HEAT TRANSFER, MASS TRANSFER AND SCALING CHARACTERISTICS IN A LONG TUBE, CLIMBING FILM, PILOT PLANT EVAPORATOR.

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Abstract

A single tube pilot plant Kestner evaporator was set up at Illovo to investigate heat transfer characteristics when simulating different evaporator effects. Tests carried out have shown that the Kestner evaporator, which is used almost exclusively as a 1st effect vessel in the South African sugarcane industry, operates most favourably at the tail of the evaporator system when compared to the conventional Roberts-type evaporator currently in use. A seven week scaling test was carried out with the pilot plant operating in parallel with the 3rd effect Roberts vessel at Illovo. It appears that the rate of scaling in the Kestner is substantially reduced due to the increased liquid and vapour velocities through the tube and the resulting decreased juice residence time.

Introduction

The Climbing-film evaporator was patented in 1899 by Paul Kestner. Long tube vertical climbing film evaporators have been employed in various industries (notably in fruit juice concentration and the European sugar beet industry) for many years. In the South African sugarcane industry this type of evaporator has been used on 1st effect only, subsequent effects usually consisting of the conventional Roberts-type evaporator which, although having several shortcomings in design, has remained popular due to its simplicity of construction and cheapness. Since the juice residence time is minimal, Kestners reduce the problems of inversion and colour formation at high temperatures. The 7m long tubes make it possible to achieve high evaporation loads on single pass operation and the high tube velocities result in high heat transfer coefficients. Easy accessibility provides simple maintenance and cleaning and if the evaporator is operating satisfactorily scale formation is slower than in conventional downflow evaporators. If the Kestner is to be provided with an independent separator more space will be taken up in the horizontal plane. As a result of the short residence time, the Kestner cannot be employed to store juice, as is often done with conventional evaporators to smooth out fluctuations in juice flow between the clarification and pan stages of the sugar manufacturing process. The Kestner requires boiling feed, i.e. boiling at the operating pressure or a supplementary juice heater and the instrumentation required for accurate control is more complex than for conventional evaporators.

Allan[1] emphasizes the advantages of using Kestners at the evaporator tail although some doubts have been expressed as to the validity of using Kestners on any effect except the first as it is believed that the increasing viscosity of the juice towards the evaporator tail will hamper the climbing ability of the film and hence prevent efficient operation. It is worth noting that there is a limit to the maximum viscosity and manufacturers[2] emphasize this when quoting Kestner specifications. This 'maximum', about 100cP, is unlikely to be reached in a sugar mill operating normally where the final syrup viscosity is typically 75cP.

The initial purpose of the pilot plant was to determine whether the Kestner can be successfully operated in the 3rd, 4th and 5th effects of an evaporator 'quin'. A simulation test was performed in parallel with the 3rd effect (of a 'quad') at Illovo so that scaling characteristics could be investigated. Finally the pilot plant was run under 1st effect conditions in order that a comparison could be made with industrial 1st effect evaporators (both Roberts and Kestner). This particular trial would also confirm or reject the validity of previous work carried out i.e., whether or not the pilot plant was a reasonable model of factory conditions.

The aims of the project were as follows:-

(i) The optimisation of heat transfer coefficient, juice flow rate and brix change across the evaporator and a study of the interdependence of these factors.

(ii) A comparison of heat transfer coefficients for Kestners and conventional evaporators and, in the case of 1st effect, a three way comparison between pilot plant, industrial Kestner and industrial Roberts vessels.

(iii) An investigation of scaling characteristics of the pilot plant compared to conventional evaporators and the effect of scale on heat transfer coefficient.

(iv) An investigation of how effectively the pilot plant can be controlled and the problems encountered by sudden fluctuations in feed, pressure and heating vapour supply for various effects.

Experimental Details

A schematic diagram of the pilot plant installed at Illovo is shown in Fig. 1.

Syrup and hot water were mixed to the required feed brix in the steam heated, stirred, mixing tank. The feed was maintained at about 10°C below the required 'boiling' temperature in the thermostatically heated feed tank and was pumped using a variable speed mono-pump via juice heater to the column. Flashing was induced by boosting the juice temperature to about 5°C above the required boiling temperature in the juice heater and also by including a small orifice in the tube feed inlet pipe. The tube was 7m long, with a 50,8 mm OD, and was constructed from stainless steel and surrounded by a 101,6 mm OD, insulated, heating chamber. The vapour-liquid mixture leaving the column was separated in an empty, centrifugal-inlet chamber, the vapour being taken off under vacuum, condensed and collected in a tank. The product juice and condensed shellside vapour were also collected in tanks. (see Fig. 1). Vacuum fluctuations were minimised using two independent product tanks, which helped to ensure steady-state conditions during the tests. Feed and product were sampled for brix determination. A factory vacuum pump was used to provide the tubeside vacuum with a direct link to the top of column 'shellside' to draw off incondensible gases. The column shell-side vacuum was provided by the conditions of the heating vapour.

Temperatures were measured using a 12 point recorder/thermocouple system. This proved useful in maintaining steadystate conditions as a continuous record of pilot plant temperatures was kept. (Generally a sharp temperature change indicated a deviation from steady-state operation). The tubeside vacuum was controlled using a pneumatic control valve linked to the cooling water outlet of the tubeside vapour condenser. The feed juice temperature was controlled using a thermostat/solenoid

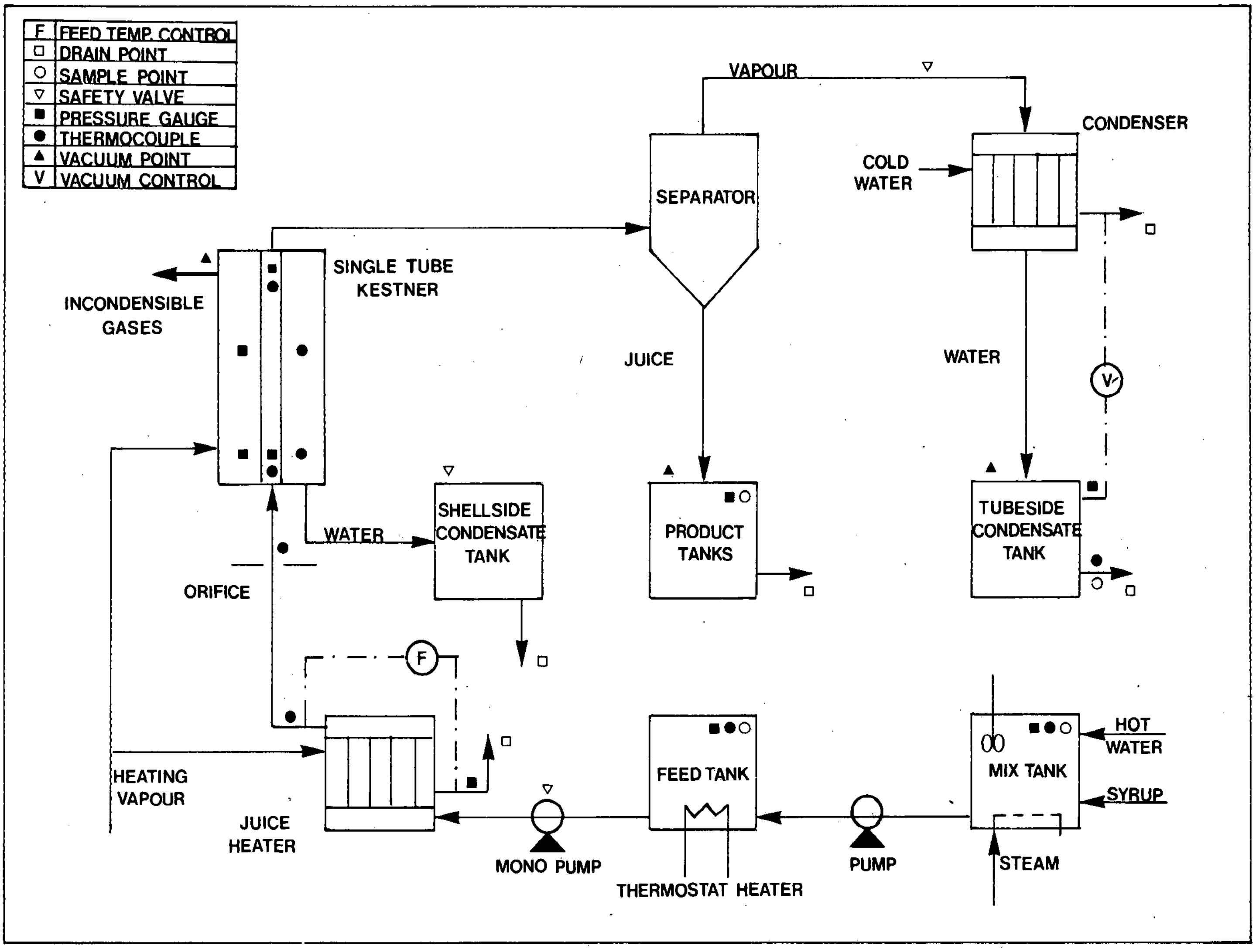

Figure 1: Schematic flow diagram of pilot plant Kestner.

valve system linked to the shellside outlet of the juice heater. Although these controllers rarely operated on a fully automatic basis they were of great assistance, when used semi-automatically, in smoothing out the fluctations in temperature and pressure. Some typical operating conditions are listed in Table 1.

Theoretical Background

A statistical approach made initially revealed that it was necessary to perform at least 8 runs at a minimum of 5 different feed rates per effect for the evaluation of any empirical equations for the investigated variables. The length of each run was also critical but by experiment it was found that a minimum run time of 15 minutes at steady state conditions was essential to procure reproducible results. The experimental error is estimated at $\pm 20\%$ maximum. The calculations were solved on the computer at the SMRI.

The basis of the calculation of heat transfer coefficients was the equation

$$Q = U{*}A{*}\blacktriangle T$$

where Q is the heat transferred per unit time (W)
U is the overall heat transfer coefficient ($W\ m^{-2}\ {}^{\circ}C^{-1}$)
A is the heat transfer surface, a constant (1,12 m^2)
$\blacktriangle T$ is the temperature difference between the 2 streams (°C)

The heat transferred per unit time, Q, was evaluated in 3 ways

(i) From the EVAPORATION RATE as determined from the difference between the measured feed and product rates.
(ii) From the evaporation rate as determined from the quantity of tubeside vapour condensed. (The vapour from the separator). This is hereafter referred to as the TUBESIDE CONDENSATION RATE and theoretically equals (i).

TABLE 1

Average Operating Conditions for Simulation Tests

Effect	Feed Temp After heater	Tubeside Temp (column base)	Tubeside Temp (column top)	Tubeside Pressure Column Base	Shellside Pressure	Shellside Temp	Vapour
	(°C)	(°C)	(°C)	(kPaAbs.)	(kPaAbs.)	(°C)	
1	116	113	109	160	195	119	Exhaust
3	103	99	96	80	110	105	2nd
4	93	86	83	50	80	95	3rd
5	80	72	69	15	50	83	4th

(iii) From the SHELLSIDE CONDENSATION RATE, this was the amount of heating vapour condensed on the outside of the tube. Theoretically this should also equal (i) and (ii).

Some adjustments were however required for (i) and (iii). In the case of (i) it was required to compensate for the evaporation caused by the flashing of the feed. This correction was approximately 1% of the total evaporation rate, the exact value depending on the precise operating conditions. In the case of (iii) it was necessary to adjust the shellside condensation rate for the effects of heat losses and water in the incoming vapour. These two factors were evaluated simultaneously by experiment and the resulting correction was typically 10-12% of the shellside condensation rate for 3rd, 4th and 5th effects, with good reproducibility in each case. For 1st effect however problems occurred due to the variable amount of water present in the incoming exhaust steam and this has certainly affected the results.

The corrected rates (i), (ii) and (iii) were converted to the required units for Q by multiplying by the relevant heat capacity figures. U was evaluated for each Q value and since good agreement was shown an average U was assumed.

Some difficulty also occurred in the evaluation of ▲T, the temperature difference. Firstly, there was a pressure drop over the tube length due to the hydrostatic head of the liquid phase (Typically 5-8 kPa caused by a depth of 0,5 m of juice at the base of the tube) and also losses due to acceleration of the vapour phase. This pressure drop resulted in a lower temperature at the top of the column than at the base. The problem was solved by selecting the lowest temperature (i.e. the product temperature) as the effective juice side temperature. Thus ▲T will be maximum (i.e. heating vapour temperature minus juice side temperature) and U will be minimum. Secondly there was the effect of boiling point elevation due to the change in concentration of the juice. This problem was overcome by ignoring the difference in boiling point elevation from feed (column base) to product (column top) — thus ▲T will again be maximum and U minimum. Typical values concerning this difference in boiling point elevation for each effect are given in Table 2. The effect of increasing juice viscosity up to tube was small enough to be neglected for all effects.

TABLE 2

The Difference in Boiling Point Elevation from Feed to Product (under ideal operating conditons)

Effect	Difference in BPE from feed to product (°C)
1	0,1
3	0,8
4	1,2
5	1,8

Results

1. *Simulation of different effects*

The results for the simulation of the various effects are presented in Figs 2 and 3. From Fig. 2 the plot of heat transfer coefficient against feed rate, it appears that heat transfer coefficient is virtually independent of feed rate except for 1st effect. The ends of the straight lines in Fig. 2 correspond to the minimum and maximum flow rates. Below the minimum feed rate crystallisation occured in some instances and no liquid product was produced. Above the maximum feed rate a negligible brix change occurred. It is meaningless to evaluate heat transfer coefficients outside these limits. For 1st effect conditions heat transfer coefficient increased linearly with feed rate, however the variable amount of water in the exhaust steam used for heating may have distorted these results.

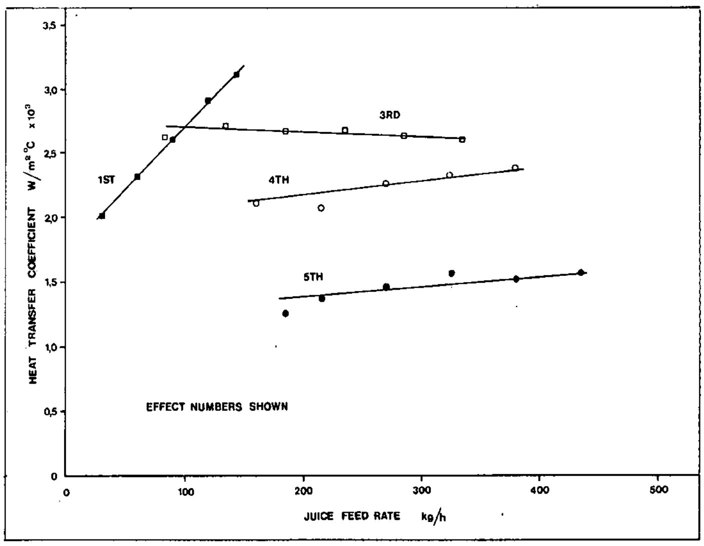

Figure 2: Relationship between the heat transfer coefficient and juice feed rate for different evaporator effects.

The plot of brix change against feed rate, Fig. 3, is as expected, that is brix change is approximately inversely proportional to feed rate. The curve obtained for 3rd effect would have been produced for all effects if larger feed rates had been used since the brix change will reach zero at a particular feed rate and not change significantly irrespective of any further feed rate increase. Since heat transfer coefficient is independent of feed rate for all effects except the first it is apparent that brix change is the most important consideration when sizing this type of evaporator. Table 3 shows a comparison between heat transfer coefficients for the Kestner pilot plant and industrial Roberts evaporators. The values for the pilot plant are taken from Figs. 2 and 3 and are chosen so that the brix change produced matches

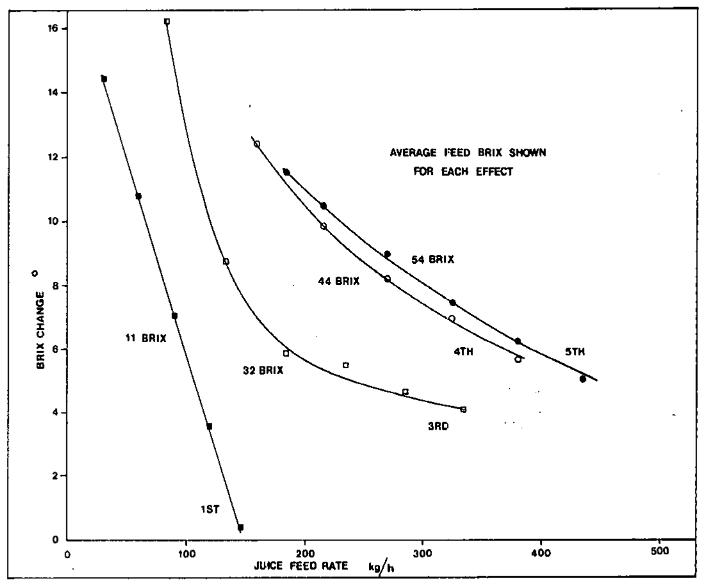

Figure 3: Relationship between brix change and juice feed rate for different evaporator effects.

that for the corresponding Roberts vessel. From Table 3 it is immediately apparent that the Kestner is most favourable, compared to a Roberts vessel, when utilised at the tail of the evaporator system. This supports the suggestions of Allan[1] previously outlined in the introductory section. It would have been highly desirable for the experiments to include work on a Roberts pilot plant so that the aforementioned phenomenon could have been given the further investigation that it deserves. Unfortunately, in the time available this was not possible. Even more interesting from Table 3 are the 1st effect results: the pilot plant value for heat transfer coefficient (2600 W m^{-2} $°C^{-1}$) shows close agreement with an industrial Roberts vessel (2400 W m^{-2} $°C^{-1}$ at Jaagbaan) and also with industrial 1st effect Kestners (2400 W m^{-2} $°C^{-1}$ at Umfolozi and Felixton and 2350 W m^{-2} $°C^{-1}$ at Illovo). These 1st effect results have confirmed and emphasised the pilot plant's ability to duplicate industrial conditions.

TABLE 3

Comparison of Heat Transfer Coefficients for Kestner Pilot Plant and Industrial Roberts Vessels

Effect	Typical Roberts Industrial Heat Transfer Coef. (Wm^{-2} $°C^{-1}$)	Measured Kestner Pilot Plant Heat Transfer Coef. (Wm^{-2} $°C^{-1}$)	Percentage increase on Roberts vessel
1	2 400	2 600	8
3	1 700	2 600	53
4	1 400	2 200	57
5	700	1 350	93

Throughout the simulation experiments it was observed that control and operation of the pilot plant was far easier towards the evaporator tail. In addition the recovery of the unit to steady state conditions after a forced stoppage, such as lack of heating vapour, was always faster under the 4th and 5th effect conditions, (typically 5 minutes). Under 1st effect conditions re-establishment of steady state conditions often took 30-60 minutes.

Figure 4, a plot of specific volume of air against absolute pressure illustrates one of the main reasons why the Kestner produces high heat transfer coefficients at the evaporator tail. Under 5th effect conditions the specific volume of air is typically 11 m^3 kg^{-1}. Since the tube cross-sectional area is fixed (0,002027 m^2) and the evaporation rate is approximately 40 kg h^{-1} the vapour velocity (V) will be:-

$$V = \frac{40}{60.60} \bullet \frac{11}{0,002027} = 60 \text{ m sec}^{-1}$$

Since heat transfer is favoured by high velocity the effect of this high value will be appreciated. The corresponding velocities for other effects are approximately as follows:-

1st effect 7 m sec^{-1}; 3rd effect 15 m sec^{-1} and 4th effect 25 m sec^{-1}. Since the value for 5th effect is 8,5 times that for 1st effect it seems surprising that Kestners are not used more frequently as final effect vessels.

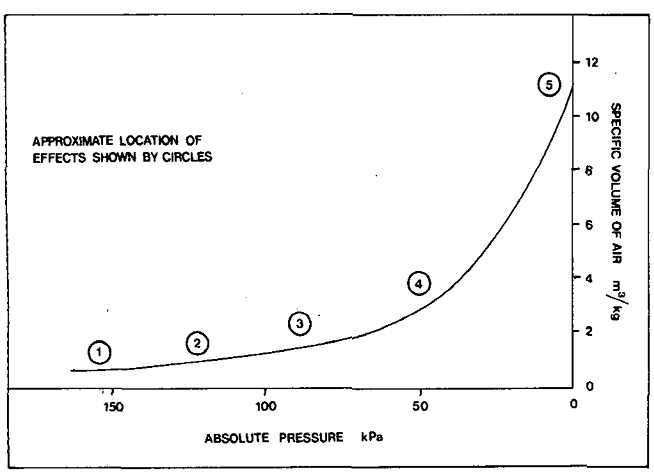

Figure 4: Plot of specific volume of air against absolute pressure.

Specific evaporation (kg m^{-2}) is a useful and popular concept when assessing evaporator performance. A typical figure for a 3rd, 4th and 5th effect Roberts vessel is 20-25 kg m^{-2}. For the pilot plant where the evaporation rate is in the range 30-40 kg h^{-1} and the tube surface area is 1,12m^2 the specific evaporation is 25-35 kg m^{-2} which is a worthwhile improvement on the Roberts figure. For 1st effect there is a negligible difference in specific evaporation values for pilot Kestner and industrial Roberts both being in the range 35-40 kg m^{-2}.

One of the problems in the simulation of conditions near the evaporator tail was the fact that a pump was used to provide the column with feed. Practically it was impossible to operate without a pump as each effect was investigated individually. The exact consequence of this pumping is not known but there is little reason to suspect that it substantially impairs or improves the operation of the plant.

Experimentally the height of liquid in the column was about 0,5 m under ideal conditions. This height rose to 1 m at the maximum practicable feed rate and fell to 0,2 m at the minimum feed rate. If the feed was not 'boiling' at the system pressure this height increased to 2 m or more, the area of tube surface available for heat transfer was thus reduced and as a result evaporator performance suffered drastically. This rise in level results from the liquid trying to reach its boiling temperature, which it does by filling space previously occupied by the climbing film and therefore is able to gain more heat and eventually 'boil'.

TABLE 4

Average Conditions during Scaling Tests

Conditions	
Effect	3rd of 4
Total duration	984 hours
Running Time	728 hours
Planned stops	210 hours
Stops due to pilot plant problems	46 hours
Feed rate	300 kg hr^{-1}
Feed concentration	27 °Bx
Product concentration	36 °Bx
Temp. after juice heater	98 °C
Temp. of tubeside at column base	91 °C
Temp. of tubeside at column top	88 °C
Tubeside pressure	50 - 65 kPa (Abs)
Shellside vapour	2nd
Shellside temperature	100°C
Shellside pressure	95 - 105 kPa (Abs.)

2. The scaling test

This test was performed under 3rd effect (of a 'quad') conditions. The pilot plant was run in parallel with the industrial 3rd effect Roberts vessel at Illovo. The conditions in the 2 evaporators were as similar as possible, i.e. the same feed juice and heating vapour were used at the same temperatures and pressures. The average pilot plant conditions are listed in Table 4, the feed rate being chosen to give an industrially comparable brix change. At the end of the test the tube was cleaned 3 times dry using a standard tube cleaner (as is used for the 1st effect Kestner at Illovo). Twenty grams of scale were collected and analysed by the laboratory at SMRI. The results of the analysis are presented in Table 5, the proportions of components in the scale being typical of an industrial 3rd effect evaporator. A graphical representation of the effect of time on heat transfer coefficient is given in Fig. 5 . There is a slow but steady fall in heat transfer coefficient but the final figure attained (2100 W m^{-2} $°C^{-1}$) is still equal to a typical design value for a 'clean' industrial 3rd effect Roberts evaporator. The drop in heat transfer coefficient over the test period was about 20% (2600 → 2100 W $m^{-2} \cdot {}^{\circ}C^{-1}$). From the results it appears that the Kestner pilot plant could operate for 6 weeks or longer without being cleaned – this compares most favourably with the current cleaning schedule of the conventional 3rd effect vessel at Illovo which is cleaned every week.

TABLE 5

Analysis of Scale from Pilot Plant Kestner

Analysis	Percent
Loss at 650°C	64,3
SiO_2	3,5
Al_2O_3, Fe_2O_3	2,6
CaO	28,0
Sulphate	Nil
Undetermined	1,6

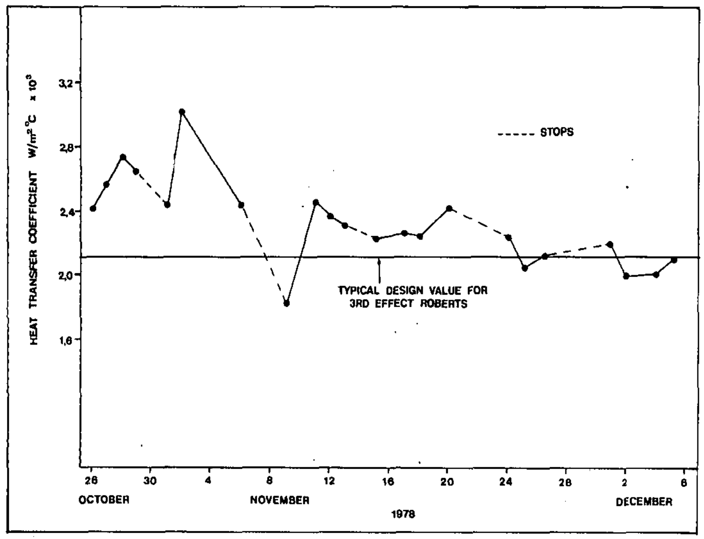

Figure 5: The effect of time on heat transfer coefficient under 3rd effect conditions.

Conclusions

From the results produced one can conclude that the Kestner pilot plant is a reasonable model of industrial conditions.

There is little advantage in using Kestner evaporators as 1st effect units since the heat transfer coefficients for pilot plant Kestner, industrial Kestner and industrial Roberts vessels are virtually identical. Investigation of other effects indicates that the pilot plant produces higher heat transfer coefficients than an industrial Roberts vessel towards the tail of the evaporator system and under final effect conditions the pilot plant value is twice that for its industrial Roberts counterpart. Control and efficient operation of the pilot plant are easiest at the evaporator tail.

From the graphical analysis of the pilot plant experiments one can conclude that (i) heat transfer coefficient is independent of feed rate except for 1st effect conditions where the amount of water in the heating vapour may have distorted the results, and (ii) brix change across the pilot plant is approximately inversely proportional to feed rate.

Scaling characteristics of the pilot Kestner are superior to an industrial Roberts vessel operating under similar 3rd effect conditions. The cleaning schedule of industrial evaporators could be reduced considerably by using Kestner evaporators.

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