3	8.28	70.9

\*Extraction times of 50, 70 and 90 min. were obtained by controlling screw speed at 3.76, 2.69 and 2.09 rpm, respectively.

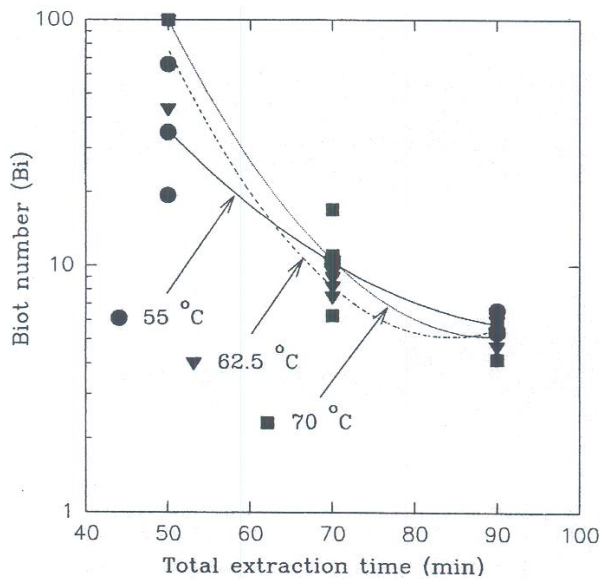


Fig. 5. Biot number vs. total extraction times and temperatures.

### 5.5 The Comparison Between the Experimental Data and the Predictions

All experimental data from 27 trials fitted well with the diffusion-backmixing model and Fig. 7 shows a sample run for comparison, the results was from trial number 1 (Table 1).

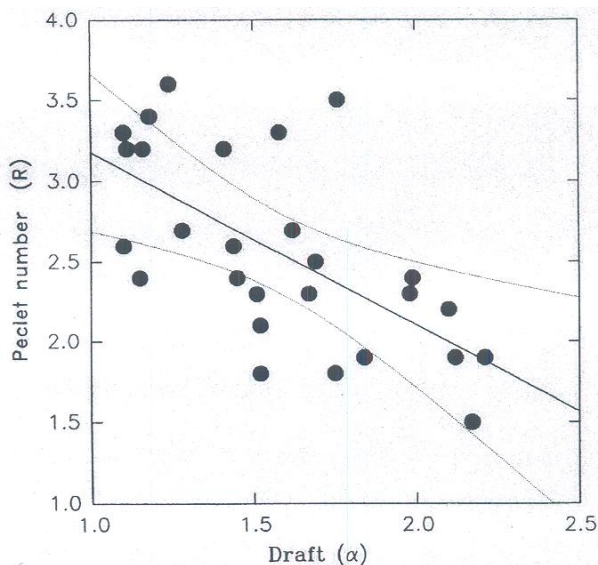


Fig. 6. The effect of draft on the Peclet number in liquid phase.

In Fig. 7, it can be seen that the model predicted the concentration profile well in both solid and liquid phases. However, the discrepancies were quite systematic. That is, the model predicted higher solute concentration in solid at near its entering end (the first 0-30 min.) but the predicted lower concentration nearby the other end (the last 30-60 min). The opposite discrepancies were found in the liquid phase. This clearly show that the assumptions of constant  $Bi$  and  $R$  along the extractor were not strictly valid. In fact, near the solid entering

end (the lower end) backmixing was considerably higher than at the liquid entering end where the flow approached plug flow. Similarly, it can be deduced that  $Bi$  was higher at the lower end than at the other end due to better fluid movement.

### 5.6 Yield

The summary of extraction yield for all 27 RCCE trials are presented in Fig. 8. In general, the higher the draft, temperature and extraction time, the higher was the yield. However, the increasing in the total extraction time too high (say > 90 min) was much less effective in obtaining greater yield.

It should be noted that the yields of all trials were not greater than 70 %. This can be clearly explained by considering simulation results of the diffusion-backmixing model as follows; Fig. 9 and 10 present the yield as related to  $Bi$ ,  $R$  and total extraction time at the temperature of 62.5 °C as predicted by the diffusion-backmixing model. Although it seems that, in general,  $Bi$  exhibits stronger effect on the yield than  $R$ , the fact that  $Bi > 4$  for all extraction trials indicate its weaker effect. On the other hand, as we found that  $R$  has the magnitude around 1.5-3.5, from Fig. 9 if we strict the total extraction time not to exceed 90 min, we will never achieve a higher yield than 80 % even  $Bi$  is infinite. Thus it was the combined effect of finite  $Bi$  and high backmixing that limited the extraction yield to about 70 % for all extraction trials.

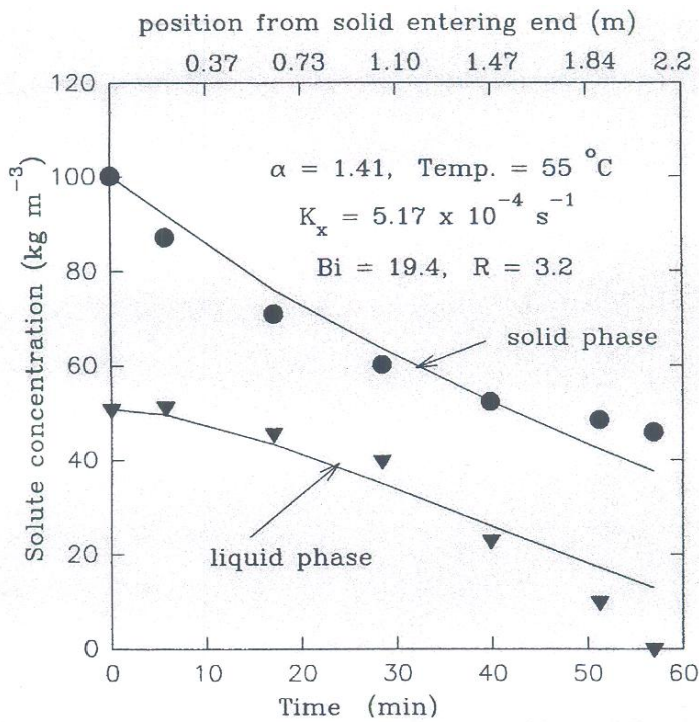


Fig. 7. A typical comparison between the experimental data and model prediction.

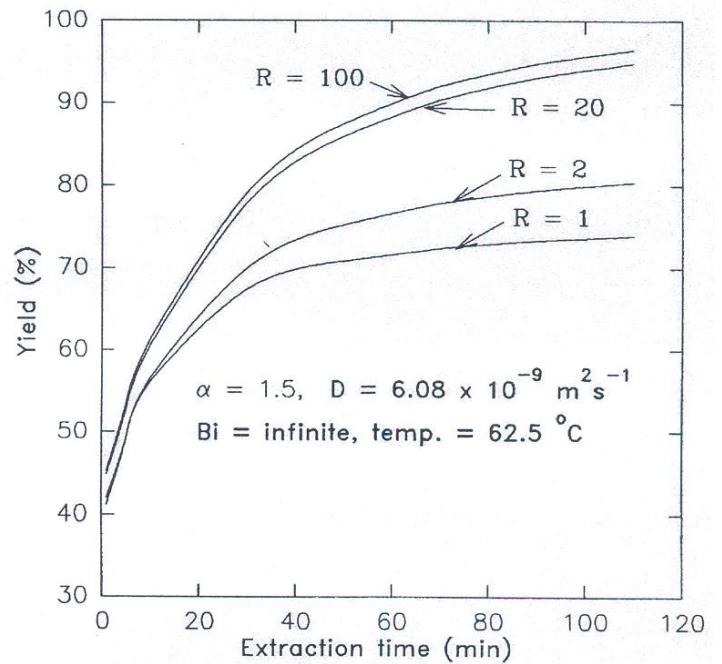


Fig 9. The effect of Peclet number ( $R$ ) and total extraction time on yield of extraction.

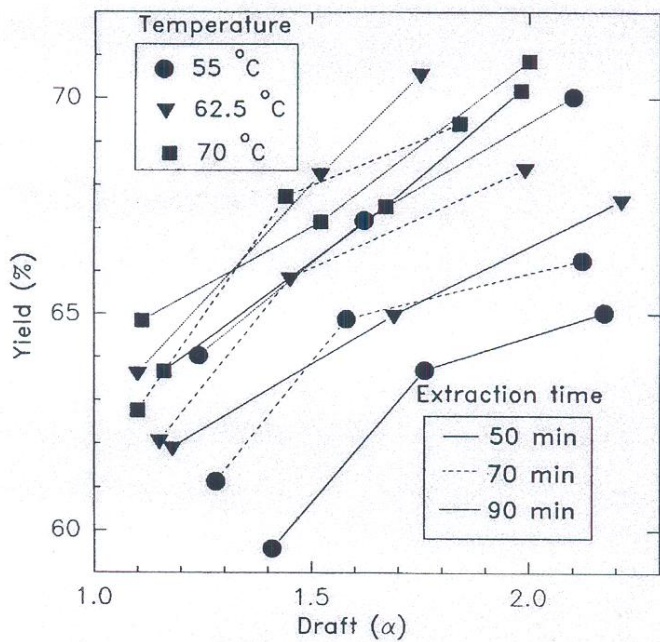


Fig. 8. Extraction yield vs. draft, temperature and extraction time.

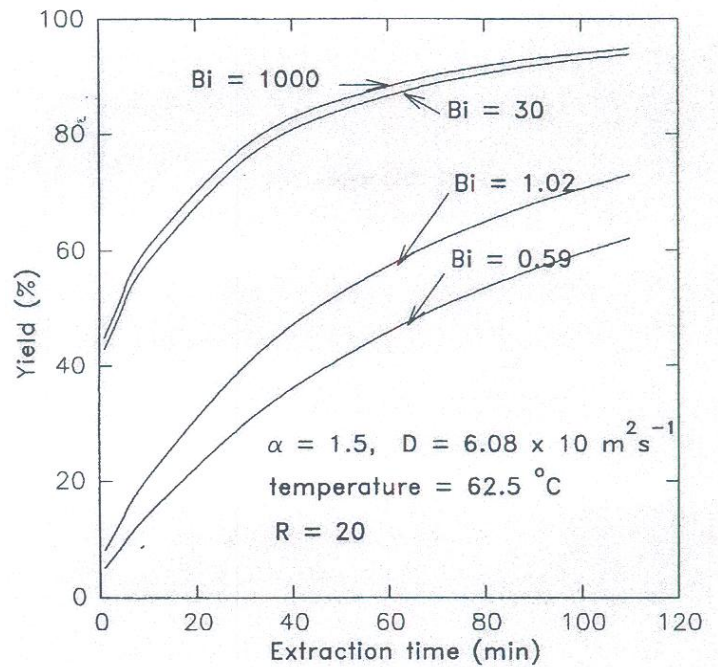


Fig. 10. The effect of Biot number ( $Bi$ ) and total extraction time on yield of extraction.



## 6 CONCLUSIONS

The study of mass transfer in the pineapple juice extraction using RCCE was critically conducted. The results indicated that the efficiency of the system can be improved by reducing axial dispersion while promoting local mixing. This may be achieved by increasing the length of the trough while keeping the total extraction time constant. However, since there are other factors which have not been investigated here such as the percent forward progression and the dimensions of raw material, further study should be carried out before the improving the design taking place.

The diffusion-backmixing model developed by Siripatana [6], Siripatana and Suparanon [18], and modified here in this work can be used to explain the experimental data satisfactorily. The model found corresponding to Rittirut *et al.*'s work for the case of steady state condition [12], [14]. Yet, it has advantage which should be noted : 1) it is simple, requiring only at hand calculation, 2) It can be used to design an extractor from fundamental parameters such as  $D$ ,  $\alpha$ , and  $t_p$ . The design requires only batch extraction trials.

## 7 NOTATIONS

$a$	= characteristic length
$A$	= area of mass transfer
$Bi$	= Biot number of mass transfer
$D$	= diffusivity from diffusion theory ( $m^2/s$ )
$D_s$	= solute diffusivity ( $m^2/s$ ); in-solid diffusivity ( $m^2/s$ )
$D_x$	= volumetric dispersion coefficient in solid phase ( $m^6/s$ )
$D_v$	= Volumetric dispersion coefficient in liquid phase ( $m^6/s$ )
$K_s$	= overall mass transfer coefficient ( $m/s$ )
$K_x$	= volumetric overall mass transfer coefficient ( $1/s$ )
$l$	= characteristic length (m) of each dimension of block (m)
$L$	= volumetric flow rate of liquid phase ( $m^3/s$ )
$L_0$	= Vol. flow rate of liquid phase at inlet of extractor ( $m^3/s$ )
$L_f$	= Vol. flow rate of liquid phase at outlet of extractor ( $m^3/s$ )
$m$	= solute equilibrium distribution coefficient
$P$	= Peclet number in solid phase
$R$	= Peclet number in liquid phase
$S$	= Vol. flow rate of liquid phase ( $m^3/s$ )
$S_0$	= Vol. flow rate of liquid phase at inlet of extractor ( $m^3/s$ )
$S_f$	= Vol. flow rate of liquid phase at outlet of extractor ( $m^3/s$ )
$Sh$	= Sherwood number
$t$	= time (s)
$t_f$	= theoretical extraction time as described by equation (4) (s)
$t_l$	= time lead due to initial solute distribution in solid (s)
$t_p$	= time delay due to plasmolysis (s)
$V$	= shape factor = $aA/V$
$V$	= ratio of liquid to solid flow rate = $L/S$
$V_x$	= accumulated vol. of solid phase from its inlet point ( $m^3$ )
$V_y$	= accumulated vol. of liquid phase from its inlet point ( $m^3$ )
$V_x$	= total volume of solid phase at position considered
$V_y$	= total volume of liquid phase at position considered
$\nabla_x$	= total volume of solid phase in extraction unit
$\nabla_y$	= total volume of liquid phase in extraction unit
$V$	= volume of mass transfer

$V_L$	= Liquid volume of mass transfer
$x$	= solute concentration in solid phase ( $kg/m^3$ )
$\bar{x}$	= average solute concentration in solid phase ( $kg/m^3$ )
$x^*$	= solute concentration in solid phase at equilibrium
$x_{in}$	= solute concentration in solid phase at inlet ( $kg/m^3$ )
$x_{out}$	= solute concentration in solid phase at outlet ( $kg/m^3$ )
$y$	= solute concentration in liquid phase ( $kg/m^3$ )
$\bar{y}$	= solute concentration in liquid phase at equilibrium
$y_{in}$	= solute concentration in liquid phase at inlet ( $kg/m^3$ )
$y_{out}$	= solute concentration in liquid phase at outlet ( $kg/m^3$ )
$z$	= relative length or dimensionless volume
$\alpha$	= draft of extraction
$\infty$	= infinite

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