REFEREED PAPER

THE LONG AND SHORT OF CVP TUBES

MOOR BStC[1], RAGHUNUNDAN A[2] AND RAMARU R[1]

[1]Bosch Projects, PO Box 2009, Durban, 4000, South Africa
[2]Illovo Sugar Proprietary Limited, PO Box 194, Durban, 4000
MoorB@boschprojects.co.za RamaruR@boschprojects.co.za
ARaghunandan@illovo.co.za

Abstract

It has long been recognised that short tubes – preferably from 600 to a maximum of 1 200 mm long – are required for good performance in batch pans. There are sound reasons for this. With the advent of vertical tube continuous vacuum pans (CVPs), this lore of short tubes has been widely extrapolated also to apply to these vessels. However, the main reason why short tubes are needed in batch pans does not apply in CVPs. Indeed, for a good circulation profile and reduced cost of construction, longer tubes are desirable, but some technologists have resisted long tubes because of the batch pan theory and because of pilot plant research by Rouillard in 1985.

In this paper, the theory of boiling in pan tubes is revisited. Thereafter, Rouillard's results for 1 000, 1 400 and 1 800 mm tubes are compared with practical results from a number of short tube (≤1 500 mm) and longer tube (1 700 mm) commercial CVPs. Conclusions are that:

- The commercial CVPs achieve heat transfer coefficients (HTCs) 30% to 200% higher than those of Rouillard.
- The differences are greatest for the long tube (1700 mm) pans.
- The long tubes perform as well as or better than the shorter tubes.

Rouillard's pilot plant HTCs are so much lower than the commercial values that his tube length conclusions should be disregarded. A possible reason is offered for the large differences. An intuitive explanation is proposed as to why long tubes perform well.

Keywords: pan tubes, boiling in pans, boiling point elevation, pan HTCs, continuous pans, CVP design.

Introduction

It is widely recognised that short tubes – preferably 600 to 1200 mm long– are required for good performance in batch pans (Rein, 2016). There are sound reasons for this, although an incorrect explanation is often quoted. When the correct reason is understood, it is apparent that the considerations for vertical tube continuous vacuum pans (CVPs) are markedly different.

In this paper, some theories of massecuite boiling in CVP tubes are reviewed. Measured heat transfer coefficients from commercial CVPs with various tube lengths are quoted and these are compared with each other and with the HTCs measured by Rouillard (1985) in his pilot plant. Conclusions are drawn and an intuitive explanation for the findings is offered.

Factors influencing heat transfer in a tube

An increase in fluid velocity in a tube promotes turbulence which reduces the laminar liquid layer and thus increases heat transfer in evaporators (Walthew and Whitelaw, 1996).

However, a point will eventually be reached whereby a further increase in fluid velocity results in minimal heat transfer benefits (Walthew and Whitelaw, 1996; Yu *et al.*, 2004).

The turbulence in a tube is also caused by bubbles rising at a relatively higher velocity than the liquid fluid because of the buoyancy effect on the bubbles (Bhagwat and Ghajar, 2012). As juice moves up the tube the proportion of vapour increases resulting in the two-phase flow velocity increasing because of the greater volume occupied by the vapour phase following evaporation, and this regime is considered to be a very efficient heat transfer regime (Peacock, 2000). In addition, bubble nucleation (i.e. formation) on the tube wall promotes convective flow of juice to the tube wall due to the differences in density between the bubbles sites and the liquid (Yu *et al.*, 2004). The same concept applies in pan boiling; high turbulence leads to good circulation which then promotes high HTCs. Rouillard (1985) noted that for massecuite boiling in a pilot plant pan the HTCs improved when the velocity in the tube and void fraction in the tube increased.

Boiling in batch pans

In a batch pan for seed or final strikes, it is necessary before commencing boiling to fill the pan to at least cover the top of the tubes. This is to ensure that circulation is possible and no part of the tube ‘dries out’. During the course of the boiling, syrup and/or molasses feed is added until the total massecuite volume is three to four times the initial fill or ‘graining volume’. For a pan with 1 000 mm long tubes and strike/graining ratio of three to four, this final massecuite level is usually about 1 500-1 800 mm above the top tubeplate. This imposes a high hydrostatic head and causes significant boiling point elevation (BPE) in the boiling zone. This in turn reduces the temperature difference (ΔT) between the calandria steam and massecuite, reduces the heat transfer rate and raises the massecuite temperature (with a risk of colour formation). The BPE is a function of the hydrostatic head, the massecuite purity and the massecuite dry solids or Brix (Rein, 2016). The greatest BPE occurs at the high Brix end of low grade batch boilings. It is at this stage that pan boilers often report that the pan has ‘died on them’. Calandria steam pressure must be increased, raising the massecuite temperature and possibly promoting a Maillard reaction and colour formation. Typical boiling point values experienced in high and low grade batch pans are shown in Figure 1.

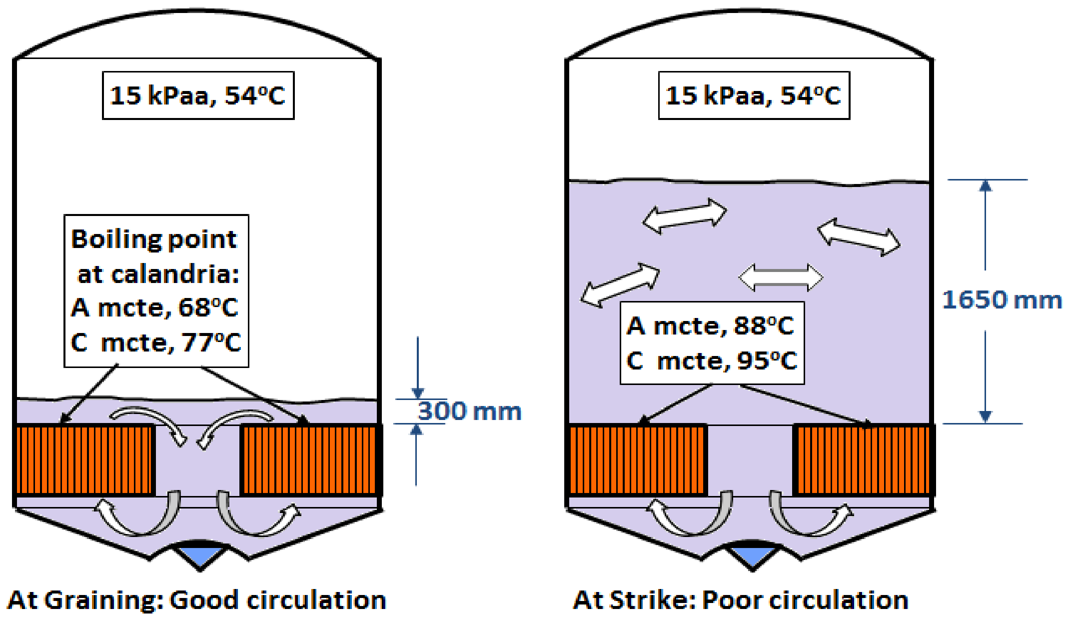

Figure 1. Boiling point is elevated by purity, density and pressure (head).

Circulation is promoted by the vapour bubbles within the tubes, but as the massecuite level increases their effect is dispersed in the open volume above the tubes. Indeed, with high hydrostatic heads towards the end of boilings, there may be no boiling within the calandria and vapour bubbles may only form in a zone above the top tube plate.

Because of the geometry, the final head above the top tubeplate is a function of the tube length: pans with wide short-tube calandrias require less additional head to accommodate the final volume than pans with narrower long-tube calandrias (see Figure 2). Short-tube calandrias therefore perform better than long-tube calandrias during the latter stages of boiling, particularly if there is no assistance from mechanical or jigger stirring.

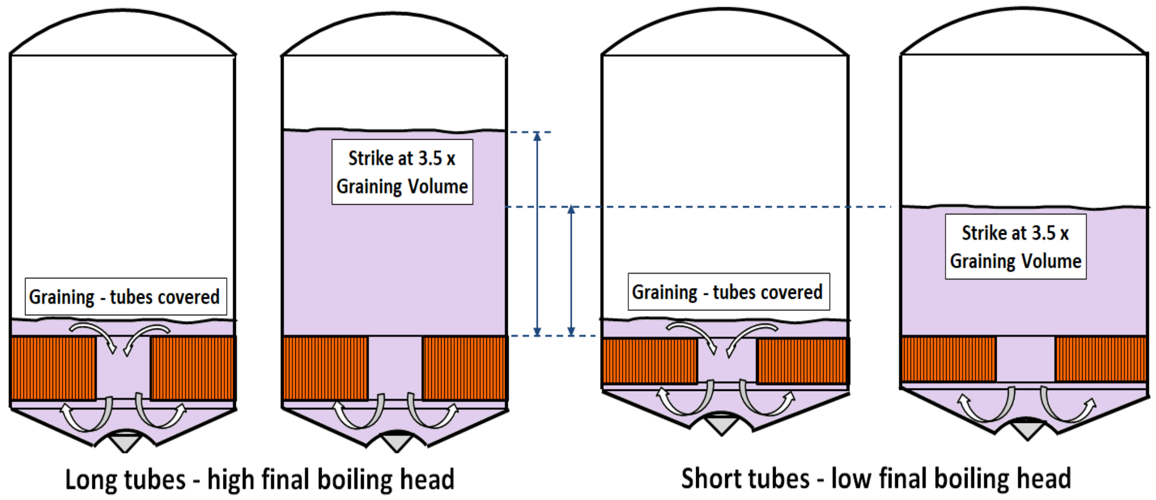

Figure 2. Effect of tube length in batch pans.

For this reason, tube length is a vital concern in batch pan design, and good designs usually limit the tube length to between 600 and 1 200 mm. However, it should be noted that this is only to address boiling at the end of the cycle. At the start, with a low head above the calandria, long tube batch pans boil at least as well as short tube pans. The length of tube below the top tube plate is of little consequence. However, as the massecuite level rises, the hydrostatic BPE raises the boiling point at the top of the calandria. It is this that suppresses boiling, and because longer tubes lead to a higher final head, shorter tubes are better in batch pans.

Experience at Maidstone during the 1960s and 1970s confirmed this. The pan station comprised 12 batch pans with tube lengths varying from 750 to 1 245 mm in length. All boiled reasonably well on a graining and all slowed as boiling progressed and the crystal content and viscosity increased. However, the pans with longer tubes and consequently higher heads sometimes stopped boiling towards the end of high brix boilings. Consequently, these pans were used only for A boilings. The problem was eased when a continuous pan was introduced for the C boilings in 1977 (the first CVP in South Africa).

Boiling in continuous pans (CVPs)

The situation within a tubular calandria horizontal continuous pan is completely different to that in a batch pan. The massecuite level is constant from the first stages to the last, usually controlled at 200-300 mm above the top tubeplate. There is thus minimal hydrostatic head at the top of the tubes, the ΔT between the calandria steam and the massecuite is not suppressed and lower pressure/temperature steam can be used. In fact, it is now well established that CVPs can be boiled on lower pressure vapours (V2 and V3) because of their low head and constant high heating surface/volume ratio (Moor, 2008).

Continuous pans generally exhibit good circulation with a relatively high proportion of vapour in the tubes. This lessens the effective pressure, even at the bottom of the tubes. The factors that suppress boiling in long tube batch pans are therefore absent and it is informative to analyse the boiling mechanism in CVPs of different tube lengths.

The boiling process in a pan tube

A comprehensive study on boiling in pan tubes was undertaken by Rouillard (1985). This research was conducted shortly after the first South African vertical tube CVP had been commissioned at Maidstone (Kruger, 1983) and large (85 m^3) batch pans had been designed for the new Felixton II factory. It was realised that the designs of this equipment relied heavily on 'intuition and feel' rather than measured relationships, and the objective of Rouillard's work was therefore to provide information 'for optimising pan design and operation'. It was to cover both batch and continuous pans.

Rouillard conducted two separate sets of experiments on two different pilot plants. The first set was to characterise the boiling process in the tubes. For this, he used a single tube fed by a positive displacement circulation pump with devices for measuring temperatures, pressures and void fractions at various positions along the tube. From this work he verified three stages of boiling in a pan tube which he described as shown in Figure 3.

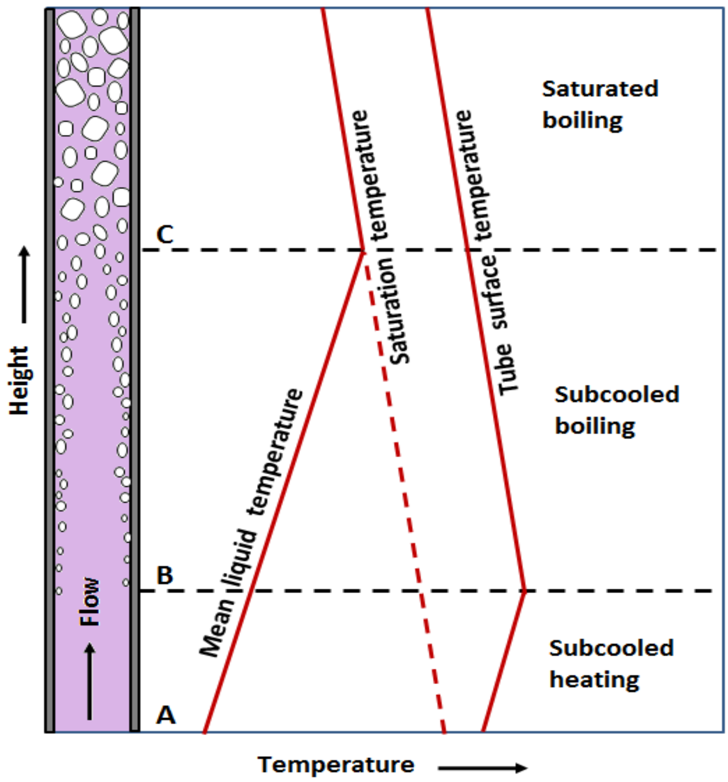

Figure 3. Boiling in a vertical tube (per Rouillard, 1985).

As the massecuite enters the tube it is subcooled, that is, its temperature is below the boiling temperature corresponding to the pressure at that point. In region AB, heat transfer is entirely by single phase forced convection. As heat is transferred to the liquid, the temperature adjacent to the walls increases until point B, at which it reaches the boiling temperature and ebullition starts, but the bulk of the liquid is still subcooled. Between B and C, the vapour bubbles grow because of the additional heat input and as a result of decompression as the liquid rises up the tube. At the same time, heat penetrates towards the centre of the tube so that the bulk liquid temperature increases. The heat transfer in region BC Rouillard called 'subcooled boiling'. Transition to full boiling occurs at point C when the combined enthalpy of the liquid and vapour equal the saturation enthalpy of the liquid. Boiling then extends across the entire sectional area of the tube and is known as saturated boiling. From this point there is a gradual decrease of the fluid temperature because of the decompression effect. Under certain conditions the transition to saturated boiling is not reached and subcooled boiling extends to the tube outlet. Rouillard (1985) reported that the length of the subcooled boiling section is a function of heat flux and velocity up the tube.

Pilot plant measurements of evaporation in tubes

Rouillard (1985) then used a different pilot plant to explore the effects of a number of variables – tube length, head above the tubes, steam pressure, vacuum and brix. For this, he used an experimental pan with four separate individually steam jacketed 100 mm diameter tubes of lengths 600, 1 000, 1 400 and 1 800 mm. Moderate brix A and B massecuites and C seed were boiled.

Because of the large number of variables, interpretation of Rouillard's results can prove confusing. To provide greater clarity they have been restated in the tables below in two ways:

1. Rouillard used two heads above the tubes – either a 'high' of 870-940 mm, which is typical of the head at the middle of a batch pan boiling, or a 'low' of 200-250 mm, which is typical of continuous pans. His data has therefore been separated into that relevant for each pan type.
2. Rouillard quoted his performances in 'evaporation rates', which are affected by the wide range of temperature differences between heating steam and massecuite. This distraction is avoided by calculating heat transfer coefficients (HTCs) from the evaporation rates.

The HTCs for these conversions (and those for Bosch CVPs quoted later) were calculated according to the method described in Appendix B. This requires a calculation of the Boiling Point elevation (BPE), using the formula of Saska (2002), described in Appendix A.

Batch pans

A full tabulation of the 'high head' evaporation rates from Rouillard's Tables 2, 3 and 4, converted into HTCs, is given in Appendix D. However, as the main concern of this paper is with continuous pans, only a summary is given in Table 1. For the reasons already mentioned, batch pans are commonly constructed with tubes of 600 to 1 300 mm length; they cannot be satisfactorily made with 1 800 mm tubes, so the 1.8 m tube results are excluded.

Table 1. 'High head' heat transfer coefficients (HTCs) from Rouillard's (1985) pilot plant for various tube lengths.

	HTCs in W/m^2.K		
	Tube length		
	0.6 m	1.0 m	1.4 m
A massecuite	501	427	400
B massecuite	272	205	180
C massecuite	119	105	84

These 'high head data consistently show HTC advantages of shorter tubes for batch pans. This conclusion has been extrapolated by some to be applicable also to low head boilings. However, as will be shown, this is highly questionable. The main reason for selecting short tubes in batch pans is the effect on the final boiling head, as already explained.

Continuous pans

The 'low head' (200-250 mm) results from Rouillard are grouped in Table 2. It is commercially not practical to use tubes of less than 1.0 m for any except very small CVPs – the calandria plan area becomes large and profile proportions are difficult. The table therefore compares only realistic CVP tube lengths.

Table 2a,b,c. Pilot plant results under continuous pan conditions (low head).

A Massecuite Estimated BPE from solution & head **12.8 °C**

Pressure kPaa		Temps.°C		Evaporatn. Rate kg/m².h			HTC (W/m².K)		
P_s	P_v	T_s	T_v	1.0m	1.4m	1.8m	1.0m	1.4m	1.8m
195	9	119	44	39	37	32	**381**	**362**	**313**
125	9	106	44	22	23	21	**278**	**291**	**266**
195	25	119	65	25	26	22	**368**	**383**	**324**
125	25	106	65	13	12	14	**287**	**265**	**309**
						Average HTCs	**329**	**325**	**303**

B Massecuite Estimated BPE from solution & head **16.0 °C**

Pressure kPaa		Temps.°C		Evaporatn. Rate kg/m².h			HTC (W/m².K)		
P_s	P_v	T_s	T_v	1.0m	1.4m	1.8m	1.0m	1.4m	1.8m
195	9	119	44	21	24	21	**216**	**247**	**216**
127	9	107	44	15	13	14	**201**	**174**	**187**
195	20	119	60	16	15	13	**226**	**212**	**183**
127	20	107	60	9	7	7	**184**	**143**	**143**
						Average HTCs	**207**	**194**	**182**

C Massecuites Estimated BPE from solution & head **20.8 °C**

Pressure kPaa		Temps.°C		Evaporatn. Rate kg/m².h			HTC (W/m².K)		
P_s	P_v	T_s	T_v	1.0m	1.4m	1.8m	1.0m	1.4m	1.8m
180	10	117	46	7.8	6.2	5.5	**96**	**76**	**67**
130	10	107	46	5.6	5.0	4.9	**86**	**77**	**76**
						Average HTCs	**91**	**77**	**71**

Suffix s = steam, v = vacuum vapour

It is seen from these tables that with low boiling heads (as in CVPs), shorter tubes usually showed slightly better HTCs, but the trend was not strong and there were some contradictory results.

Comparison of pilot plant with commercial results

How does the performance of the pilot plant compare with that of commercial CVPs?

HTC data are available from three types of commercial vertical tube CVPs – Tongaat Hulett/Fletcher Smith, Bosch and SRI. The averages of each type are compared with those from Rouillard's pilot plant for similar length tubes.

Tongaat-Hulett /FS HTCs.

Table 3 shows ranges of HTCs from Tongaat-Hulett/FS CVPs reported by Rein and Msimanga (1999). Tubes of these pans ranged from 1.3 to 1.5 m in length. The HTCs are therefore compared to the pilot plant results for 1.4 m tubes.

Table 3. Average heat transfer coefficients (HTCs) for Tongaat Hulett continuous vacuum pans (CVPs) vs. pilot plant (W/m^2.K).

Massecuite grade	A	B	C
Averages for TH/FS pans	413	212	115
* Pilot plant averages for 1.4 m tubes	278	159	77
T-H/FS HTCs are higher by	***49%***	***33%***	***49%***

(*Commercial CVPs are generally operated with calandria steam pressures between 70 and 140 kPa abs. Rouillard's results for 195 kPa abs have therefore been excluded from the averages). For ranges of the T-H/FS HTCs, see Table 7 in Appendix 3.

Data from other Tongaat Hulett CVPs within Bosch's experience have also yielded HTCs within the ranges quoted by Rein and Msimanga (1999) – considerably higher than those of Rouillard's pilot plant.

Bosch HTCs

In Table 4, a similar comparison is made comparing the HTCs from Bosch 1.7 m tube CVPs (seven results from A pans, six from Bs and four from Cs) with those of the pilot plant's 1.4 m and 1.8 m tubes.

Table 4. Heat transfer coefficients (HTCs) for Bosch 1.7 m tube continuous vacuum pans (CVPs) vs. Pilot plant (W/m^2.K).

Massecuite grade	A	B	C
Average for Bosch pans, 1.7 m tubes	473	359	212
Pilot plant averages for 1.4 m tubes	278	159	77
Bosch HTCs higher by	***70%***	***126%***	***175%***
Pilot plant averages for 1.8 m tubes	277	159	70
Bosch HTCs higher by	***71%***	***126%***	***202%***

Ranges of the Bosch HTCs are given in Table 8 in Appendix 3.

Bosch has also supplied CVPs with 1.5m tubes. The HTCs of the A pan at Nakambala and the C pan at the NA&TL factory in Vietnam were both higher than the 1.7 m averages but that of the Maragra C in Mozambique was considerably lower. There are too few results to draw any conclusions.

It should be noted that in calculating the Bosch HTCs, the under-base heating area, which is equivalent to 4 to 6% of the tube area, has not been taken into account.

Australia's Sugar Research Institute CVP HTCs

The SRI has supplied CVPs with tube lengths varying from 1 200 up to 2 000 mm, including stepped calandria pans where tubes of different length are used in the same pan. Broadfoot (1999) has indicated that SRI does not consider tube length *per se* to be an important factor in performance and that they have been happy with the results from 2 000 mm tubes (Broadfoot, 2015).

Guiding values used by SRI for the average HTCs from commercial SRI continuous pans are reflected in the recommended design data quoted by Broadfoot *et al (*2018).

Table 5. Heat transfer coefficients (HTCs) for the Australian Sugar Research Institute (SRI) continuous vacuum pans (CVPs) vs. Pilot plant (W/m^2.K).

Massecuite grade	A	B	C
Averages for SRI pans	473	359	212
Pilot plant averages for 1.0 m tubes	283	193	91
Pilot plant averages for 1.4 m tubes	278	159	77
Pilot plant averages for 1.8 m tubes	277	159	70

The SRI HTCs are significantly higher than those of Rouillard's pilot plant for any tube length.

Overall, the HTC performances from commercial CVPs are seen to be much higher than those from the pilot plant. The information provided in Rouillard's paper provides no obvious explanation for these differences. However, one factor that could possibly account for Rouillard's low HTCs is inadequate incondensables venting. His illustration shows a steam entry point part-way up each steam-jacketed tube and a single vent at the top of each jacket, as in Figure 4.

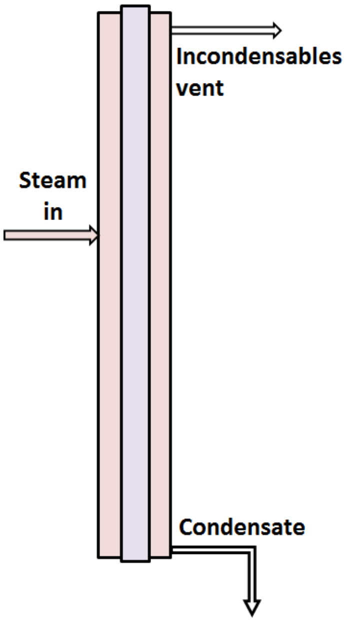

Figure 4. Rouillard's venting.

This could have allowed accumulation of incondensables (which are generally heavier than steam) in the quiescent zone at the bottom of each jacket. If so, this would have lowered the steam partial pressure and hence the steam temperature in these zones, with the effect probably greater in the longer tube jackets. Whatever the reason, the differences between

Rouillard's HTCs and practical experience are so large as to cast serious doubt on the relevance of the pilot plant conclusions regarding tube lengths for commercial CVP designs.

It is not possible to examine boiling inside commercial pan tubes in the way that Rouillard did in his pilot plant. However, his theory can be used to give a plausible explanation of why the measured HTCs in the long tube CVPs are as good as or better than those from shorter tube CVPs.

Massecuite boiling in a CVP with 1500 mm tubes

Consider first the boiling in a single tube of a CVP with 1500 mm tubes. For this purpose, the following assumptions are made:

Pan boiling pressure	15 kPa abs
Then temperature of vapour in pan	53.9°C
Massecuite surface height above tubes	300 mm
Brix of massecuite in tube	92.3%
Purity of massecuite (B pan)	79%
Then BPE due to solution is	13.1°C (a function of brix and purity)
Temperature of massecuite at surface	53.9 + 13.1 = 67°C
Density of massecuite	1.44 t/m^3
Hydrostatic BPE per 0.1 m depth	0.19°C (a function of density)

These assumptions lead to the situation depicted in Figure 5. The boiling temperatures shown on the right ignore the reduction in hydrostatic head due to vapour bubbles within the tube:

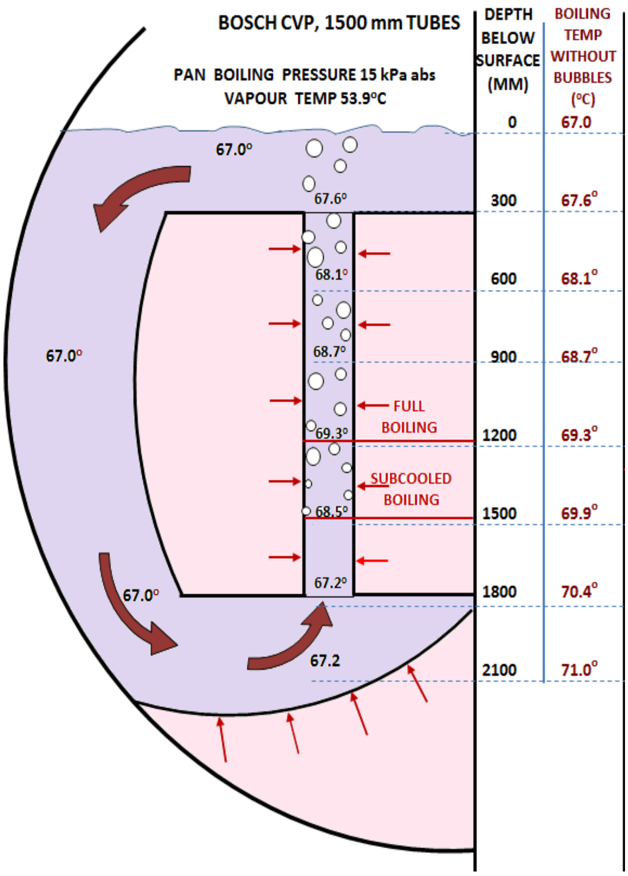

Figure 5. Boiling in a single tube; temperatures are ignoring the effect of bubbles on boiling point elevation (BPE).

- After exiting the top of the tubes, turbulence exposes the massecuite to the surface where it will flash down to the surface temperature of 67°C. Ignoring losses, it then remains at this temperature in the down take, until it reaches the heated floor zone.
- The area of this heated floor is equivalent to 4 to 6% of the walls of the tubes, so that some heating occurs before entering a tube, depending on the tube position. In Figure 5, a 0.2°C temperature increase is assumed.
- At tube entry, the massecuite is below its boiling temperature so is not boiling (‘subcooled heating’ in Rouillard’s terminology).
- As it climbs inside the tube, enthalpy (heat) is transferred into the massecuite, vapour bubbles are formed adjacent to the tube walls and subcooled boiling commences.
- The ebullition increases turbulence and heat transfer rates and full saturated boiling soon commences. As more and larger bubbles are formed, the process is accelerated.
- The vapour bubbles reduce the bulk density and hydrostatic boiling point elevation, so that boiling is initiated at a lower level within the tube.
- The difference in density between that in the down take and in the tube promotes circulation. The volume occupied by the bubbles also increases turbulence and the velocity up the tube, increasing the heat transfer rate. Because of surface tension, massecuite adheres to the tube walls as a rising film.

If jigger steam or superheated (flashing) feed is injected below the calandria, this will provide additional bubbles in the tube, further reduce the hydrostatic head and improve circulation.

Boiling in a CVP with 1700 mm tubes

To assess the effect of longer tubes, the above analysis is compared to a 1 700 mm tube pan in Figure 6.

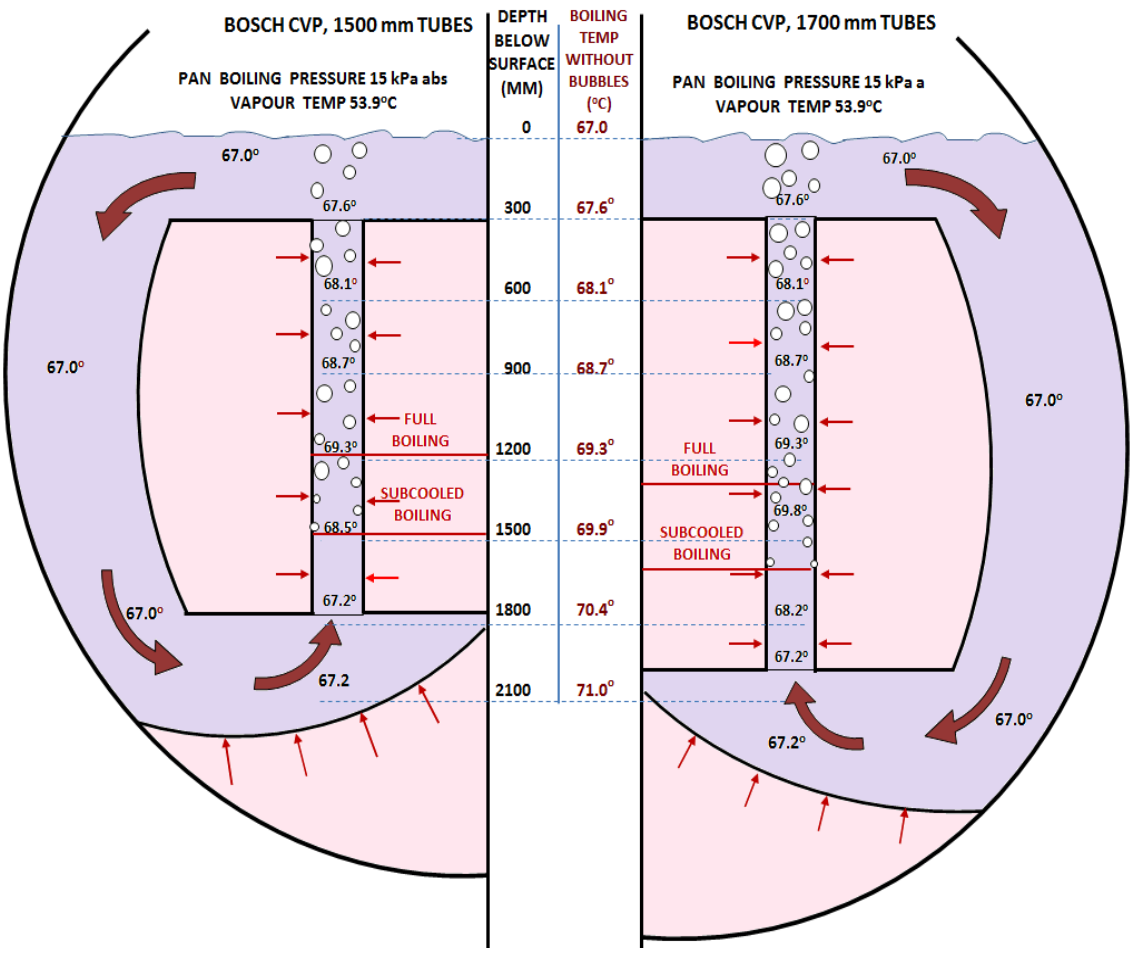

Figure 6. Comparison of boiling in 1500 mm and 1700 mm tubes.

In Figure 6, a 1 700 mm tube design is set opposite the 1 500 mm design analysed earlier. In the 1 700 mm tube pan on the right, the massecuite surface temperature is again 67°C. Again, it receives some heating from the floor, before it climbs up the 1 700 mm tube and is further heated by the calandria steam.

Compared to the 1 500 mm case:

- The hot tube walls start to form bubbles slightly further from the base of the tube (subcooled boiling), but further below the massecuite surface.
- Full boiling also commences further up the 1 700 mm tube, but below the level in the 1 500 mm tube.
- In total, more vapour bubbles are generated, the overall bulk density difference is greater, the circulation force is greater by more than the additional friction of the longer tube and the circulation rate is higher.
- The higher circulation velocity increases the HTC.
- The extra bubbles further reduce the hydrostatic BPE and thus increase the ΔT between calandria steam and massecuite.
- As in the 1 500 mm tube, if jigger steam and/or superheated feed is used, this will provide additional bubbles and enhance performance even more because of the longer tube.

For these reasons, a 1700 mm tube pan can be expected to perform better than the 1 500 mm tube pan. There are counteracting effects on HTC between an increase in BPE due to hydrostatic head and an increase in turbulence (or fluid circulation). The benefit of high turbulence (or better circulation) on HTC in a tube can reach a plateau, as observed by Walthew and Whitelaw (1996) and Yu *et al.* (2004) in evaporator tubes. However, in the typical CVP tube length range, it is likely that the effect of high fluid turbulence and natural circulation due to formation of more bubbles in the tube has a more dominant effect on HTCs compared to the effect of hydrostatic head.

Conclusions

Many variables influence measured HTCs, including massecuite quality (viscosity), brix profile along the pan, cleanliness of tubes, ΔP (because higher ΔP produces better circulation rates), throughput (Moor, 2007) and whether there is any circulation assistance such as jigger steam. As is to be expected, this results in wide ranges in the performances of CVPs of similar design (as seen in Appendix C), but not too wide to draw some conclusions. The results quoted can be summarised thus:

Table 6. Comparative heat transfer coefficients (HTCs) from pilot plant and commercial continuous vacuum pans (CVPs).

	Short tubes			Long tubes	
Equipment	Rouillard	TH/FS	SRI	Rouillard	Bosch
Tube length	1400	1300-1500	1270-2000	1800	1700
Avge 'A' HTCs	278	413	428	277	473
Avge 'B' HTCs	159	212	278	159	359
Avge 'C' HTCs	77	115	162	70	212

From this it is seen that:

1. The HTCs from all the commercial pans are considerably higher than the figures for similar length tubes in Rouillard's pilot plant experiments – by so much that his conclusions should not be extrapolated for commercial designs.
2. The averages in Tables 3 and 4 suggest that 1 700 mm tubes perform at least as well if not better than tubes of 1 300-1 500 mm.

In selecting the optimum tube length for a particular CVP design, the most important considerations should be a good circulation profile, an acceptable aspect ratio, an appropriate heating surface to volume ratio and a high HTC. Selecting short tubes 'to minimise the BPE at the bottom of the tube' is not a valid consideration.

REFERENCES

Bhagwat SM and Ghajar AJ (2012). Similarities and differences in the flow patterns and void fraction in vertical upward and downward two phase flow. *Experimental Thermal and Fluid Science* 39: 213-227.

Broadfoot R (1999). Energy efficient design and operation of continuous pans. *Proc Aust Soc Sug Cane Technol* 21: 373-380.

Broadfoot R (2015). Tube lengths in SRI continuous pans. Personal communication.

Broadfoot R, Rackemann D and Moller D (2018). Why the emerging strong interest in vertical continuous pans? *Proc Aust Soc Sug Cane Technol* 40: 512-525.

Kruger GPN (1983). Continuous 'A' pan boiling trial at Maidstone sugar factory. *Proc S Af. Sug Technol Ass* 57: 46-51.

Moor BStC (2007). Successful Innovation: Results from the Bosch continuous pan. *Proc Int Soc Sug Cane Technol* 26: 1669-1675.

Moor BStC (2008). Modern sugar equipment assisting cogeneration. *Proc S Afr Sug Technol Ass Workshop* 81: 1-13.

Peacock SD (2000). Mathematical modelling of climbing film evaporators. MSc Thesis, University of Natal, Durban, 153 pp.

Rein PW (2016). *Cane Sugar Engineering*. Bartens, Berlin. 2nd Edition, pp 411, 426, 439.

Rein PW and Msimanga MP (1999). A review of continuous pan development in the southern African sugar industry*. Proc Int Soc Sug Cane Technol* 23: 124-136.

Rouillard EEA (1985). Massecuite boiling. *Proc S Afr Sug Technol Ass* 59: 43-47.

Saska M (2002). Boiling point elevation of technical sugarcane solutions and its use in automatic pan boiling. *Int Sug J* Vol 104, No. 1247*:* 500-507.

Walthew DC and Whitelaw RW (1996). Factors affecting the performance of long tube climbing film evaporators. *Proc S Afr Sug Technol Ass* 70: 221-224.

Yu H, Sheikholeslami R and Doherty OS (2004). The effects of operating parameters on composite fouling in sugar mill evaporators by calcium oxalate and amorphous silica. *Proc Aust Soc Sug Cane Technol* 26: 14 pp.

Appendix A
Calculation of Boiling Point Elevation (BPE)

In a boiling massecuite, the boiling temperature is higher than that of water at the same pressure due to two effects:

- BPE due to dissolved solids in the solution (BPE_{sol})
- BPE due to hydrostatic head (BPE_h)

In this paper, the BPE_{sol} for the Rouillard results and the Bosch CVPs have been calculated according to the formula established by Saska (2002). The BPE_{sol} is that which applies to the mother liquor or Nutsch component of the massecuite, i.e. the massecuite minus the crystals. Saka's formula is:

$$BPE_{sol} = 0.1660 \times \left[\frac{W_{DS}}{100-W_{DS}}\right]^{1.1394} \times \left[\frac{273 + t_{bw}}{100}\right]^{1.9735} \times \left[\frac{Q}{100}\right]^{0.1237}$$

where W_{DS} = mass concentration of dissolved solids (g/100 g solution)
T_{bw} = boiling point of water
Q = purity.

Thus, for a massecuite with Nutsch dissolved solids/water of 86%, water boiling temperature of 54.0°C (15 kPaa) and Nutsch purity of 57%, BPE_{sol} = 12.7 K.

True DS% is generally a little less than the commonly quoted refractometric Brix%, particularly at lower purities. Using Brix% instead of DS% in the formula would slightly overstate the BPE_{sol}. Therefore, where only Brix figures have been available, these have been converted to DS using a typical pol/ sucrose ratio for the purity concerned.

The BPE_h for the Bosch pans has been calculated at the top tube plate, assuming the normal boiling level of 0.30 mm above the top tube plate. Thus, for example:

Density of massecuite		1440 kg/m^3
Pan boiling pressure (in vapour space)		15.0 kPa abs
Additional pressure at top tube plate		= 9.81 * 1440 * 0.3/1000
		= 4.2 kPa
Actual boiling point of water is that at		= 19.2 kPa abs
And BPE_h	= Water Boiling point at 19.2 kPaa – Boiling point at 15 kPaa	
	= 59.2°C – 54.0°C	= 5.2 K
For this pan,	BPE_{sol} + BPE_h	= (12.7 + 5.2) K = 17.9 K

Appendix B
Example of an HTC calculation

This example is for a 280 m^3 B CVP, but the methodology is generally applicable.

It is assumed that the massecuite condition progresses linearly from seed at pan inlet to the final massecuite. 'Average' dissolved solids% and purities are thus the average of seed and final massecuite values.

Inputs

Pan boiling pressure	15 kPa abs
Seed brix	90%
Final massecuite Brix	94.5%
Average Nutsch DS/water ratio in the pan	6.47
Average Nutsch purity in the pan	69.0%
Boiling head above tubes	0.3 m
Supply steam pressure to calandria (at saturated)	120 kPa abs
Steam pressure inside calandria	82 kPa abs
Calandria heating surface	2724 m^2
Condensate flow from pan	25200 kg/h

Calculation

Vapour temperature in pan (from pressure)	54.0°C
Pressure at top tube plate (0.3m below surface)	19.3 kPa abs
Water boiling point at this pressure	59.3°C
BPE of solution (per Saska)	13.9 K
Boiling temperature of massecuite at top of tube	= (59.3 + 13.9) = 73.2°C
i.e. Combined BPE_h and BPE_{sol} is	= (73.2 - 54.0) = 19.2 K
Temp. of satd. steam/condensate in calandria	94.2°C
Hence ΔT between steam and massecuite	= (94.2 – 73.2) = 21.0 K
Specific enthalpy of supply steam (at sat temp) h_g	2683 kJ/kg
Specific enthalpy of condensate h_f	394.5 kJ/kg
Enthalpy transferred to massecuite	= (2683 – 395) = 2288 kJ/kg
Heat transfer coefficient	= 2288*25200/3600/2724/21.0 kJ/s.m^2.K
	=280 W/m^2.K

Appendix C
Ranges and averages of T-H/FS and Bosch HTCs

Table 7. HTCs for Tongaat Hulett CVPs (W/m^2.K).

Massecuite grade	**A**	**B**	**C**
Maximum	492	254	170
Minimum	304	173	65
Average for TH/FS pans	**413**	**212**	**115**

Table 8. HTCs for Bosch 1.7m tube CVPs (W/m^2.K).

Massecuite grade	**A**	**B**	**C**
Maximum	592	461	258
Minimum	300	279	177
Average for Bosch pans	**473**	**359**	**212**

Appendix D
Rouillard's pilot plant HTCs with high heads (batch pan conditions)

The 'high head' (870-940 mm) evaporation results from Rouillard's (1985) Tables 2, 3 and 4 have been converted to HTCs and are reproduced in Table 1 below. The figures for 1.8 m tubes are omitted as this is too long for satisfactory batch pans.

A Massecuite BPE from solution & 0.91 m head 23.0K s = Steam, v = Vac. vapour

Pressure kPaa		Temps. °C		Evaporation Rate kg/m².h			HTC (W/m².K)		
P_s	P_v	T_s	T_v	0.6m	1.0m	1.4m	0.6m	1.0m	1.4m
195	9	119	44	47	36	37	**549**	**421**	**432**
125	9	106	44	31	28	26	**495**	**447**	**415**
195	25	119	65	28	20	19	**546**	**390**	**370**
125	25	106	65	12	13	11	**415**	**450**	**380**
							501	**427**	**400**

B Massecuite BPE from solution & 0.91m head 25.2K

Pressure kPaa		Temps. °C		Evaporation Rate kg/m².h			HTC (W/m².K)		
P_s	P_v	T_s	T_v	0.6m	1.0m	1.4m	0.6m	1.0m	1.4m
195	9	119	44	19	13	15	**232**	**159**	**183**
127	9	107	44	16	11	11	**267**	**183**	**183**
195	20	119	60	15	12	10	**269**	**215**	**179**
127	20	107	60	11	9	6	**321**	**263**	**175**
							272	**205**	**180**

C Massecuites BPE from solution & 0.91m head 27.1 K

Pressure kPaa		Temps. °C		Evaporation Rate kg/m².h			HTC (W/m².K)		
P_s	P_v	T_s	T_v	0.6m	1.0m	1.4m	0.6m	1.0m	1.4m
180	10	117	46	8.2	7.0	6.4	**115**	**98**	**90**
130	10	107	46	6.7	6.1	4.3	**122**	**112**	**79**
							119	**105**	**84**