SOME DATA ON HEAT TRANSFER IN MULTIPLE EFFECT EVAPORATORS

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Abstract

Heat transfer data from measurements covering 15 Roberts evaporator vessels at 3 mills (82 data sets total) is presented. Heat transfer coefficient (HTC) is plotted against brix, temperature, temperature difference (Δt) and viscosity respectively. It is shown that the influnce of Δt on HTC cannot be deduced from this type of data. The pronounced difference in operating conditions between last effect and all earlier effects, due to viscosity levels which are orders of magnitude higher in the last effect, is highlighted. HTC values from second to penultimate effect nearly all fell into the range of 1,8 to 3,5 kW m^{-2} $°C^{-1}$, with no pronounced dependence on effect number. It is shown that very high last vapour vacuum levels are detrimental to evaporation capacity and optimum last vapour saturation temperature is in the range 55 to 60°C. A heating surface distribution wherein last effect is double the size of intermediate effects would provide 6% more evaporation than the conventional arrangement of all tail vessels having equal area. HTC values are presented for evaporator sets where this has been achieved by paralleling existing vessels on vapour. It is concluded that for series juice feed, the advantage of lower brix in the first vessel of the pair is offset by the absence of flash in the second vessel.

Introduction

In recent years a number of surveys have been undertaken to determine heat transfer coefficients (HTC) of the individual vessels in multiple effect evaporators in Huletts mills. The main purpose of this work has been to identify and quantify instances of sub-optimum performance. In some cases severe steam-side tube fouling conditions were found which were corrected by chemical cleaning (Lewis *et al*[6]).

In the course of this work a fairly substantial body of heat transfer data has been accumulated. The purpose of this paper is to discuss certain indications that can be derived from these results, in respect of both operating conditions and evaporator configuration. Some cautionary remarks on the dangers associated with interpretation are also included.

Measurement and Evaluation Procedures

The procedure used for determination of heat transfer data varied somewhat between different investigations, but the following basic approach was used to derive all the results presented here.

The main data required for calculation of evaporator overall heat transfer coefficients are heating surface, evaporation rate and temperatures on both shell and tube side of the calandria.

The most practical procedure for deriving evaporation rates is to determine flow rate of solids or brix through the train and measure juice concentration (as brix %) at all terminal and intermediate points. This was the approach adopted here. Brix rates were derived from mixed juice mass and analysis data, adjusted for loss of brix in the filter cake output from the clarification stage. Test run duration was typically two to four hours to enable fluctuations in flow rates and pressure conditions to be averaged out. Juices and syrup were sampled continuously, through cooled copper coils to eliminate sample flash or evaporation, into large glass containers which were under partial vacuum where necessary.

Temperatures can be measured either directly or indirectly via pressure readings. Our experience with direct temperature measurement has not been good. Values obtained were frequently lower than saturation temperatures derived from simultaneous vapour pressure measurements, even though thermometer pockets in juice and condensate outlet lines were located as close to the vessel as possible in order to minimise the effect of flash cooling. Therefore only data derived from measurements of pressure in the calandria and the evaporator body has been included here. Pressure measurements were made with a mercury manometer in most cases, or alternatively with calibrated gauges, and corrected for barometric pressure. Corresponding saturation temperatures were adjusted for boiling point elevation and hydrostatic head on the juice side, but not for condensate subcooling on the steam side.

All data processing was done using the evaporator calculation program PEST (Hoekstra[2]). In addition to the factors already mentioned, this program takes into account temperature/enthalpy relationships, juice superheat, vapour bleeding and the return of condensate flash in the precise configuration applicable to each installation.

Results and Discussion

Data from 19 test runs covering a total of 15 vessels at three mills is shown in the figures discussed here. This data has been screened to exclude results for vessels which are known to suffer from steamside tube fouling, and for cases of two-vessel effects. The latter situation is abnormal if juice feed is in series since there is no flash in the second of the pair, and if juice is fed in parallel then there is a problem with quantification of the feed split and hence determination of evaporation rates. A few more results were rejected because of suspected dilution or contamination of juice samples.

Long tube climbing film evaporators are also excluded because of their fundamentally different design. Many of the first effect vessels in the installations studied here are of this type. It so happened that data for most of the remaining first effect vessels had to be excluded for one or other of the reasons mentioned above. The evaluation has thus been restricted to *tail* vessels only, that is from second to last effect. This has the further advantage that, with one exception, no vapour bleed is withdrawn from the vessels considered and one potential source of error is thus avoided.

Of the five evaporator trains represented, four are quadruple effect and one quintuple effect. The vessels have been divided into three groups, represented by different symbols in the figures, as follows :

O Second effect

Δ Intermediate effects — third effect, plus fourth effect of quin

□ Last effect

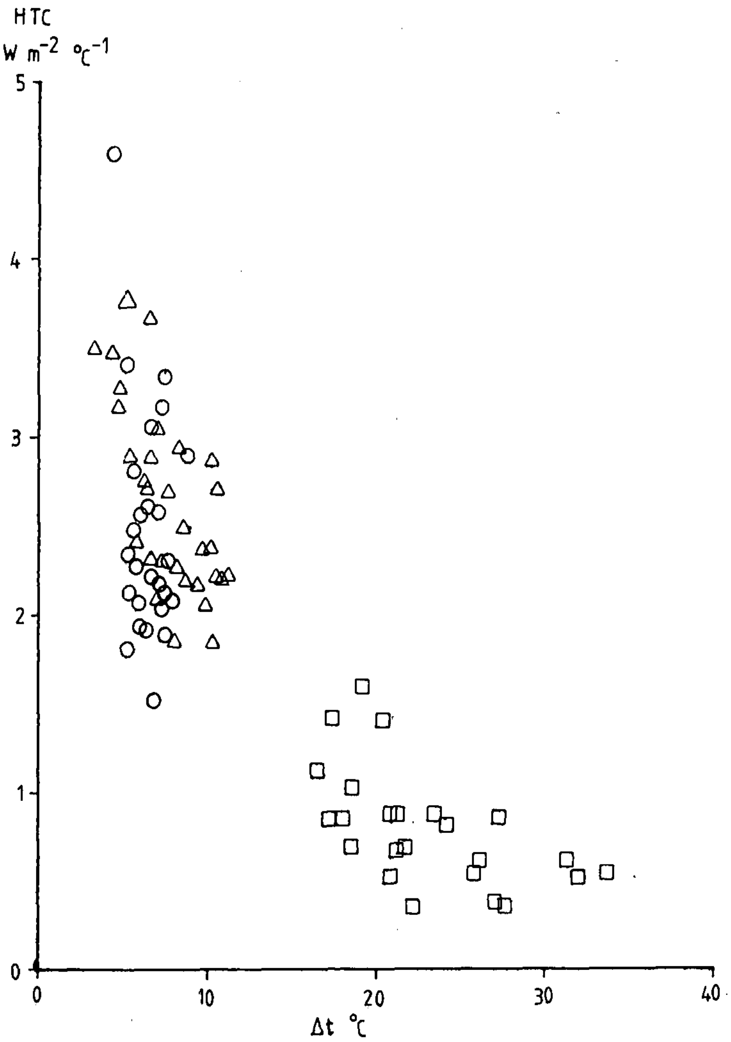

FIGURE 1 HTC versus temperature difference.

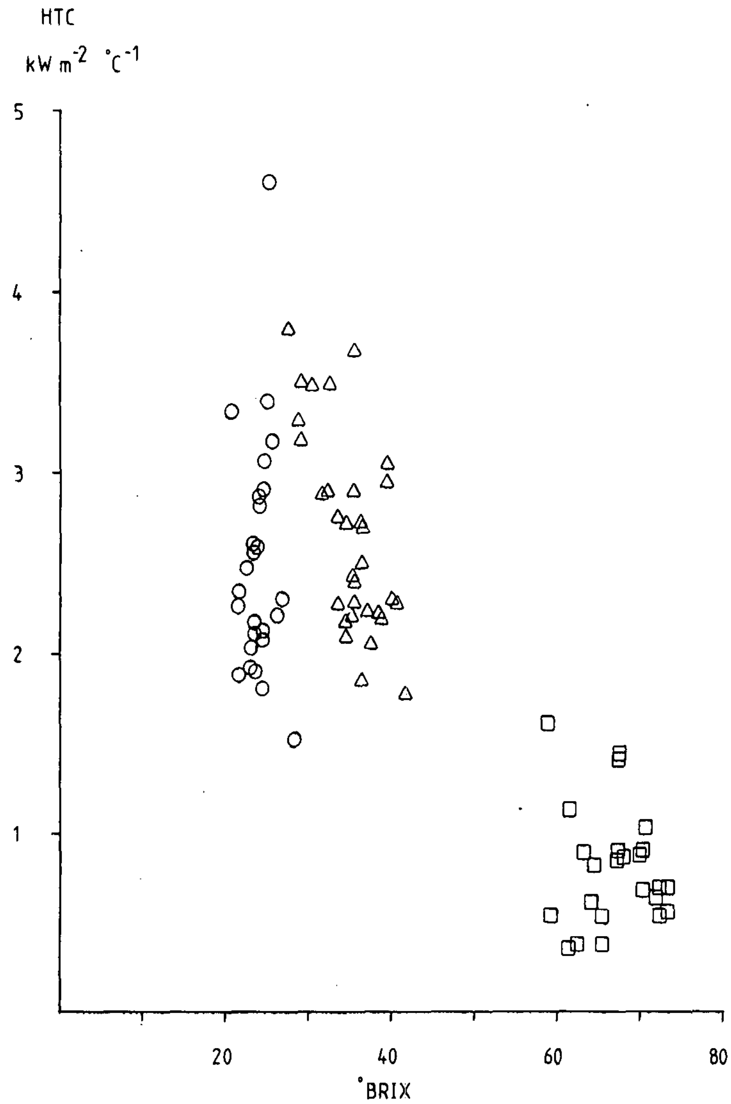

FIGURE 2 HTC versus outlet brix.

Relationship between HTC and temperature difference

A formula for prediction of HTC from operating variables such as brix, temperature and temperature difference across the heating surface (Δt) would be valuable in evaporator design and evaluation. Various relationships along these lines have been proposed (Batstone and Prince[1], Zadrodzki and Kubasiewicz[9]), which imply a positive effect of Δt on HTC. It may be thought that statistical regression analysis on data like that presented here could provide such a formula. It is worth mentioning why this approach is not feasible.

Figure 1 is a plot of HTC against Δt for the 82 data sets considered here. The picture shown is one of an inverse relationship between HTC and Δt, for the following reasons.

(a) There is a strong intercorrellation between the two variabes due to the definition of HTC, as expressed in the well-known formula

$$U = \frac{Q}{A\ \Delta t}$$

where U = Heat transfer coefficient, kW m^{-2} °C^{-1}.
Q = Heat flow rate, kW.
A = Heating surface, m^2.
Δt = Temperature difference, °C.

Thus in the case of a measurement error which results is an underestimation (for instance) of Δt, the calculated HTC will be an overestimate. Any errors in temperature estimation will therefore contribute to the indication of an inverse Δt/HTC relationship.

(b) The second, and overriding, reason relates to the way in which sugar factory evaporators are designed and operate. Conventional design practice is to manufacture all tail vessels in the same set with equal heating surface, and this applies in most cases for our data. Thus for the different effects of each set, A in the above equation is constant. Likewise, without vapour bleed the heat transfer rate Q is substantially equal for each effect. Thus HTC must be approximately inversely proportional to Δt. Since the set must balance itself out, the conditions which give rise to low HTC — high brix and low temperature in the last effect — inevitably also result in high Δt. It is therefore HTC which governs Δt in this situation, at least to an extent which overwhelms any influence in the reverse direction which Δt may have on HTC independently on these constraints.

For these reasons the type of data presented here can give no information on the effects of Δt in isolation. Another related factor is the greater reliability of HTC values for last effect due to the high Δt values (between 15 and 35°C here). In contrast the great majority of Δt values for earlier effects are below 10°C. An error of 1°C would thus have a significant influence on the calculated HTC value, and it is not easy to achieve this accuracy under typical sugar factory conditions of varying temperature due to changes in juice feed and vapour bleed rates, and steam pressure.

Effects of brix and temperature on HTC

HTC is plotted against exit juice brix and juice temperature in Figures 2 and 3.

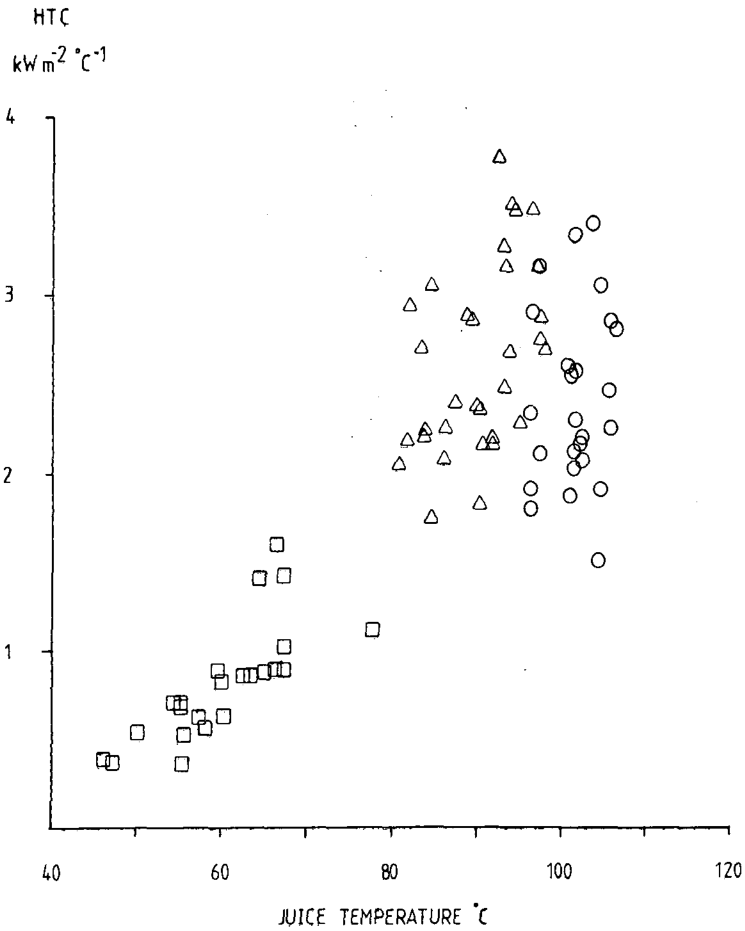

FIGURE 3 HTC versus juice temperature.

There is a distinct separation between second plus intermediate effect and last effect groupings, in respect of both HTC and the two independent variables. The influences of brix and temperature can be combined by considering the variable which most directly affects HTC, namely viscosity. A regression equation was fitted to data in tables of viscosity of pure sucrose solutions against brix and temperature (Norrish[7]) and the equation then used to calculate viscosity for each data set. HTC is plotted against these values in Figure 4. The viscosity values will not be accurate in absolute terms because of the influence of impurities in the juice, but they are adequate for the comparative evaluation presented here.

Taking into account the logarithmic viscosity scale, the split into early/intermediate and last effect groupings is still more pronounced. From the pattern on the graph this is clearly not suitable material for any attempt at statistical regression, but a few generalised observations can be made on each of the two groupings referred to above.

Prior to last effect, viscosities seldom exceed 1 centipoise and there is little evidence of reduction in HTC with increasing effect number. This is contrary to accepted technology where a linear decrease is usually assumed (Hugot[3]) but in line with measurements carried out by the Sugar Milling Research Institute on a pilot climbing film evaporator (James *et al*[5]). As regards the actual HTC levels found here, little can be said in that nearly all lie between 1,8 and 3,5 kW m^{-2} °C^{-1} with the few outliers probably due to undetected determination problems. The measurement techniques used are clearly insufficiently accurate for closer quantification, perhaps largely due to the sensitivity to temperature estimation mentioned earlier, although they have succeeded in pinpointing areas of gross malfunction. We believe that procedures used in future should be based on continuous instrumented monitoring of individual vessels, rather than sequential measurements on all vessels (often 10 or more) which limits the number of readings that can be taken and thus the accuracy of resultant averages.

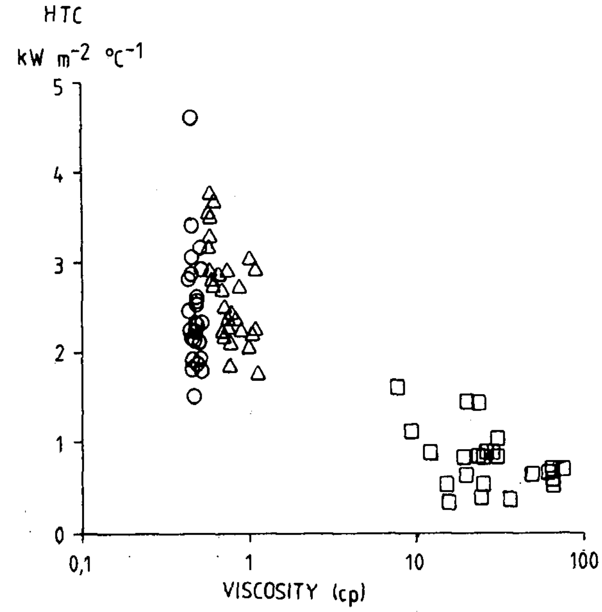

FIGURE 4 HTC versus juice viscosity.

The position is somewhat different for last effect where large brix as well as temperature differences result in HTC values of much greater reliability. Two aspects, relating to evaporator operation and sizing respectively, are discussed separately below.

Optimum last vapour pressure/temperature

Reduction of last vapour pressure and thus temperature increases the overall temperature difference across the set, which tends to increase total evaporation. At the same time viscosity of syrup rises, which lowers last effect HTC. From the data presented here it has been possible to quantify this interaction. A significant linear correlation was found between the HTC and last effect temperature values which comprise some of the plots in Figure 3.

This formula : HTC = (0,034 x Last vapour temperature, °C − 1,13) kW m^{-2} °C^{-1} was used to calculate HTC values at 5°C intervals of last vapour saturation temperature. A series of PEST runs was then done with data from one of the Darnall investigations as base, to calculate total evaporation for each of these last effect HTC/temperature combinations. The output is summarised in Table 1.

The data in Table 1 demonstrates clearly the undesirability of operating at the high vacuum conditions which can be achieved with modern countercurrent rain condensers. In such a situation the drastic reduction in last effect HTC due to cool and viscous syrup can reduce vaporator capacity by 10% or more relative to more moderate conditions. The optimum operating range indicated is from 16 to 20 kPa absolute corresponding to saturation temperatures between

TABLE 1
Change in total evaporation at different last vapour temperature and corresponding last effect HTC levels

Run No.	Last effect: Pressure kPa abs.	Vapour sat. temp. °C	HTC kW $m^{-2}°C^{-1}$	Overall Δt °C	Total evaporation tons/hour
1	10	45,8	0,40	78,4	196,3
2	12	49,4	0,56	74,8	227,4
3	16	55,3	0,74	68,9	240,1
4	20	60,1	0,90	64,1	241,9
5	25	65,0	1,08	59,2	237,0
6	31	69,9	1,24	54,3	224,8

55°C and 60°C. In this case our findings are substantially in line with recognised practice (Hugot[4]).

Appropriate distribution of heating surface

As mentioned earlier, standard design practice is for all tail vessels to have the same heating surface. This is not the optimum situation due to the significantly lower HTC of last effect relative to earlier effects. The procedure of a series of PEST program runs was again used to investigate this aspect, the variable quantities here being the heating surfaces of last effect and of intermediate effects. A quintuple effect evaporator of 8 000 m^2 total heating surface was considered. First effect heating surface of 4 000 m^2 or 50% of the total (fairly typical because vapour bleed greatly increases evaporation here relative to later effects) was assumed. Distribution of the remaining 4 000 m^2 between the other four effects was varied, with the base case being 1 000 m^2 for each. The comparison cases assumed the same area for the intermediate effects but different ratios of last to intermediate effect areas — thus for the ratio of 2 for last effect heating surface (HS_5) to intermediate effect heating surface (HS_i), the second, third and fourth effects are of 800 m^2 each while fifth effect is 1 600 m^2. HTC values assumed were 2,7 kW m^{-2} $°C^{-1}$ for first effect, reducing in 0,3 unit intervals to 1,8 for fourth effect and 0,6 for fifth effect; being a compromise between the conventional HTC decrease assumption and the indications from our work of little diferece before last effect. Results are shown in Table 2.

TABLE 2
Change in total evaporation for different ratios of last effect to intermediate effect heating surface (constant total area for set)

Run No.	Area ratio HS_5/HS_i	Last effect Δt °C	Total evaporation: Tons/hour	% Run 2
1	0,5	31,6	162	85
2	1	25,6	190	100
3	1,5	21,0	200	105
4	2	18,0	202	106
5	3	14,1	196	103

The indication is that total evaporation is 6% higher if last effect is twice the area of intermediate effects than the conventional situation of the same heating surface for all tail vessels. For a grassroots installation the extra design and fabrication costs of a different sized vessel must be offset against this advantage. The potential to improve heating surface distribution in the case of expansion of an existing set has however been recognised (Van Hengel[8]). The final two vessels can be parallel on vapour to form a double-sized last effect and a new vessel installed (or vessels rearranged) earlier in the tail. Modifications along these lines have in fact been carried out to five of the eight evaporator trains in the Huletts mills over the last 10 years. In all cases juice feed is in series between the two paired vessels. This provides a further advantage in that brix in the first vessel of the pair is well below final (syrup) brix level and thus HTC is higher. However another factor tends to negate this advantage, namely absence of flash in the second vessel of the pair. Table 3 gives HTC values for such two vessel last effects. It must be noted that most of these figures were calculated using short-cut manual methods and are thus not directly comparable with the rigorous PEST values discussed earlier.

TABLE 3
Measured HTC values for two-vessel last effects with juice flow in series

Mill	Year	HTC kW m^{-2} $°C^{-1}$ 1st vessel	2nd vessel
FX	1974	0,62	0,25
AK	1974	0,70	0,36
	1979	0,70	0,37
DL (Fletcher tail) ..	1976	0,57	0,53
	1980	0,93	0,65
ME	1975	0,81	0,45
	1976	0,90	0,40

HTC values for the second vessel of the pair are invariably lower than for the first vessel, to the tune of some 50% except in the case of the small Fletcher tail at Darnall. Overall HTC for the last effect as a whole is no better, and perhaps somewhat lower, than for a single-vessel last effect. Parallel juice feed would equalise the HTC values but would introduce split control problems as well as eliminating the low-brix advantage in the first vessel of the pair. In general, operating results from this expansion route philosophy have been a trifle disappointing. This is not easy to say that the original decision was wrong since material savings in capital cost have also accrued.

Conclusions

Our involvement with evaporator testing over the past few years has perhaps taught us more about the difficulties and dangers of this activity than it has produced concrete and reliable data. Nonetheless we now have a better "feel" for what happens in this important unit operation of sugar manufacture. In particular, the overriding importance of the last effect has been highlighted.

The conventional wisdom of backing off from maximum attainable vacuum conditions has been confirmed. The magnitude of the disadvantage in the traditional sugar industry forward feed arrangement for both juice and vapour, resulting in last effect liquid viscosity in a much higher order of magnitude than those earlier in the train, has been demonstrated. Perhaps the standard practice in this area should be reviewed in the light of today's much higher capital costs.

Plant measurements do not provide a suitable basis for determining the relationship between HTC and operating conditions, and this type of information should be derived from pilot plant investigations, where the relevant parameters can be varied at will. As regards future plant test work, we favour an approach involving intensive measurements on individual vessels rather than the extensive evaluation of entire evaporator installations as reported here.

Acknowledgements

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