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PREDICTION OF THE EXTRACTION PERFORMANCE OF A DIFFUSER USING A MATHEMATICAL MODEL

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ABSTRACT

The application of a mathematical model previously developed and tested on a pilot plant scale to a full-scale diffuser of the moving bed type is described. Together with a simplistic model of a de-watering mill, this model may be used to predict the extraction performance of a diffuser as a function of the diffuser operating conditions. The predictions of the model are compared with the actual extraction performance of a full-scale diffuser and satisfactory agreement between theory and practice is achieved. It is then shown through the use of this model how extraction is influenced by changes in process variables. The important effect of the juice flow system in a diffuser is established. The degree of fineness of bagasse is found to be the most important operating variable, and the effect of other variables on extraction is clarified.

INTRODUCTION

With the increasing attention being paid to extraction of sugar by diffusion the need for some means of predicting the extraction performance of a diffuser as a function of the particular diffuser operating conditions has assumed greater importance. For this reason, a research project was undertaken with the prime object of developing a mathematical model of the extraction process to represent the performance of a diffuser. This model was comprehensively tested by means of laboratory and pilot plant tests, the results of which have been published elsewhere.[3] This paper describes the application of the model to a full-scale diffuser of the moving bed type, the testing of the model by comparing actual diffuser results with the model, and the prediction of the effect of changes in process conditions on extraction in a diffusion system.

The mechanism of extraction of sugar from bagasse has been clearly established.[2] Extraction occurs by a combination of displacement-washing and molecular diffusion. It has been shown that in a packed bed of bagasse, extraction performance is influenced by liquid hydrodynamics.[3] High liquid flow rates promote the rate of mass transfer by improving the liquid-solid contact efficiency. In particular, the presence of static liquid trapped in and between bagasse particles significantly lowers the rate of extraction, since some juice in ruptured cells, which should be easily removed by a percolating liquid has to find its way by a slow diffusional process through the static liquid to the displacing liquid.

In essence, the diffuser model postulates are: Juice in broken cells on or near bagasse particle surfaces is readily extracted by a washing process at a rate K_1, while the remaining juice in unbroken cells and in broken cells in the interior of particles is extracted by a slower diffusional mechanism at a rate K_2. Static liquid is considered as part of the juice holdup in bagasse. The fraction of juice extracted at a rate K_1 is designated by α (where $0 \leqslant \alpha \leqslant 1$), and $(1 - \alpha)$

represents the remaining less accessible juice.

The model has been validated on a pilot plant scale and has been shown to provide an accurate description of extraction behaviour. The 3 parameters of the model, K_1, K_2 and α, have been evaluated over a wide range of operating conditions in a pilot plant diffuser.[3]

FORMULATION OF THE MODEL FOR A MOVING BED DIFFUSER

The majority of diffusion installations are of the moving bed type, and therefore the model has been formulated specifically for this type of diffuser.

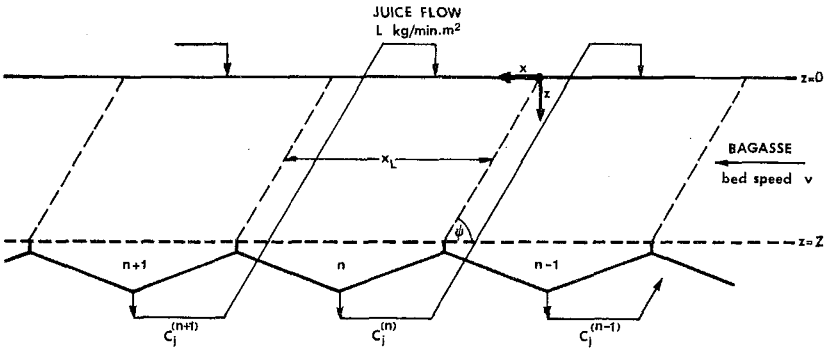

FIGURE 1. Basis for formulation of model of a single stage in a moving-bed diffuser.

In the first instance the model is formulated for a single stage within the diffuser. Consider the nth stage shown in Fig. 1. Juice percolating through the bed is displaced in the x-direction due to the movement of the bagasse. In order to formulate the model it is necessary to make the following assumptions:

1) Steady-state operation, so that all quantities concerned are independent of time.
2) Liquid is uniformly applied over the top of a stage, and flows in plug flow through the bed.

Mass balances over an element within the bed of height dz, length dx and unit width yield the following differential equations:
for the percolating juice,

$$-L\frac{\delta C_j}{\delta z} - \frac{L}{\tan\psi}\frac{\delta C_j}{\delta x} + K_1 (C_{b1} - C_j) + K_2 (C_{b2} - C_j) = 0 \qquad (1)$$

and for the juice in bagasse,

$$-\alpha H v \frac{\delta C_{b1}}{\delta x} = K_1 (C_{b1} - C_j) \qquad (2)$$

$$-(1-\alpha) H v \frac{\delta C_{b2}}{\delta x} = K_2 (C_{b2} - C_j) \qquad (3)$$

Boundary conditions are given by the concentration of juice applied to this stage and the concentration of juice in bagasse entering the stage. To preserve mathematical tractability, it is assumed that C_j varies in the x and z directions, while C_{b1} and C_{b2} vary only with x and not z. Therefore equations (2) and (3) need to be replaced by 2 slightly modified integro-differential equations. Details of the solution of these equations are published elsewhere,[3] and are not given here. The final solutions have the following form:

$$C_j^{(n)} = g_o C_j^{(n+1)} + g_1 C_{b1}^{(n-1)} + g_2 C_{b2}^{(n-1)} \tag{4}$$

$$C_{b1}^{(n)} = e_o C_j^{(n+1)} + e_1 C_{b1}^{(n-1)} + e_2 C_{b2}^{(n-1)} \tag{5}$$

$$C_{b2}^{(n)} = d_o C_j^{(n+1)} + d_1 C_{b1}^{(n-1)} + d_2 C_{b2}^{(n-1)} \tag{6}$$

where the coefficients g_i, e_i and d_i are functions of K_1, K_2, α, Z, L, x_L, H and v (nomenclature used is listed at the end of the paper). Since the juice in bagasse is divided into 2 fractions, 2 quantities C_{b1} and C_{b2} are required to represent the brix of each fraction. The average brix of juice in bagasse C_b is given by

$$C_b = \alpha C_{b1} + (1 - \alpha) C_{b2} \tag{7}$$

Superscripts are used to denote stage numbers. $C_j^{(n)}$ represents the brix of juice in the nth tray, and $C_b^{(n)}$ represents the brix of juice in bagasse leaving the nth stage.

Once the brix of percolating juice and juice in bagasse entering the nth stage are known, the brix of juice in the nth catch-tray and juice in bagasse can be calculated using equations (4) to (6). Thus a progressive stagewise calculation procedure is required. In addition, the overall mass balance must be satisfied; referring to Fig. 2, this may be expressed as:

$$Lx_L W C_j^{(1)} + M_b HC_b^{(N)} = M_b HC_b^{(o)} + \frac{PW}{100} M_b C_{pw} \tag{8}$$

Note that $C_j^{(N+1)} = 0$ (imbibition water) and so does not appear in equation (8). Also, evaporation in the diffuser is neglected.

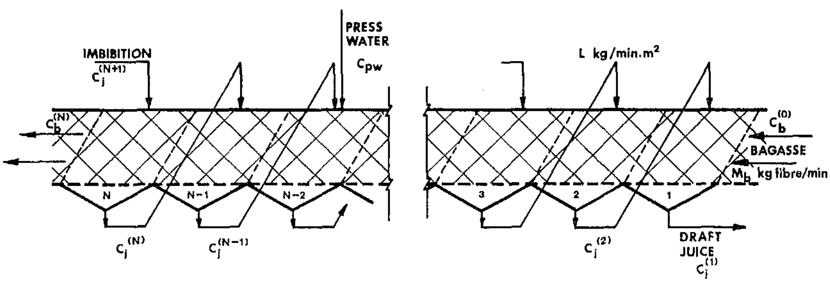

FIGURE 2. Schematic diagram of diffuser.

Allowance for recirculation of juice within a stage

The stagewise calculation procedure referred to above assumes that all juice applied to a given stage finds its way into the correct catch-tray. Clearly this is not always true in practice. Even if the sprays are correctly positioned relative to the catch trays, a small amount of liquid will find its way into the preceding and succeeding trays. This however would be balanced by corresponding flows from adjacent stages, of similar brix, and the overall effect is likely to be negligibly small.

If juice sprays or distribution weirs are not optimally positioned relative to their corresponding catch trays, recirculation or by-passing of liquid will occur, depending on whether the juice spray is too little or too far advanced ahead of the catch tray. By-passing is a complicated effect which reduces extraction performance by lowering juice flow rates, and should not occur in well-designed diffusers. Recirculation however can be used to improve extraction performance.

A diagrammatic sketch of juice recirculation within a stage is shown in Fig. 3. A slightly different computational procedure is required to handle this situation.

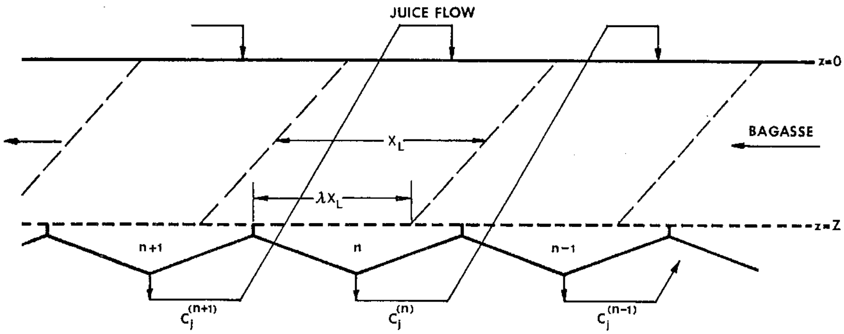

FIGURE 3. Schematic representation of juice recirculation within a stage of the diffuser.

The degree of recirculation is characterised by the parameter λ. Referring to Fig. 3, it can be seen that λ is defined as the fraction of liquid applied to the nth stage which finds its way into the nth catch-tray. The case of $\lambda = 1$ represents zero recirculation.

It can be easily shown that the liquid flow rate increases from the value L calculated with no recirculation to a value L_r where

$$L_r = \frac{L}{\lambda} \tag{9}$$

The greater the degree of recirculation, the lower is λ and the higher the flow rate. Excessive recirculation may lead to an increase in the value of L_r to such an extent that flooding occurs.

The model has been modified to handle the case of constant recirculation in every stage. However, different values of λ in the stages before and after the presswater entry stage can be handled by the model. Since the juice in each tray now consists of 2 components, firstly a fraction λ of the juice applied to the nth stage and secondly a fraction $(1-\lambda)$ applied to the $(n-1)$th stage, both components need to be evaluated and suitably combined to calculate $C_j^{(n)}$.

Equation (4) for $C_j^{(n)}$ is obtained by integrating the value of C_j at $z = Z$ over the range $x = 0$ to $x = x_L$ to obtain $C_j^{(n)}$ as the average value of C_j at the bottom of the bed. In the case of recirculation the 2 components of $C_j^{(n)}$ are evaluated by integrating over the ranges 0 to λx_L and λx_L to x_L separately. The algorithms required are easily derived.

Model of the de-watering stage

Calculation of the composition of wet bagasse leaving the diffuser is of little utility unless this can be related to the analysis of final bagasse. A simplistic model of a de-watering mill was therefore developed to predict the composition of final bagasse.

Briefly, this model assumes that all readily available juice is expressed preferentially in a de-watering mill before any of the tightly-held juice. Since the readily available juice is more accessible to the percolating liquid, it must physically constitute surface juice, and should therefore be removed first in a mechanical expression process.

The postulates of the model are:

i) all readily available juice and only part of the tightly-held juice is expressed.
ii) the remaining tightly-held juice is not disturbed and retains its identity as such,
iii) subsequently, a certain amount of juice is reabsorbed. The amount of reabsorption is represented by a reabsorption factor k, as defined by Murry and Holt.[1]
iv) all expressed juice is well-mixed, and the reabsorbed juice has the same brix as the press water.

This model has one parameter k which has been found to have a value of the order of 1,3 in Australia.[1] The value to be used in this case should be found by comparing predictions of the model with experimental data.

The fact that the readily available juice is expressed first in a de-watering mill has been substantiated by measurements made at Empangeni,[4] which showed that the brix of adhering juice in discharge bagasse is very close to the brix of juice taken from the front roller of the first de-watering mill. The figures are shown below (16 data points):

	mean brix	std deviation
Adhering juice in discharge bagasse	1,42	0,29
Juice from front roller of de-watering mill	1,40	0,27

The equations required to predict brix % bagasse and presswater brix are derived using the above assumptions and a volumetric extraction theory approach utilised by Murry and Holt.[1]

CALCULATION PROCEDURE

The model has been formulated for a bagasse diffuser. First mill extraction has to be specified in the model, and together with values of fibre % cane, brix % cane and moisture % 1st mill bagasse, is used to calculate the composition of bagasse entering the diffuser. Specification of the fibre content of diffuser bagasse and final bagasse is required to compute the amount of presswater, and a value of the reabsorption factor is required for the de-watering stage model. Values of the parameters of the diffuser model i.e. K_1, K_2 and α as well as the static juice holdup in bagasse are calculated from correlations developed from the results of a pilot plant investigation,[3] as a function of liquid flow rate, liquid properties, fibre density, PI and bagasse specific surface S. Values of S were obtained from sieve analysis, and are calculated as:

$$S = 60 \; \Sigma \frac{w_i}{x_i} \tag{10}$$

where w_i is the weight % retained on the ith screen and x_i the corresponding mean particle size.

An iterative computation scheme is required for the diffuser model. Stage calculations using equations (4) to (6) are started at stage 1 (feed end) using an assumed value of $C_j^{(1)}$ and continued stage by stage to the discharge end. Allowance is made at the point of presswater entry for the brix and amount of presswater and at the discharge end for juice carry-over. If the mass balance across the diffuser is not satisfied, a new value of $C_j^{(1)}$ is assumed and the calculation procedure repeated.

If the presswater brix is not specified another iteration loop is required. A value of presswater brix is assumed, calculations for diffuser and de-watering stages are carried out and the predicted presswater brix is used for a repeat calculation. This procedure terminates when assumed and predicted brix values agree within a specified tolerance (0,01).

For the case where juice recirculation occurs, the value of λ must be specified for the programme. Alternatively, the liquid flow rate within each stage can be nominated to be a given percentage of the flooding flow rate, and the value of λ is then computed.

COMPARISON OF MODEL PREDICTIONS WITH ACTUAL DIFFUSER EXTRACTION RESULTS

For comparison purposes, average results for a week's operation were obtained from the BMA diffuser at the Huletts Empangeni Mill. Weekly data were used to eliminate the effect of large random fluctuations observed over shorter time periods. Some of the data from Empangeni used as input to the computer is shown in Table 1; this also illustrates what information is required for the model.

Such data is however subject to certain inconsistencies which arise particularly where sampling and analysis of bagasse is concerned. It is because of

TABLE 1. Data from Empangeni diffuser used in diffuser simulation.

Number of stages = 12. Presswater entry stage = 10. Width, W = 4,80 m. Length/stage, x_L = 3,25 m.

	Week 23 1969 - 70	Week 30 1969 - 70	Week 12 1970 - 71	Week 21 1970 - 71	Week 22 1971 - 72
TCH	199	204	181	171	210
F_c	17,2	17,5	18,5	19,0	15,9
Bx_c	15,9	15,5	17,0	18,5	15,7
E_1	57,1	59,3	53,6	54,1	56,4
M_1	60,6	59,5	57,2	54,1	56,9
IMB%F	287	278	309	318	288
T	73	73	75	78	76
S	3 000	3 200	3 300	3 300	3 300
q	73,7	69,5	79,4	83,5	72,1
v	0,94	0,98	0,91	0,79	0,94
F_D	18,6	15,8	18,7	17,0	17,3
C_{pw}	3,7	2,8	1,9	1,8	1,9
F_B	45,0	44,3	43,8	43,4	43,9

such unreliability of factory data that the model was validated on a pilot plant scale. Such comparisons serve only therefore to confirm that the model can adequately represent full-scale diffuser performance, and to evaluate the reabsorption coefficient k in the de-watering stage model.

Values of extraction calculated by the model for the input data of Table 1 are compared with actual recorded values of extraction in Table 2. Different results are obtained depending on the values assumed for λ and k.

TABLE 2. Comparison of model predictions with actual Empangeni extraction results.

		E_D	E	LAJ	C_{pw}	λ	k
Week 23	actual	23,5	94,0	28,3	3,7		
1969 - 70	predicted	22,5	93,9	29,4	4,2	1,0	1,3
		23,9	93,2	32,9	3,6	0,8	1,1
		23,9	94,0	28,6	3,8	0,8	1,3
Week 30	actual	22,4	93,5	29,7	2,8		
1969 - 70	predicted	19,7	93,1	32,4	3,1	1,0	1,1
		21,1	93,3	31,8	2,8	0,8	1,1
		21,1	94,3	27,0	3,0	0,8	1,3
Week 12	actual	34,4	94,3	25,0	1,9		
1970 - 71	predicted	32,7	93,9	27,0	2,3	0,8	1,1
		32,7	94,9	22,8	2,6	0,8	1,3
		34,8	94,1	25,9	1,7	0,6	1,1*
Week 21	actual	34,4	95,1	20,9	1,8		
1970 - 71	predicted	31,1	94,3	24,5	2,5	1,0	1,1
		31,1	95,1	20,9	2,7	1,0	1,3
		32,6	94,4	23,8	2,1	0,8	1,1*
Week 22	actual	20,0	94,9	26,6	1,9		
1971 - 72	predicted	27,5	94,7	28,1	3,0	0,8	1,3
		29,6	93,9	32,1	2,2	0,6	1,1
		29,6	94,9	26,6	2,5	0,6	1,3

* Indicates flooding predicted from Fig. 5.

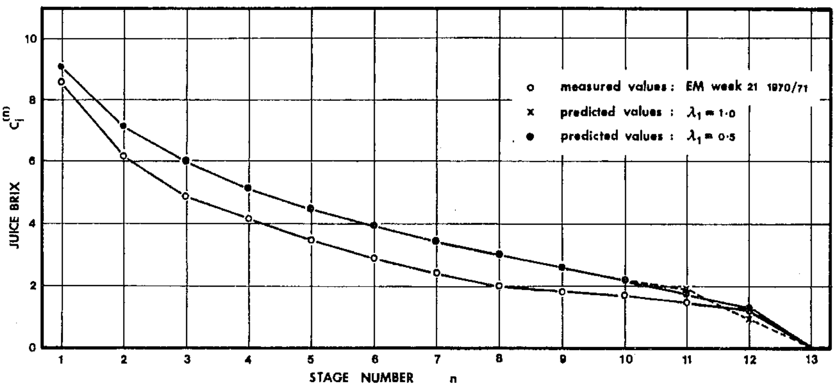

FIGURE 4. Comparison of measured and predicted values of $C_j^{(n)}$.

The diffuser spray system at Empangeni is set up to promote recirculation in the stages after the presswater entry stage. Thus a value of λ in these stages, λ_1, of 0,5 was used (calculated value), and the effect on the brix profile in the diffuser is shown in Fig. 4. Clearly, if no recirculation is assumed ($\lambda_1 = 1{,}0$) unrealistic results are obtained.

The fact that the predicted juice profile shown in Fig. 4 is higher than observed values is due to the fact that the maceration stage was neglected in the model. Thus the predicted value of $C_j^{(1)}$ represents the diffuser draft juice brix, but the actual draft juice brix is higher than the value measured on juice from tray 1.

Evaporation in the diffuser would tend to inflate values of C_j. However these values are still lower than the predicted values. The net effect on the model predictions is that the model should underestimate extraction performance slightly, because the higher juice brix implies a reduced driving force between percolating juice and juice in bagasse.

Reference to Table 2 shows that if no recirculation is assumed, i.e. $\lambda = 1$, the model predictions are generally lower than actual results. However, if recirculation of the order of 20% ($\lambda = 0{,}8$) is assumed, agreement between observed and predicted brix extraction in the diffuser E_D is good.

As a consequence of recirculation, the juice flow rate within each stage is increased as shown by equation (9). In general, the flow rate must be kept below the flooding flow rate L_f; a correlation for values of L_f as a function of bagasse

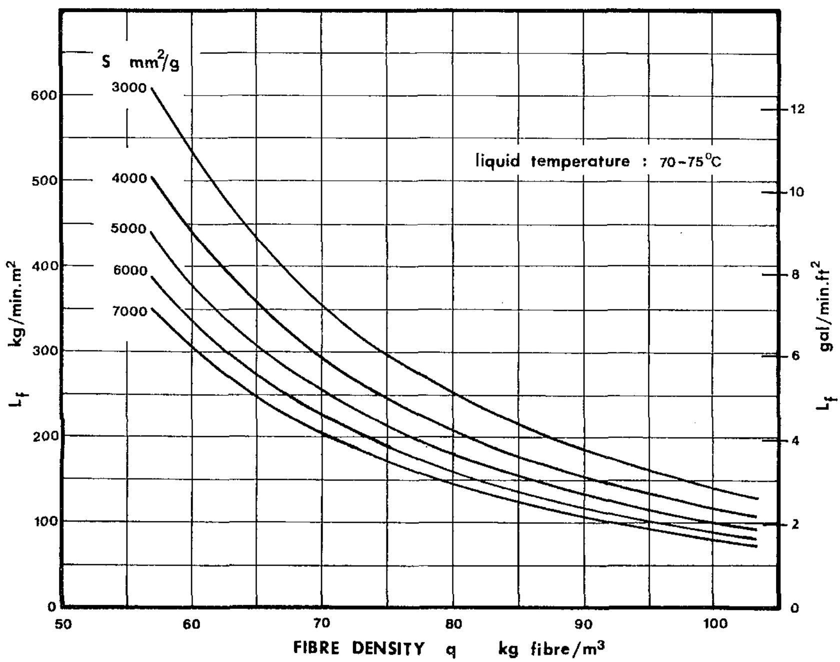

FIGURE 5. Effect of fibre density and preparation on the flooding flow rate L_f.

specific surface S and fibre density q was developed in a pilot plant diffuser[3] and is shown graphically in Fig. 5. In the absence of recirculation, values of L are generally considerably lower than L_f. In practice, the diffuser was normally operated at a condition approaching flooding which was achieved by recirculation of liquid, as confirmed by tracer tests.

Table 2 shows that the use of values of reabsorption coefficient of ≈ 1,2 leads to a good representation of the de-watering stage. It should be remembered that the values obtained here represent a 2-stage de-watering process rather than a single mill.

The results of Table 2 show that the model is capable of providing a satisfactorily accurate representation of extraction performance. Predicted and observed values of E_D are comparable because the same value of fibre % diffuser bagasse is specified in each case. However, it should be remembered that observed values of E_D are the least accurate because of the difficulty of sampling and analysing wet diffuser bagasse.

DEPENDENCE OF EXTRACTION ON DIFFUSER OPERATING CONDITIONS

Having established that the model does provide a valid description of extraction performance, each primary variable may now be varied to see how it affects extraction. In this way the quantitative dependence of extraction on operating variables may be obtained. It is however seldom possible to change only 1 variable without changing other conditions in the diffuser, since the process variables often interact. In particular, the liquid flow patterns in the diffuser are sensitive to changes in diffuser operating conditions, which is an aspect of diffuser operation which has often been neglected in the past.

It is necessary to choose a base case from which deviations in operating conditions can be made. Details of the base case conditions are given in Table 3. Under these conditions, a brix extraction of 95,33% is predicted, or a value of LAJ% fibre of 24,5. The flooding flow rate L_f in Table 3 is obtained from Fig. 5 for the values of S and q given in Table 3.

TABLE 3. Specifications of diffuser base case conditions.

Diffuser:			
number of stages = 10	W	=	4,80 m
presswater entry stage = 9	x_L	=	3,90 m
Cane analysis:			
F_c = 16,0 %	Bx_c	=	15,3 %
First mill:			
E_1 = 55,0 %	M_1	=	57,5 %
Diffuser operating conditions:			
IMB % F = 300	v	=	0,88 m/min
TCH = 200	T	=	75 C
S = 4 000 mm²/g	Z	=	1,57 m
L = 147 kg/min m³	L_f	=	205 kg/min m²
λ = 1,0	q	=	80,1 kg/m³
De-watering stage:			
F_D = 17,5%	k	=	1,2
F_B = 45,0%			

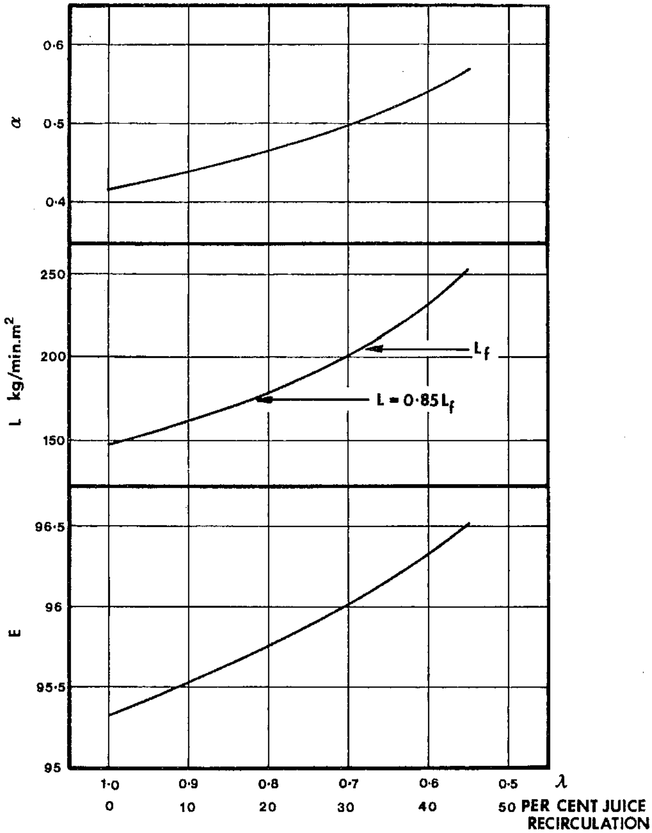

FIGURE 6. Effect of recirculation of juice within a stage.

Effect of juice flow pattern in the diffuser

The base case considered above assumes that no recirculation of liquid occurs within a stage. The effect of juice recirculation on the base case is shown in Fig. 6. This shows that extraction can be significantly improved by inducing recirculation to promote higher liquid flow rates. The effect of higher flow rates is to improve the liquid-solid contact efficiency, thus allowing more of the sucrose to be extracted by washing and less by diffusion, thereby increasing the overall rate of extraction. Fig. 6 shows how the value of α (representing the fraction extracted by washing) increases as the recirculation and flow rate increase. However if the degree of recirculation is increased too much, flooding will occur (in this case at $\lambda = 0{,}68$ or 32% recirculation).

Obviously it would be beneficial to operate at as high a flow rate as possible (i.e. flooding) without ever increasing the flow rate above this value. In practice this is not normally possible.

The value of λ depends not only on the displacement between juice sprays or weirs and catch-trays, but also on cane throughput and liquid holdup which in turn depends on bagasse particle size, fibre density and imbibition level. Changes in cane quality for instance from one consignment to the next would cause both the degree of recirculation and the flooding flow rate to change.

Effect of degree of preparation

In measuring the degree of preparation of 1st mill bagasse, the use of PI can be misleading as its value depends on the efficiency of the 1st mill in separating juice from fibre. Therefore in this case particle size obtained from sieve analysis is used as the primary measure of preparation. It is expressed as specific surface S in mm²/g, so higher values of S imply finer preparation. Typically, diffusion factories in South Africa operate at values of S of the order of 3 500.

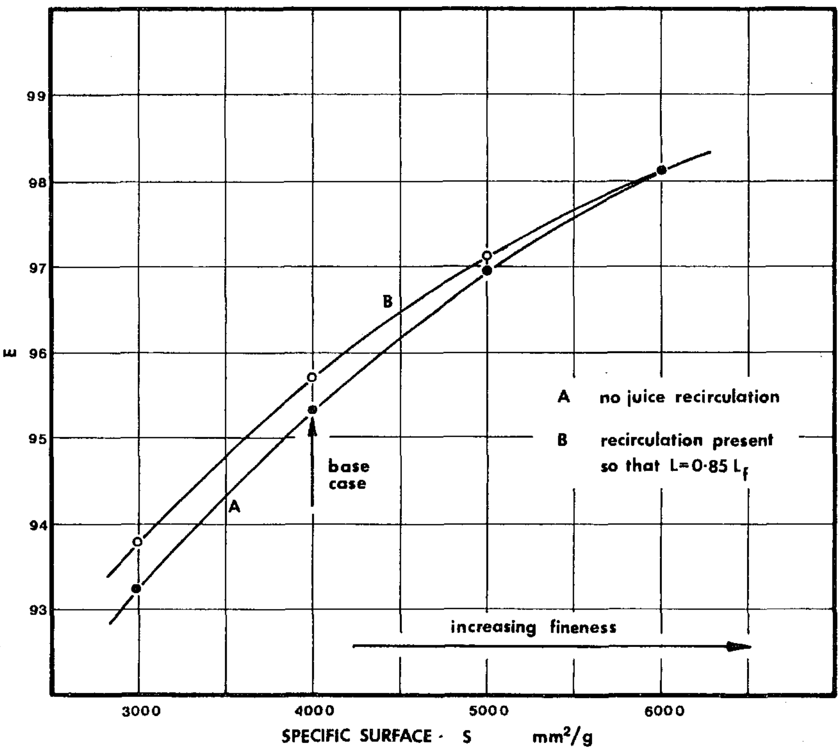

FIGURE 7. The dependence of extraction on fineness of bagasse.

Fig. 7 shows how extraction depends on the fineness of bagasse. Two curves are shown: the first marked A is obtained assuming in each case that no juice recirculation occurs; the second B is obtained assuming that sufficient recirculation occurs so that in each case the flow rate through each stage is 85% of the flooding rate. This curve therefore takes into account the fact that higher flow rates are possible with coarser preparation. It is still very obvious however that the net effect of increasing the fineness of preparation is an increase in extraction.

A strong dependence of extraction on degree of preparation is demonstrated. It should also be remembered that these results assume a constant 1st mill extraction. In fact a finer preparation may well increase 1st mill extraction and thus inflate the trend with degree of fineness somewhat.

Effect of other changes in diffuser operation

The effect of imbibition level on extraction is two-fold. Firstly higher levels result in generally lower juice brix values leading to an increased driving force (brix difference) for mass transfer; secondly higher imbibition levels lead to higher interstage flow rates in the absence of juice recirculation.

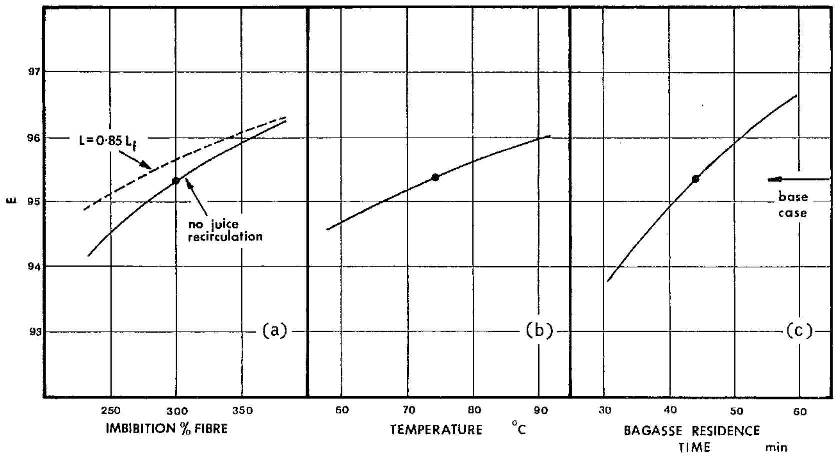

FIGURE 8. Effect of changes in operating conditions on extraction.

The effect of imbibition on extraction is shown in Fig. 8a. If the degree of recirculation can be varied the second advantage of higher interstage flow rates falls away as the recirculation can be adjusted to maintain a constant liquid flow rate. The dotted curve in Fig. 8a represents the case of constant flow rate maintained at 85% of flooding.

Higher temperatures are advantageous in diffusion since they promote the rate of extraction. Apart from the fact that the rate of molecular diffusion is greater, reduced liquid viscosities at higher temperatures lead to better liquid-solid contacting and therefore more efficient mass transfer.[3] The effect of temperature on extraction is represented in Fig. 8b.

Fig. 8c shows the dependence of extraction on bagasse residence time. Clearly some effort should be made to run at maximum possible residence time subject to constraints of bed height, bed speed and crushing rate, as an increase of only a few minutes implies a significant increase in extraction.

Table 4 summarises the effect of these variables on extraction, and in addition includes the effect of 1st mill extraction, fibre % final bagasse, brix % cane and fibre % cane. The value of each variable x_i in the base case and the change in overall extraction ΔE as a result of a change in each variable Δx_i are shown. Negative values in the table imply an inverse relationship between E and x_i (e.g. E increases as fibre % cane decreases). It should be remembered

TABLE 4. Sensitivity of extraction E to changes in variables x_i.

Variable x_i	Value of x_i in base case	Δx_i	ΔE	Remarks
specific surface S (fineness)	4 000	500	0,92	Constant 1st mill extraction
imbibition % fibre	300	30	0,41	
juice flow rate L	147	24	0,33	
recirculation (%)	0	16		
temperature	75	5	0,22	constant q, F_D, F_B
bagasse residence time	44,1	5	0,50	
1st mill extraction %	55	5	0,4	constant M_1
fibre % bagasse	45	1	0,27	
brix % cane	15,3	1	0,07	
fibre % cane	16	1	−0,06 −0,23	constant imb % fibre constant imb % cane

that the results above are obtained by changing one variable at a time, with all others held constant. Possible interactions such as between fibre % cane and 1st mill extraction may introduce slight differences in practice.

CONCLUSIONS

The utility of a mathematical model in simulating the extraction performance of a diffuser has been demonstrated. Through the use of the diffuser model quantitative predictions of the effect of changes in the operation of the extraction system have been obtained. The results summarised in Table 4 have identified the major variables affecting extraction and serve as a guide to possible gains to be achieved in improving the operation of a diffuser. The model also enables the design of a diffuser to be carried out on a rational basis.

Although the model has been validated on a pilot plant scale, a rigorous comparison between model predictions and actual diffuser extraction results cannot be achieved because of the doubtful accuracy of the factory data. Nonetheless it appears that the model can be used with some confidence to predict extraction performance under different operating conditions or in a diffuser of different configuration. Even if absolute levels of extraction predicted can be faulted, the model should give a reliable indication of the relative effect of changes in diffuser design or different operating policy.

NOMENCLATURE

Bx_c	brix % cane
$C_b^{(n)}$	brix of juice in bagasse leaving nth stage
C_{b1}	brix of readily available juice in bagasse
C_{b2}	brix of tightly held juice in bagasse
$C_j^{(n)}$	brix of percolating juice in nth tray
C_{pw}	presswater brix
E	overall brix extraction (%)

E_1	brix extraction in 1st mill (%)
E_D	brix extraction achieved in diffuser (%)
F_B	fibre % final bagasse
F_c	fibre % cane
F_D	fibre % diffuser discharge bagasse
H	juice holdup in bagasse (kg/kg fibre)
K_1	mass transfer coefficient (kg/min m^3)
K_2	mass transfer coefficient (kg/min m^3)
k	reabsorption coefficient
L	liquid mass flow rate (kg/min m^2)
L_f	flooding liquid flow rate (kg/min m^2)
LAJ	lost absolute juice % fibre
M_1	moisture % 1st mill bagasse
M_b	mass flow fibre in bagasse (kg fibre/min)
N	number of stages in diffuser
PI	preparation index
PW	presswater % fibre
q	fibre packing density (kg fibre/m^3)
S	specific surface of bagasse particles, defined by equation (10) (mm^2/g)
T	temperature (C)
TCH	crushing rate (tons cane/hour)
v	bagasse bed velocity (m/min)
W	width of diffuser (m)
x	co-ordinate in direction of movement of bagasse (m)
x_L	length of 1 stage in diffuser (m)
Z	height of bagasse bed (m)
z	distance from top of bagasse bed (m)
α	fraction of juice in bagasse readily available for extraction
λ	parameter characterising degree of circulation
λ_1	value of λ in post-presswater entry stages

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PREDICCION DEL TRABAJO DE EXTRACCION DE UN DIFUSOR UTILIZANDO UN MODELO MATEMATICO

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RESUMEN

Se describe la aplicación de un modelo matemático previamente desarrollado y probado en una planta piloto a un difusor de escala commercial del tipo de colchón móvil. Junto con un modelo simplista de un molino desaguador, este modelo puede ser usado para predecir la operación de un difusor como una función de las condiciones de operación del difusor. Las predicciones del modelo son comparadas con los resultados reales de un difusor comercial y se logra un acuerdo satisfactorio entre la teoría y la práctica. Se demuestra luego con la utilización de este modelo como varía la extracción con cambios en las variables del proceso. Se establece la importancia del efecto del sistema de flujo del jugo en el difusor. Se encuentra que el grado de finura del bagazo es la variable de operación de mayor importancia, y se aclara el efecto de otras variables en la extracción.