SOME KEY PRINCIPLES FOR THE DESIGN OF ROBERT EVAPORATORS

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Abstract

The performance of an evaporator station cannot be attributed solely to the vessel design, but to the design of the station as a whole. This paper aims to summarise some key principles pertinent to the South African design of Robert evaporators, with a particular focus on feed distribution and piping design. The pressure drop associated with the piping design can alter the hydraulics within a system and result in preferential flow, which can prevent adequate distribution of juice, the venting of incondensible gases or the draining of condensate. Advances in down-take design have resulted in a move toward semi-sealed down-takes which, coupled with the installation of feed rings, allows for improved evaporator performance through increased recirculation. The presence of flash vapour in the feed stream, as a result of the decreasing pressure profile, assists in circulation of juice through the tubes and thus increases the heat transfer coefficient. However, excess vapour in the feed stream can result in violent eruptions, spouting and entrainment, all of which can be reduced with the installation of a partial flash tank. Conflicting research regarding the optimum distribution of feed and flash vapour has resulted in the development of a feed distribution model to quantify the brix effect. A trade-off appears to exist between optimum juice distribution and optimum flash distribution, where an even distribution of the flash vapour is considered to be most advantageous.

Keywords: Robert evaporator, incondensible gases, feed ring, down-take

Introduction

The evaporator station is an integrated system in which the poor performance of one vessel disrupts the operation of the station as a whole. The driving force for heat transfer lies in the pressure profile across the station, fixed by the exhaust steam pressure and final effect vacuum and, as a result, allows juice to flow from one effect to another with no mechanical assistance. The intermediate evaporator conditions are thus dependent on the effective heat transfer occurring in each vessel.

Irregularities in the brix and pressure profile of an evaporator station are indicative of suboptimal performance. Poor performance of an evaporator often results in scrutiny of the equipment design. However, the problem often lies with anomalies within the piping design. The pressure drop associated with the piping design can alter the hydraulics within a system to cause problems such as the drying of tubes, spouting, violent eruptions of juice, poor boiling and high juice levels, all of which are commonly experienced within the South African sugar industry.
This paper aims at highlighting some key principles pertinent to the South African design of Robert evaporators, as shown in Figure 1, with a particular focus on feed distribution and piping design.

![Figure 1. The Robert evaporator (Rein, 2007).](image)

### Juice flow

The distribution of evaporative heating surface area between effects is dictated by the evaporator capacity and vapour bleed requirements. Additional evaporator area in the first two effects allows for improved vapour 1 and vapour 2 bleed pressures, whereas additional area in the tail effects allows for increased evaporator capacity. The installation of multiple vessels in parallel may thus be required to achieve the necessary heat transfer area.

**Splitting of juice flow between parallel vessels**

The distribution of steam to multiple evaporators in parallel can be carried out in the piping network as the specific volume of steam is large, and thus the hydraulic changes in the piping are normally not significant enough to interfere with the distribution. Liquids, however, possess a smaller specific volume and require careful piping design.

Juice is often traditionally split between parallel evaporators by a split in the juice piping, as shown in Figure 2. When two evaporators are fed from the same pipe, the only way to ensure an even split of feed is to design the piping such that the pressure drop, or the resistance to flow, in each branch is identical, as in Figure 2.
If the pressure drop is greater in one pipe, the evaporator with the higher pressure drop, and therefore higher resistance to flow, will receive less flow and this may result in poor boiling and possible drying out of the tubes, as shown in Figure 3.

A higher pressure drop on one branch of a feed line could be as a result of differences in the pipe length, diameter, elevation and number of fittings or bends. Uneven fouling or partial blockage of the feed lines could also lead to an unequal flow split between vessels. The distribution of juice to multiple evaporators within the same effect is thus best achieved by the installation of a splitter box.

**Splitter box design**

In order to ensure adequate distribution of juice using a splitter box, the design of the box itself, as well as the surrounding piping design, needs to be correct. In the Tongaat Hulett design, juice enters the bottom of the splitter box and overflows a number of carefully designed weirs to the juice outlet pipes. Each juice outlet pipe is sized for self-venting flow, to allow for adequate disengagement of vapour from the juice, thereafter reducing to the normal pipe size.
The splitter box vapour space is balanced to the source of the juice feed to ensure adequate drainage of juice. As a result of the pressure difference between the splitter box and the subsequent vessels, a liquid seal is required in the form of a U-leg or control valve on each outlet pipe to prevent the blow-through of juice and vapour from one vessel to the next.

Flow pattern
Evaporators can be designed to achieve either plug flow or perfect mixing. In an ideal plug flow evaporator, there would be no down-take and the juice above the top tube plate would have no return path to the bottom tube plate, resulting in a once-through flow of juice. With a single juice pass through the tubes, plug flow would be best, as the tubes are always in contact with juice at the lowest brix, resulting in the highest heat transfer coefficients (Pennisi et al., 2003). This flow pattern is associated with a Robert evaporator with no down-take or a sealed down-take design. Depending on the juice flow and concentration change required, once-through evaporators are prone to poor wetting rates and excessive fouling due to tube dry-out.

In order to achieve a ‘series’ of well-mixed cells to approximate plug flow in a Robert vessel, multiple down-takes can be positioned around the calandria, which will allow a significant improvement in the heat transfer coefficient when compared with a vessel with no down-take (Watson, 1986). The presence of multiple down-takes is believed to enable a concentration profile to be developed within the vessel, approximating plug flow (Steindl, 2003).

However, work done by Pennisi et al. (2003) has shown that the presence of a return path back to the bottom tube plate promotes extensive mixing of the recirculating juice with the incoming feed, resulting in the bottom calandria being in contact with juice with properties close to those of the exiting stream, thus destroying any attempt to simulate plug flow.

The South African design of Robert evaporators includes a large central down-take, allowing high brix juice to flow to below the tube plate and pass, once again, through the tubes. This flow pattern is associated with the semi-sealed down-take design, and the recirculation path improves wetting rates and fouling characteristics.

By good mixing between the high brix recirculating stream and the fresh feed, this design aims to achieve a lower average brix within the evaporator tubes, thereby improving heat transfer.

Feed ring design

Feed distribution pattern
The distribution of feed within an evaporator has a great impact on the flow pattern within the Robert evaporator (Pennisi et al., 2003) and is best achieved by means of a feed ring. Two popular designs are shown in Figure 4, the first design representing a peripheral distribution of feed, reported to be desirable in an evaporator aiming to achieve plug flow (Steindl, 2003), and the second design representing a more even distribution of feed, found in evaporators aiming to achieve perfect mixing.
Figure 4. Peripheral feed vs a more even feed distribution.

The feed to a Robert evaporator is comprised of both juice and flash vapour, and the distribution of both streams is to be optimised.

- Distribution of juice
  Robert evaporators within the South African industry commonly aim to achieve an even distribution of feed to optimise adequate mixing of the incoming juice and the recirculating stream. However, Rein (2007) recommends a peripheral distribution of juice feed and an even distribution of the flash vapour.

- Distribution of flash vapour
  An even distribution of flash vapour promotes the circulation of juice through the calandria tubes (Wright et al., 2003), thus increasing the juice velocity at the entrance to the tubes, which in turn results in an increase in the heat transfer coefficient, and increases the heat transfer rate significantly (Rein, 2007).

A trade-off appears to exist between optimum juice distribution and optimum flash distribution; however, only a single distribution pattern can be achieved for both streams. Whilst there appears to be consensus on the even distribution of flash vapour, the opposing recommendations for the distribution of juice result in confusion as to which design will offer optimum heat transfer. In an attempt to quantify the effects of progressively more even mixing on evaporator performance, a model was developed to investigate the brix effect of juice feed distribution.

- Feed distribution model
  A heat and mass balance model was developed to ascertain the brix effect of increasingly more even mixing between fresh feed and recirculating juice in a set of Robert evaporators. A more detailed explanation of the model can be seen in Appendix 1. Results of the mixing model, as shown in Table 1, indicate a very slight preference for uneven mixing (peripheral feeding) rather than even mixing.

<table>
<thead>
<tr>
<th>Percentage of more even mixing</th>
<th>Change in outlet brix %</th>
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<tbody>
<tr>
<td></td>
<td>2nd effect</td>
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<tr>
<td>25% mixing</td>
<td>10.38</td>
</tr>
<tr>
<td>50% mixing</td>
<td>10.17</td>
</tr>
<tr>
<td>75% mixing</td>
<td>10.09</td>
</tr>
<tr>
<td>100% mixing</td>
<td>10.07</td>
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</tbody>
</table>
As the extent of even mixing between the incoming feed and the recirculating juice is increased, the change in outlet brix, in all effects, is seen to decrease marginally, where the difference between the outlet brix of the 25% mixing and 100% mixing case varies between 0.06 and 0.31 units within the different evaporator effects.

It must be noted that the full effect of the flash vapour could not be quantified as the increase in juice velocity associated with flashing, and thus the subsequent effect on the heat transfer coefficient, has been ignored. This effect is considered to be substantial.

Seeing that the brix benefit associated with the peripheral feed arrangement is marginal, an even distribution of the flash vapour is recommended as being more likely to maximise heat transfer, and an even distribution of feed is therefore recommended.

**Feed ring riser pipe orifice plates**

Juice typically enters a feed ring in a Y-like connection, splitting half the flow to each side of the ring. A feed ring is often designed to provide an even distribution of juice across the evaporator via a series of nozzles with installed orifice plates. The juice flow within a pipe is inversely related to the pressure drop and, as a result, the juice flow will take the path of least resistance.

Considering the flow to two nozzles on one side of a feed ring, as shown in Figure 5, and assuming the pressure drop associated with the two flow paths to be 0.1 and 0.2 kPa, the pressure drop associated with the first nozzle is 50% of that of the second nozzle, resulting in the first nozzle receiving two-thirds or 66.7% of the flow, whereas the second nozzle would receive 33.3% of the flow.

![Figure 5. Effect of orifice plates on feed distribution.](image)

With the installation of orifice plates in the riser pipes of a feed ring, an external pressure drop is imposed on the system, allowing a more even distribution of flow. Assuming a pressure drop of 5 kPa across the orifice, the pressure drops associated with the two flow paths are now 5.1 and 5.2 kPa, resulting in a more even flow of juice. The pressure drop at the first and second nozzles is now almost identical, with the first nozzle receiving 50.5% of the flow and the second nozzle receiving 49.5% of the flow.

**Orientation of splash plates**

The installation of splash plates above the juice feed enables the flash vapour and jet of liquid to be dissipated in the outward direction, thus preventing the localised spouting of juice and damage to the bottom tube plate (Pennisi *et al.*, 2003). Often, little attention is paid to the orientation of the splash plate. The incorrect orientation of the splash plate, as shown in Figure 6, positioned parallel to the bottom saucer, could lead to the preferential escape of
flash vapour from one side of the plate causing localised spouting. The splash plate should be installed parallel to the bottom tube plate, with sufficient clearance from the tube plate to allow for the adequate distribution of juice and flash vapour.

![Incorrect splash plate angle](image1) ![Correct splash plate angle](image2)

**Figure 6. Splash plate orientation (adapted from Rein, 2007).**

**Partial flash tank**

As a result of the decreasing pressure profile across the evaporator station, a Robert evaporator accepts feed at a higher temperature and pressure which, upon entering the vessel, flashes to produce a mixture of juice and vapour. The flash vapour serves to promote circulation and increase the juice velocity at the bottom tube plate, thus increasing the rate of heat transfer (Rein, 2007).

While a small amount of flash vapour is desirable, anomalies within the evaporator station often result in abnormally large pressure drops existing between evaporator effects. By design, larger pressure drops are also commonly found between the last two effects (Muzzell et al., 2013). As a result, the excessive flashing of juice feed can result in spouting of the syrup in the final effect, thus leading to entrainment and sucrose losses (Muzzell et al., 2013).

The installation of a partial flash tank before the final effect allows excessive flash vapour to be removed and has been found to be successful at the Felixton Mill (Bindoff and Dlamini, 2013) and at the Hippo Valley Mill in Zimbabwe. Some flashing of the feed entering the final effect vessel is still allowed, in order to assist with heat transfer.

The critical parameter in the design of the partial flash tank is the actual pressure drop between effects, where a large drop could result in a flash tank of up to 3 m in diameter.

**Down-take design**

The rate of heat transfer within a Robert evaporator is influenced by the down-take design, where three basic designs exist, as shown in Figure 7. According to Hugot (1972), the diameter of the down-take itself should be between one-eighth to one-quarter of the vessel diameter, with the actual design value taking into consideration the evaporator effect, optimum vessel diameter and desired heating surface area.

The open down-take allows for 100% recirculation of juice. However, this design is associated with poor level control as well as bypassing of juice from the inlet directly to the outlet pipe. The sealed down-take eliminates the problem of bypassing. However, this arrangement allows no recirculation of juice and is also associated with poor level control and high levels of fouling due to the drying out of tubes. A compromise between the two designs, viz. a semi-sealed down-take design with an off-take cone, is believed to achieve optimum heat transfer (Rein, 2007). Proper positioning of the off-take cone allows for the physical (overflow weir) control of the juice level and no level control valves are required.

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1Personal Communication, Edmore Dzirove, Tongaat Hulett Sugar, Hippo Valley Sugar Mill.
Minimum wetting rate
Adequate recirculation of juice allows an appropriate wetting rate to be achieved, resulting in higher juice velocities at the tube inlet, thus increasing the heat transfer coefficient and evaporation rate.

In 1999, the practice within Tongaat Hulett was to design the off-take cone for 50% recirculation\textsuperscript{2}. Rein (2007) later suggested that the off-take cone should occupy one-third of the down-take area, allowing for 66.7% recirculation. The most recent practice at Tongaat Hulett has been to design the off-take cone for a minimum recirculation rate of at least 1.2 kg/min/tube\textsuperscript{3} (Shah and Peacock, 2013).

Whilst these figures have provided useful guidelines for down-take design, an equation for the minimum wetting rate has recently been developed by Shah and Peacock (2013) and is dependent on the temperature driving force between the heating medium and juice. The wetting rate can be calculated according to equation 1, with a minimum wetting rate of at least 1.2 kg/min/tube being recommended (Shah and Peacock, 2013).

\begin{equation}
WR \ (\text{Wetting Rate}) = 0.0000979(\Delta T)^2 - 0.002966\Delta T + 0.0572 \tag{Eq 1}
\end{equation}

where \( WR \) = wetting rate (kg/s/m wetted perimeter)  
\( \Delta T \) = temperature driving force between the steam and juice (°C).

The wetting rate expressed in equation (1) can be multiplied by the tube perimeter to yield the wetting rate per tube. A more detailed calculation based on heat transfer theory can be carried out in terms of the Grashof and Reynolds numbers as shown by Shah and Peacock (2013).

\textsuperscript{2}Meadows DM (1999) Internal Memorandum, Tongaat Hulett Sugar, Durban, South Africa.
\textsuperscript{3}The recirculation rate is actually accommodated by sizing the area for recirculating flow around the outside of the off-take cone.
**Minimum liquid level**

A minimum liquid level is required to ensure the onset of recirculation and an optimum heat transfer rate. The calculation of the minimum liquid level aids in the vertical positioning of the down-take cone and can be calculated according to equation 2.

\[ h_{\text{minimum}} = -0.0149 \Delta T + 0.5816 \quad \text{Eq 2} \]

where \( h_{\text{minimum}} = \) liquid level as a fraction of the total tube height
\( \Delta T = \) temperature driving force between the steam and juice (°C).

A minimum liquid level of at least 33% of the tube height is recommended in order to err on the side of caution, as a higher liquid level is associated with a more gentle drop-off in the heat transfer coefficient, as compared with a lower liquid level (Shah and Peacock, 2013). In addition, a higher liquid level prevents the excessive fouling that could occur at low levels due to tube dry-out.

**Incondensible gases**

Incondensible gases are introduced into the steam system mainly upon boiling of the sugarcane juice, when gases dissolved in the juice are released, and are thus more of a problem from the second effect onwards (Hugot, 1972). Incondensible gases may also enter the exhaust steam range in the form of dissolved air within the desuperheater water.

The accumulation of incondensible gases through the flow path reduces the heat transfer coefficient primarily due to a reduction in the rate of mass transfer. The incondensible gases diffusing away from the condensing layer hinder the diffusion of steam to the condensing tube surface (Lienhard and Lienhard, 2008) and can therefore be considered as an additional resistance to mass transfer (Standiford, 1979).

To a smaller extent, the accumulation of incondensible gases result in a reduction of the steam condensing temperature due to the partial pressure effect.

Equation 3, adapted from Dalton’s law of partial pressures, illustrates that the total pressure of a system is equal to the sum of the partial pressure of the gas species present in the system. Therefore, as the quantity of incondensible gases increases, their partial pressure is increased, whereas the system pressure in the calandria remains constant. As a result, the effective pressure of the vapour, and temperature, is reduced, lowering the temperature driving force for heat transfer.

\[ P_{\text{total}} = P_{\text{vapour}} + P_{\text{incondensible gases}} \quad \text{Eq 3} \]

A calandria should be designed with a dedicated vapour flow path, preferably tapered to prevent the accumulation of air pockets, with incondensible gas vents located at the end of the flow path (Standiford, 1979).

Incondensible gases have the ability to drastically reduce the heat transfer coefficient in an evaporator. Allan (1979) reported a 13% reduction in the heat transfer rate, with 0.5% of air
in a calandria (by mass). Lienhard and Lienhard (2008) reported a 20% reduction in the heat transfer coefficient in the presence of 5% air by mass.

**Incondensible gas piping design**

For Robert evaporators operating under sub-atmospheric conditions, the incondensible gases from each effect are usually piped to a larger common manifold, which is then routed to the final effect vapour space.

The incondensible gases present in the steam are often classified as either light or heavy, and vents are thus required at both the top and bottom tube plates to ensure adequate removal. The incondensible gas take-offs are designed with orifices that allow the correct quantity of gases to flow freely to the common manifold. In practice, these pipes are often incorrectly combined with only a single, smaller pipe routed to the common manifold. This arrangement can sometimes interfere with effective venting, and it is therefore recommended that the light and heavy gas take-offs be routed separately to the common manifold (Rein, 2007), as shown in Figure 8.

![Figure 8. Incondensible gas piping (adapted from Rein, 2007).](image)

The risk associated with the combining of the incondensible gas vents increases with time, as the orifice holes erode and exert less control over the gas flow. Depending on the piping layout, the flow of one of the gas streams then has the potential to prevent adequate venting of the other gas stream, resulting in poor evaporator performance with time, as reported by Muzzell *et al.* (2013).

The vent holes on the heavy incondensible gas pipe inside the calandria should be facing in an upward direction, as a level of condensate could easily develop within the calandria and prevent adequate venting.
Condensate removal

In order to maximise the area available for heat transfer within the calandria, adequate drainage of condensate is required. Condensate is often cascaded down to flash pots at lower pressure, allowing the flashing of condensate and the recovery of the flash vapour for improved energy efficiency. The blow-through of steam to a pot of lower pressure can be prevented with the installation of a steam trap, U-leg or a level-controlled condensate pot.

Condensate pot design

The elevation of the condensate pot relative to the evaporator is critical as it should be situated below the bottom tube plate, with the condensate off-take positioned as near the bottom tube plate as possible.

It is common for condensates from multiple vessels within an effect to drain to a common pot. Condensate pots accepting multiple streams risk hold-up within one of the evaporator calandrias as a result of slight differences in calandria pressure. Banging can also occur due to possible temperature differences, causing flashing and subsequent condensation. Condensate piping should be carefully engineered to ensure adequate drainage from all vessels under all operating conditions.

Condensate piping design

Condensate piping should be carefully designed for two-phase flow. As the condensate flows and the pressure is reduced due to frictional losses, the condensate begins to flash in the line. The velocity of steam is much greater than that of the liquid. As a result, the flash vapour accelerates the liquid and causes a large pressure drop in the piping, which can result in poor condensate drainage, as a greater amount of head is required to overcome the larger pressure drop.

The presence of inverted U-legs should be avoided within condensate piping in particular, as vapour collects in the inverted U, causing a vapour lock and preventing adequate drainage of condensate.

All condensate pipes should drain separately into the condensate pot. The mixing of condensates within the piping network should be avoided, as this can lead to hammer.

Conclusion

Research has shown that the inclusion of a down-take within the calandria of a Robert evaporator greatly increases the rate of heat transfer and promotes extensive mixing, therefore hindering the achievement of true plug flow. The South African design of Robert evaporators includes a large central down-take based on the principle of perfect mixing to achieve a minimum juice brix within the evaporator tube.

The principles that are key to Robert evaporator design relate to heat transfer theory and are based on maximising the available heat transfer area and heat transfer coefficient. The heat transfer area is maximised through adequate removal of condensate and incondensible gases. The heat transfer area is most effectively used by evenly distributing the juice flow to parallel vessels with the aid of a splitter box. The heat transfer coefficient is then maximised by increased recirculation.
Optimum recirculation can be achieved with the incorporation of a semi-sealed down-take design, together with the installation of a feed ring, fitted with carefully designed orifice plates and splash plates. An even distribution of feed and flash vapour is recommended, where excessive flash vapour can be removed with the installation of a partial flash tank, preventing spouting and entrainment. Employing these key design checks on an evaporator can allow for improved evaporator performance not only for one vessel, but for the station as a whole.

Acknowledgements

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APPENDIX 1:
FEED DISTRIBUTION MODEL

A heat and mass balance model was developed to ascertain the effect of increasingly more even mixing between fresh feed and recirculating juice in an evaporator. Input data from a factory steam balance model was used to provide the following inputs to the model:

- Area of evaporator
- Number of tubes
- Incoming juice brix
- Incoming juice flow
- Temperature of steam in
- Vessel pressure and temperature
- Feed pressure and temperature.

Model description

The evaporator model assumed a partitioning of the evaporator, as shown in Figure 9, where a certain percentage of the vessel would receive fresh juice feed and the rest would receive recirculating juice only.

The optimum wetting rate was calculated from equation 2, and the total juice flow was appropriately divided. In both parts of the vessel, this wetting rate is to be maintained. Thus, should the feed flow alone not satisfy the wetting rate in the first portion of the vessel, the juice flow would be ‘topped up’ with juice from the recirculating stream.

Feeding of 25, 50, 75 and 100% of the evaporator area with fresh feed was investigated, representing progressively more even mixing between the feed and recirculating juice.
Calculation method

**Flashing of evaporator feed and promotion of circulation**
Assuming the vessel pressure to be fixed, the incoming feed will flash to the boiling point of the juice at the base of the evaporator, which is at the highest temperature as a result of the boiling point elevation (BPE) due to brix and hydrostatic head.

As the juice rises toward the bottom tube plate, the BPE due to hydrostatic head is reduced, allowing further flashing of juice, after which the juice comes into contact with the calandria tubes and the further effects of flashing can be ignored for the purpose of this model.

The flash vapour formed upon entering the vessel accelerates the juice at the bottom tube plate and results in an increase in the heat transfer coefficient. However, in order to simplify the model, this effect on the juice velocity was ignored and only the brix effect of feed distribution on heat transfer was considered.

**Juice flow through the tubes**
Juice flow within an evaporator tube is preferentially in the upward direction (Pennisi et al., 2003). As the juice rises through the tubes, the transfer of heat energy from the steam to the juice causes evaporation. The temperature difference between these two streams forms the driving force for heat transfer, as shown in equation 4 below.

\[
Q = U \cdot A \left( T_{\text{steam}} - T_{\text{juice}} \right)
\]

where \( A = \) area available for heat transfer (m\(^2\)).

In a mixed-flow vessel such as a Robert evaporator, \( T_{\text{juice}} \) is often taken to be the juice boiling point at its exit concentration. Alternatively, average conditions within the tubes can be used.

The heat transfer coefficient, \( U \), was calculated according to the correlation of van der Poel (1998), recommended as a correlation that best represents experimental heat transfer data gathered between 1964 and 1981 (Peacock, 2007),

\[
U = \frac{k \cdot T_{bp}}{\text{brix}}
\]

where \( T_{bp} = \) boiling point of the juice within the evaporator vessel
\( k = 0.573518 \), a constant adjusted by Peacock (2007) to provide a better fit to the data.

The average brix was used in the calculation of the heat transfer coefficient, representative of the product within the tubes.

**Absorption of flash vapour by recirculating stream**
The mixture of juice and vapour bubbles rises up the tubes to the top of the tube plate, after which a portion of the juice will be recycled through the downtake to the bottom tube plate to provide recirculation. The remaining juice exiting via the off-take cone is fed forward to the next effect.
At the top of the tube plate, there is a BPE due to brix only with no hydrostatic head contribution, resulting in a juice temperature lower than that of the juice below the bottom tube plate.

Upon contact between the fresh feed and the cooler recirculating juice, now below its saturation temperature due to its submergence below the liquid level, some of the flash vapour is able to condense, giving off latent heat to the recirculating stream until the saturation temperature is reached. As a result, the enthalpy of the combined juice feed increases, contributing significantly to the heat transfer process.

The latent heat was assumed to be absorbed only by the recirculating juice mixing with the feed stream. As the percentage of mixing increased, the quantity of recirculating juice in contact with the fresh feed increased; however, the absorption was limited by the saturation point of the recirculating stream. The vapour not able to be condensed by the recirculating stream was assumed to exit via the evaporator tubes, as shown in figure 10, resulting in a slight increase in the juice brix.

A summary of the model can be seen in Figure 11.
Figure 11. Evaporator model developed to investigate the brix effect of feed distribution.