OPTIMAL EVAPORATOR OPERATION

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Abstract

Factors influencing the heat transfer coëfficient in an evaporator are discussed, particularly the effects of concentration and scale formation. Theoretical policies for maximizing the average heat transfer coefficient are suggested. Based on data collected from a real evaporator, recommendations are made for applying these policies in practice.

Introduction

Inefficient heat transfer in a sugar factory implies unused production capacity. Attempts have been made to predict the performance of industrial evaporators with a view to optimizing the variables affecting the process. However, the results obtained are usually applicable only in particular circumstances and are largely dependent on the properties of the cane. The most reliable way to assess the efficiency of a given evaporator appears to be to perform tests on it under normal operating conditions. With the advent of minicomputers and the availability of sophisticated on-line measuring instruments, this becomes a feasible undertaking.

In 1974 a data-logging minicomputer was installed at Jaagbaan to sample data from the fourth effect of their multiple effect evaporator. The major aim of the exercise was to observe the time trend in the heat transfer coefficient due to the accumulation of scale within the calandria tubes. Exit concentration, liquid level and temperatures were also monitored.

An optimum cleaning policy was derived from an expression for the theoretical rate of scale growth reported in the literature. The data collected from the real evaporator were then analysed in the light of this policy.

Optimum cleaning schedule

Assume that the production rate is approximately proportional to the sum of the total evaporation rates from the multiple effect evaporator and the vacuum pan station, and that the former decreases with time due to the accumulation of scale on the evaporator tubes, while the latter remains approximately constant.

Now, if E_i is the vapour rate from the ith effect, then, by the nature of the operation of a multiple effect evaporator:

$$E_1 + E_2 + E_3 + E_4 \approx 4E_4 + \text{vapour bleed rate.}$$

Further, since $E \approx UA\ \Delta T/H$ (ignoring flash evaporation), if ΔT in the fourth effect is constant E_4 is linearly proportional to the heat transfer coefficient, U(t), in that effect.

Lastly, assume a negligible variation in the vapour bleed rate. Then we can express the production rate P(t) in the form:

$$P(t) = k_0 + k_1\,U(t)$$

where k_0 and k_1 are constants.

In 1924 McCabe and Robinson[1] derived the expression:

$$U(t) = U(0)\,(1 + at)^{-\frac{1}{2}}. \qquad (1)$$

This gives the theoretical heat transfer coefficient as a function of time, assuming that the rate of scale growth is directly proportional to the heat flux. U(0) is the heat transfer coefficient for clean tubes and "a" is a constant. Fig. 1 shows a typical graph following this function.

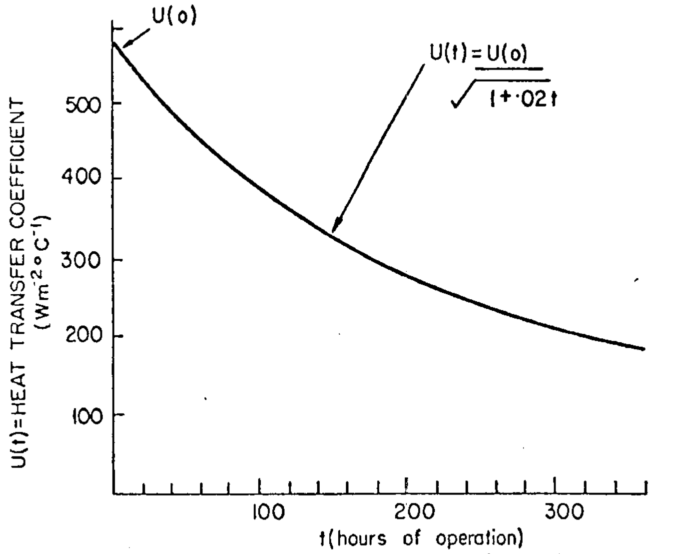

FIGURE 1 The change in the overall heat transfer coefficient of an evaporator with time of operation.

If the heat transfer coefficient decreases in the fashion shown in Fig. 1, it is desirable to know the optimum time to shut the plant down for cleaning. This decision should be made with a view to maximising the average evaporation rate for each production cycle. This will maximize total production for a fixed season or minimize seasonal operating time for a fixed production target.†

Let t_p be the production time and t_c the cleaning time for one cycle. The cleaning time is assumed to be constant; it was about 16 hours at the time of the investigation. The situation is depicted graphically in Fig. 2. For short cycles the time spent on cleaning is relatively large and represents wasted production time. On the other hand, long cycles allow more scale to accumulate so that heat transfer coefficients are low for much of the time. Clearly, there is an optimum which lies between these extremes.

The average production rate $\overline{P}$ is given by

$$\overline{P} = \frac{\int_0^{t_p} P(t)\,dt}{(t_p + t_c)}$$
$$= \int_0^{t_p} [k_0 + k_1\,U(t)]\,dt/(t_p + t_c)$$
$$= \frac{k_0\,t_p}{t_p + t_c} + k_1\,\overline{U}$$

where $\overline{U}$ is the average heat transfer coefficient.

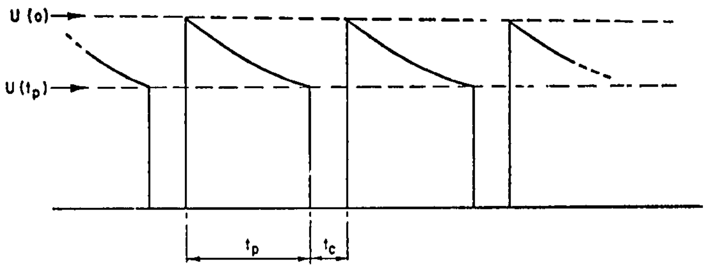

FIGURE 2 Variation of the heat transfer coefficient
t_p = Production time
t_c = Cleaning time

† Assuming all cycles are uniform.

If t_c is small compared to t_p, $\overline{P}$ can be approximated by the expression:

$$\overline{P} \approx k_o + k_1 \overline{U}$$

Clearly to maximize $\overline{P}$, $\overline{U}$ must be maximized. Mathematically this is expressed as "maximize the average heat transfer coefficient, $\overline{U}$, by choice of the production time, t_p, i.e.

$$\max_{t_p} \{\overline{U}\} \quad (2a)$$

$$\text{where } \overline{U} = \frac{1}{(t_p + t_c)} \int_0^{t_p} U(t)\, dt \quad (2b)$$

If the heat transfer coefficient for clean tubes and the rate of decline are the same for every cycle, then the optimal cycle time will be constant and this policy will maximize the average production rate for the whole season. This will not be the case for a non-uniform trend but since we have no knowledge of the future, we must work on the former assumption.

To find t_p^*, the optimum value of t_p, we differentiate (2a) with respect to t_p, equate the result to zero and solve for t_p. Let $\int_0^{t_p} U(t)dt = F(t_p) - F(0)$. Then we arrive at an equation of the form:

$$\frac{U(t_p^*)}{(t_p^* + t_c)} - \frac{F(t_p^*) - F(0)}{(t_p^* + t_c)^2} = 0. \quad (3)$$

Equation (3) can be re-written as:

$$U(t_p^*) = \overline{U}. \quad (4)$$

Equation (4) implies that as long as the actual heat transfer coefficient is larger than the average value, calculated as if the plant were to be shut down immediately, then it is optimal to continue production.

If U(t) is a McCabe-Robinson type of function, the average heat transfer coefficient $\overline{U}$ calculated from (2b) follows a curve similar to that depicted in Fig. 3. The values U(0) = 800, "a" = 0,02 and t_c = 16 gave a maximum at about t_p = 70 hours.

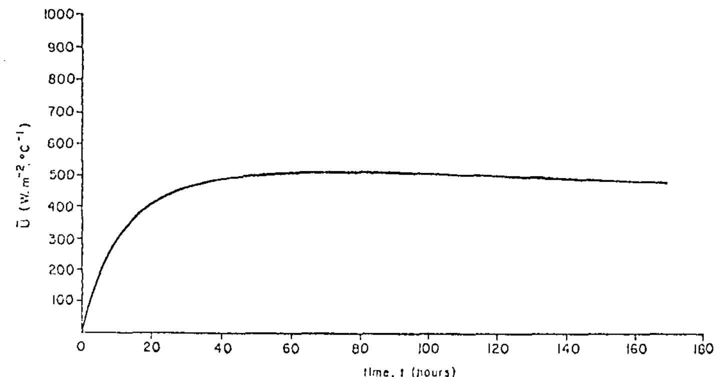

FIGURE 3 $\overline{U} = \left(\frac{1}{t_p + 16}\right) \int_0^{t_p} U(t)\, dt$ where $U(t) = 800\,(1+02t)^{-\frac{1}{2}}$

By substituting equation (1) into (2b) and performing the above mathematical calculations, we arrive at the following expression for equation (4):

$$\frac{(1 + at_p^*)^{-\frac{1}{2}}}{(t_p^* + t_c)} - \frac{2[(1 + at_p^*)^{\frac{1}{2}} - 1]}{(t_p^* + t_c)^2} = 0. \quad (5)$$

The solution of equation (5) is

$$t_p^* = t_c + 2\sqrt{\frac{t_c}{a}} \quad (6)$$

Using the above values for t_c and 'a' gives t_p^* = 72,57 hours with an average heat transfer coefficient equal to 63,87% of the value for clean tubes. If the plant were operated for 152 hours (i.e. a cycle time of one week) the average heat transfer coefficient would be only 60,12% of the value for clean tubes. The former policy thus gives a 6,24% higher total production.

The shape of the curve indicates that it would be preferable to operate too long rather than for too short a period. Also, the shut-down time need not be very close to the optimum: a deviation of up to one shift (8 hours) would probably be satisfactory. If the time trend in U(t) is approximately linear, of the form U(t) = U(0) (1 − bt) ("b" a constant), the optimum production time is given by:

$$t_p^* = \left(t_c^2 + \frac{2t_c}{b}\right)^{\frac{1}{2}} - t_c. \quad (7)$$

The variables influencing the evaporation process

The evaporator examined was of the calandria type. The heat transfer coefficient U depends on the geometry of the vessel and the physical properties of the syrup being evaporated. The latter can be related to the syrup concentration or brix, the syrup and steam temperatures, and possibly the liquid level. The resistance to heat transfer increases with time t as scale builds up on the surface of the tubes.

It is well known that an increase in the brix causes a decrease in the heat transfer coefficient. Saronin & Jenkins[2] (Fig. 4) found a concave relationship between B and U, while Cordiner[3] found an almost linear relationship (Fig. 5).

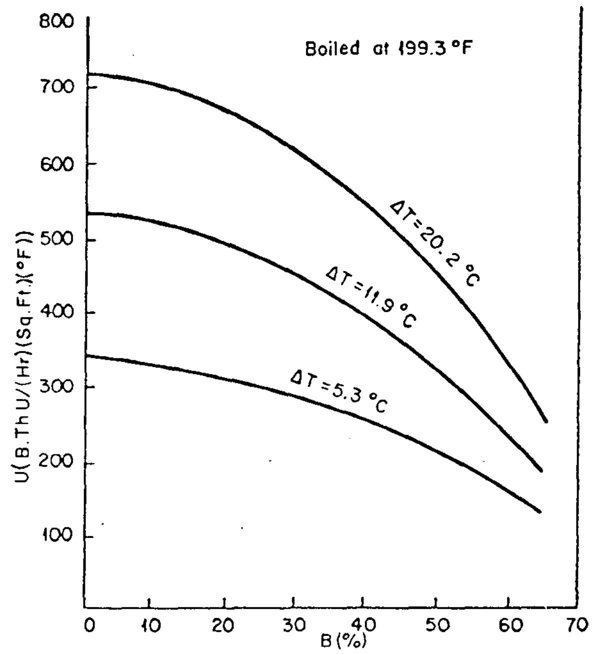

FIGURE 4 Data of Saranin and Jenkins: U vs B.

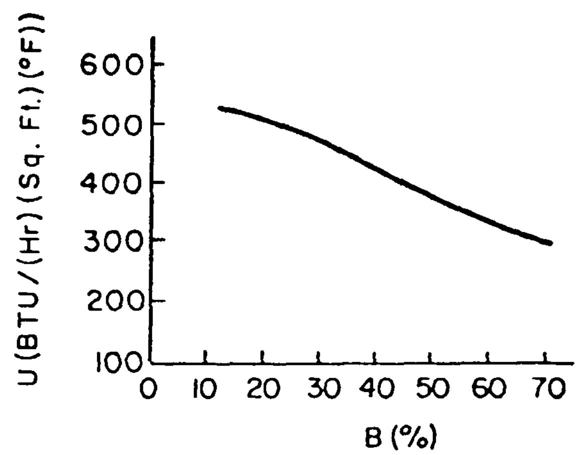

FIGURE 5 Data of Cordiner.

If dependence of U on B is linear, then a uniform disturbance of B about its mean will have no effect on $\overline{U}$. However, if the function is concave a uniform disturbance in B will result in a smaller $\overline{U}$ than that value corresponding to the mean B. In this case it would be desirable to control B at a steady value.

Reports differ as to the effect of liquid level L on the heat transfer coefficient. Some workers[4] have found that the optimum level for highly concentrated sugar solutions lies above 50% of the tube height but is not critical. If fluctuations in the liquid level influence the residence time of syrup in the vessel, this could possibly cause undesirable variations in the concentration.

The variables in an industrial evaporator interact to some extent so that the net effect of a change in a particular variable need not be the same as that found under controlled laboratory conditions. Indeed, the plant could shift to an entirely new set of operating conditions. For example, a step change in the feed brix alters the exit brix, which causes a change in U. This results in a different vapour rate and influences the pressure within the vessel. The syrup therefore boils at a different temperature, which alters ΔT. The change in the vapour rate could also cause a fluctuation in the level, necessitating an adjustment of the flow rate to correct it.

Many of the interactions are favourable because they tend to oppose changes in the operating conditions and lead to stable operation of the process. One of the aims of the present exercise was to establish where control was necessary for improving the average heat transfer coefficient.

Experimental results

In each run the computer sampled the process automatically every two minutes for about five days. Every five consecutive samples were averaged and the values, converted to the correct units, were recorded on magnetic and paper tape. Later the values of U were calculated from these data and plotted against time. In an attempt at linearization of the data, B, L, T and ΔT were plotted against $\frac{1}{U}$ which, according to the classical concept, is a linear function of each of the resistances to heat transfer. Least squares regression was used to find simple mathematical models for the data and statistical tests were applied to these to assess their significance. From the results of the tests, deductions were made about the influence of each variable on the overall heat transfer coefficient.

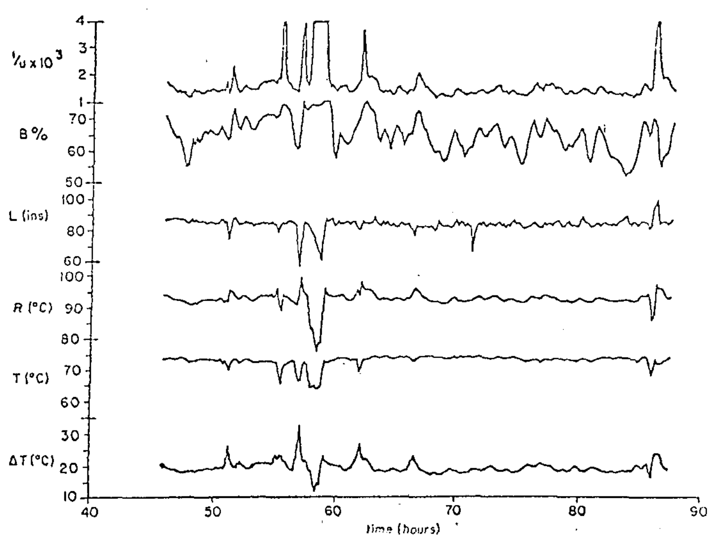

FIGURE 6 Portion of data from set 1.

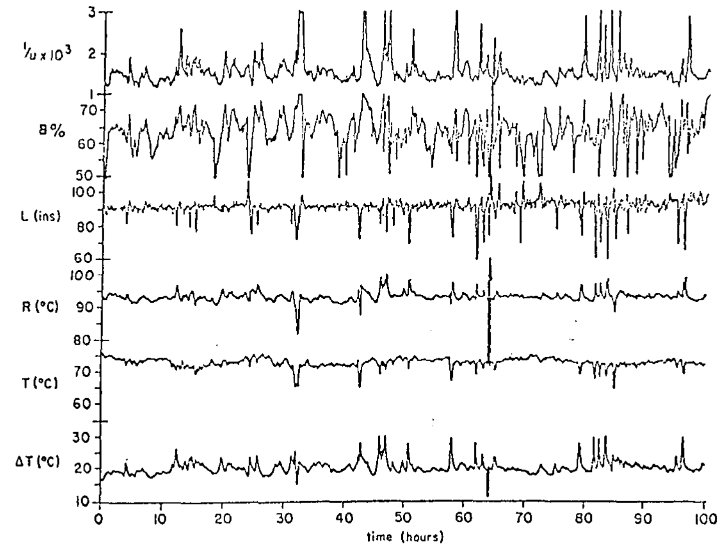

FIGURE 7 Data from set 2.

Figs. 6 and 7 show plots of the resistance to heat transfer 1/U, brix B, level L, steam temperature R, syrup temperature T and temperature difference across the tubes ΔT, against time, for runs 1 and 2. Some trouble was experienced with instruments, particularly those in contact with syrup. Flow and brix measurements became progressively more noisy, and the instruments were out of order for most of the runs (Figs. 6 and 7 were selected as the best records).

There is a clear correlation between B and 1/U and between ΔT and 1/U. However, 1/U and L seem to be correlated only when a major change occurs such as that at 58 hours for run 1, when boiling appears to have stopped temporarily.

Figs. 8 to 11 show plots of U against time. The outliers lie mostly below the mean heat transfer coefficient and correspond to peaks on the brix and ΔT curves in Figs. 6 and 7.

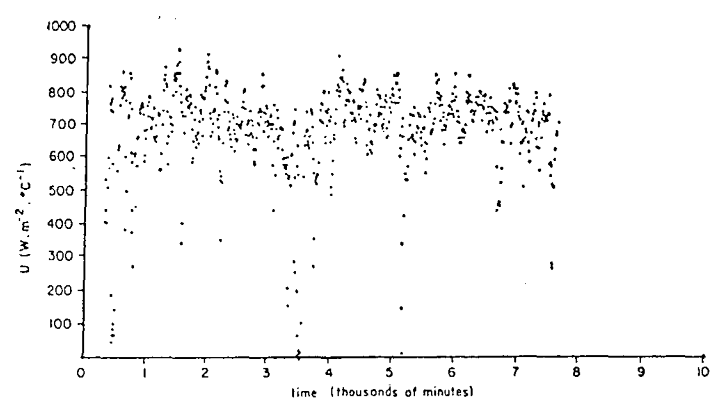

FIGURE 8 Heat transfer coefficient, U, versus time for data set 1.

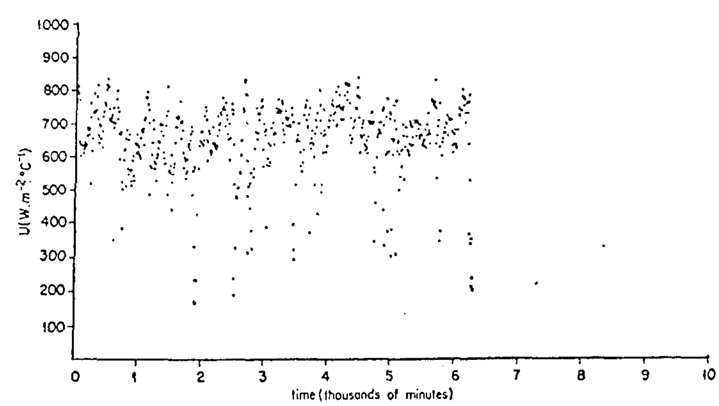

FIGURE 9 Heat transfer coefficient, U, versus time for data set 2.

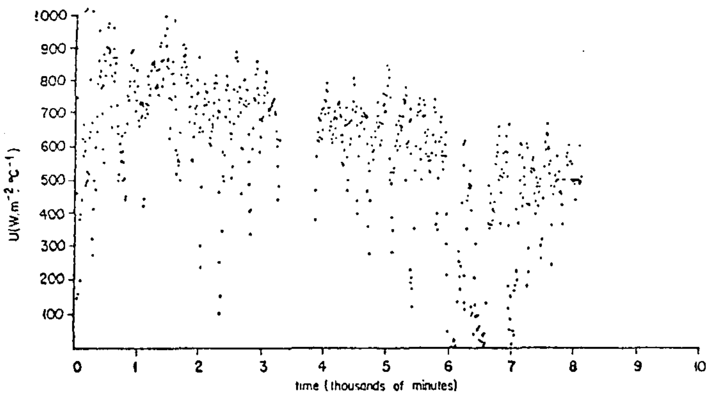

FIGURE 10 Heat transfer coefficient, U, versus time for data set 3.

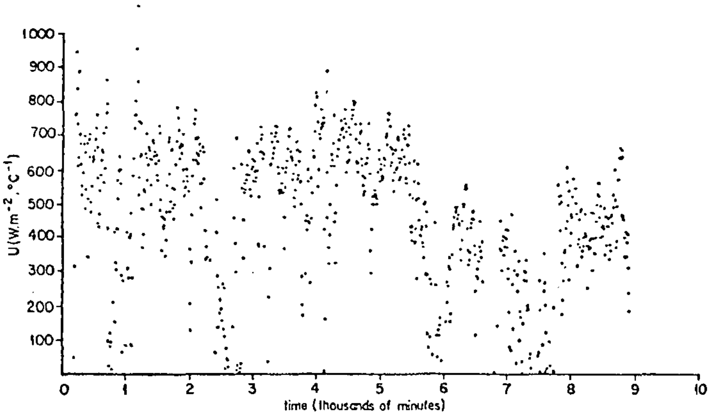

FIGURE 11 Heat transfer coefficient, U, versus time for data set 4.

Fig. 8 appears to have a slightly downward trend, particularly near the end of the run. Fig. 9 has no noticeable trend while Fig. 10 has a pronounced downward trend. Fig. 11 varied unpredictably and seemed to be tending upwards near the end.

Figures 12 to 15 show cumulative time average heat transfer coefficients for each run using the formula

$$\overline{U}_t = \frac{1}{(t + 16)} \sum_{i=1}^{n} U_i \, \delta t,$$

where U_i was the heat transfer coefficient of the ith record, and $\overline{U}_t$ was the cumulative average heat transfer coefficient calculated over the n records taken up to time t. The cleaning time was assumed to be 16 hours. The time between records, δt, was equal to one sixth of an hour. $\overline{U}$ for run 1 was still increasing at the end of the run after about 130 hours of operation. $\overline{U}$ for run 2 was increasing rapidly after 105 hours. $\overline{U}$ for run 3 contains a definite peak at about 100 hours, which would have been the optimal time to shut down. Run 4 has a peak at about 90 hours and could have been shut down then.

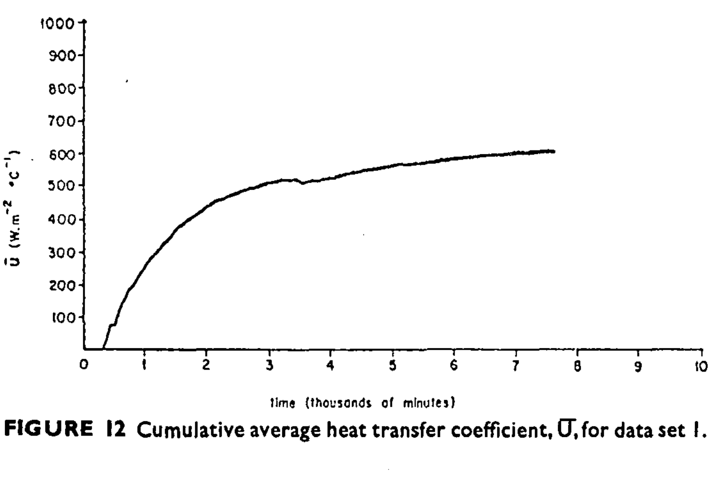

FIGURE 12 Cumulative average heat transfer coefficient, $\overline{U}$, for data set 1.

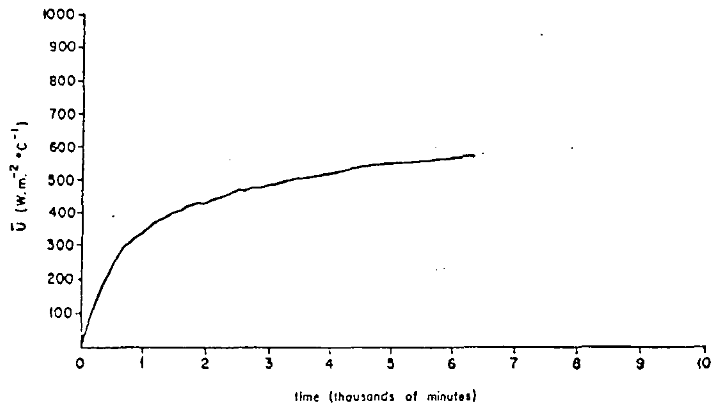

FIGURE 13 Cumulative average heat transfer coefficient, $\overline{U}$, for data set 2.

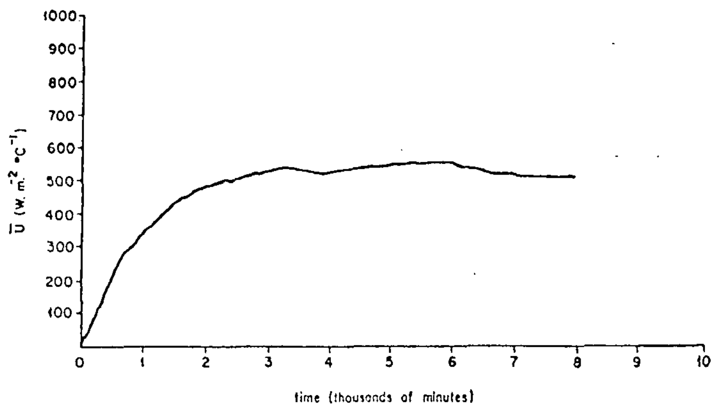

FIGURE 14 Cumulative average heat transfer coefficient, $\overline{U}$, for data set 3.

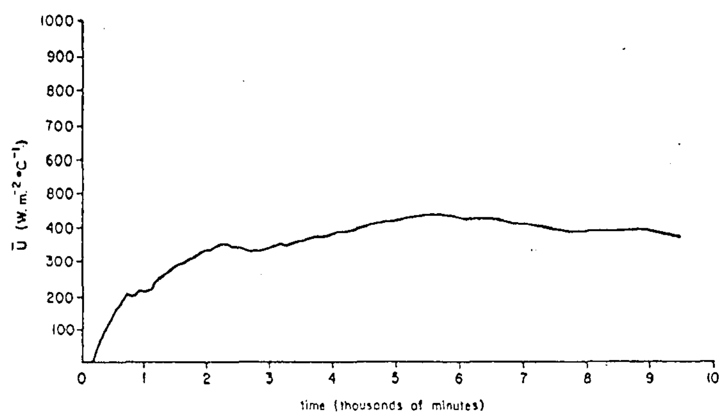

FIGURE 15 Cumulative average heat transfer coefficient, $\overline{U}$, for data set 4.

We were unsuccessful in fitting a McCabe-Robinson type of function to the data. However, the data were regressed against a quadratic function of time and the statistical significance tests applied to the parameters of the regression equations showed the following. Run 1 had no significant time trend and run 2 had a linear upward time trend. Runs 3 and 4 were concave from below with downward trends. Therefore none of the runs followed a McCabe-Robinson type of trajectory, which is convex from below. However, this does not disprove their theory since the true trend could have been obscured by the higher frequency variations in U.

The last 60 hours of run 1 were regressed against a linear function of time and the equation was used to predict the optimum shut-down time assuming that this trend continued. Accounting for the variance of the data and its effect upon the accuracy of the regressed parameters, the predicted optimum shut-down time was 208 ± 21,5 hours.

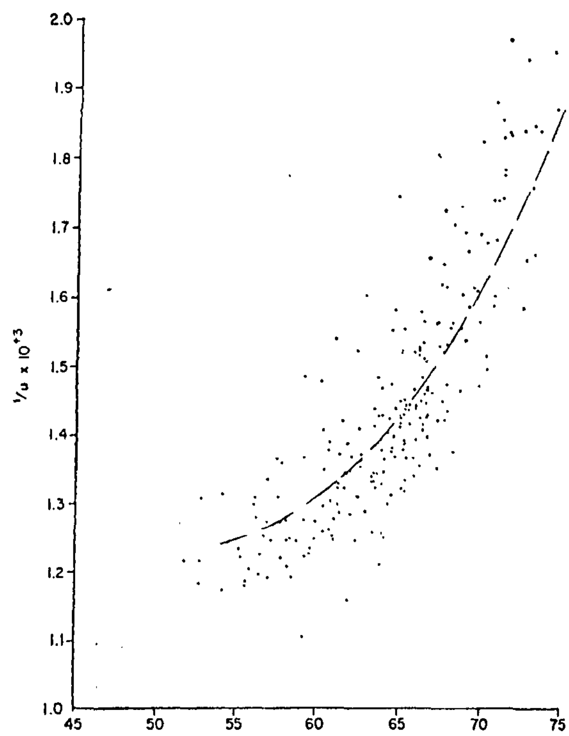

FIGURE 16 $1/_u$ versus B for data set 1.

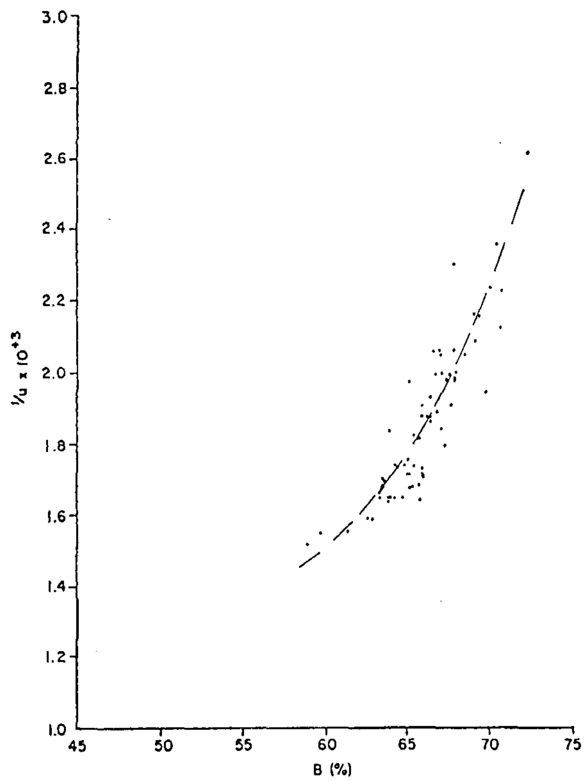

FIGURE 17 $1/_u$ versus B for data set 5.

Figs. 16 and 17 show the relationship between B and 1/U, the total resistance to heat transfer, for runs 1 and 5. (The brix values for run 2 were very noisy and the meter was out of order for runs 3 and 4.) Regression analysis accounted for about 59% of the variation of 1/U in run 1 and 72% in run 5. The dotted lines represent the function which best fitted the data in each graph. A plot of the inverse of these functions gave non-linear curves similar to those in Fig. 4.

Fig. 18 is typical of the plots obtained for 1/U against ΔT and shows that 1/U and ΔT increase together. Intuitively, the change in ΔT must be the reaction to a change in 1/U as described earlier.

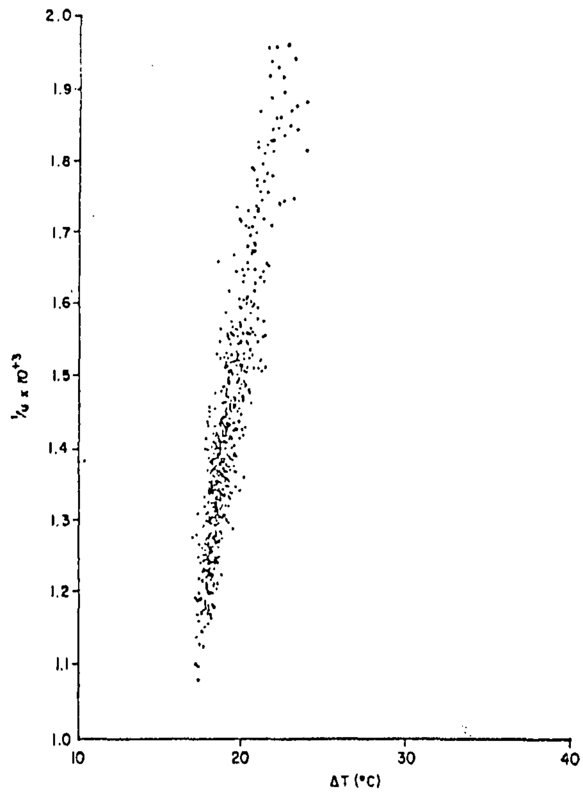

FIGURE 18. $1/_u$ versus ΔT for data set 1.

No correlation could be detected between 1/U and the level, nor between brix and level.

Conclusions and recommendations

The time trend in the heat transfer coefficient was not the same for each production cycle. Therefore determination of an optimum shut-down time under the conditions stated above would require continuous monitoring of the heat transfer coefficient (possibly by a data-logging minicomputer). The cumulative average heat transfer coefficient could be calculated and an optimum shut-down time predicted from a linear regression of the data.

This operating policy will not necessarily be optimal if there is some means of improving the evaporation rate without shutting down, e.g. increasing ΔT. If ΔT is not constant throughout the production cycle, it would be better to monitor U ΔT or E_4 and maximize $\overline{E}_4$.

When measurement error or disturbances in the process give rise to uncertainty in the value of U(t) or E_4, it should be remembered that operating too long is better than operating for too short a period.

There was a non-linear correlation between the concentration and the heat transfer coefficient over the range of variation encountered during each run. This indicates that the average heat transfer coefficient could be increased by tighter control of the concentration.

The wide variation in the heat transfer coefficient seemed to be due to changes in the concentration. These in turn could have been the result of fluctuations in the feed brix, but this cannot be verified as measurements were not taken. This should be done and control implemented if necessary.

There appeared to be no correlation between the brix and the liquid level, neither could a relation be detected between the level and the resistance to heat transfer. Therefore tighter control of the level would apparently hold no benefits.

Nomenclature

A = heat transfer area (m^2)
B = juice concentration (°Brix)
E_i = vapour rate from ith effect (t/h)
H = latent heat of evaporation (J/t)
L = liquid level (% tube height)
P = production rate (t/h)
t = time (h)
t_c = cleaning time per cycle (h)
t_p = production time per cycle (h)
T = temperature (°C)
ΔT = mean temperature drop across tubes (°C)
U = heat transfer coefficient ($W.m^{-2}\,°C^{-1}$).

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