

# PILOT PLANT INVESTIGATIONS INTO A BALCKE DÜRR FALLING FILM PLATE EVAPORATOR AT GLEDHOW SUGAR FACTORY

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## Abstract

Compact plate evaporator designs offer high rates of heat transfer, with a number of advantages which have been well researched in the beet industry. However, such designs are not well established in the cane sugar industry, mainly because of the fears of high rates of fouling and consequent blockages. During the 1996-97 season, a Balcke Dürr falling film plate evaporator pilot plant was installed at the Gledhow sugar factory. Tests were carried out with the objectives of assessing the performance of the unit under factory conditions, and identifying potential problems of operating such a unit in the cane sugar industry. Some results obtained which are of particular interest to the South African sugar industry are considered.

*Keywords:* evaporator, falling film plate, Gledhow

## Introduction

For more than a century evaporation technology was dominated by variations of the short tube (Robert) evaporator. The need for large units and more compact design lead to the introduction of the long tube rising (Kestner) and falling film evaporators. Plate evaporators, which can take the form of rising or falling film designs, were introduced to the sugar industry in 1987 (Licha *et al.*, 1989; Punter and Christopherson, 1992a). Plate evaporators have been studied extensively in the beet industry and many of the benefits achieved may be transferrable to the cane industry. The claimed advantages of plate evaporators are as follows (Austmeyer *et al.*, 1996; Licha *et al.*, 1989, 1994; Morgenroth *et al.*, 1995; Morgenroth, 1996; Morgenroth *et al.*, 1996a, b):

- Shorter juice residence times leading to lower colour formation and reduced sugar degradation.
- Designs that offer higher heat transfer coefficients (HTCs) resulting in greater rates of evaporation for the same installed surface area.
- A compact construction that uses less material and may thus result in lower capital cost.

Of all designs available, the falling film plate evaporator offers the largest heating surface per single unit (Austmeyer *et al.*, 1996; Licha *et al.*, 1989, 1994; Morgenroth *et al.*, 1995; Morgenroth, 1996; Morgenroth *et al.*, 1996a, b).

The falling film evaporator principle makes it possible to operate at lower temperature differences than for the equivalent rising film unit. This is an advantage when dealing with viscous solutions at high brix and has to be weighed against the disadvantage of the costs of additional pumping capacity.

The need for expansion of evaporator station in South Africa has generated interest in technologies that can offer high heat transfer coefficients. To date in the southern African industry only two mills have installed plate evaporators, Glendale (GD) and Ubombo Ranches (UR), and these have been of the rising film type (Alfa Laval). The installation at GD has been considered previously (Walthew and Whitelaw, 1996), and some fouling and cleaning problems have been identified in the UR installation (Walthew *et al.*, 1997). Prior to the tests reported here, most work on the falling film plate evaporator (FFPE) has been carried out using beet juice, and there is a fear that the compact evaporator designs may result in higher rates of fouling and difficult cleaning, with consequently longer down times negating the benefits listed above. To examine some of the possible advantages, problems and general suitability of the FFPE for cane, a pilot plant was set up at Gledhow (GH) during the 1996 season. Some results and observations from this installation form the subject of this paper.

## General description of the Balcke Dürr falling film plate evaporator

The main characteristic of the (FFPE) in comparison with other units is its compact design (i.e. large surface area in a single unit), and the relatively low wetting rates required. The wetting rate indicates the degree of liquid coverage per surface area. For the Balcke Dürr evaporator the wetting rate (0.6-1.0 L/cm of circumference/h) has been found to be approximately 10-20 times less than that for the tube-bundle falling film evaporators (typically 8-16 L/cm of circumference/h). This indicates that high recirculating liquid loads are not required to prevent 'dryout' of the heating surface.

The falling film plate evaporator can be regarded as a combination of plate and tube-bundle falling film evaporators. The plate pack is completely welded and is constructed to form an elliptical, vertical 'pseudo tube' on the juice side, as shown in Figure 1. The steam flows across the surface of the plate pack (cross flow principle). The 'tubes' are joined horizontally at

intervals to aid in juice distribution, as shown in Figure 2. The tube diameters can vary from 5 to 11 mm. The geometry of the corrugated steam-side combined with the crossflow principle produces a uniform flow that results in a smooth flow profile and is claimed to leave no dead spaces. The plate modules are 32 cm high and can be stacked one above the other to achieve the required surface area.

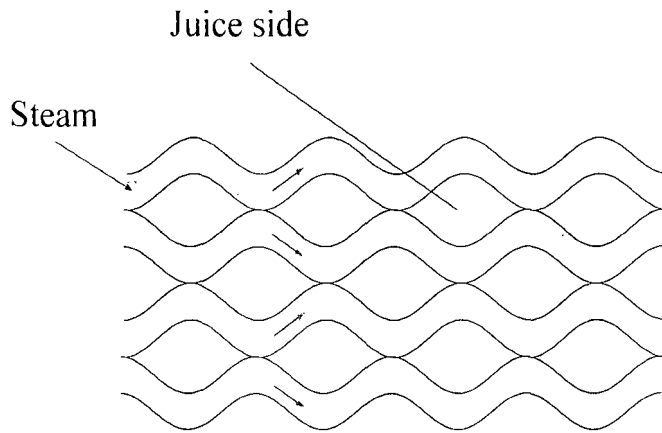


Figure 1. Cross section of the plate showing the pseudo tube formed by the plates and the corrugated side that carries the steam.

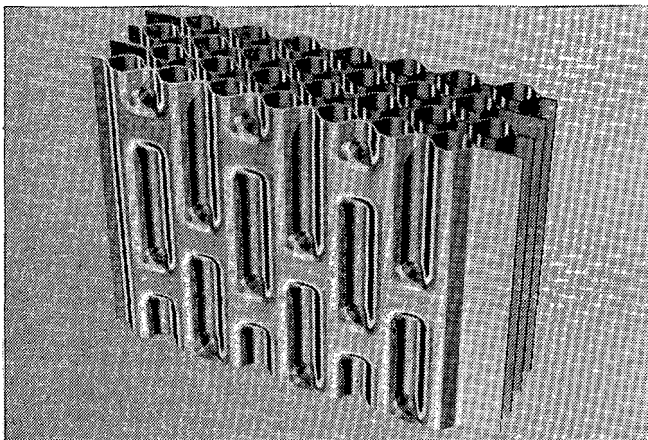


Figure 2. Three dimensional view showing the interconnection of the pseudo tubes formed.

The juice is typically distributed by means of simple trays in a diagonal cross pattern, followed by multiple perforated plates. The plate pack design also serves to distribute juice and, where several packs are located above one another, the packs themselves act as distributors. The low flowrate to heating surface ratio means that usually only a single pass is used, unlike the tube-bundle type, where a large amount of liquid is recirculated.

Plate packs can be installed in purpose made vessels or retrofitted inside an existing Robert vessel after the tubes have been removed. In one instance the individual modules have been fitted into the evaporator as removable cupboard drawer-like segments (Licha *et al.*, 1994). While offering the advantage of removable/replaceable heating surfaces, this configuration adds significantly to the cost of the installation.

The design of the plate evaporators in general and the FFPE in particular make certain other demands on the factory:

- Because of the small gap on the juice-side it is necessary for the feed juice to be subjected to screening to remove coarse impurities (>2-3 mm). This is a requirement of all plate evaporators as well as for the distribution systems of tube-bundle falling film evaporators.
- Unlike the Robert-type evaporator, which is easy to control via the juice level and can tolerate juice flow fluctuations relatively well, plate evaporators have a significantly smaller buffer volume and react more sensitively to fluctuations in the operating parameters. This problem is not new to the industry, having been identified in the operation of Kestner evaporators (Walthew and Whitelaw, 1996) and is overcome by careful control of the juice flow.
- The nature of the unit's construction is such that mechanical cleaning is not possible and the plates must be cleaned chemically. This has the potential to increase costs and down time (Walthew *et al.*, 1997). The design is such that a build-up of scale over time can be detrimental to performance. During cleaning the entire plate pack can be immersed if necessary to ensure thorough cleaning or to deal with blockages.

### Pilot plant description

The pilot evaporator with a heating surface of 7,2 m<sup>2</sup> was installed at the GH factory for the 1996-97 season. A single plate pack was used as a heating chamber. This allowed good access to the surface for visual inspection of the scaling and cleaning effects. Figure 3 shows a simplified diagram of the installation. Juice from any effect from the main plant can be used as feed. Exhaust steam is used for heating, suitably throttled as required, and vapour one for the preheating of the juice (using a small plate heat exchanger).

The vapour outlet was connected to the main plant condenser line and the product syrup was discharged to the fifth effect vessel. Juice flow, temperatures and pressures were measured on-line and the data logged by computer. The pressures and the juice level in the entrainment separator could be controlled, and the plant could run for extended periods (overnight) unattended. An on-line refractometer (IPR II, Schmidt & Haensch) was used to monitor the outlet brix.

### Operation, measurements and calculations

The primary aim of this work was to examine the performance of the FFPE under normal plant conditions. The plant was operated with clear juice (to simulate first effect) and first and fourth effect juice (to simulate second and fifth effect operations, respectively). Trials of 100 to 300 hours were carried out to determine the fouling rate. The plant was operated with a juice flow of between 1 400 and 2 500 L/h - about 20 to 35 t juice/h/[100 m<sup>2</sup>] or a juice wetting rate of between 0,7 and 1,2 L/h/(cm

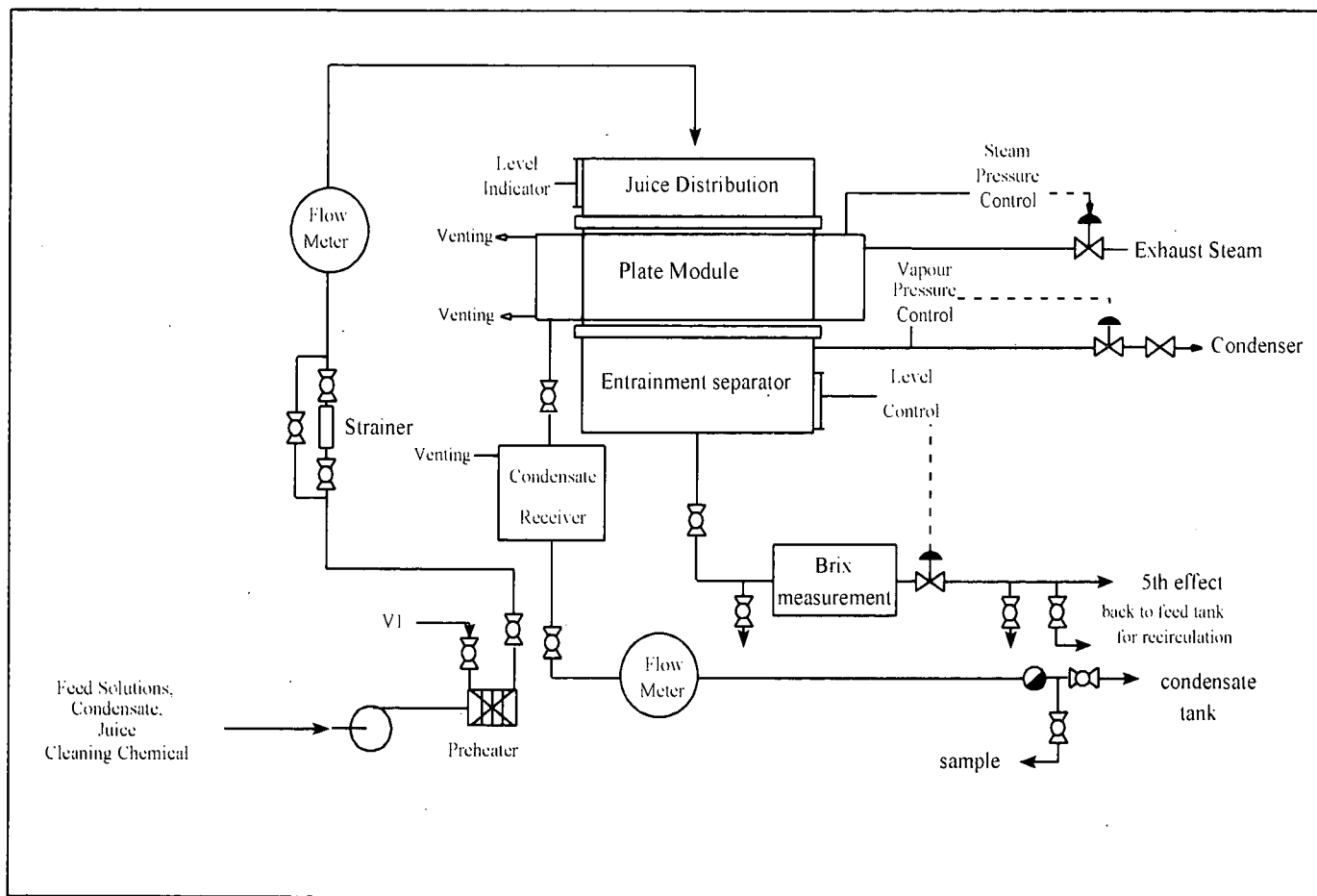


Figure 3. Simplified equipment layout showing the main features of the pilot plant installation.

of circumference). (The flowrate was higher because a single plate pack was used. It is normally 5 to 8 ton/h/[100m<sup>2</sup>]). The conditions of operation are summarised in Table 1.

Performance was measured through calculation of the heat transfer coefficient (HTC). This was calculated using both the brix balance and condensate flow methods, and details of the calculation have been presented previously (Walthew and Whitlaw, 1996). A brief description of the method of calculation is given in Appendix 1. The effective temperature difference (DT) used was determined from the measured pressures of the steam and the temperature of the outlet juice. The calculated HTC values by both methods were found to be in acceptable agreement. A heat loss of 5% of the total heat flow was taken into account. The determined HTC value has a maximum uncertainty level of 10%.

**Results**

*Heat transfer measurements*

Figure 4 shows HTC values obtained with pure sucrose solutions (made up using refined sugar) and the factory cane sugar juice, with the FFPE clean. Using factory cane juice, higher HTC values were determined in comparison with pure sucrose solution. In addition, it was observed that juice flow through

**Table 1**  
Summary of experimental runs reported.

Run	Brix in (%)	Brix out (%)	ΔT (K)	Vapour pressure (kPag)
Clean evaporator. Synthetic and factory juices to determine the relationship between HTC and brix	10-60	10-65	0,5-11	60(-40)
First effect simulation	11,5 (av)	12,7 (av)	3 (av)	50 (av)
Second effect simulation	25 (av)	32 (av)	9 (av)	20 (av)
Fifth effect simulation	52 (av)	57 (av)	12 (av)	-30 (av)

the plate pack became more uneven when factory juice was used, as opposed to synthetic sugar solutions. Reasons for this are not clear at this stage since no such effect has been found on a tubular rising film pilot evaporator at Felixton (FX). This effect will be examined in future work. Both sets of results showed very high HTC values obtained using the equipment under the conditions tested. These results are largely of academic value since the DT values used were small (<4 K).

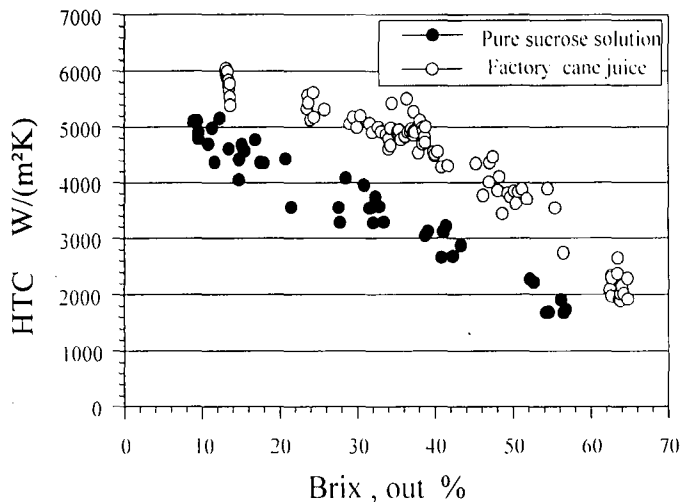


Figure 4. HTC measurements for a range of brix solutions. The measurements were all carried out with the evaporator surface clean.

In practice, the HTC value will be determined largely as a result of the level of fouling, and therefore long term tests were undertaken to examine the change in the HTC with time, due to fouling. Figure 5 shows the results with clear juice simulating a first effect, and trials with first effect juice simulating a second effect operation. After each trial the evaporator was cleaned and restarted. Only a very small drop in the HTC value was found when clear juice was used and, on inspection, only small amounts of scale were formed and most of the heating surface was free of scale. This result is not unexpected since the brix change was quite small for these runs. When run as a second effect under more realistic conditions, a significant drop in HTC value was found in the first 24 hours, after which the HTC value remained steady. In all cases, the HTC values obtained when the evaporator was fouled was markedly higher than any obtained for clean factory vessels.

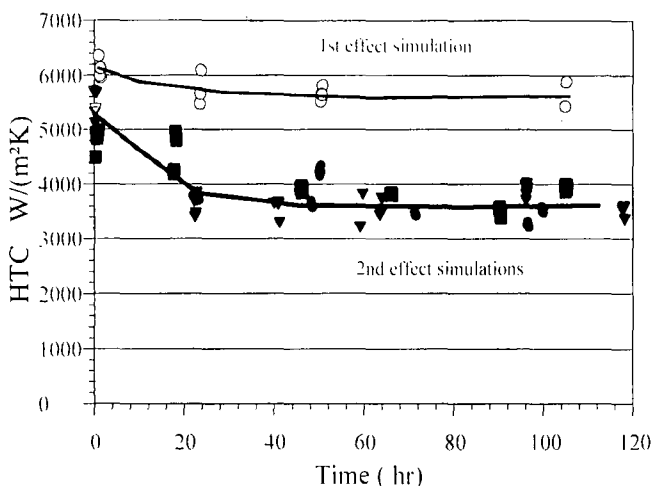


Figure 5. Long runs using clear juice (first effect simulation) and several runs using first effect juice (second effect simulation), showing the change in HTC due to fouling.

Figure 6 shows the deterioration of the HTC value over 300 hours of operation when the evaporator was run as a fifth effect, using fourth effect juice as a feed. Due to alterations in the brix of the incoming juice, the HTC values show relatively large

variations. At the end of the trial a scale thickness of about 0,4 to 0,5 mm was attained. Even when fouled, the HTC value was much higher than expected from Robert evaporators.

Throughout the work no blocked 'tubes' were found, despite occasional low or no juice flow due to variations in the factory. The only problem experienced was that of the distributor holes blocking due to excessive clarifier mud carry-over, combined with the action of large flakes of rust, presumably the result of corrosion after the strainer. Such distributor fouling occurred under unusual conditions and the distributor has been redesigned to prevent a recurrence. Furthermore, such distributor blocking could have occurred with many different falling film distributor designs, and this aspect will need to be examined in more detail in the coming season.

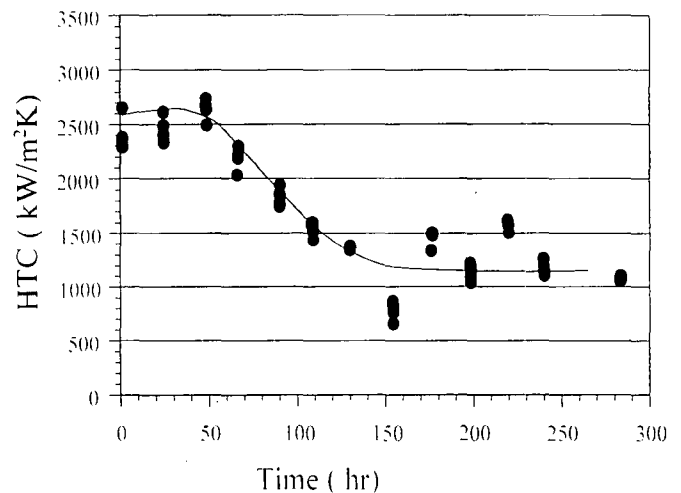


Figure 6. Change in HTC value with time due to fouling when the FFPE was operated as a fifth effect.

#### Cleaning procedure

The evaporator was cleaned chemically eight times within the four months of operation. The cleaning procedure was as follows:

- An alkaline solution was circulated for four to five hours through the evaporator under boiling conditions (steam pressure = 26 kPag; vapour pressure = 20 kPag; boiling temperature = 105°C), using the following mixture: 0,5% dispersant (Kebosol PM); 0,2% wetting agent (Kebosol VD); 1,5% NaOH (caustic soda) and 3,5%  $Na_2CO_3$  (soda ash). This step was designed to attack all amorphous organic material and silica. The sodium carbonate should convert all calcium salts to calcium carbonate.
- The evaporator was washed with water (condensate).
- An acid cleaning was carried out by circulating a 3% formic acid solution (inhibited with lithosolvent CS) at 60 to 75°C for three hours. This served to dissolve any carbonates formed.
- A neutralisation step to avoid corrosion. The following compounds were added: 1,5% NaOH; 0,15% wetting agent (Kebosol VD) and 0,2% dispersant (Kebosol PM). The

solution was circulated for one hour through the evaporator under boiling conditions (steam pressure = 26 kPag; vapour pressure = 20 kPag; boiling temperature = 105°C).

- The evaporator was washed with clear water.

The above procedure restored the HTC values satisfactorily. After the 300 hour run using fourth effect juice, and after the alkaline (caustic) part of the cleaning, the evaporator was opened and inspected. It was found that caustic alone had resulted in large flakes of scale, loosened by the caustic, blocking the plates. These were broken up when an acid cleaning was carried out and the HTC was fully restored, demonstrating the need for a two-step cleaning. In practice, the use of formic acid may be limited by the cost, and a cheaper acid may need to be substituted (Walthew *et al.*, 1997). The cleaning time (about 10 hours in all) was also much longer than is normal in South African factories.

*Scale composition*

Scale from the plate evaporator and the GH factory evaporators were analysed using X-ray fluorescence (XRF), X-ray diffraction (XRD) and high performance liquid chromatography (HPLC), and from these results the main components were able

to be determined (Walthew, 1996). Table 2 shows the composition of scale in the main plant, and shows the scale to be typical of the local industry, with silica and oxalate rising with increasing effect and phosphate and amorphous organic compounds highest in the early effects (Walthew and Turner, 1995). The loss on ignition (LOI) figure 'estimated from composition' refers to the LOI calculated from the proposed composition, and comparison with the measured LOI figure indicates the accuracy of the estimated amount. Table 3 shows an analysis of the samples of scale removed from the pilot plant. Inspection of the scale samples showed that the scale formed was different from that from the main plant effect being simulated. The differences could have arisen from:

- The lower residence time in the FFPE compared with the main plant evaporators, since phosphate and amorphous organic compound fouling are kinetically controlled. The negligible oxalate formed in the pilot plant would be consistent with evidence that oxalate forms as a result of aconitic acid degradation, and would be inhibited by short residence times (Walford and Walthew, 1996).
- The brix change across the evaporator was not as great as that of the main plant.

**Table 2**  
Scale composition, factory evaporants (per cent of dry scale).

Effect	Amorphous calcium phosphate	Magnesia hydrate	Silica hydrate	Oxalate	Lime hydrate	Amorphous organic	Loss on ignition	
							Calculated	Measured
1	64,9	7,2	6,0	0,0	0,0	15,2	32	40
2	34,3	3,0	5,8	3,1	0,0	49,2	60	62
3	19,5	2,3	14,7	19,0	16,7	26,1	50	50
4	2,5	0,3	31,8	17,9	21,5	25,4	50	50
5	11,8	2,1	40,7	14,9	8,1	18,2	42	44

**Table 3**  
Scale from the plate evaporator (per cent of dry scale).

Effect simulated	Amorphous calcium phosphate	Magnesia hydrate	Silica hydrate	Oxalate	Lime hydrate	Amorphous organic	Loss on ignition	
							Estimated from composition	Measured
First	49,6	6,5	5,9	1,8	0,1	36,0	50	47,8
Second	8,1	2,0	22,7	1,8	0,9	64,5	73	57,5
Fourth effect scale samples								
Distributor	11,9	0,6	6,2	0,9	0,0	80,4	85	85,4
'Flakes' after caustic only	19,8	2,3	23,6	1,8	1,9	50,6	62	59,9
Before/during cleaning	16,2	3,7	21,7	3,6	13,4	40,5	56	54,1
After cleaning	12,3	1,8	30,1	5,5	69,6	0,0	31	10,0

Although these tests suggest changes in scale composition that could result from a FFPE installation, the scale is sufficiently similar to make the cleaning procedure of interest.

Included in Table 3 are a number of special samples of interest:

- ‘Flakes’ refer to the loose scale removed from the plates after caustic cleaning only. This shows that the flakes still contained substantial quantities of silica and amorphous organic material. This suggests that the caustic solution was not 100% efficient in those areas, due either to insufficient caustic soda, passivation or insufficient wetting. The fact that a follow-up acid wash was successful suggests that insufficient wetting was not the cause. Acid cleaning following a caustic boil has been found necessary on the plate evaporators at UR (Walthew *et al.*, 1997).
- ‘Distributor’ scale refers to scale that was found to have blocked the feed distributor holes. Analysis showed that this scale was mainly clarifier mud, and the fact that this mud had passed through the screening system gives some cause for concern. If falling film evaporators are installed in the industry, then steps need to be taken to ensure either that no mud is carried over, or that the distributor design can cope with random occurrences. Although iron was removed from the analysis to avoid confusion, inspection showed the presence of rust flakes that may have initiated the blockage.
- ‘Before/during cleaning’ sample refers to samples taken from the separator before and during the early stages of the cleaning operation with caustic. Compared with samples taken ‘after’ cleaning, it can be seen that the amorphous organic matter was entirely leached during the cleaning operation, but that silica and free lime were not attacked to an appreciable degree.

*Scale quantity*

The quantity of scale can be estimated to some extent by consideration of the fouling resistance ( $R_f$ ) which is calculated from measurements of the HTC when the evaporator is clean ( $U_0$ ) compared with when the evaporator is fouled ( $U_1$ ), according to the following equation:

$$R_f = 1/U_1 - 1/U_0$$

Fouling resistance is related to scale thickness by:

$$R_f = x / k$$

where  $x$  = scale thickness and  $k$  = thermal conductivity. The fouling resistance indicates the quantity of scale formed. However, this is a rough measurement only since, ideally, the HTCs should be measured under the same conditions. This is not always possible in a factory since, as the evaporators become fouled, both flowrate and DT change. It is also assumed that the evaporator is perfectly clean at the ‘clean’ stage of measurement, a difficult achievement in practice without extensive cleaning procedures. For this pilot plant work the calculated fouling resistances are given in Table 4 along with some others obtained in the industry, as well as from work on the pilot plant at FX (Walthew and Whitelaw, 1996). Table 4 shows that the FFPE

produced encouragingly low fouling resistance when compared with both the industry and pilot plant work at FX. The larger fouling resistance values in the fifth effect compared with the second effect are expected to be due to higher fouling rates at the lower temperatures and higher brix levels (Walthew, 1994).

Scale formed in the pilot FFPE had a thickness of 0,1 to 0,2 mm for the 100 h trials with clear and first effect juices. In the 300 h trial (which is longer than normal factory operation between cleaning), with fourth effect juice as feed, the scale thickness was in the range 0,4 to 0,5 mm, which gives a calculated thermal conductivity to the scale of about 0,8 W/m<sup>2</sup>/(°K/m). This is well within the values reported in the literature (Walthew, 1994). The scale formed in the pilot plant was thinner than that generally produced by the typical factory scale from the fifth effect. In terms of scale quantity and composition, it would seem that the FFPE fouling produced scale comparable with that experienced by the main plant.

**Table 4**  
Some typical figures for fouling resistance  
(EFR = estimated fouling resistance).

Mill	Evaporator type	EFR* (m <sup>2</sup> K/kW)
Sezela	First effect Kestner	2,0
Pongola	First effect Kestner	0,4
Pongola	First effect falling film	0,3
GD/Aust mill	Second effect rising film	0,1
Union Co-op	First effect semi-Kestner	0,7
FX main evap	Second effect Kestner (1995)	0,2-0,3
FX main evap	First effect Kestner (1996)	0,07
FX main evap	Second effect Kestner (1996)	0,11
FX main evap	Fourth effect Robert	0,2-0,6
FX main evap	Fifth effect Robert	0,6-1,0
FX pilot plant	Second effect simulation	0,1-0,3
GH FFPE	Second effect simulation	0,09
GH FFPE	Fourth effect simulation	0,6

**Conclusions**

Work to date on the FFPE has been encouraging, in that the plant operated in a relatively trouble free manner for the season. HTC values were excellent and, even when fouled, the pilot plant HTC values were much greater than the best recorded figures for main plant evaporators. The rate of fouling appeared acceptably low even with extended operation as a final effect, with periodic flowrate fluctuations. The type and quantity of scale formed seemed to be an improvement over those currently experienced by the main plant, although higher performance is a typical characteristic of pilot plant operations. The distributor

blockage occurred under exceptional circumstances and does not represent an insurmountable problem. This will be investigated further. A similar problem with distributor blockage on a 1 047 m<sup>2</sup> plant at Cantly was easily dealt with by 'backflushing', i.e. raising the juice level to the top of the plate pack (personal communication). Should a move to falling film operation be anticipated by the industry then distributor design, as applied to cane industry evaporators, is an important area for further investigation. Although the operation was more time consuming than current practice, chemical cleaning was effective. It is thought that, even in the case of extreme scale formation and tube blockages, the unit can be cleaned without dismantling. Aspects of chemical cleaning are considered separately (Walthew *et al.*, 1997). The use of plate packs with larger gaps will also reduce the possibility of blockages.

Further work during the 1997-98 season will include investigations into cleaning using other, cheaper, chemicals, reduced cleaning times and the use of different plate packs.

### Acknowledgments

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### APPENDIX 1

#### Calculation of the HTC

The heat transfer coefficient was calculated from the modified Fourier equation:

$$HTC = \frac{Q_c}{A * dT}$$

where:

HTC = apparent heat transfer coefficient, kW/(m<sup>2</sup> °K)

A = surface area (m<sup>2</sup>)

Q<sub>c</sub> = rate of heat energy transfer from latent heat  
(calandria condensate flowrate x latent heat of vapour) [calandria condensate method]

or

(latent heat of vapour x calculated mass difference between flowrates in and out as calculated from the brix change) (kW)

dT = T<sub>s</sub> - T<sub>j</sub> = temperature difference across the heat transfer surface,

where: T<sub>s</sub> = steamside temperature (°C)

T<sub>j</sub> = juice temperature out (°C).

Juice temperature at the outlet was determined by adding the saturated steam temperature at the measured vapour pressure to the boiling point evaluation at the outlet brix. The formulae used to calculate the properties needed to obtain the HTC from the measured data are those recommended by Peacock (1995). Methods of calculation and their implications are given in more detail elsewhere (Walthew, 1994; Smith and Taylor, 1981).

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