DESIGN AND EVALUATION OF A REFINED PAN STIRRER AT HULETT REFINERY

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Abstract

Test work was conducted at Hulett Refinery to establish the benefits of using mechanical stirrers in ribbon calandria pans boiling white sugar. Initially tests were carried out using an Ekato 5 bladed Kaplan type stirrer and immediate benefits in lower colour formation and higher yields were found. Increasing the speed of this stirrer, up to the limit of the shaft design, gave a further improvement in results.

In order to improve performance further, a stirrer with a 4 bladed adjustable pitch helical screw propeller was designed to maximise flow rather than back mixing. This, together with a more rigid tubular section shaft allowed experimentation with different blade pitches and rotational speeds up to the maximum power output of the drive motor.

The paper discusses the logic behind the design used and provides the results obtained during evaluation testwork. It is concluded that the best results in a ribbon calandria pan were obtained with a 4 bladed helical screw impeller, with a pitch ratio of 0,8 and a speed of 147 rpm. The testwork shows that a 60% reduction in colour formation and an 8% increase in crystal yield were achieved when compared with an unstirred pan.

Introduction

During 1987, an Ekato 5 bladed Kaplan type stirrer was installed in No. 2 refined pan at Hulett Refinery. The two primary objectives were firstly, to allow the use of exhaust 125 kPag steam instead of 330 kPag steam without sacrificing production capacity and secondly, to reduce the conglomerate level of sugar. These improvements would allow an optimisation of the Refinery steam balance and a reduction in conditioning time giving increased bulk silo capacity.

Although these objectives were not met, benefits in lower colour formation during boiling and higher yields were found. Modifications were made to allow the stirrer speed to be varied to establish the optimum speed.

A new stirrer of the marine helical screw type with variable pitch was designed. This was to allow higher speeds and maximise axial flow as opposed to the mixed flow characteristics of the Kaplan turbine.

An Ekato Kaplan stirrer was fitted to a newly installed tubular calandria pan for use on 4th boilings. The stirrer was of greater diameter than those in the ribbon calandria pans and consequently operated at lower rpm.

This paper discusses the logic behind the design and the results obtained during evaluation. The performance of the unstirred pan, the Kaplan stirrer and the helical screw stirrer are compared under a range of conditions.

Theory

Rheology

Determination of massecuite viscosity is important when designing stirrers for pans. With low grade boilings this can be done with a fair measure of accuracy (Austmeyer and Kipke[1]). However in refined sugar pans with up to 60% crystal content, the massecuite shows a greater degree of non-Newtonian behaviour. Some information is available of estimated viscosities of refined massecuites. These (Hill *et al*[3]) are as follows:

Table 1
Typical Massecuite Viscosities (Pa.s)

Refined Sugar Boilings		5
Recovery Sugar Boilings	1st Crop	17
	2nd Crop	100
	3rd Crop	3 000

A Newtonian fluid is one in which the viscosity of the fluid remains constant with variation in shear rate. For an impeller moving in a fluid, this would mean that the viscosity would remain constant as the speed of the impeller changes.

One type of non-Newtonian behaviour where the viscosity of the liquid decreases with increasing shear rate is called pseudoplastic. Initial results obtained with varying the speed of the Kaplan impeller could be explained by this type of behaviour. Advantage of this observation was taken into consideration in the design of a high speed helical screw impeller.

Kaplan vs Helical screw

The blades of the Kaplan turbine tend to be wide, with rectangular edges and with the same blade angle at the root and the tips. The model originally fitted at the Refinery had three planes as shown in Figure 1.

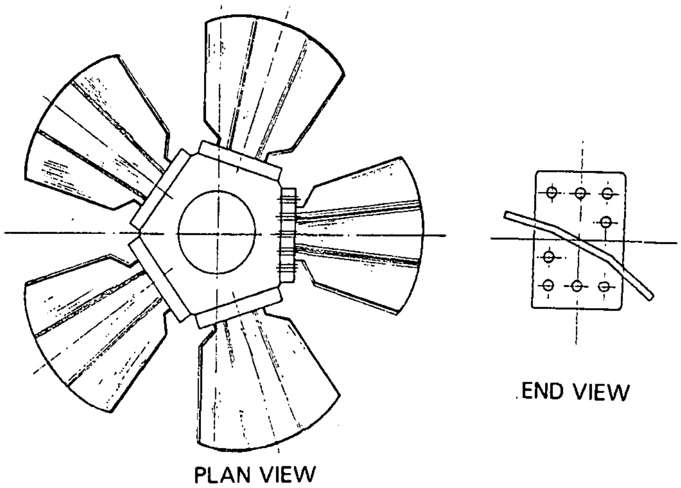

FIGURE 1 Kaplan Turbine Impeller

This design gives a mixture of axial and radial flow and promotes back mixing in the downtake. It produces excellent thrust and flow characteristics at low speeds. The Kaplan impeller is suited to tubular calandria pans where there is relatively high resistance to flow. It finds widespread use in both refined and low grade boilings.

The helical screw impeller has narrower elliptical shaped blades that have a greater angle at the root than at the tips. This gives a higher proportion of axial flow and performs better at higher speeds (Kuijvenhoven[5]).

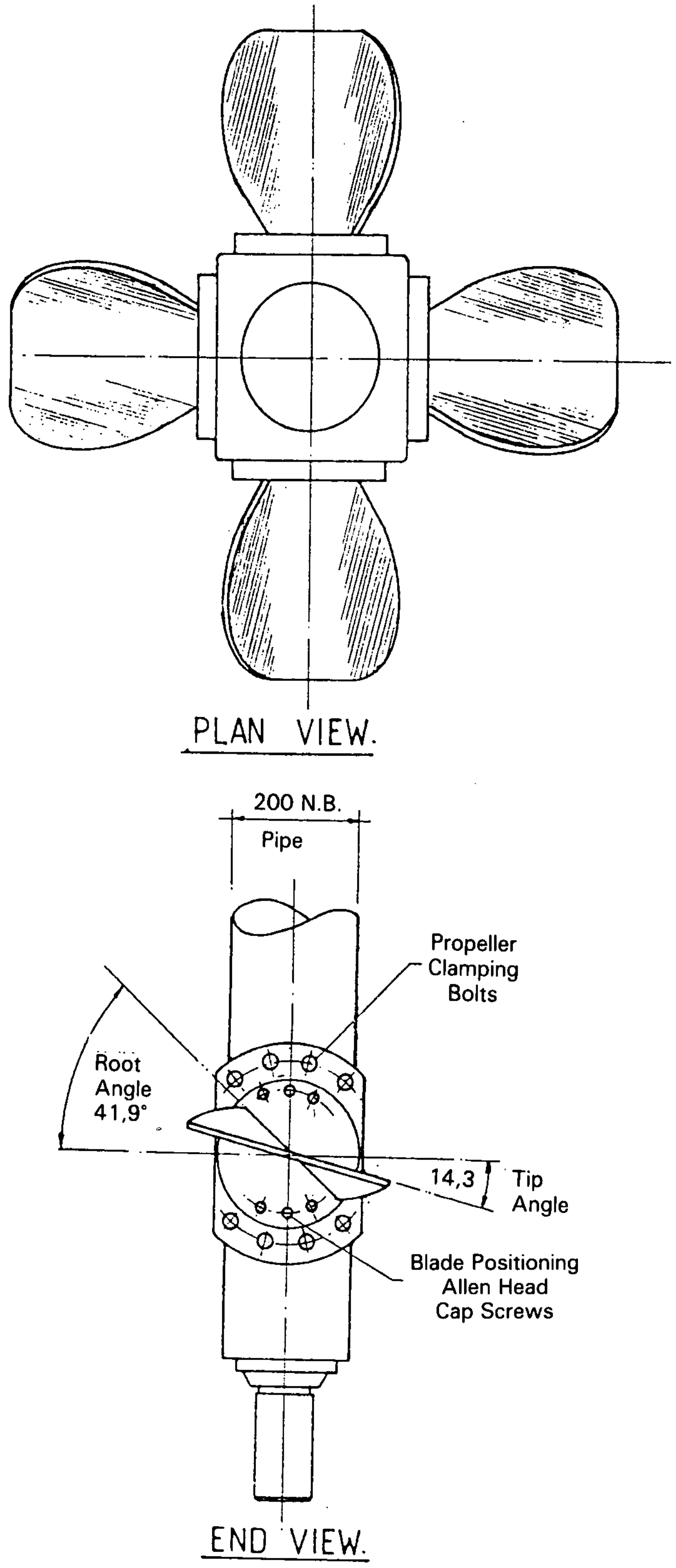

FIGURE 2 Helical Screw Impeller

According to Kuijvenhoven,[5] at low speeds and with a Reynolds No. less than 5 000, the Kaplan turbine is better than the helical screw with regard to flow. At higher speeds, with relatively low massecuite viscosities, the flow characteristics of both impellers become almost the same. However, the power requirement for the helical screw is substantially lower. With a given amount of power available, increasing the speed of the helical screw will give maximum axial flow per unit power input.

Table 2

Reynolds No. for Refined Massecuites at Different Impeller Speeds

Speed (rpm)	*Reynolds No.*
90	2 860
120	5 080
135	6 420
147	7 620

Density of massecuite taken as 1,4 kg/m^3
Viscosity of massecuite taken as 5 Pa.s

It can be seen from Table 2 that the region of operation at the higher speeds of 135 and 147 rpm is where the helical screw impeller has an advantage over the Kaplan impeller with regard to axial flow per unit of power.

Tip Clearance

Kipke[4] shows that at a tip clearance of 5% of the impeller diameter, the volumetric pumping efficiencies of both types of impeller are the same. The Kaplan impeller gives a fairly constant rate up to a clearance of 20% of the impeller diameter, whereas the marine impeller loses efficiency rapidly as the clearance increases. At a tip clearance of 20% of the impeller diameter the marine impeller has lost half of its pumping efficiency.

The Huletts impeller has a tip clearance of 6,9% of the impeller diameter, the actual clearance being 82,5 mm and the diameter 1 200 mm.

The ability to operate with large tip clearances explains why Kaplan stirrers can be designed without a bottom bearing, while marine impellers need steadying at the bottom of the shaft.

Materials and Methods

Scope

The evaluation extended for more than a year covering a range of massecuites with and without stirrers. The main variables included stirrer type, stirrer speed, blade pitch, steam pressure and sugar grade.

Initially the performance of pans 1 and 2 without stirrers was monitored. Both pans were boiling 1st sugar from the same feed material and these data gave a base level from which to measure improvements. Fourth boilings were also evaluated in pans 3 and 4 without stirring.

Pan 2 was then fitted with a Kaplan stirrer running at 120 rpm and a series of tests using different steam pressures was run. The feed inlet in pan 2 was modified after these tests to feed directly into the downtake above the impeller in order to improve mixing. The stirrer was then tested at different speeds (68, 120 and 133 rpm).

The Refinery Engineering Department then designed and fitted a helical screw stirrer to pan 2 in place of the Kaplan stirrer. This stirrer was tested at different speeds (120, 133 and 147 rpm), and coarse and fine blade pitches. All the testwork on the Huletts stirrer was on 1st boiling sugar and with the normal steam pressure of 330 kPa.

A tubular calandria pan (pan 5) was installed at the Refinery during the test period. The pan was fitted with an Ekato stirrer running at 60 rpm and was designed to boil 4th sugar using low pressure steam. The performance of this pan and stirrer combination was measured.

Pan and Stirrer Details

Pans	*Pan 2*	*Pan 5*
Strike Volume (m^3)	52	43
Diameter (m)	4,4	4,8
Calandria	Ribbon	Tube
Number of ribbons/tubes	12	1 112
Downtake Diameter (m)	1,37	1,87
Calandria open area (m^2)	9,0	6,0
Downtake open area (m^2)	1,7	3,1
Heating Surface (m^2)		
Elements	181	263
Botton Jacket	7	
Side Jacket	113	
Total	201	263

Stirrer Motor

Dual speed, high speed 1450 rpm/55 kw and low speed 960 rpm/38 kw. The motor was set to trip at 70 Amps in both cases.

Gearbox

Radicon CS17, 12,5:1
1000 Nm at 80 rpm, V-belts from motor to gearbox.

Stirrers (pan 2)

	Ekato	*Huletts*
Type	Kaplan Turbine	Helical Screw
Shaft Length (m)	12,3	12,3
Diameter (mm)	140	200
Type	Solid	Pipe
Blades Number	5	4
Diameter (m)	1,2	1,2
Pitch	Fixed	Variable

Stirrer Design

In choosing a helical design, the following points were considered:

- In the Huletts ribbon calandria pans, the ratio of the area of the downtake to the open area of the calandria is 0,18. This is a much lower ratio than that of a typical tubular calandria pan of about 0,5. Maximising flow through the downtake is thus an important requirement, especially at the end of the boiling cycle.
- The viscosity of refined massecuites at about 5 Pa.s is relatively low compared to low grade massecuites
- The non-Newtonian behaviour of the massecuite as described previously indicated a design for maximum rotational speed.

Helical Screw Design

Information on the design of the impeller was obtained from design data for marine impellers (Baker[2]).

The best efficiency is obtained with three blades for moderate loading and four blades for heavy loading. A choice of four blades was made. A width to diameter ratio of 0,3 at 0,7 diameter was chosen, giving a blade width of 370 mm at 0,7 diameter. This is substantially thinner than a Kaplan turbine.

A diametral pitch of 0,8 was chosen and provision was made for varying the pitch exactly 10° either way.

Table 3
Impeller Pitch Values

	−10°	Normal	+10°
Tip Angle	4,3°	14,3°	24,3°
Root Angle	31,9°	41,9°	51,9°
Diametral Pitch	0,45	0,8	1,35

During the tests only two blade positions were used, the normal position referred to as "fine" and the + 10° setting, referred to as "coarse".

Blade Profile

The blade profile chosen was that of a distorted ellipse given by:

$$Bw = \left[\frac{1}{2} + \frac{2d}{D}\right] + \text{Width of Elliptical Outline}$$

As the propeller would be working inside the downtake, the elliptical outline of the blade tip was truncated, making the blade width 200 mm at the tip. From this a formula of blade width was derived:

$$Bw = 205\left[\frac{1}{2} + \frac{2d}{D}\right] + \left[1 - \frac{(325 - d/2)^2}{(300)^2}\right]^{\frac{1}{2}}$$

Drive Shaft Design

One of the limitations of the first stirrer installation was the flexibility of the shaft. As the speeds were increased the shaft displayed a tendency to whip. Calculations were done for the transverse vibration of the shaft, considering it as a simply supported beam and neglecting the weight of the impeller. The natural frequency at the lowest mode was found to be 15 Hz, which represents a rotational speed of 143 rpm.

The shaft was redesigned using a thick walled pipe. Although, in theory, a 150 mm nominal bore schedule 80 pipe would have been sufficient, it was decided to use a 200 mm schedule 80 pipe. This had a natural frequency of 28,5 Hz (276 rpm) at the lowest mode and gave a safety margin to allow an additional increase in shaft speed.

Kuijvenhoven[5] states that the tip speed should be less than 10 m/sec to prevent crystal damage. This would limit a 1200 mm diameter stirrer to 150 rpm.

Installation

The general arrangement of the Hulett stirrer in pan 2 is shown in Figure 3. The impeller position is wholly within the downtake. The shaft is supported below the impeller to restrict any sideways movement.

The feed pipe is not shown, but is 150 mm in diameter and runs across the top of the calandria with a short bend into the downtake. This is to ensure that the feed is thoroughly mixed into the massecuite as quickly as possible.

Results and Discussion

Power Consumption

A typical power consumption curve of the stirrer during a panboiling cycle is shown in Figure 4.

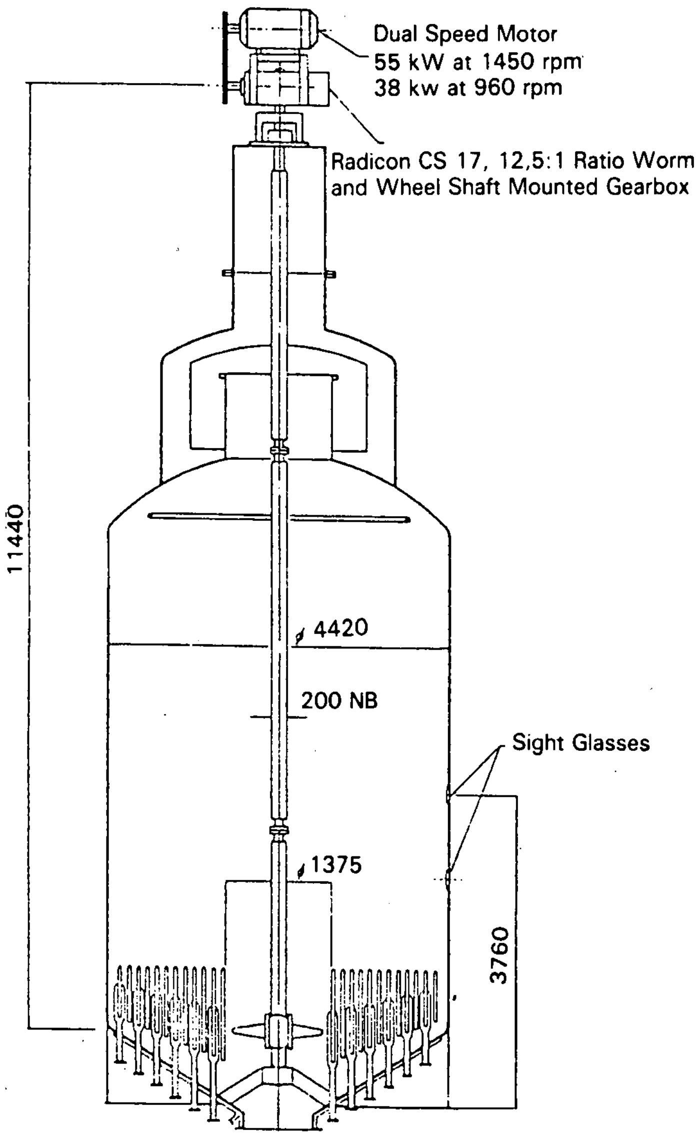

FIGURE 3 General Arrangement of the Hulett Stirrer

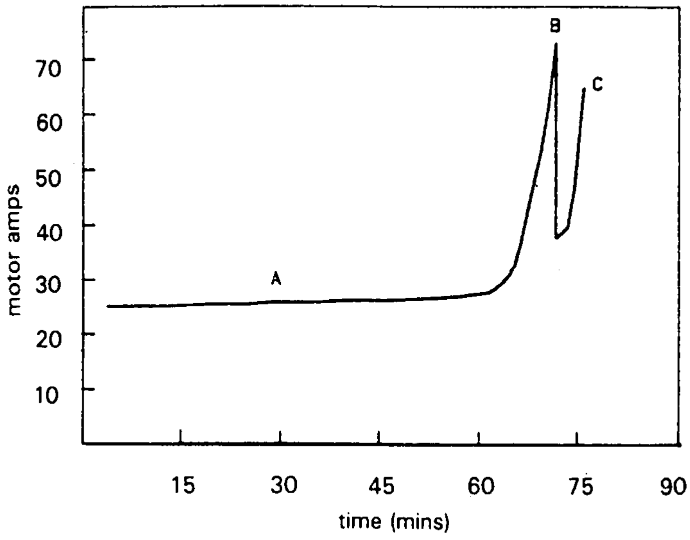

FIGURE 4 Typical Power Consumption Curve

The main features are a fairly flat section (A) where the power requirements during the crystallisation process are more or less constant, followed by two sharp peaks (B) and (C). The peaks are during the brixing up stage. The point (B) is where the motor trips from high speed down to low speed, reducing the current drawn by the motor and (C) where the final low speed trip occurs, terminating the boiling.

A summary of the results found for the two impellers is given in Table 4.

Table 4
Stirrer Power Consumption

Stirrer	Pitch	(A) Boiling kW	(B) HS Trip kW	(C) LS Trip kW
Kaplan				
68 rpm		6,3	58,1	46,8
120 rpm		22,6	55,6	48,0
133 rpm		32,4	57,7	48,5
Huletts				
120 rpm	fine	14,5	63,4	44,0
133 rpm	fine	18,3	63,4	45,9
133 rpm	coarse	31,8	63,4	45,1
147 rpm	fine	25,0	63,4	43,1
147 rpm	coarse	42,0	64,4	52,7

As the stirrer speeds are increased, the power consumption during the crystallisation period shows a steady increase as would be expected. The difference between the two high pitch settings on the Huletts stirrer is very marked, the coarse pitch having a greater power consumption at the same speed.

The results found during the evaluation of the stirrers show that in all instances and contrary to expectations, the massecuite brix was slightly higher at higher rational speeds for any given impeller type and pitch.

Taking into considerations the fact that the power drawn is approximately proportional to the cube of the speed and the power is limited due to the motor trip, the higher strike brixes indicate lower viscosity of the massecuite at higher stirring speeds.

This observation was borne out by the observed time interval between the stirrer tripping out at high speed and low speed. Once again, contrary to expectation, the time interval at low speed increased with increasing impeller speed. This indicates pseudoplastic non-Newtonian behaviour.

Typical results for the Hulett stirrer are:-

stirrer speed	rpm	120	133	147
time from B to C	secs	63	80	99

Colour Formation

With the use of high pressure steam in the ribbon calandria pans, high temperatures are reached on the heating surfaces. Poor circulation can cause considerable colour formation. Colour profiles of the massecuite during the boiling cycle with and without a stirrer were measured in order to find where in the cycle the colour was being formed.

Several points are clear from Figure 5. Very little colour formation occurs before the feed is stopped and the pan is tightened. The colour changes during this part of the boiling cycle are almost entirely due to variations in the feed colour. Colour formation begins immediately the brixing up period starts, with no stirrer, and continues at an increasing rate until the pan is emptied. With a stirrer colour formation is minimal during brixing up.

Reduced colour formation is apparent in moving from no stirrer to the Kaplan stirrer and then to the Huletts stirrer with increasing rotational speeds.

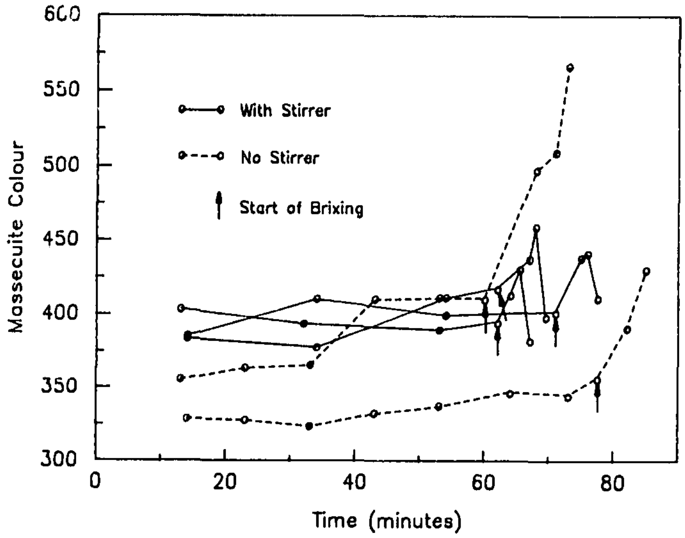

FIGURE 5 Massecuite Colour Profiles During Boiling

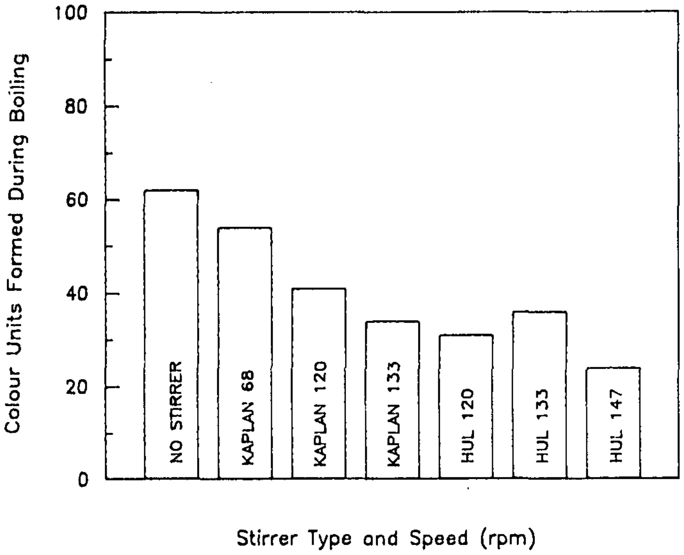

FIGURE 6 Colour Formed During Boiling

Sugar Quality

The colour of the sugar produced is probably the most important quality factor for the Refinery. The total colour is a combination of colour occluded in the crystal and colour in the syrup film on the crystal surface.

Table 5
Sugar Colours and Colour Removals

Stirrer	Feed Colour	Sugar Colour		% Colour Rem.	
		Prod	Aff	Prod	Aff
No Stirrer	324	32	25	90,1	92,3
Kaplan 68	380	33	24	91,3	93,7
Kaplan 120	386	30	26	92,2	93,3
Kaplan 133	332	28	20	91,6	94,0
Huletts 120F	262	26	21	89,7	92,0
Huletts 133F	242	22	17	90,7	93,0
Huletts 133C	293	22	15	92,3	94,9
Huletts 147F	279	16	12	94,1	95,7
Huletts 147C	317	23	19	92,6	94,0

The varying feed colours make direct comparisons of the sugar colours difficult, but it is clear from Table 5 that a marked improvement in both product and affinated sugars has been achieved. This confirms the supposition that both types of colour, occluded and surface, are lowered by the use of stirrers.

The most significant factor in reducing colour transfer to the crystal is stirrer speed. At the same rpm there is little difference between the Kaplan and Huletts impellers on the basis of % colour removal.

One of the prime objectives was to improve the quality of the crystal produced without affecting the production rate. This is difficult as the high crystallisation rates in the white pans, especially immediately after seeding, encourage irregular crystal formation.

Table 6
Sugar quality for different stirrer combinations

Stirrer	MA microns	CV %	Conglom. Count %	Bulk Density g/l
No Stirrer	565	39	77	848
Kaplan 68	608	34	72	861
Kaplan	557	37	72	860
Kaplan 133	554	34	77	855
Huletts 120F	567	32	72	857
Huletts 133F	611	34	74	846
Huletts 133C	567	37	70	849
Huletts 147F	551	33	69	859
Huletts 147C	569	31	61	853

What can be seen from Table 6 is a trend towards lower conglomerate counts and CV values as the stirrer development proceeded. The Huletts stirrer with the blade set at the coarse angle seems to be the best configuration for minimum conglomerate formation, again with increasing speed also an important factor. The individual Pan Boiler's techniques will have had a far larger effect on the crystal quality than on the other measured parameters and this may explain why the improvements are not as clear cut as was hoped for.

Pan Yields and Massecuite Brix

At the same time as the colour profile results were being obtained, brix profiles were measured during the boiling cycle. These are shown in Figure 7.

While the stirrer appears to have very little effect in the early part of the boil, it increases the final strike brix. Stirring allows the viscosity to be increased past the point where natural circulation would stop. This enables heat transfer and thus evaporation to continue, giving a higher brix at strike.

The massecuite strike brixes obtained with unstirred pans were ± 89,5. This increased to ± 90,5 with the Kaplan stirrer and to ± 90,7 with the Huletts stirrer. Previous work at the Refinery has shown a strong relationship between strike brix and pan yields. Without stirrers it was not possible to apply this fact to improve yields as poor circulation at high brixes caused low heat transfer, local overheating and excessive discharge times.

The use of stirrers overcomes all of these problems and a considerable increase in crystal yields was found as the test programme proceeded. This is illustrated in Figure 8.

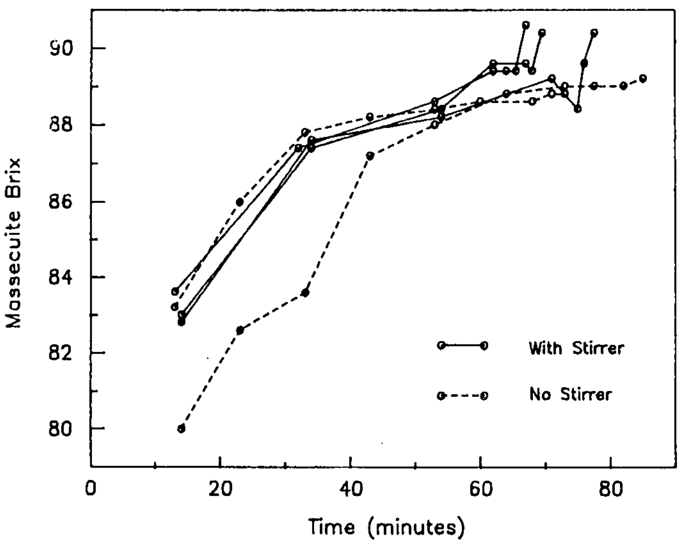

FIGURE 7 Massecuite Brix Profiles During Boiling

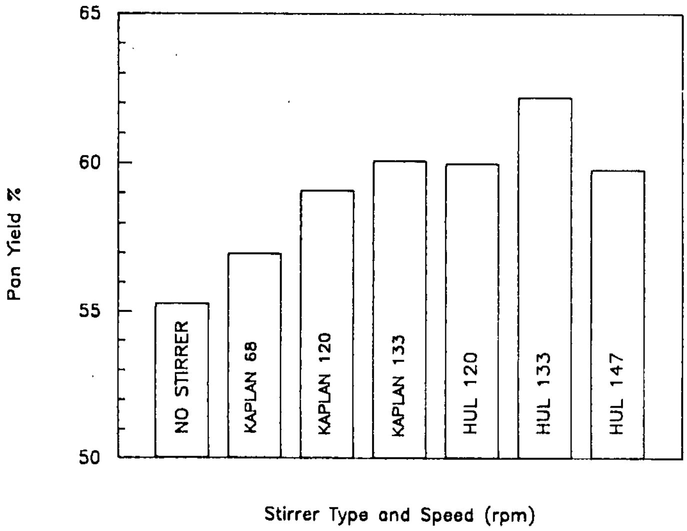

FIGURE 8 Pan Yields

Evaporation Rates and Heat Transfer Coefficients

Table 7
Evaporation Rates and HTC's

Stirrer	Steam Pr kPa	Feed Brix	Av Evap Rate kg/m²/hr	Av HTC kW/m²/°C
No Stirrer	310	76,3	56,4	0,78
Kaplan 68	349	75,6	66,4	0,87
Kaplan 120	336	76,5	62,6	0,83
Kaplan 133	346	75,7	64,8	0,85
Huletts 120F	310	74,8	64,9	0,90
Huletts 133F	311	74,2	65,2	0,90
Huletts 133C	309	74,5	64,0	0,89
Huletts 147F	321	73,5	67,2	0,95
Huletts 147C	345	74,0	68,6	0,94

Table 7 shows the improvement in heat transfer due to the use of mechanical circulation. The Huletts stirrer gives better heat transfer than the Kaplan stirrer, which could be due to the higher proportion of axial flow and less back-mixing. Faster stirrer speed also improves matters which again is probably due to more efficient removal of heat from the calandria surface by the faster circulating massecuite.

4th Boiling Sugar

A tubular calandria pan (Pan 5) was installed at the Refinery to boil 4th boiling sugars. Previously these sugars had been boiled in one of the ribbon calandria pans. The pan was fitted with a Kaplan type stirrer of 1800 mm diameter running at 60 rpm.

Tests on 4th Boilings were done in Pans 3 and 4 (ribbon calandria) without stirring and on Pan 5 with stirring. A direct comparison is difficult because of the very different conditions in the pans, but it does show the considerable improvement in Pan Yield and Colour Removal due to stirring and lower calandria temperatures.

Table 8
Comparison of 4th Boilings

	Pans 3 & 4	Pan 5
Steam Pressure (kPag)	280	28
Cycle Time (min)	117,3	197,5
Pan Yield (%)	51,0	59,5
Colour Removal (%)	94,5	95,5
Evap. Rate (kg/m²/hr)	47,8	20,1
HTC (kW/m²/°C)	0,70	0,79

The use of low steam pressure in pan 5 is made possible by the tubular calandria. The colour formation during the boiling cycle is about 10% higher in pan 5 and is due in part to the much longer cycle time.

Refinery Ash and Colour Balance

A refinery mass balance model was used to calculate the anticipated benefits to the refinery if all the pans were fitted with stirrers. The basic assumptions used in the model were:

Melt Rate	85 tons/hr
Colour removal over carbonatation	54%
Colour removal over ion exchange	65%
Colour gain over evaporation	5%

Table 9
Predicted Results With and Without Stirrers

	No Stirrers	Stirrers
Colours 1st Boiling Sugar	30	20
2nd Boiling Sugar	72	48
3rd Boiling Sugar	98	91
4th Boiling Sugar	158	148
Refined Sugar Blend	56	40
tons/hr 1st massecuite	101	98
2nd massecuite	48	42
3rd massecuite	23	18
4th massecuite	13	9
Ash % Return Syrup	2,1	3,3

There is a marked improvement in the colours of the white sugars, most noticeable in the 1st and 2nd boilings. The lesser effect in the 3rd and 4th boilings is due to the concentration effect of the higher pan yields.

The tonnage of refined massecuites is predicted to drop from 185 to 167 tons/hr, giving a considerable steam saving and increase in pan capacity.

Conclusions

1 The fitting of a mechanical stirrer to Pan 2 has led to a number of improvements in the performance of the pan in respect of sugar quality, exhaustion efficiency and sugar production rate.

2 The most consistent improvement in terms of sugar quality has been in lower colours. Reduced colour formation in the pan, and more homogeneous crystallisation conditions have lowered the colour in the crystal and in the syrup film. The colour removal has improved from 90% without a stirrer to 94% with the marine stirrer.

3 Sugar conglomerate counts and CV's have been reduced by about 20%, with the MA remaining unchanged.

4 Pan Yields and Heat Transfer Coefficients have improved and taken together are worth approximately 10% additional pan capacity.

5 A marine type helical screw impeller is the most suitable impeller for ribbon calandria pans boiling refined massecuites. This is due to the non-Newtonian behaviour of the massecuite at high brixes.

6 The most successful operating parameters for the helical screw stirrer were a rotational speed of 147 rpm and a blade root angle of 42° and tip angle of 14°.

7 A Kaplan turbine stirrer running at 60 rpm gave excellent results in a tubular calandria pan used for 4th boiling sugar, enabling low calandria pressure to be used and giving a high colour removal and a high yield.

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APPENDIX 1

Results Summary

	No Stirrer			Kaplan Stirrer			
	Pan 1	Pan 2	Pans 3/4	120 Rpm Side Feed Pan 2 LP	68 Rpm Centre Feed Pan 2	120 Rpm Centre Feed Pan 2	135 Rpm Centre Feed Pan 2
Stirrer	No	No	No	Yes	Yes	Yes	Yes
Data Sets	28	20	5	5	6	15	10
Sugar Boiling	1st	1st	4th	1st	1st	1st	1st
Steam on Time (min)	74,4	74,1	108,9	146,8	67,7	67,4	68,6
Strike Time (min)	6,3	7,4	3,5	3,9	6,8	5,6	5,4
Steam Pressure (kPa)	306	313	280	125	349	336	346
Strike Temp (°C)	83,9	82,8	87,2	82,5	84,2	82,7	84,1
Feed Brix	76,2	76,4	74,9	76,1	75,6	76,5	75,7
Massecuite Brix	89,4	89,9	92,0	90,7	90,3	90,3	90,5
Pan Yield (%)	55,2	55,3	51,0	57,0	57,0	59,1	60,1
MA	565	564	563	587	608	557	554
CV	38	39	32	32	34	37	34
Conglomerate Count	76	77	86	70	72	72	77
Bulk Density (g/l)	850	846	782	869	861	860	855
Feed Colour	328	319	3423	381	380	386	334
Massecuite Colour	390	380	3539	444	434	427	368
Sugar Colour	33	30	187	28	33	30	28
Colour formed in Pan	62	61	116	63	54	41	34
Colour Removal (%)	89,9	90,3	94,5	92,7	91,3	92,2	91,6
Evaporation Rate (kg/m²/hr)	55,6	57,1	47,8	29,9	66,4	62,6	64,8
Sugar Production (tons/hr)	31,1	29,8	19,7	16,8	33,0	35,2	34,7
H.T.C. (kW/m²/°C)	0,78	0,78	0,70	0,68	0,87	0,83	0,85

APPENDIX 1 (cont)

Results Summary

	Helical Screw Stirrer						Kaplan
	120 Rpm Draught Tube Pan 2	Fine 120 Rpm Centre Feed Pan 2	Coarse 135 Rpm Centre Feed Pan 2	Fine 133 Rpm Centre Feed Pan 2	Fine 147 Rpm Centre Feed Pan 2	Coarse 147 Rpm Centre Feed Pan 2	60 Rpm Centre Feed Pan 5
Stirrer	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Data Sets	12	8	9	7	9	8	11
Sugar Boiling	1st	1st	1st	1st	1st	1st	4th
Steam on Time (min)	74,9	72,4	77,2	78,6	79,4	75,5	170,7
Strike Time (min)	4,5	5,0	4,2	3,9	3,1	4,1	17,2
Steam Pressure (kPa)	309	310	309	311	321	345	28
Strike Temp (°C)	83,5	81,4	82,3	77,2	82,3	79,3	82,7
Feed Brix	75,8	74,8	74,5	74,2	73,5	74,0	74,1
Massecuite Brix	90,1	90,5	90,7	90,7	90,9	90,7	91,9
Pan Yield (%)	56,7	60,0	61,2	63,1	59,8	59,7	59,5
MA	557	567	567	611	551	569	577
CV	38	32	37	34	33	31	35
Conglomerate Count	78	72	70	74	69	61	82
Bulk Density (g/l)	839	857	849	846	859	853	789
Feed Colour	273	262	293	242	279	317	2478
Massecuite Colour	310	293	324	283	301	343	2610
Sugar Colour	24	26	22	22	16	23	112
Colour formed in Pan	37	31	31	41	22	26	132
Colour Removal (%)	90,9	89,7	92,3	90,7	94,1	92,6	95,5
Evaporation Rate (kg/m²/hr)	57,7	64,9	64,0	65,2	67,2	68,6	20,1
Sugar Production (tons/hr)	31,5	33,2	32,6	33,7	31,2	32,5	11,3
H.T.C. (kW/m²/°C)	0,80	0,90	0,89	0,90	0,95	0,94	0,79

APPENDIX 2

Calculation of Performance Parameters

$$\text{Pan Yield \%} = \frac{(\text{Nutsch Ash} - \text{Massecuite Ash}) \times 100}{(\text{Nutsch Ash} - \text{Crystal Ash})}$$

$$\text{Massecuite Production m}^3\text{/hr} = \text{Strike Vol} \div \text{Cycle Time}$$

$$\text{Crystal Production t/hr} = \frac{\text{M/C Production} \times 1.47 \times \text{Yield}}{100}$$

$$\text{Water Evap. tons} = \text{Strike Vol} \times 1.47 \times \left[\left[\frac{\text{Feed Brix}}{\text{M/C Brix}}\right] - 1\right]$$

$$\text{Evap Rate kg/m}^2\text{/hr} = \frac{\text{Water Evap} \times 1000 \times 60}{\text{Heating surface area} \times \text{Steam on Time}}$$

$$\text{H T C kW/m}^2\text{/°C} = \frac{\text{Evap Rate} \times \text{Vapour Enthalpy}}{(\text{Steam Temp} - \text{M/C Temp}) \times 3600}$$