HEAT TRANSFER COEFFICIENT CORRELATIONS
FOR ROBERT JUICE EVAPORATORS

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Abstract

CORRELATIONS FOR heat transfer performance in sugar factory juice evaporators are important for benchmarking and prediction of evaporator set performance. The correlation recently proposed by Broadfoot and Dunn (2007) (B&D) is examined and comparisons are made between it and formulae currently being used, in particular with the ‘Australian typical’ correlation, using data from Australian and overseas evaporator sets. When used on the available data from Australian evaporator sets the B&D formula was found to give a poorer fit than did the ‘Australian typical’ formula, particularly so where the conditions involve a high juice concentration. A modified correlation is derived which retains a positive power of the temperature difference term gives a reasonable fit to the extended factory data for conventional Robert Australian evaporator vessels. The new correlation is named the ‘Austyp08’ formula and it is argued that this should be used in the future for conventional Robert evaporator vessels. The ‘Austyp08’ formula takes the form: $HTC = 0.00049 \times (110 - B_j)^{1.1616} \times T_j^{1.0808} \times \delta T^{0.266}$ where: $HTC$ is the heat transfer coefficient, $kW/(m^2 K)$; $B_j$ is the concentration of the juice leaving the vessel, °Bx; $T_j$ is the temperature of the exit juice, estimated as sum of the saturated vapour temperature in the headspace and the boiling point elevation of the juice, °C. $\delta T$ is the temperature difference between the condensing temperature of the heating steam/vapour in the calandria $T_s$ and $T_j$, K. For the SRI radial flow design Robert evaporators, it is recommended at present that the constant in the ‘Austyp08’ equation be increased by 30%. It is obvious, however, that more measurement data are required on these designs, especially when they are operating at low $\delta T$ values and as final stage vessels.

Introduction

In a recent paper Broadfoot and Dunn (2007) presented pilot and factory evaporator performance results and developed an improved correlation for heat transfer coefficients (HTC) which included a term for the vapour condensation loading. They claimed that the use of the new correlation would improve the reliability of modelling and of investment decisions. In the discussion on this paper it was agreed that having a vapour condensation loading term in the correlation, though it is in part correlating the HTC with a HTC-dependent term, would help explain some experiences with evaporator predictions. The 2007 paper did not detail any of the older HTC correlations, or highlight any comparisons with
them. The well used ‘Australian typical HTC equation’, a regression on data taken from 185 run-of-the-mill values matched for the 1980s Australian conditions of about 6–13 days operation between cleans, was not described or compared. A section of the paper dealt well with the evaluation of a time-dependent term to estimate the effect of fouling on the HTC.

It was considered that the B&D correlation needed to be examined and comparisons made between it and formulae currently being used overseas and in particular with the ‘Australian typical’ correlation, using data from Australian and overseas evaporator sets. It is also considered appropriate that a modified formulation of the present Australian formula for conventional Robert evaporators be derived, if necessary, using the extended Australian factory data.

Heat-transfer and evaporation coefficients

Hugot and Jenkins (1986) discussed several parameters for comparing the performance of the evaporators. Following their discussion we can distinguish the evaporation coefficient, the mass of vapour furnished by the vessel per hour per unit heating surface (kg/[m² h]). We can vary this by using the vapour condensation coefficient (VCC), the mass of heating vapour condensed by the vessel per hour per unit heating surface (kg/[m² h]), as a more accurate method of estimating heat flux, as used by Broadfoot and Dunn (2007).

Another Hugot and Jenkins term is the uncorrected heat-transfer coefficient, the heat flux per unit heating surface and per degree drop in temperature (kW/[m² K]). Here it is necessary to distinguish between values according to whether the coefficient uses the following methods for evaluating the temperature driving force for heat transfer, viz:

1. the apparent temperature drop where the juice temperature is taken as equal to that of the vapour produced from it; or
2. the net temperature drop (δT) where the juice temperature is taken as equal to the headspace vapour saturated temperature plus the boiling-point elevation due to the brix; or
3. the real temperature drop where the juice temperature is obtained by adding to the headspace vapour temperature the boiling-point elevation due to the brix, as well as the elevation due to hydrostatic pressure.

Australian technologists have generally preferred to take the temperature drop (or temperature driving force for heat transfer) as the ‘net’ value with the juice temperature defined by method 2 above. Their reasoning is that the hydrostatic pressure effect is difficult to quantify and is, in any case, a characteristic of the design of the vessel itself, assuming that the boiling height in the tube is kept close to that giving maximum performance. When making comparisons with some published HTC data, however, one must be careful to note and to make allowance for the method of calculation which has been used.

Data and correlations for heat transfer coefficients

For the purposes of the present paper the literature on heat transfer coefficients was reviewed to focus on design values for multiple effect evaporator vessels operating on cane clarified juice¹ cane for a normal operating period (one or two weeks) between cleans, and

¹ Clarified juice from simple defecation clarification.
where the given values can be adjusted to reflect the ‘net’ temperature drop obtained by Method (2) above. It is noted that almost all the available data and regressions use brix as a measure of concentration rather than dry substance².

**General heat transfer coefficient values**

General values for ‘apparent’ HTC (by Method 1) are given for cane juice quadruple effects by Jenkins (1966), Hugot and Jenkins (1986), and the Sugar Technology Manual (Bubnik et al, 1995), and others.

In order to convert ‘apparent’ to ‘net’ HTCs, it was first necessary to assume typical operating conditions for multi-effect sets (obtained by simulating typical sets using overall conditions proposed by Hugot and Jenkins (1986), viz. heating steam temperature is 120°C, final effect vacuum at 13kPa, juice and syrup brix respectively 12° and 65°, and then to estimate the conversion ratio \((T_s-T_v)/(T_s-T_v-T_{hpe})\).

These values were 1.034, 1.053, 1.088 and 1.112 for stages #1 to #4 of a quadruple set. The corresponding conversion ratios for a quintuple set have been estimated as 1.037, 1.057, 1.080, 1.127 and 1.129 for stages 1 to 5 respectively.

Apparent HTC for overseas cane juice Robert multiple effects, originally listed by Hugot and Jenkins (1986), have been included with other values and tabulated by Bubnik et al. (1995) in their Sugar Technology Manual.

The same values are currently listed on the Sugar Engineers’ Library (2007) page, along with some data from Tongaat-Hulett Sugar derived from plant measurements over a number of years. Some of these values have been averaged and converted to net coefficients in SI units as in Table 1.

The fit of some HTC formulae to these values are shown in Figure 1. The details of the formulae are discussed in a later section. It is seen that three of the values attributed to Clavijo, and all four Baloh (1991) values appear to be higher than the formula predictions.

**Table 1**—Mean overseas design HTCs taken from the literature.

<table>
<thead>
<tr>
<th></th>
<th>Overall net (Method 2) Heat Transfer Coefficient, kW/(m² K)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Tongaat-Hulett*</td>
</tr>
<tr>
<td>#1 stage</td>
<td>2.587</td>
</tr>
<tr>
<td>#2 stage</td>
<td>2.317</td>
</tr>
<tr>
<td>#3 stage</td>
<td>1.850</td>
</tr>
<tr>
<td>#4 stage</td>
<td>0.779</td>
</tr>
</tbody>
</table>

* Values converted from averages from the Sugar Engineers’ Library (2007), and from Bubnik et al (1995)

Other average values are given for Robert and other types of evaporator by Rein (2007). Where the sulphitation process is applied to juice the fouling scale in evaporators and juice heaters is more difficult to clean than for juice from the simple defecation process, and design HTC values for evaporator stages are usually reduced by about 30%.

² While dry substance is the preferred method of analysis for concentration in raw sugar factories, brix is satisfactory for high purity juices and syrups. Brix is usually measured by spindle or by refractive index methods, while dry substance is determined by vacuum drying.
HTC values for Australian Robert evaporators

Values for Robert evaporators as designed and operated in the Australian sugar industry\(^3\) are available from the work of Selman and Plomley (1950). The ‘standard’ values obtained there were set in the top 10% of the values measured and applied to the concentration of a relatively high brix clarified juice up to a syrup of 70 Brix. The estimated conditions and standard HTC used are given in Table 3.

<table>
<thead>
<tr>
<th>Stage</th>
<th>(T_s) °C</th>
<th>(T_v) °C</th>
<th>(T_j (=T_v+T_{bpe})) °C</th>
<th>Bj Brix</th>
<th>VCC, kg/(m(^2) h)</th>
<th>(\delta T), K (Method 2)</th>
<th>(HTC), kW/(m(^2) K)</th>
</tr>
</thead>
<tbody>
<tr>
<td>#1</td>
<td>110.28</td>
<td>103.33</td>
<td>103.73</td>
<td>19.02</td>
<td>39.56</td>
<td>6.55</td>
<td>4.129</td>
</tr>
<tr>
<td>#2</td>
<td>103.33</td>
<td>96.39</td>
<td>96.99</td>
<td>24.88</td>
<td>28.94</td>
<td>6.34</td>
<td>2.801</td>
</tr>
<tr>
<td>#3</td>
<td>96.39</td>
<td>87.78</td>
<td>88.87</td>
<td>36.24</td>
<td>29.38</td>
<td>7.52</td>
<td>2.422</td>
</tr>
<tr>
<td>#4</td>
<td>87.78</td>
<td>60.00</td>
<td>64.54</td>
<td>70.00</td>
<td>29.42</td>
<td>23.24</td>
<td>0.761</td>
</tr>
</tbody>
</table>

It is seen that the HTC values for the first three stages are well above the overseas design values given in Table 1. This is partly due to the choice of values as the better 10% of the measured values, and partly because of the superior arrangements for venting and design which were then becoming established in Australian installations.

McGrath and Webster (1982) recorded the HTC values for the quintuple set at Pioneer Mill, and noted the beneficial effect of the scale inhibitor Busperse-49 in slowing the fouling rate. The average of three sets of HTC values in the 1980 season taken 2–3 days after

\(^3\) For evaporators in the Australian raw sugar industry the clarified juice is usually relatively high in purity and relatively high in concentration (14–16 Brix), and the syrup concentration is usually set at 68 to 72 Brix.
cleaning was 2.43, 2.38, 2.35, 2.18 and 0.51 kW/(m² K) for stages #1 to #5 respectively. The syrup was maintained at 71 Brix. Comparative values taken in the 1982 season, where 12 ppm of Busperse-49 was added, were 3.39, 2.84, 2.69, 2.55 and 0.66 units.

De Viana and Coleman (1995) carried out trials in 1994 to assess the effect of a plate heat exchanger (PHE) on the performance of the quintuple set at Mossman Mill. Without the PHE addition they reported HTC values some 66 h after the set was cleaned, and decreasing values up to 316 h of operation. The set had been modelled by the present author in 1992, with the conditions defined as in Table 4. The 1994 results are also included. The HTC values from Mossman are seen to be well above average.


<table>
<thead>
<tr>
<th>Stage</th>
<th>$T_s$, °C</th>
<th>$T_v$, °C</th>
<th>$T_j = (T_v + T_{bpe})$, °C</th>
<th>$B_j$, Brix</th>
<th>VCC, kg/(m² h)</th>
<th>$\delta T$, (Method 2), K</th>
<th>HTC, 1992, 66h, kW/(m² K)</th>
<th>HTC, 1994, 316h, kW/(m² K)</th>
</tr>
</thead>
<tbody>
<tr>
<td>#1</td>
<td>120.00</td>
<td>113.00</td>
<td>113.38</td>
<td>17.83</td>
<td>27.50</td>
<td>6.62</td>
<td>3.387</td>
<td>3.400</td>
</tr>
<tr>
<td>#2</td>
<td>113.00</td>
<td>104.00</td>
<td>104.49</td>
<td>21.42</td>
<td>42.18</td>
<td>8.51</td>
<td>2.888</td>
<td>2.970</td>
</tr>
<tr>
<td>#3</td>
<td>104.00</td>
<td>94.00</td>
<td>94.68</td>
<td>27.14</td>
<td>36.92</td>
<td>9.32</td>
<td>2.357</td>
<td>2.490</td>
</tr>
<tr>
<td>#4</td>
<td>94.00</td>
<td>81.00</td>
<td>82.14</td>
<td>37.94</td>
<td>39.33</td>
<td>11.86</td>
<td>1.999</td>
<td>2.250</td>
</tr>
<tr>
<td>#5</td>
<td>81.00</td>
<td>54.00</td>
<td>57.95</td>
<td>68.10</td>
<td>38.76</td>
<td>23.05</td>
<td>1.010</td>
<td>1.170</td>
</tr>
</tbody>
</table>

Snoad (1997) discussed experiences with on-line evaporator performance monitoring at Mulgrave Mill, and recorded HTC values for the 3rd, 4th and 5th stages of a quin set as 2.55, 1.55 and 0.65 (in SI units) respectively, changing to 2.50, 1.35 and 0.60 units after ~240 h operation.

The author assembled a data set of 181 measurements taken in the 1970s by matching simulations on Australian evaporator sets. At that time there was a weekly cleaning regime, and the data were obtained by approximate matching of the typical performance, representing conditions appropriate for design predictions. These data can now be augmented by a further 67 measurements taken by the author in evaporator matching exercises carried out in the 1980s and early 1990s.

The measurements cover Australian factories in all cane growing regions, and all stages of quadruple and quintuple evaporator sets. One of the original points was rejected as a gross outlier, and some five of the #5 stage values were rejected on for having final vacuum saturated temperatures below 52°C.\(^4\)

Broadfoot and Dunn (2007) presented some 90 measurements of factory data for clean operation obtained from evaporator stations at Tableland, Isis and Broadwater Mills in the early 2000s. In the present examination, the points relating to the SRI Radial Flow technology evaporator were omitted, one point #4 stage (of quin) was rejected as an outlier for having a improbable $\delta T$ value and five #5 stage values were omitted for having final headspace saturation temperatures below 52°C.

\(^4\) Such low final headspace saturation temperatures, with pressures below 13.5 kPa abs, are feasible only in southern mills sited on cold water streams. Use of such low pressures augments the adverse effects of hydrostatic head in the tubes of Robert evaporators and can be considered counterproductive as far as evaporator performance is concerned.

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The combined HTC data have been tabulated along with the associated operating conditions. The table is too large to be presented in this paper. The fact that operating conditions for run-of-mill factory feed-forward evaporator sets are highly correlated can be clearly seen by the plot (Figure 2) of the conditions for all the data. Exponential trend curves are drawn through the data to guide the eye. Regressions on these data and indeed on all factory-derived data are complicated by the fact that the later evaporation stages are at the lowest juice boiling temperatures and the highest temperature differences. It is difficult to separate the individual effects of these parameters in accounting for the low HTC in the later stage vessels.

![Image of Figure 2: Operating conditions in the factory data set.](image)

**Formulae for design heat transfer coefficients**

Formulae for design heat transfer coefficients have been reviewed by Jenkins (1966), Hugot and Jenkins (1986), and in a number of Sugar Technology texts.

**The Dessin formula**

The French engineer Dessin proposed a formula permitting the evaporation coefficient to be calculated for any vessel of a multiple effect. This was modified by Hugot and Jenkins (1986) to take the juice concentration as the average of the inlet and outlet brix, and to reduce the multiplier to allow for average fouling during an operational cycle. There appears to be some confusion in Hugot and Jenkins’ equations, so after converting the Dessin formula as give by Jenkins (1966) the modified Dessin expression for HTC with reduction to 85% to allow for fouling becomes:

\[
HTC = 2.2 \times 10^{-7} \lambda_s (100-B_{jav}) (T_s-54) \tag{1}
\]

where the symbols are defined in the Nomenclature.
Another formula for overall HTC is used in an online design of a Robert evaporator in the Sugar Engineers' Library web page (2007). It is a function of juice or syrup temperature and brix\(^5\) as:

\[ HTC = 0.465 \cdot T_j \cdot B_j^{-1} \quad (2) \]

where the symbols are defined in the Nomenclature.

**The ‘Australian Typical’ formula**

Watson (1987) cited an earlier ‘typical’ HTC correlation from the original Australian data set assembled by the present author. Difficulties arise in the separation of the individual effects of the conditions. It was realised that the negative power term of the \(\delta T\) parameter was against the indications of theories of heat transfer in boiling tubes and so, after about 1987 the correlation was re-done with the power term of \(\delta T\) set to zero. This gave the ‘Australian Typical’ formula, which has served well in providing a useful benchmark for HTC against which evaporator performance can compared. The ‘Australian Typical’ formula then became:

\[ HTC = 0.01694 T_j^{1.0174} \left( B_j/(86–B_j) \right)^{-0.2695} \quad (3) \]

where the symbols are defined in the Nomenclature.

**The B&D correlation**

Broadfoot and Dunn (2007) presented pilot and factory evaporator performance results and developed a correlation for heat transfer coefficient which included a term for the vapour condensation coefficient (VCC). They operated a pilot evaporator to evaluate HTC over a range of processing conditions (factorial design) and found that the HTC at optimum juice boiling level was dependent on VCC to the power 0.21 and on the juice viscosity \(\mu\) to the power −0.5.

The VCC term varies linearly with HTC (by definition, \(VCC = (3.6 \delta T \ HTC)/\lambda_s\)) and this substitution can be made. It is then found that the HTC would vary as \(\delta T\) to the power 0.266 and viscosity \(\mu\) to the power −0.633.

When, however, these authors performed regressions on data obtained from Robert evaporator stations at Tableland, Isis and Broadwater Mills in the early 2000’s, they gave the formula for the HTC of ‘clean’ conventional Robert evaporators (termed herein the B&D formula) as:

\[ HTC = 21.6 \times 10^{-6} VCC^{0.4} \lambda_s \mu^{-0.34} \quad (4) \]

where the symbols are defined in the Nomenclature.

After substituting the expression \([3.6 \delta T HT C / \lambda_s]\) for VCC the B&D equation shows a higher \(\delta T\) power than in the pilot evaporator correlation. The B&D formula then becomes:

\[ HTC = 3.935 \times 10^{-6} \lambda_s \delta T^{0.6667} \mu^{-0.5667} \quad (5) \]

The value for ‘clean’ evaporators is adjusted for the influence of scaling using an exponential decay term, where the decay parameter (the scaling factor, ‘a’) reflects the % reduction in HTC per hour of operation time. Broadfoot and Dunn give values of ‘a’ as ~0.09 h\(^{-1}\) for the first and last stages and ~0.05 h\(^{-1}\) for the intermediate stages of the set. Thus the

\(^5\) A version of this formula quoted by Jenkins (1966) as ‘used in Sweden’ is the same as Equation 2 except that the converted value of the multiplier is given is 0.503. The formula is taken from van der Poel et al. (1998).
HTC estimate for a design value of ~240 h would be reduced to the range 0.80 to 0.88 of the clean value. Therefore a value which is 85% of the B&D formula (termed here ‘B&D 85%’) would give a more suitable benchmark for evaporators with a 1–2 week cleaning schedule.

The claim was made that the use of the new B&D correlation would improve the reliability of modelling and of investment decisions. However, the high value of the power of the $\delta T$ term in this correlation is of concern, as it is 2.5 times that found in the B&D pilot correlation and was not included at all in the much-used ‘Australian typical’ formula. As well, there are some problems in simulation due to the strong variation of HTC with the $\delta T$. It is considered necessary to validate the new B&D formula against available industry data before it can be recommended for adoption.

**Formulae based on regressions on the combined Australian data set**

With the 67 new measurements of Australian factory performance available to be added to the original 181 measurements, an approach was made to Broadfoot and Dunn, who kindly supplied the factory data used in the 2007 paper.

A combined data set was assembled to include the 67 new measurements and 75 of the Broadfoot and Dunn measurements. With the latter the HTC values had to be multiplied by 0.85 to allow for the fact that they were carried out on clean vessels and the other data were based on 1–2 weekly-cleaned vessels. Some measurement omissions were made as discussed above.

The combined data set was processed in a spreadsheet designed to derive the required parameters such as brix ratios, log values as well as viscosity values calculated from the brix and temperature by the expression of Broadfoot and Steindl (1980). Regression fits to the measured data were made using various combinations of the derived parameters using both the inbuilt EXCEL regression function, as well as a Fortran-compiled routine for minimisation of nonlinear least squares. The latter used the Levenberg-Marquardt method for nonlinear least squares and a trust region Newton method for minimisation.

After many variations $^6$ were tried it was considered that the near-to-best fit of to the HTC data in this set was as given as the simple equation of Equation 6:

$$HTC = 0.00056 (110–Bj)^{1.0025} T_j^{0.8294}$$

(6)

where the symbols are defined in Nomenclature.

For each measurement the deviation from unity of the ‘ratio of predicted to measured value’ was calculated. The root mean square (RMS) deviation of the 328 measurements was then estimated. It was 20.25% for Equation 6.

An exercise was then made to force into the regression the Broadfoot and Dunn (2007) finding from the pilot evaporator results that the HTC varied as $\delta T^{0.266}$. The regression fit of the term $HTC \delta T^{0.266}$ to the parameters $T_j$ and $(110–Bj)$ resulted in an expression which fitted the data with a somewhat poorer RMS error of 22.5%.

After communications with Dr Ross Broadfoot it was concluded that, even though the fit is poorer, it would be preferable to use the separately-evaluated positive value of $\delta T$ in

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$^6$ The addition of a viscosity term in the regression did very little to improve the fit using the $T_j$ and $(110–Bj)$ terms. The viscosity was calculated as an exponential function involving the ratio $B_j/(116.8–B_j)$, with $B_j$ corrected for juice temperature. The addition of a $\delta T$ term did improve the fit by 0.5%, but its power was negative and it was preferred not to use a negative power on $\delta T$. 

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the expression, so that the new formula might better predict possible future conditions of vessel operation with lower VCC and low $\delta T$ values. The expression is shown in Equation 7.

$$ HTC = 0.00049 (110 - B)_{j}^{1.1616} T_{j}^{1.0808} \delta T^{0.266} \quad (7) $$

where the symbols are defined in Nomenclature.

Equation 7 is here put forward as the ‘Austyp08’ formula for conventional Robert evaporators.

The B&D formula (Equation 5), discounted to 85% to allow for the \sim 11 days operation, has a rather poorer fit to the combined data set, with an RMS deviation close to 32.6%.

On the other hand the old ‘Australian Typical’ formula (Equation 3) has a fit approaching that of Equation 6, with an RMS deviation of 22.6%.

**Check of Australian HTC formulae against factory data**

Figure 3 shows the plot of the ‘Austyp08’ formula HTC estimates against brix together with all the combined data points. The trendline (as an exponential equation) for the data points and the ‘Austyp08’ points are also shown.

![Figure 3](image)

Fig. 3—Fit of the ‘Austyp08’ formula (Equation 7) to the combined data set.

Figure 4 is similar to Figure 3 except that the B&D formula (Equation 5) estimates, discounted to 85% to allow for 1 to 2 weeks operation time, are compared with the combined data points.

It is seen that the fit of the ‘B&D 85%’ points is reasonable in the low brix region, but is quite poor in the high brix region.
Fig. 4—Fit of the B&D formula (Equation 5) to the combined data set.

It is seen that the ‘Austyp08’ formula fits the measured data much better in the high brix region than does the B&D formula. It also fits the overseas data plotted earlier in Figure 1 rather better. Therefore the ‘Austyp08’ formula, Equation 7, is preferred as a performance benchmark for conventional Robert evaporator vessels.

Modification of HTC formula for SRI radial flow Robert evaporators

The main features of these SRI designs, as described by Wright et al. (2003), are the radial flow of heating vapour and condensate from the outer diameter to the inner tube of the calandria, and the approximate plug flow of juice from its entry around the periphery to its overflow to a central inner tube. These features give improved heat transfer conditions on the steam side and a wider level range for optimum heat transfer. The more defined concentration gradient on the juice side allows a lower average juice viscosity and improved HTC, especially where a large concentration step takes place across the vessel. Therefore there is a case for the use of the average concentration rather than the exit concentration with the new design vessels. However, this will have to be tested further, especially on final stage vessels, before such a change can be incorporated.

From the available data it is suggested here that for SRI–RF designs the exit brix value be retained for the time being in the ‘Austyp08’ expression (Equation 7), but that the constant be increased by a factor of 1.3 to 0.00064.

Conclusions

Correlations for heat transfer performance in sugar factory juice evaporators are important for benchmarking and prediction of evaporator set performance.

When used on the available data from Australian evaporator sets the correlation recently proposed by Broadfoot and Dunn (2007) was found to give a poorer fit than did the
‘Australian typical’ formula, particularly so where the conditions involve high juice concentration. A modified correlation is derived which gives a better fit to the extended factory data for conventional Robert Australian evaporator vessels, and yet incorporating a positive effect of $\delta T$ as found for the Broadfoot and Dunn pilot evaporator experiments. It is argued that this ‘Austyp08’ formula should be used in the future for conventional Robert evaporator vessels. The ‘Austyp08’ formula takes the form:

$$HTC = 0.00049 \ (110 - B_j)^{1.1616} T_j^{1.0808} \delta T^{0.266}$$

where the symbols are defined in Nomenclature.

After an examination of the available data on the performance of SRI radial flow design Robert evaporators, it is obvious that more investigations are required on the radial flow designs, especially when they are operating at low $\delta T$ values and as final stage vessels. At this stage it is recommended that for SRI–RF designs the exit brix, $B_j$, be retained in the ‘Austyp08’ expression given in Equation 7, with the constant be increased by 30%. Thus the tentative formula for radial flow Robert evaporators becomes:

$$HTC = 0.00064 \ (110 - B_j)^{1.1616} T_j^{1.0808} \delta T^{0.266}$$

**Nomenclature**

- $a$ the % reduction in HTC per hour of operation since the last cleaning.
- $B_j$ The concentration (brix) of the juice leaving the vessel. %
- $B_{jav}$ Average of the brix of the juice entering and leaving the vessel, %
- $B&D$ The HTC correlation proposed by Broadfoot and Dunn (2007), Equation.2
- $HTC$ Heat transfer coefficient, kW/(m$^2$ K)
- $HTR$ Ratio of the actual HTC to a value predicted from the operating conditions
- $T_s$ Condensing temperature of the heating steam/vapour in the calandria, °C
- $T_v$ Condensing temperature of vapour in the headspace of the calandria, °C
- $T_{bpe}$ Boiling point elevation of the juice at exit conditions, K
- $T_j$ Juice temperature (estimated as $T_v + T_{bpe}$), °C
- $VCC$ Mass of vapour condensed per hour per unit heating surface, kg/(m$^2$ h)
- $\delta T$ Temperature difference between $T_s$ and $T_j$ (Method 2 system), K
- $\mu$ Estimated viscosity of the exit juice, Pa.s
- $\lambda_s$ Latent heat of condensation of the steam in the calandria, kJ/kg

**REFERENCES**


