REFEREED PAPER

# DESIGN, INSTALLATION AND OPERATION OF PARTIAL FLASH TANKS IN THE EVAPORATOR STATION AT HIPPO VALLEY SUGAR MILL

DZIROVE E[1], DAVID W[2,3], MUGADHI A[1] AND SHEAHAN W[2,4]

[1]*Tongaat Hulett Sugar, PO Box 1, Chiredzi. Zimbabwe*
[2]*previously at Tongaat Hulett Sugar, Private Bag 3, Glenashley, 4022, South Africa*
[3]*presently at Fluor Igoda Projects, Private Bag X24, Umhlanga Rocks, 4320, South Africa*
[4]*presently at Fives Fletcher, 33 Brunel Parkway, Pride Park, Derby, Derbyshire, DE24 8HR, UK*
*edzirove@hippo.co.zw warren.david.email@gmail.com amugadhi@hippo.co.zw*
*sheahanwarren@gmail.com*

## Abstract

Sucrose losses in a sugar factory as a result of entrainment into the vapour phase during evaporation or boiling can be a significant source of undetermined loss. This risk is higher in the final evaporator vessel, where sucrose can be entrained with the vapour into the condenser and is subsequently lost to the injection water. This is normally observed by a spouting or 'fountain effect' of the boiling syrup or liquid. The effect is created by a large pressure drop, or pressure differences between the final and preceding effects, which then causes significant flashing of the liquid entering the final effect vessel. The flashing effect has been a problem at Hippo Valley Estates sugar mill and has resulted in significant carry-over of syrup and losses to injection water. In order to reduce the flashing effect, partial flash tanks were designed and installed into each train of the evaporator station. This paper discusses the design, installation and operation of the flash tanks, and highlights the immediate benefits that the installation brought to the factory.

*Keywords:* evaporator, flash, entrainment, losses, Robert, spouting

## Background

Hippo Valley Estates (HV) sugar mill operates two identical sets of evaporators in a quadruple effect set-up. Each set has two identical parallel Kestners as the first effect, with the last three effects being Robert vessels. The two sets are designed to handle a total crush rate of 440 tons cane per hour (tch). The evaporator sets are designed to bleed vapour 1 (V1) for use in batch pan boiling, secondary mixed juice heating and scalding juice heaters. Vapour 2 (V2) is bled for use in primary mixed juice heating, B-massecuite boiling in the continuous vertical crystallisation tower (VKT) and C-station horizontal continuous pan. No vapour bleeding is done from the third effects.

Evaporators are considered to be the heart of a sugar factory, as their performance has a huge bearing on overall plant performance. The evaporator needs to generate adequate vapour to drive the boiling house and, in the process, concentrate syrup to 65-70 brix so as to minimise evaporation load on the pans. The evaporator has to return condensate for boiler feed water and supply adequate volumes of condensate for process use. HV evaporator performance has not been satisfactory over the years, due to a number of problems. A series of modifications and improvements were implemented over the years and this improved performance in various facets (Muzzell *et al*., 2013). The main challenge which remained hidden was loss of sucrose through entrainment, a problem which contributed significantly to higher levels of undetermined losses (UDL). Several modifications were done over the years to reduce entrainment. This included installation of save-alls on evaporators as well as vertical chevron plates (VCPs), or louvres, as entrainment separators. At some stage in the past, HV operated with three sets of entrainment separation systems in the evaporator fourth effects. The entrainment systems was comprised of VCPs as first stage separators, standard top hats (flow reversal separator) as the secondary entrainment separators and the Still-man save-alls or catch-alls as the last stage separators (Figure 1).

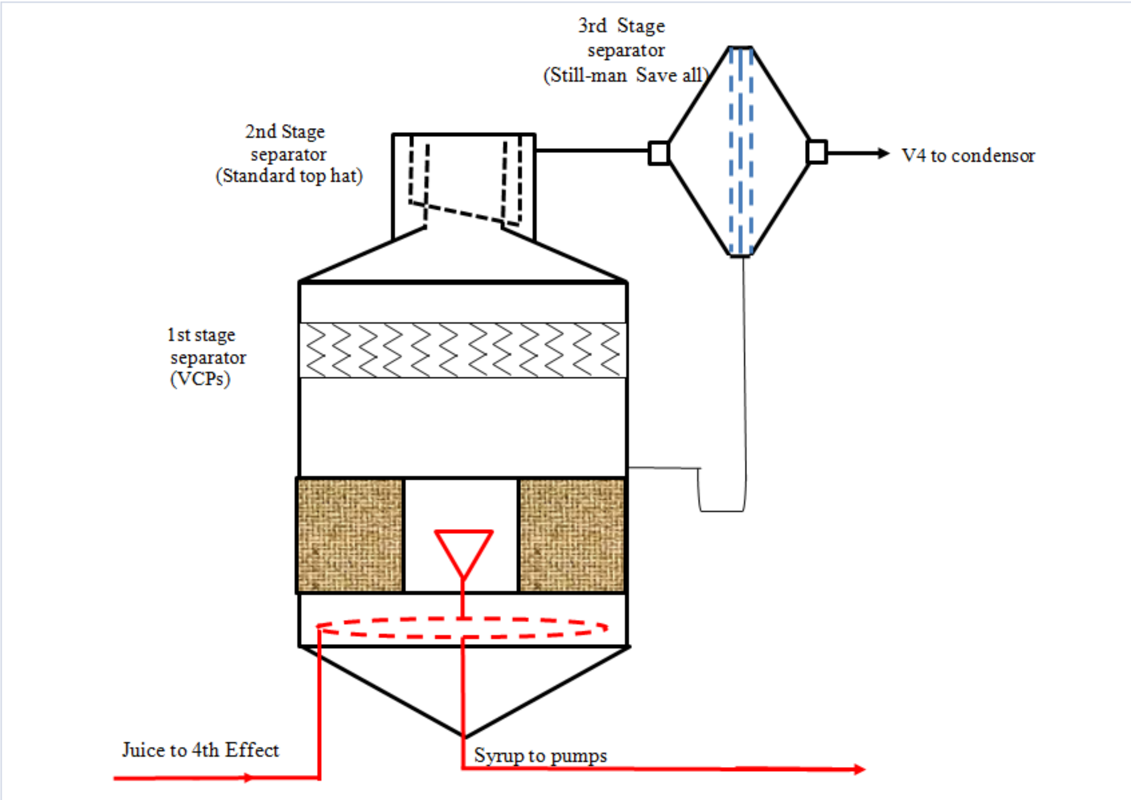



**Figure 1. Historical set-up of Hippo Valley sugar mill fourth effect evaporator vessels.**

Due to frequent blockage problems incurred on the VCPs, these were decommissioned and the top hats and the Still-man save-alls remained. Still-man save-alls were removed in mid-2011 as a result of vacuum challenges (aged equipment, leaks, and their high replacement cost with stainless steel material). It was also considered that this was old technology, long phased out in many factories. From mid-2011, only one entrainment separation device, i.e. the flow reversal separator (standard top hat), remained on the evaporator fourth effects. Undetermined loss still remained a challenge and investigations continued. One main issue of concern was persistent high levels of entrainment in cooling water (>500 ppm) and in factory effluent, hence focus remained on identifying the source of this entrainment. A decision was taken to enhance visibility of entrainment return on pans and evaporator vessels through installation of enlarged 150 mm sight glasses on entrainment return pipes. It was at this point that the magnitude of juice feed flash-induced entrainment was observed and modifications were pursued to resolve this.

## The Evaporator Station

*Evaporator station description*

Figure 2 shows the plan view of the evaporator set-up at HV. The two A and B sets are identical in design and each is designed to handle 220 tch.

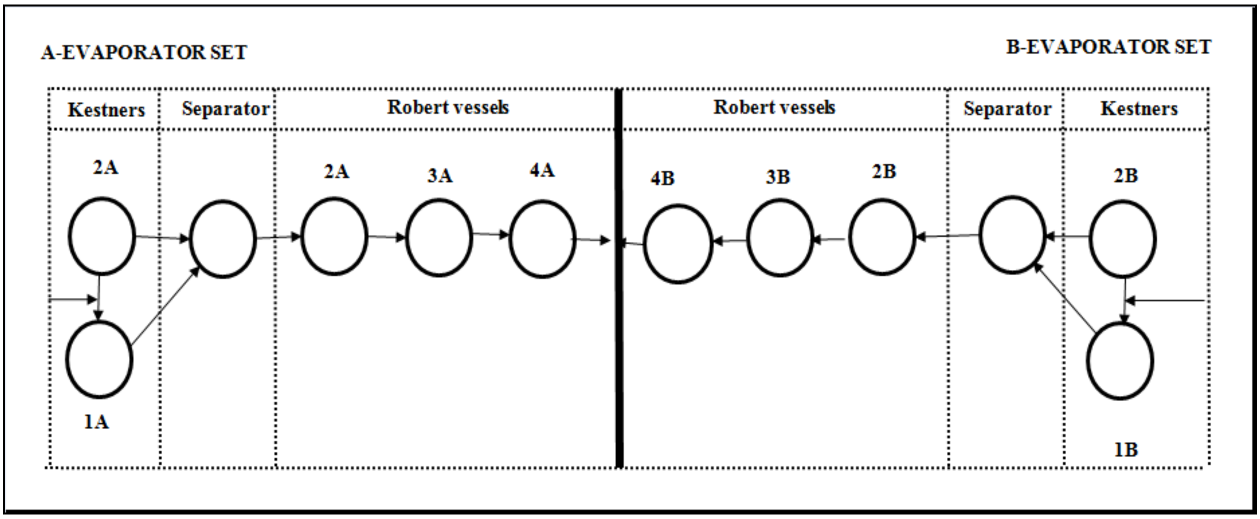



**Figure 2. Hippo Valley sugar mill evaporator station plan view**.

Table 1 shows the design parameters for the quadruple effect.

**Table 1. Design parameters for the evaporators, and 2012 average values.**

| | Design | | Actual (2012) | |
|---|---|---|---|---|
| | Pressure (kPa (g)) | Temperature (°C) | Pressure (kPa (g)) | Temperature (°C) |
| Exhaust steam | 100 | 120.2 | 117 | 123* |
| Vapour 1 (V1) | 50 | 111 | 48 | 111 |
| Vapour 2 (V2) | 20 | 105 | 20 | 104 |
| Vapour 3 (V3) | -20 | 93.5 | -19 | 94 |
| Vapour 4 (V4) | -85 | 54 | -77 | 63 |

*Temperature measurement is located near evaporators. Exhaust steam close to saturation conditions.

Each evaporator set is designed to handle 220 tch for a combined capacity of 440 tch. Clear juice is heated in shell and tube juice heaters (two heaters per set) and fed to the parallel set of Kestners for each set. The Kestners are of the rising film design with external feed rings. The Kestner separators have vertical chevron separators (VCPs) installed as entrainment separation devices. Downstream, the concentrated juice is fed into Robert vessels for the 2nd, 3rd and 4th effects, both sets with standard top hats as entrainment separators. The Robert vessels are basically the same in design, varying only in tube length and diameter, the 2nd effect of which is 38.1 mm o/d and the rest 50.8 mm o/d. The Robert vessels have internal circular feed rings with square shaped slots, the number of slots decreasing downstream. The 4th effect vessel has a standard top hat entrainment separator (flow reversal separator) and a rain type condenser.

Typical design and operational pressure profiles achieved show the highest pressure drop to be between 3rd and 4th effects and this causes high levels of flash as juice is fed into the fourth effects. A quintuple effect evaporator may reduce the pressure differential; however, it would not be a cost effective solution.

*Evaporator modifications during the 2012 season*

The 2012 crushing season start-up, characterised by very high levels of entrainment and factory undetermined loss (UDL), was abnormal. Investigations were carried out and several operational interventions were made, with limited success, to halt this UDL, evidently caused by entrainment. Vessel inspections were made. Anomalies such as blockages of entrainment return lines were cleared, and improperly fitted feed rings were adjusted. Splash plates were installed in the 4th effects at a disengagement height of more than 4 m above the top tube sheet, but entrainment levels remained abnormally high. Entrainment sight glasses on 4th effect entrainment return pipes were enlarged, so as to fully view entrainment returns during the boiling process. It was observed that the amount of entrainment return from evaporators far exceeded the expected returns, hence more observations were made to determine the root cause. An abnormal boiling pattern was observed in the 4th effects, indicating high levels of spouting, entraining concentrated syrup to the evaporator top hat and causing significant losses of syrup to the cooling water. Experiments were conducted to check changes in boiling patterns when feed to the 4th effect was taken off, and it was observed that the spouting effect significantly reduced. A decision to temporarily feed juice through the top of the calandria was done and observations were made. The modifications resulted in a significant reduction in the spouting induced-entrainment, as evidenced by returns from the top hat. A decision was taken to run the evaporator feeding juice into the 4th effect above the calandria for the remainder of the 2012 season. Figures 3 and 4 show the evaporation station set-up pre- and post-4th effect modifications to manage entrainment resulting from the spouting effects induced by juice flashing (Muzzell *et al*., 2013).

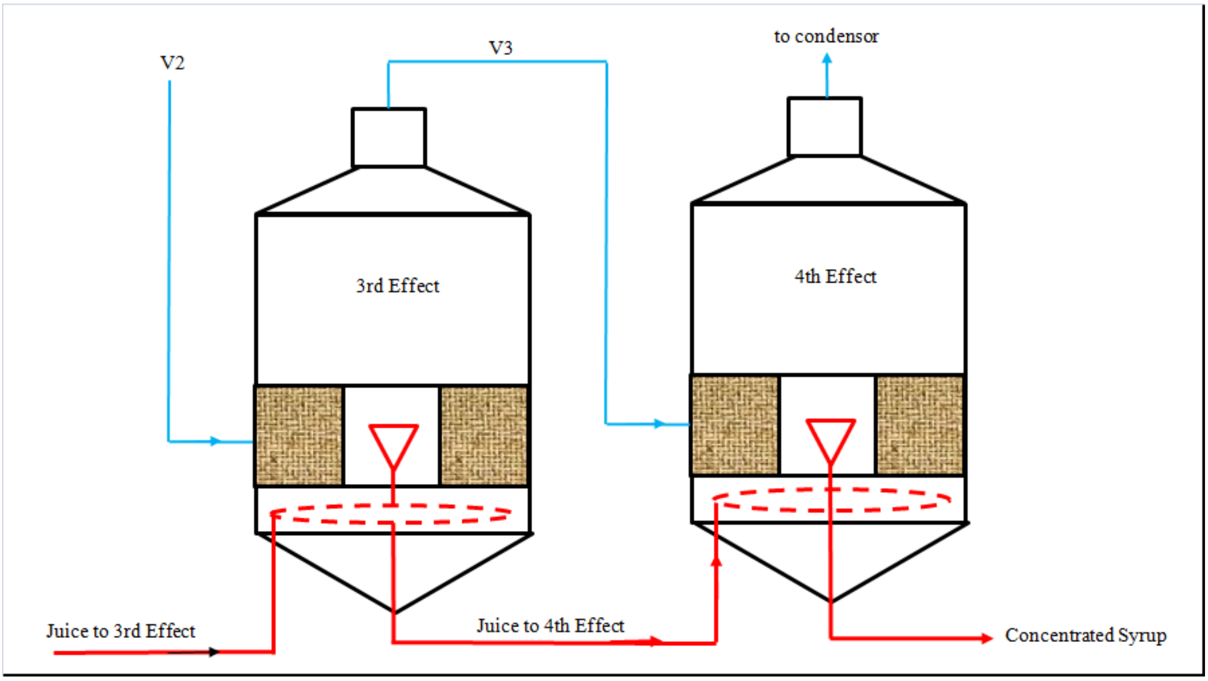



**Figure 3. Set-up showing juice feed into 4th effect before 2012 modifications.**

The modifications entailed juice being fed into the 4th effect just above the top tube plate of the calandria through a 150 mm diameter pipe which was carefully terminated inside the vessel to avoid juice by-passing directly to the centre of the vessel (down-take position). The pipe section protruding into the 4th effect was slotted at the bottom and blanked at its end to ensure that juice was directed downwards as it was fed into the 4th effect.

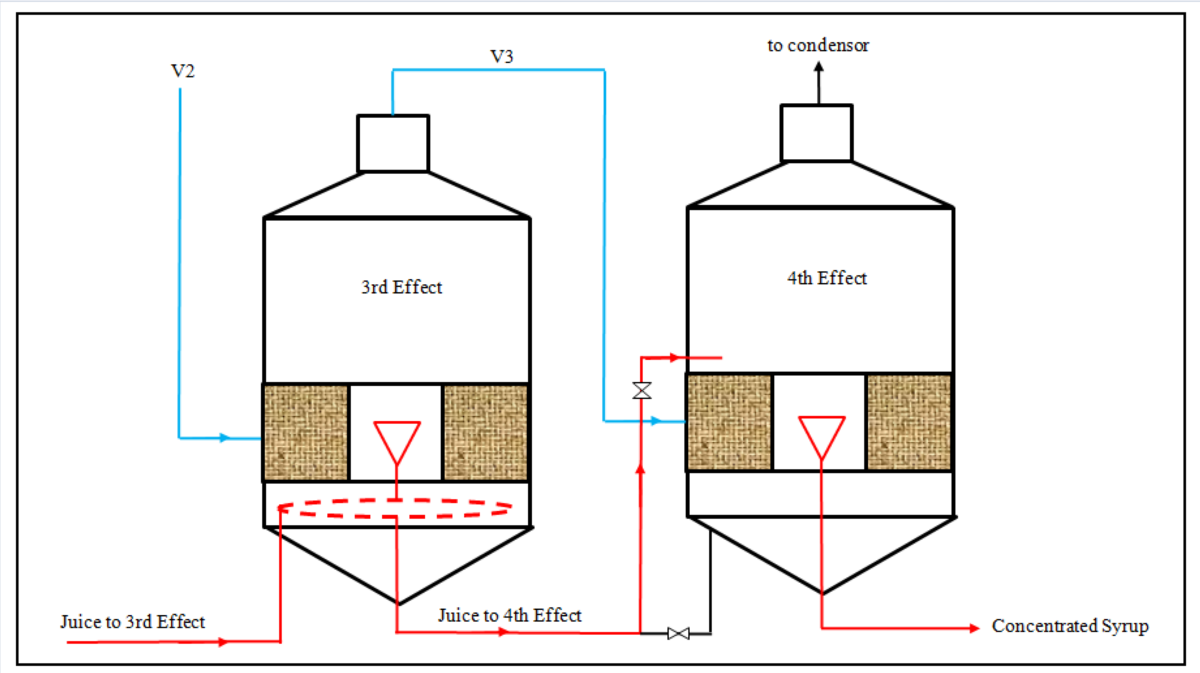



**Figure 4. Juice feed into 4th effects after modifications (above top tube plate of calandria).**

The 2012 modifications successfully reduced the high losses incurred through entrainment, and UDL was significantly reduced. However, performance of the evaporators was not satisfactory as both sets could not brix up to expected levels for the season. This resulted in limited use of imbibition, which negatively affected efforts to optimise extraction. An increase in the steam pressure required to drive evaporators was also noted, as the evaporators consistently failed to brix up to required levels at the usual recommended steam pressures. The factory was slowed down on several occasions due to high syrup levels. The sub-optimal evaporator performance at throughputs below design capacity brought the need to restore normal operation (4th effect feeding through the feed rings) whilst eliminating entrainment-induced flashing through installation of an intermediate flash tank.

*Partial flash tank installation during 2013 off-crop*

The partial flash tank installation was done as a long-term solution during the 2013 off-crop. The schematic layout is shown in Figure 5. Each set of evaporators had a partial flash tank installed.

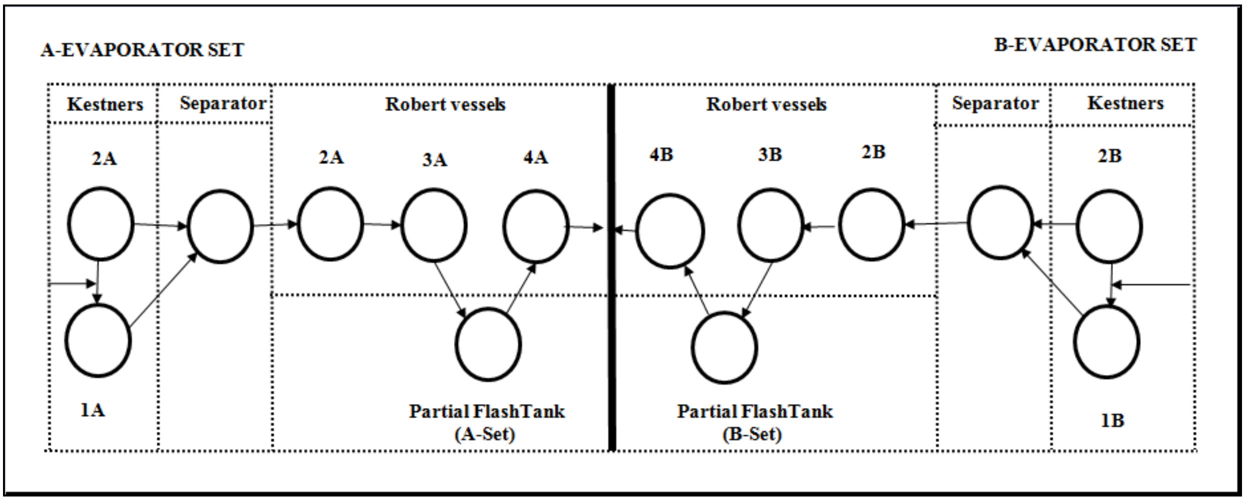



**Figure 5. Modified plan view of set-up showing positioning of partial flash tanks.**

## Design

*Theory*

Partial flash vessels have been used with some success within Tongaat Hulett Sugar (Bindoff and Dlamini, 2013). In the case of Felixton sugar mill, the installation was primarily aimed at eliminating banging noises experienced in the final effect vessels. In the case of HV sugar mill, the primary aim was to eliminate spouting in the final effect vessels. The overall driving force in an evaporator station is set, in addition to other factors, by the differential pressure of the exhaust steam entering the 1st effect and the vacuum in the final effect. This differential pressure is profiled along the train and tends to be more significantly stepped towards the end of a 4th effect evaporator station, as is the case at HV.

Flash vapour, generated from the juice feed stream entering an evaporator vessel feed system, is important in ensuring adequate circulation and heat transfer within the vessel. Too much flash vapour can disrupt juice circulation and heat transfer, and entrain juice into the vapour and subsequent condensate system. This is particularly problematic on the final effect vessel, as the combination of high pressure drop and low absolute operating pressure generate a high volumetric flow rate of flash vapour. In addition, sucrose lost to the condenser tail-pipe cannot be recycled into the process via diffuser imbibition.

*Partial flash system*

The solution was to reduce flash vapour by exposing the juice feed stream to conditions at or near the operating pressure of the final effect. This was achieved by installing a partial flash tank between the final effect and the preceding effect, as can be seen in Figure 6.

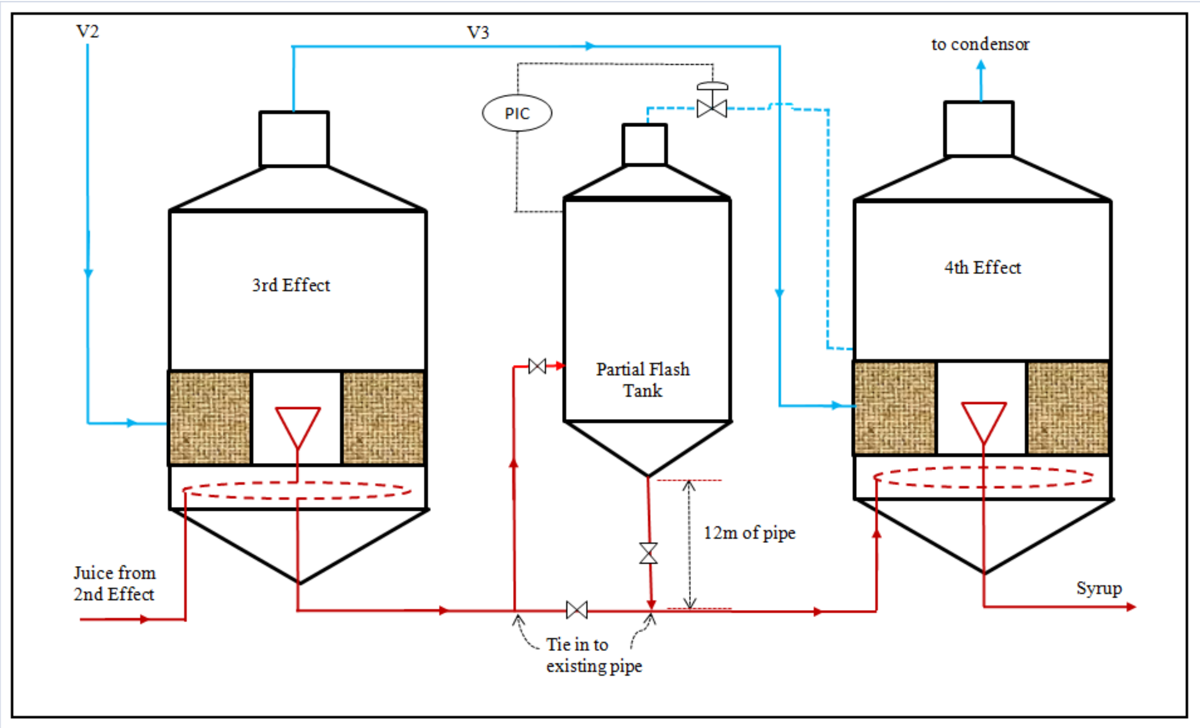



**Figure 6. Schematic presentation of the partial flash tank installation.**

In a quadruple effect evaporator station, the juice leaves the 3rd effect and is routed to enter the flash vessel. The vapour line of the flash vessel is connected to the vapour space of the final effect, upstream of the entrainment separator. The vapour control valve on this line allows the pressure in the flash vessel to be adjusted. These conditions cause the juice stream entering the vessel to flash, with the vapour moving upward and out through the vapour outlet, and the remaining juice moving downward and out through the juice outlet. The juice moves forward to the final effect, entering the vessel through the juice feed system. The vapour moves across the pressure control valve and enters the final effect in the vapour space, then moves on to the condenser. A line was installed to allow the system to be bypassed, and also for transitioning during start-up or shut-down. A pressure transmitter was installed on the vessel to allow the pressure to be measured and controlled via the pressure control valve.

*Design approach*

The intent of the design was to install the above system into the evaporator station at HV. A key consideration of this type of design is constructability, operability and maintenance of the system. The system was fabricated and installed in-house, with each evaporator train having its own partial flash tank installed. The layout of the station allowed the system to be designed ergonomically as the available space was sufficient.

The installation was designed for a future expanded HV throughput of 530 tch. This was to ensure that the installation would not limit the future expansion of the evaporator station and the mill.

Data was obtained from a combination of plant data and available in-house model output. Plant data included clear juice flow rates, operating pressure profiles, brix profiles and exhaust steam pressure. Output from the in-house Program for Evaporator Simulation and Testing (PEST) model (Hoekstra, 1981) provided juice and vapour flow rates between effects. The selection of the data depends on how conservative an approach is taken, the operating envelope of the installation and information from previous installations. For the installation at HV, a rather conservative approach was taken as less conservative approaches at other mills proved less effective. In addition, the mill requested a large operating envelope as the evaporator station was the bottleneck in the process.

*Hydraulic calculations*

The system required hydraulic calculations to be undertaken to size the necessary juice pipe lines and vapour control valve. A tie-in just before the 4th effect juice feed ring was made and this pipe-run was directed to the partial flash tank body tangentially. The juice return from the flash tank travelled along the same path and tied into the 4th effect feed ring. Hence two parallel pipe lengths of roughly 12 m were used. Elevations involved were roughly 4.5 m from the 3rd effect exit to partial flash tank feed.

A simple single phase calculation could not be undertaken in this system as the juice process fluid is near flash point. A more rigorous 2-phase approach was adhered to.

The significance of pressure drop with flashing systems is evident when a simple single phase approach is compared to the 2-phase approach. This vast difference in pressure drop is due to the boundary layer forces that are interacting between the less dense, accelerating vapour and the denser, slower liquid layer. The boundary layer is not of a laminar but rather turbulent regime, which causes unique interaction with the friction surface of the pipe walls.

The Lockhart and Martinelli semi-empirical correlation (Lockhart and Martinelli, 1949) was used to calculate the gas-liquid 2-phase pressure drop. Details of the calculation equations are presented in Appendix A. For the single phase pressure drop calculations, the Churchill equation (Churchill, 1977) was used for the friction factor $f$, and pressure drop for pipe fittings were incorporated using the 2-K method (Hooper, 1981). Details of the calculation equations are presented in Appendix B.

As is evident from the velocities in Table 2, the flashing produced is aggressive and is a significant factor in the pressure drop of the system. When compared to the single phase liquid only assumption which provided a pressure drop of 3.59 kPa before the manual valve which incorporates roughly 15 m of pipeline, the step jump in pressure drop after the valve, which is only 0.730 m in length, is staggering.

**Table 2. Pressure drop and flash vapour produced within the juice inlet line at various feed valve openings for the design flow rate of 189 t/h/train.**

| | Valve at 50° | Valve at 60° | Valve at 90° |
|---|---|---|---|
| Flash vapour (t/h) | 3.73 | 1.63 | 0.47 |
| Pressure drop (kPa) | 54.50 | 24.94 | 12.32 |
| Liquid velocity (m/s) | 2.85 | 2.89 | 2.92 |
| Vapour velocity (m/s) | 172.62 | 60.16 | 15.42 |

Even though the manual valve was placed as close to the opening of the flash tank as possible to limit the pressure drop due to 2-phase flow, the values in Table 2 prove and confirm the necessity for the more rigorous procedure.

A similar procedure was followed for the juice return line from the partial flash tank to the 4th effect vessel feed ring.

The vapour or flash return line to the 4th effect vessel, along with the control valve, had to be sized for maximum flash. With the inlet conditions of flow, brix, temperature and pressure into the flash as well as brix, temperature and pressure of the final effect, a mass and energy balance yielded the maximum flash to be 11 t/h/train. The vapour line was sized to be a 600 NB with a pressure drop of 1.634 kPa. The type of valve selected was a butterfly control valve as this proved most cost effective. The selected control valve was a full line size (600 NB) as the pressure drop was minimal, therefore making available the largest operating pressure envelope of the partial flash tank, as shown in Figure 7. In addition, a smaller size valve would not be able to maintain control at the lower pressures due to its rapid opening characteristic.

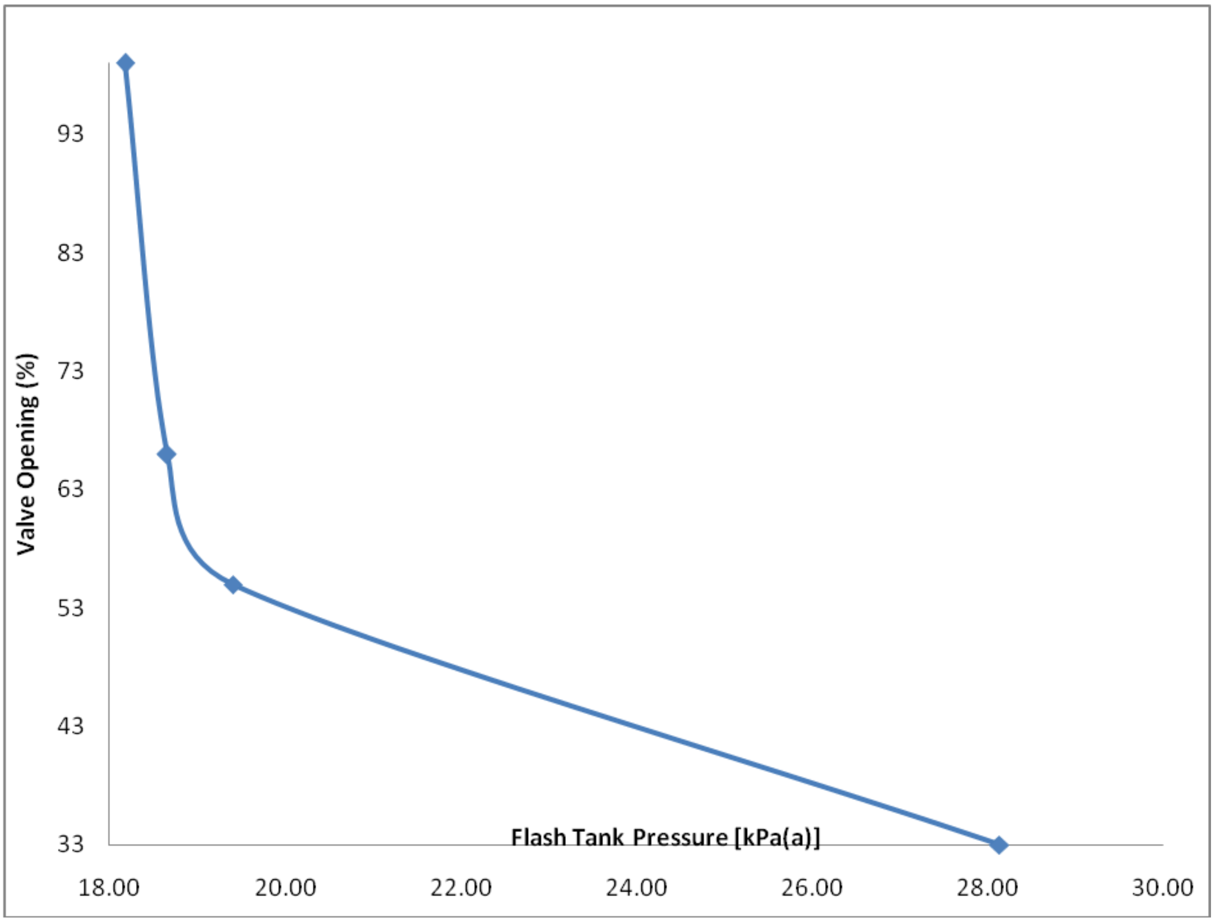



**Figure 7. Relationship between flash tank pressure and pressure control valve opening for the 600 NB butterfly valve.**

*Partial flash tank*

The sizing of the partial flash tank is driven primarily by the volumetric flow rate of vapour. The diameter is determined by the selected velocity criteria. All other dimensions are then related to the vessel diameter. A suitable velocity is determined by the level of acceptable entrainment of liquid droplets in the vapour stream.

Initially the approach of the mixed juice flash tank sizing was taken, but this proved to be impractical and uneconomical. For this application a vessel with a diameter of >6 m would have been required (Rein, 2007). This approach allows a minimum of entrained liquid, as the discharge is to the atmosphere, and would present a potential safety risk and sucrose loss. However, this application is not as constrained, since the vapour is routed to the final effect vessel. The tolerable entrainment is much higher as it will be removed in the entrainment separator of the final effect vessel.

A different approach (as detailed in Coulson and Richardson (Sinnott, 1993)) was then taken. This method tolerated more liquid entrainment in the vapour stream, which seemed more suitable to the application. The resulting vessel diameter was 3 m. This size of vessel was more practical and economically feasible to install.

The vessel contains no internal elements. It has nozzles for inlet and outlet streams, and pressure measurement. Sight and light glasses were installed to observe the operation of the vessel.

The elevation positioning of the vessel is constrained by upper and lower limits. The upper limit is the position at which the juice boiling level in the 3rd effect starts to increase in order to drive the juice into the flash vessel. The lower limit is the liquid level in the flash tank relative to the inlet nozzle, which needs to be above the juice level for adequate separation to occur. Both limits are dependent on the operating pressure of the partial flash tank. Hence the operating pressure envelope needs to be considered when determining the vessel elevation.

## Commissioning

The system was designed during the 2012 season. The procurement and fabrication were executed during the 2013 off-crop. The system was commissioned and started up as part of the 2013 season operation and operated for the full 2013 season.

The A-set was commissioned first and showed a significant rise in the 3rd effect operating liquid level. The set was shut down and the B-set was started up. This also showed an increase in the 3rd effect operating level. Investigations showed that the pressure drop through the items in the line were potentially under-estimated, resulting a larger than expected pressure drop. This was being compensated for by an increase in static head in the 3rd effect.

The solution was to lower the feed nozzle to the partial flash tank and relocate the feed valve to the nozzle itself to reduce the length of pipe experiencing 2-phase flow. This solution was applied to both A and B-sets, which solved the static level increase in the 3rd effect vessels.

Once the system was operating without any problems, the benefits to the evaporator station were immediately noticeable. The ‘fountain effect’ previously experienced in the final effect was no longer there. It was evident that a ‘normal’ circulation and boiling pattern had been restored. The downstream entrainment separator was virtually drip-free. Brix results of the final syrup showed consistently high values to the point where the exhaust steam pressure had to be reduced because the brix was too high.

The need for the pressure control valve loop proved not necessary as, even with nearly equal pressure between the final effect evaporator and the partial flash tank, the liquid level was sufficient to overcome the pressure drop of the feed system. However, this is a feature that should be retained for similar installations elsewhere as this flexibility may be required.

## Results

Evaporator performance before commissioning of the partial flash tanks is shown in Table 3.

**Table 3. Data for the 2012 and 2013 seasons.**

| | A-Evaporator set | | | | B-Eevaporator set | | | |
|---|---|---|---|---|---|---|---|---|
| | 2012 | | 2013 | | 2012 | | 2013 | |
| | kPa (g) | °C | kPa (g) | °C | kPa (g) | °C | kPa (g) | °C |
| Exhaust steam | 117 | 123 | 101 | 121 | 117 | 123 | 101 | 121 |
| V1 | 47 | 111 | 39 | 109 | 48 | 111 | 41 | 110 |
| V2 | 19 | 104 | 10 | 102 | 21 | 104 | 11 | 102 |
| V3 | -19 | 94 | -25 | 93 | -20 | 94 | -25 | 93 |
| PFT | N/A | | -76 | 61 | N/A | | -76 | 61 |
| V4 | -77 | 63 | -78 | 59 | -77 | 62 | -79 | 58 |

The vacuum remained sub-optimal over the two years although comparisons were done under fairly identical conditions (Table 4). Work to optimise evaporator final effect vacuum to less than 20 kPa abs is ongoing.

**Table 4. Average brix profiles for the 2012 and 2013 seasons.**

| | A-Evaporator set | | B-Evaporator set | |
|---|---|---|---|---|
| | 2012 | 2013 | 2012 | 2013 |
| Average CJ brix (%) | 13.5 | 13.17 | 13.5 | 13.17 |
| Average juice flow ($m^3/h$) | 244.0 | 257 | 244.0 | 257 |
| Ex-first effect brix (%) | 23.4 | 24.66 | 24.8 | 23.73 |
| Ex-second effect brix (%) | 30.1 | 30.81 | 33.2 | 31.26 |
| Ex-third effect brix (%) | 39.4 | 41.17 | 41.8 | 40.96 |
| Ex-partial flash tank brix (%) | N/A | 43.70 | N/A | 42.65 |
| Syrup brix (%) | 63.4 | 68.54 | 64.0 | 68.23 |

## Discussion

*Partial flash tank operational issues*

Installation and commissioning of the flash tank was well received by the mill operational team as it was fairly easy to commission, easy to operate and brought immediate visible benefits to the factory operation on various fronts. Ideally, partial flash tank operation entailed adjusting the vapour valve opening to control the amount of flashing whilst monitoring the boiling pattern in the 4th effect. However, the vacuum in the flash tank approached, or operated very close to, that of the 4th effect even with the isolation valve fully closed, hence operation was theoretically more of a full flash tank rather than a partial flash tank.

*Partial flash tank immediate benefits (2013 season)*
Several changes were noticed immediately following installation of the partial flash tank, as described below.

(a) Reduction in overall entrainment losses to the injection cooling water

Installation of partial flash tanks resulted in a significant reduction in entrainment losses to the injection water thus reduced losses to effluent. This was observed through a significant reduction if not close to nil entrainment return from the fourth effects and a significant drop in sucrose contamination levels in factory injection water (Figure 8).

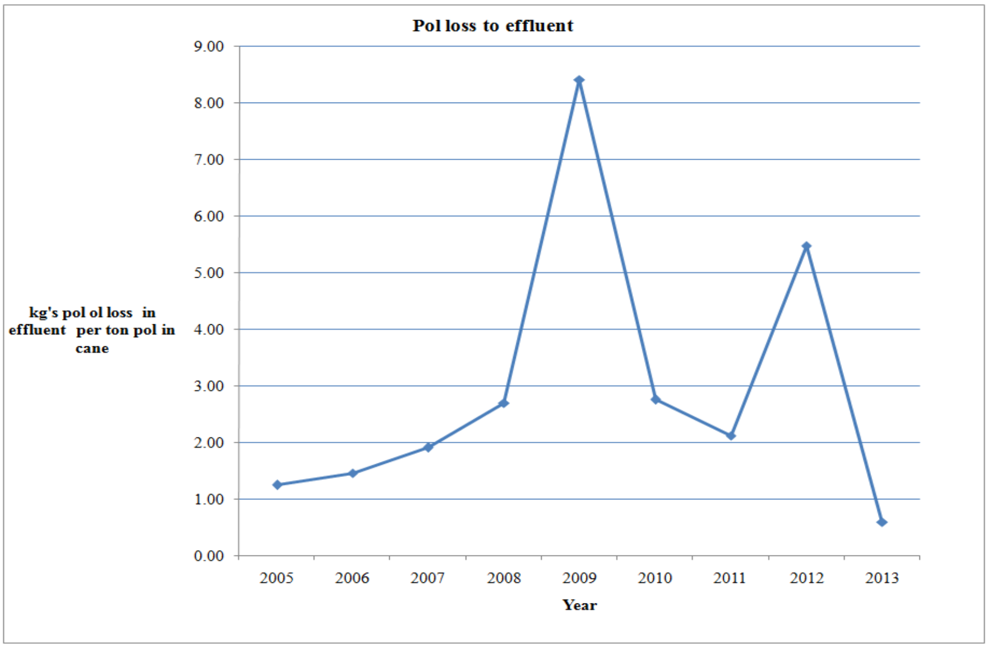



**Figure 8. Pol loss to effluent (2005-2013).**

A comparison of pol losses per week is shown in Figures 9 and 10 for years 2012 and 2013. The comparison shows that the partial flash tank was more effective in reducing entrainment than the modifications done in 2012, which entailed juice feeding above the calandria top tube plate.

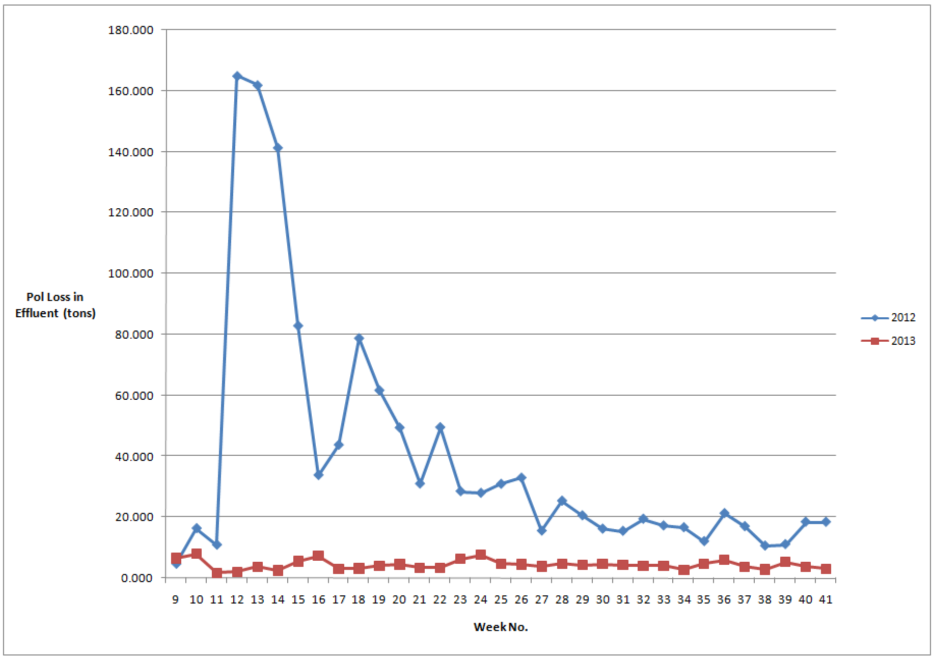



**Figure 9. Weekly comparisons of tons lost to effluent (2012 and 2013).**

As a result of the significant drop in entrainment contribution to undetermined loss, UDL for 2013 was a record low over a 10-year period.

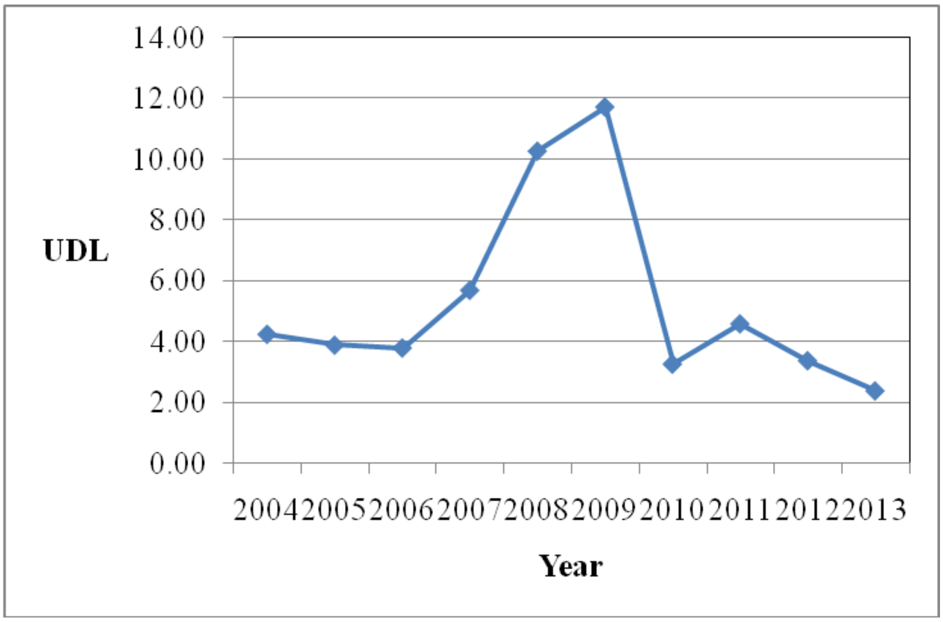



**Figure 10. Undetermined loss (UDL) trends (2004-2013).**

(b) Improved evaporator performance (high syrup brix)

The flash tank installation brought immediate benefits to evaporator brixing performance which was realised immediately on commissioning of the flash tanks. Syrup brix profiles were measured on both sets, and it was observed that around 2 units brix gain were realised across the flash tank, corresponding to about 4 t/h evaporation rate at average throughputs of 250 $m^3$/h clear juice into one set (refer to Table 4). This improvement was attributed to the introduction of the partial flash tank as no other modifications or operational changes had been made to the evaporators during the off-crop or season. Elimination of localised spouting assisted in heat transfer only where wetting was occurring. The partial flash tank resulted in more uniform juice distribution taking place thus resulting in increased utilisation of the vessel heating surface area in heat transfer (Figure 11).

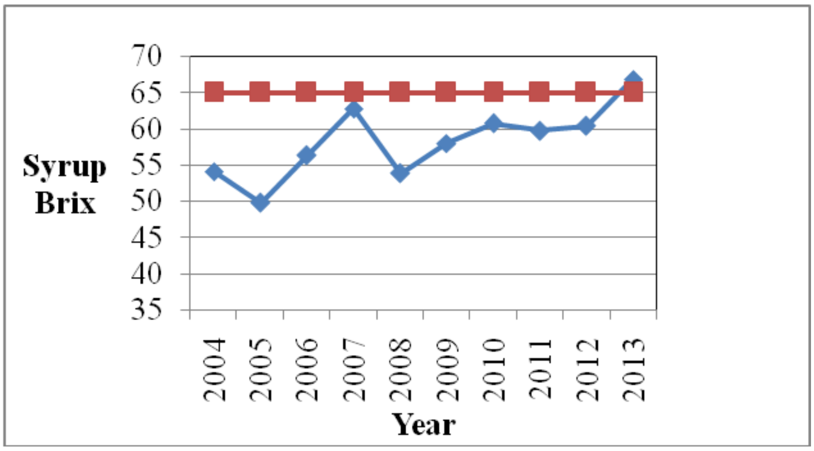



**Figure 11. Average syrup brix attained by A and B sets (2004-2013).**

A comparison of 2012 and 2013 weekly syrup brix values (Figure 12) clearly shows that 2013 performance was consistent in achieving the minimum required brix of 65% for almost all of the season weeks. For 2012 and 2013 seasons, a reduction in the steam pressure required to operate evaporators was noticed as the evaporators could attain and exceed the minimum brix of 65 at less than 200 kPa absolute pressure exhaust steam, compared to more than 210 kPa utilised in 2012 even though the targeted syrup brix was not being achieved at that time.

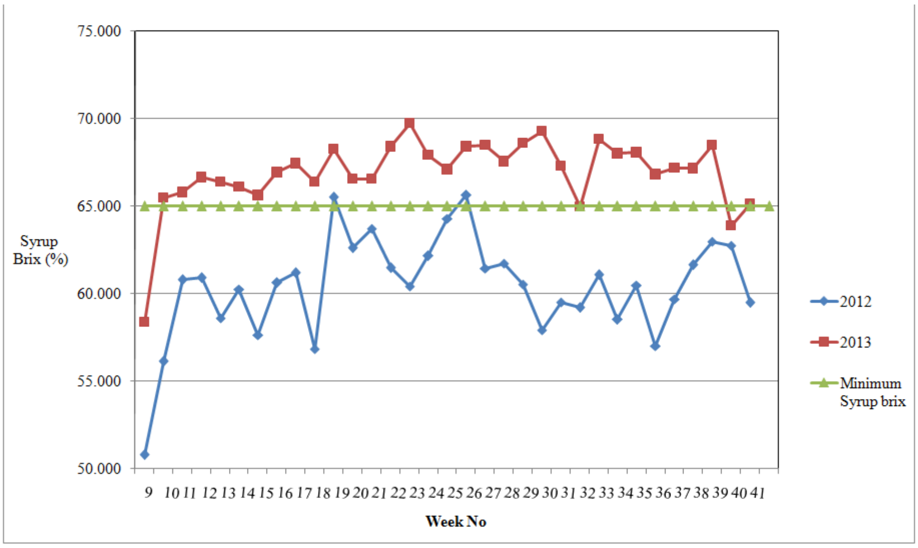



**Figure 12. Weekly comparison of syrup brix (2012 and 2013).**

(c) Improved operation of the diffusers (high imbibition rates and improved extraction)

Over the years, factory throughput was limited by the evaporators, as they could not achieve or handle design throughputs whilst still attaining satisfactory syrup brix. Improved evaporators performance resulted in flexibility to increase imbibition water usage, hence an improvement in operations of the diffuser (Figure 13). This enabled optimisation of diffuser operations thus realising an improvement in extraction over the season. Imbibition per cent fibre of 373% was achieved at an average 419 t/h (design capacity 440 tch) breaking an eight year record with extraction ending at 97.15% for the 2013 season (Figure 14).

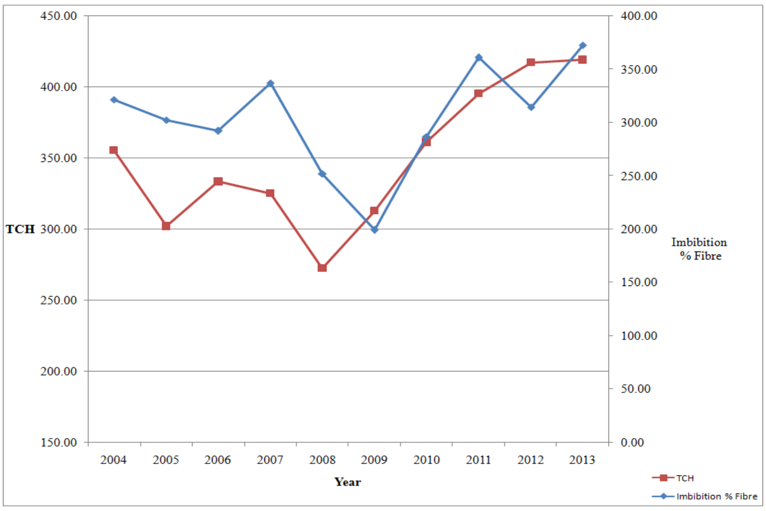



**Figure 13. Crush rate and imbibition % fibre trends (2004-2013).**

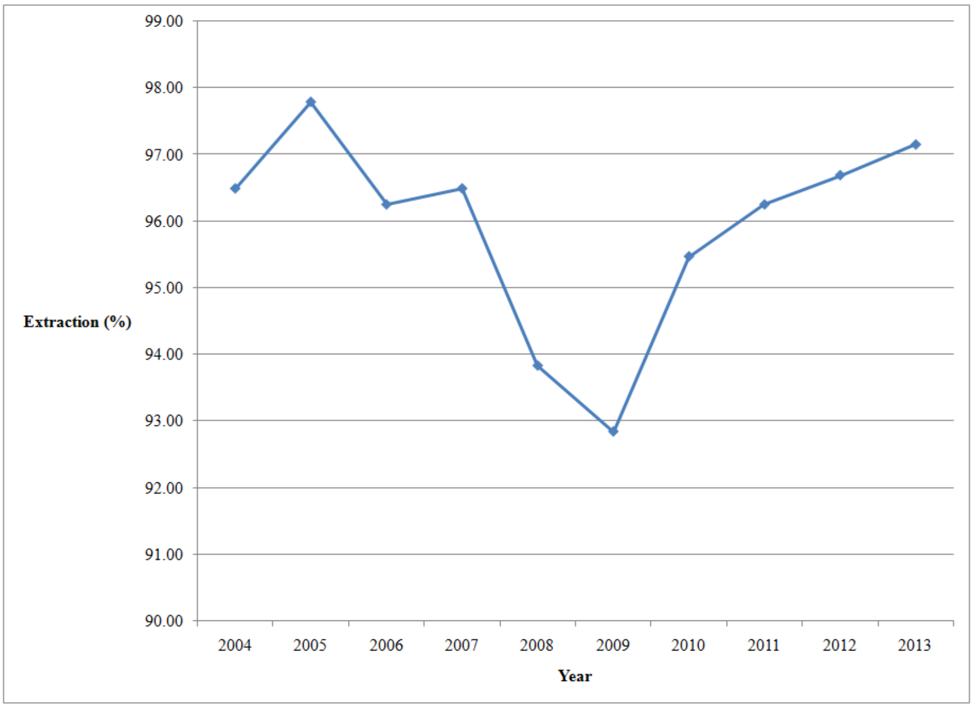



**Figure 14. Extraction performance (2004-2013).**

## Conclusion

The installation of partial flash tanks on the evaporator sets at Hippo Valley Estates sugar mill has proved to be of major benefit to the mill. Most noticeable benefits have been:

(i) Significant reduction in sugar loss through entrainment to injection water.

(ii) Restoration of the boiling pattern in the final effect evaporators with a reduction in the exhaust steam pressure required.
(iii) Removal of the bottleneck at the evaporator station as the factory constraint and improving the extraction potential of the mill.

While not without its challenges, the installation has proved that if designed, commissioned and operated well, this can bring a step change improvement to the performance of a sugar factory with a similar problem.

## REFERENCES

Allen RG (1996). Relating the Hazen-Williams and Darcy-Weisbach friction loss equations for pressurized irrigation. *Am Soc Agric Eng* 12(6): 685-693.

Bindoff A and Dlamini W (2013). Felixton mill evaporator station – Partial flash tank. *Proc S Afr Sug Technol Ass* 86: *375- 383*

Churchill S. W., (1977). Friction-Factor Equation Spans all Fluid-Flow Regimes, *Chem. Eng*. 84, pp 91- 92.

Hoekstra RG (1981). A computer program for simulating and evaluating multiple effect evaporators in the sugar industry. *Proc S Afr Sug Technol Ass* 55: 43-50.

Hooper WB (1981). The two-K method predicts head losses in pipes. *Chemical Engineering* August 24, 1981, pp 96-100.

Lockhart, R. W., and Martinelli, R. C., (1949), "Proposed Correlation of Data for Isothermal Two Phase Flow, Two Component Flow in Pipes," *Chem. Eng. Prog*., 45, pp. 39–48

Muzzell DJ, Dzirove. E, Peacock. S, Mugadhi. A (2013). Solutions to Hippo Valley evaporator problems. *Proc S Afr Sug Technol Ass* 86: 384-403.

Perry JH (1963). *Chemical Engineers' Handbook*. 4th Ed. Mcgraw-Hill Book Company Inc. New York, USA. pp 538-539.

Rein P (2007). *Cane Sugar Engineering*. Verlag Dr Albert Bartens KG, Berlin, Germany. 214 pp.

Sinnott RK (1993). Coulson and Richardson's Chemical Engineering. Vol 6, p 417.

## Appendix A. Calculation method of the 2-phase pressure drop

The Lockhart and Martnelli semi-empirical correlation is based on the single phase pressure drop for either phase multiplied by a factor which is a function of the single phase pressure drop of the two phases (Perry, 1963).

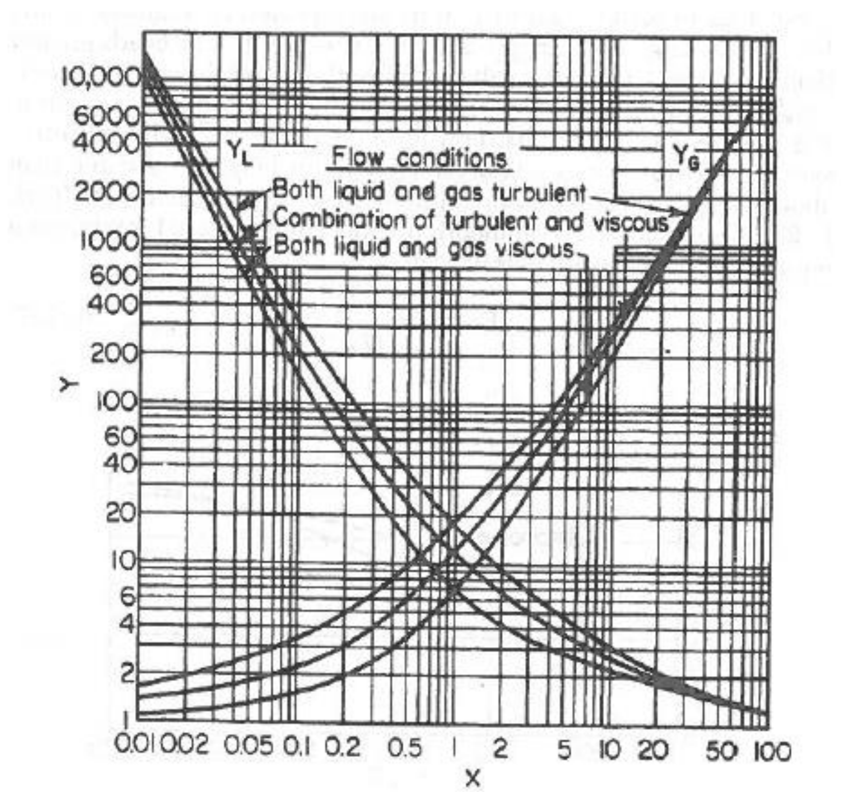
$$\left(\frac{\Delta P}{\Delta L}\right)_{T,P} = Y_L \left(\frac{\Delta P}{\Delta L}\right)_L \tag{A1}$$

$$\left(\frac{\Delta P}{\Delta L}\right)_{T,P} = Y_G \left(\frac{\Delta P}{\Delta L}\right)_G \tag{A2}$$

where $Y_L = F_1(X)$
$Y_G = F_2(X)$
and

$$X = \left[\frac{\left(\Delta P / \Delta L\right)_L}{\left(\Delta P / \Delta L\right)_G}\right]^{1/2} \tag{A3}$$

The functions $F_1$ and $F_2$ (eqns X and X) are plotted on the graph below and are used to determine $Y_L$ or $Y_G$ (Perry, 1963).




Work was carried out in Matlab to determine the equation for $F_1$. The resulting equation can then be used to determine the $Y_L$ factor. The equation obtained is as follows:

$$\log(Y_L) = -0.0167 \times \log(X)^3 + 0.216 \times \log(X)^2 - 0.9796 \times \log(X) + 1.2522 \qquad \text{(A4)}$$

A mass and energy balance around the control valve leads to the following equations which are used to solve for the liquid and vapour flowrates caused by the pressure drop across the control valve:

$$m_2 = \frac{m_1 \, x \, (H_1 - H_3)}{(H_2 - H_3)} \qquad \text{(A5)}$$

$$m_3 = \frac{m_1 \, x \, (H_1 - H_2)}{(H_3 - H_2)} \qquad \text{(A6)}$$

where:

$m_1$ is the liquid flowrate into the control valve in kg/h
$m_2$ is the vapour flowrate after the control valve caused by the pressure drop in kg/h
$m_3$ is the liquid flowrate after the control valve caused by the pressure drop in kg/h
$H_1$ is the enthalpy of the $m_1$ stream in kJ/kg
$H_2$ is the enthalpy of the $m_2$ stream in kJ/kg
$H_3$ is the enthalpy of the $m_3$ stream in kJ/kg.

## Appendix B. Single phase pressure drop calculations using the Churchill equation for the friction factor

A flowing fluid in a pipe loses energy due the contact between the fluid and the pipe wall. These energy losses due to friction are termed head losses and are calculated as follows.

$$H = \frac{fLu^2}{2gd} \tag{B1}$$

where
$H$ is the head loss in metres
$f$ is the friction factor for the pipe
$L$ is the length of pipe in metres
$g$ is gravity in m/s$^2$
$d$ is the diameter of the pipe in metres
$u = \frac{Q}{\pi r^2}$ = the superficial velocity of the fluid in the pipe in m/s
$Q$ is the volumetric flowrate in m$^3$/h.

The head loss is terms of pressure drop across the pipe can also be calculated as follows.

$$\Delta P = \frac{H\rho g}{1000} \tag{B2}$$

where
$\Delta P$ is the pressure drop in kPa
$\rho$ is the density of the fluid in kg/m$^3$.

The friction factor *f* can be determined in many ways for a particular pipe, one of which is by using the Churchill equation. The equation is valid for both rough and smooth pipes and for the full range of laminar, transition and turbulent flow regimes (Allen, 1996).

$$f = 8\left[\left(\frac{8}{\mathrm{Re}}\right)^{12} + \frac{1}{(A+B)^{1.5}}\right]^{1/12} \tag{B3}$$

where

$$A = \left(-2.457\ln\left[\left(\frac{7}{\mathrm{Re}}\right)^{0.9} + 0.27\left(\frac{e}{d}\right)\right]\right)^{16} \tag{X}$$

$$B = \left(\frac{37530}{\mathrm{Re}}\right)^{16} \tag{X}$$

where

$$\mathrm{Re} = \frac{\rho v d}{\mu} = \text{the Reynolds number} \tag{X}$$

$e$ is the roughness factor for a particular pipe material in metres.