IMBIBITION OPTIMISATION AT MOUNT EDGECOMBE

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Abstract

The Mount Edgecombe (ME) sugar mill normally operates at an imbibition level of about 300% on fibre in bagasse. At this level there is sufficient bagasse to supply fuel for the boilers. To optimise the imbibition, mass and steam balances, based on the 1991/92 season, were carried out at various imbibition levels. They show that at an imbibition level of 320% additional fuel in the form of coal would be required. Furthermore they show that when coal is considered as the only marginal cost an imbibition of more than 470% will no longer be cost effective.

Introduction

The South African sugar industry has the highest extraction in the world with close to 98% for the season. At the same time the imbibition levels are amongst the highest in the world at, on average, 375% on fibre in bagasse, with individual mills being well above 400%. This does not necessarily mean that the imbibition levels in South Africa are too high or that in other countries they are too low. There is no doubt that the optimum level is different for each country and even each mill and depends on factors such as the financial structure of the industry, length of season, the sugar price, the coal price etc. It is felt however that the present imbibition levels in South Africa might well be too high while at the same time the optimum levels are not known. The reason for the latter is most likely the difficulty in finding that optimum. This paper describes an attempt to calculate the optimum imbibition for the ME sugar mill with respect to the use of coal. It is assumed that the existing equipment and manpower are sufficient to process the present amount of cane irrespective of changes due to variations in imbibition, and that the only cost that must be offset by the production of any additional sugar and molasses is the cost of coal.

Factory mass balances are carried out and the effect of imbibition on the output of sugar and molasses is discussed. In addition factory steam balances are performed and the influence of imbibition on steam and fuel consumption is investigated. Finally the financial implications of increasing imbibition are considered showing the optimal imbibition at the current sugar price and the effect of that price on the optimal imbibition. All the factory balance data refer to the 1991/92 season whereas the financial data are based on current prices. The 1991/92 factory data have been used because the two subsequent seasons were abnormal due to severe drought.

Factory mass balance

The extraction plant at ME is a standard milling tandem consisting of seven mills. Prepared cane is fed into the first mill and subsequently into the following units. The expressed juice from the first and second mill is screened before it is pumped away as mixed juice for further processing. The juice from the other mills is pumped in front of the preceding mills as maceration juice. Imbibition water is applied just before the last mill. The boiling house at Mount Edgecombe is typical of most other South African factories. Mixed juice is heated before going to the clarifier. The clear juice is again heated and subsequently evaporated in a quintuple effect evaporator. Pan boiling is done using a standard three pan boiling scheme with A-sugar and C- or final molasses as output from the centrifigals. Figure 1 shows a diagram of a mass balance at 300% imbibition.

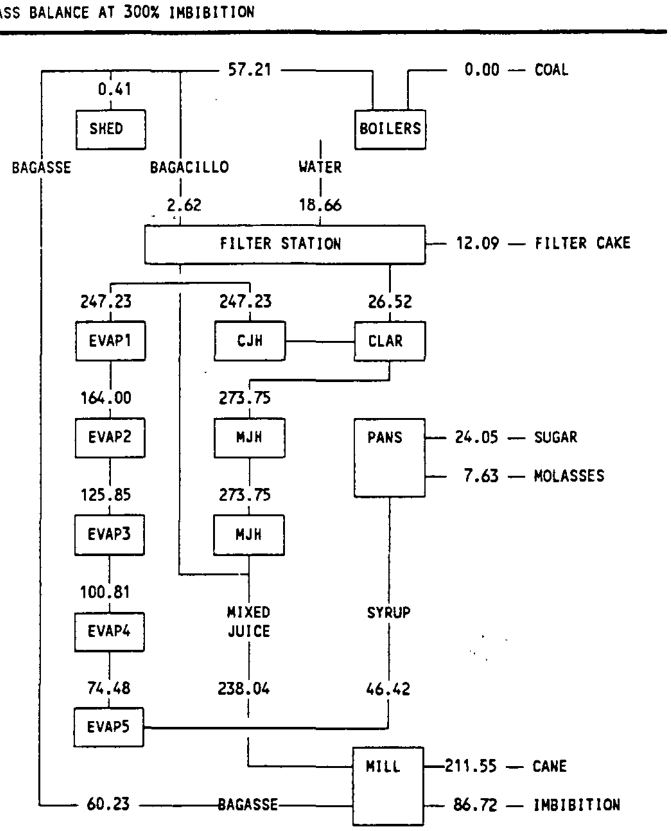

FIGURE 1 Factory mass balance.

Extraction

In order to analyse the effect of imbibition on extraction a base case mass balance was carried out on brix, fibre and water for each individual mill in the tandem. From the figures published in various reports it is easy to derive a mass balance for the tandem as a whole. The data required for a mass balance around each mill are not as easily available and are less reliable. The data concerned are the moisture % bagasse, the brix % bagasse and the fibre (suspended solids) % expressed juice for each mill. The first two are measured at the factory on a daily basis. For the fibre % expressed juice no figures were available and so some simple tests were performed. Samples of expressed juice were put through a standard sieving procedure using a 600 μm sieve. The mud that stayed behind in the sieve was expressed as a percentage of the juice. Control tests by Cane Testing Services showed that this mud consisted of roughly 90% liquid which was the figure used to derive the dry solids % expressed juice. The base case mass balance at 300% imbibition based on these figures is given in Appendix A. Although the mass balance results in a brix extraction it does not give a sucrose extraction. Lionnet (1981) however has established an empirical relationship between the two which has the following form:

Sucrose extraction

$$= \sqrt{1919 + 187{,}6 * \text{Brix extraction}} - \sqrt{1919}$$

The mixed juice purity was calculated from the information related to the brix as supplied by the mass balance together with the sucrose % cane (13,02) and the sucrose extraction (97,39) based on the above equation.

From the base case mass balance some efficiency parameters were calculated such as imbibition efficiency, separation efficiency and reabsorption coefficient. These coefficients are defined as:

$$\textit{Imbibition efficiency} = 100 * \frac{\textit{Brix \% liquid in juice}}{\textit{Brix \% liquid in cane}}$$

Separation efficiency

$$= 100 * \frac{\textit{Fibre \% cane - Fibre \% juice}}{\textit{Fibre \% cane}}$$

Reabsorption coefficient

$$= \frac{\textit{Fibre \% bagasse in discharge roll}}{\textit{Fibre \% bagasse after discharge roll}}$$

Mass balances for other imbibition levels were carried out assuming these parameters to be constant and independent of the level of imbibition. According to Murry and Holt (1967) this assumption is not entirely correct but is acceptable within certain limits and for the purpose of this exercise. Based on these assumptions, and as is to be expected, the tons brix in mixed juice increase but the brix % mixed juice decreases with increasing imbibition. The latter is due to an increase in the quantity of mixed juice. For the same reason the fibre % mixed juice will decrease. Figure 2 shows the brix % mixed juice and the fibre % mixed juice for imbibition levels from 0 to 500%.

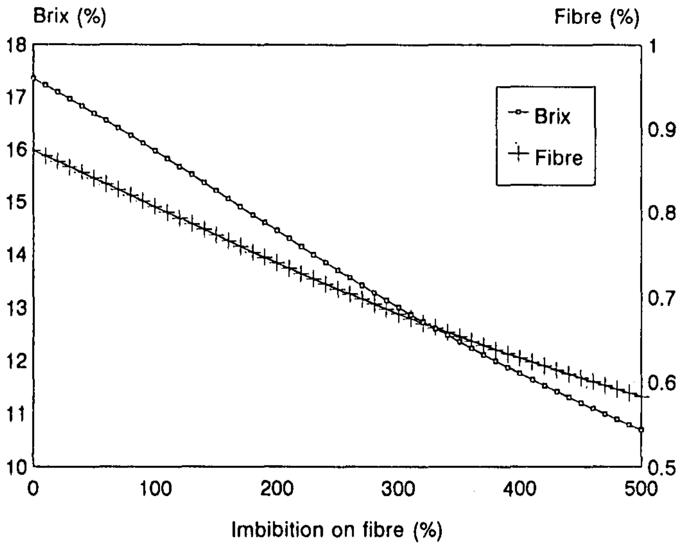

FIGURE 2 Brix and fibre % mixed juice.

With increasing imbibition the quantity of mixed juice will obviously also increase and be more dilute. This increases the load on the evaporators which is the main reason for the increase in exhaust steam. The purity of mixed juice shows a decrease with increasing imbibition. This is supported by practical experience where mixed juice purity drops as extraction increases. Figure 3 shows the quantity and purity of mixed juice for imbibition levels from 0 to 500%.

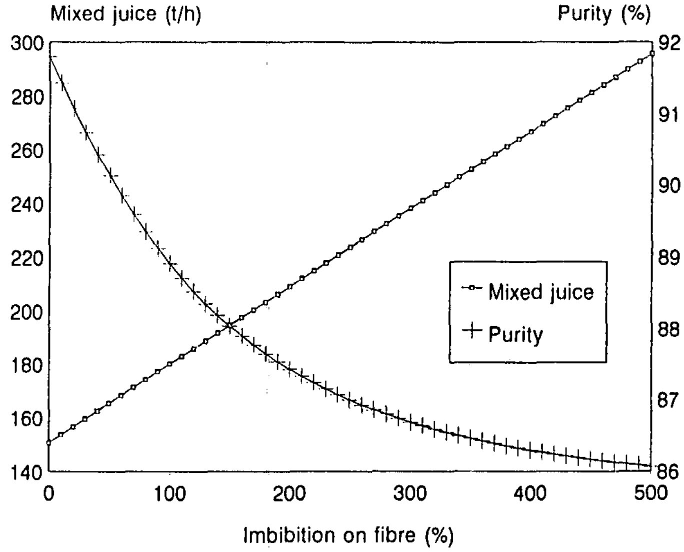

FIGURE 3 Tons and purity of mixed juice.

Boiling house recovery

Assuming unlimited evaporator capacity the brix % syrup (66,56) should be constant and independent of the level of imbibition. From a known mixed juice quantity and brix % mixed juice, the quantity of syrup can be calculated using a simple mass balance on brix. The purity difference between mixed juice and syrup is normally insignificant and can be ignored. The quantity and purity of syrup at various imbibition levels is given in Figure 4.

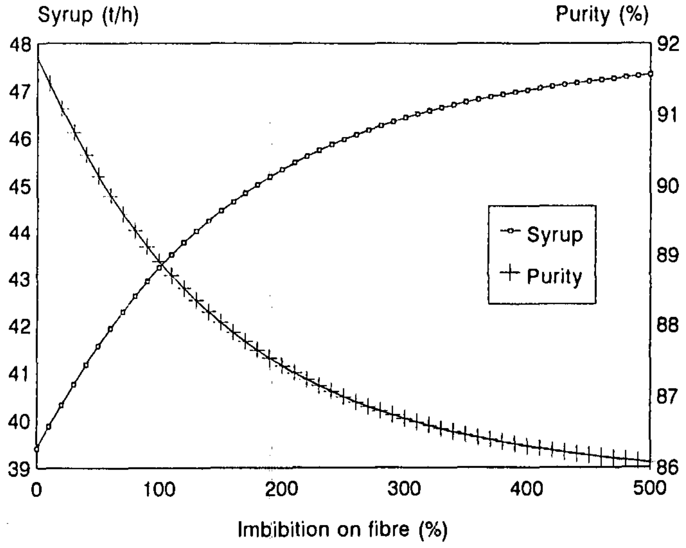

FIGURE 4 Tons and purity of syrup.

For comparative purposes the sugar quality and the final molasses quality should be constant and independent of the level of imbibition. This means that the sugar purity (99,40), the brix % sugar (99,90), the molasses purity (37,83) and the brix % final molasses (83,61) should all be constant. Under these conditions it is possible to calculate the quantities of sugar and molasses produced using a simple mass balance on sucrose and brix. To allow for undetermined losses the sugar output was reduced by 2%. The output in tons of molasses and sugar as a function of the imbibition is given in Figure 5.

The extraction increases with increasing imbibition but the boiling house recovery will decrease due to the drop in

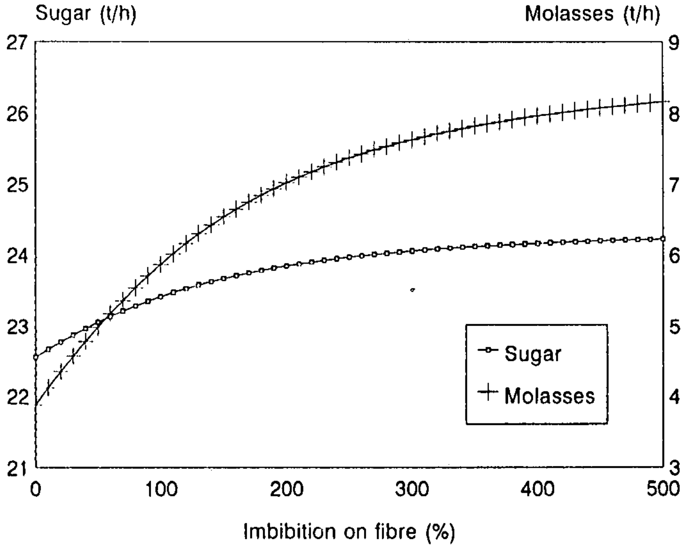

FIGURE 5 Tons of sugar and molasses.

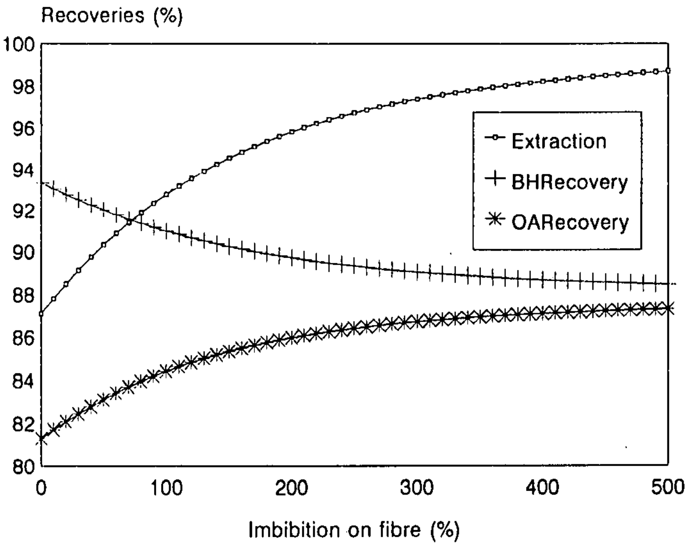

FIGURE 6 Extraction, boiling house recovery and overall recovery.

syrup purity. The combined effect expressed in the overall recovery shows an increase. The three recoveries are shown in Figure 6 as a function of the imbibition.

Factory steam balance

ME has six boilers of which four usually operate at the same time. Only one boiler can burn coal as well as bagasse. Live steam pressure is quite low at 1500 kPa(abs) at a temperature of 270°C. In the power station there is ample capacity with a mixture of back pressure and passout condensing turbines. The total required electrical power is estimated to be 8,5 MW. Two mills are driven by turbines while the shredder and other mills are electrically driven. Mixed juice heating is done in two stages the first one using vapour 2 and the second one vapour 1. Clear juice is heated using exhaust steam and the pan floor uses vapour 1. Figure 7 shows a diagramatic steam balance at an imbibition level of 300% on fibre in bagasse.

While most data for the steam balance were available at the factory some had to be estimated. However knowing some important parameters such as the total live and exhaust steam made it is possible to make these estimates

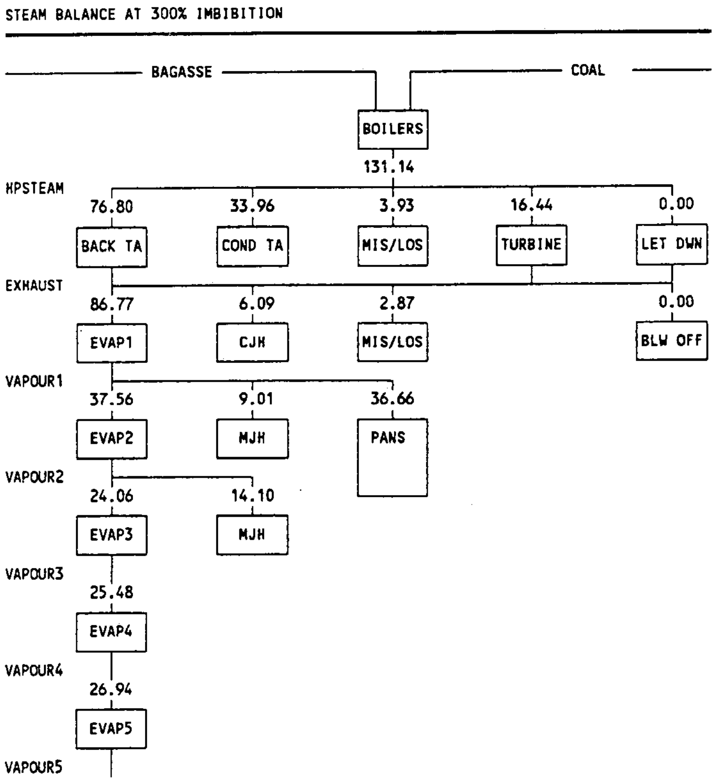

FIGURE 7 Factory steam balance.

reasonably accurately and calculate a steam balance suitable for the purpose at hand. The base case steam balance at an imbibition level of 300% is given in Appendix B. Allowance is made for 2 3% heat loss on the evaporators and pans. The common data for the mass and steam balances are the imbibition % fibre, the brix % bagasse, the moisture % bagasse, the fibre or suspended solids % mixed juice and the brix % syrup. Any change in these data in the mass balance as a result of a change in the imbibition has to be carried forward to the steam balance. In this way a whole series of steam balances was carried out, each of which related to a mass balance with the same imbibition. Figure 8 shows the quantities of live steam, exhaust steam and exhaust blow off as a function of the imbibition. At imbibition levels up to 280% the steam required for the generation of electrical power is greater than that required for process. This means that the

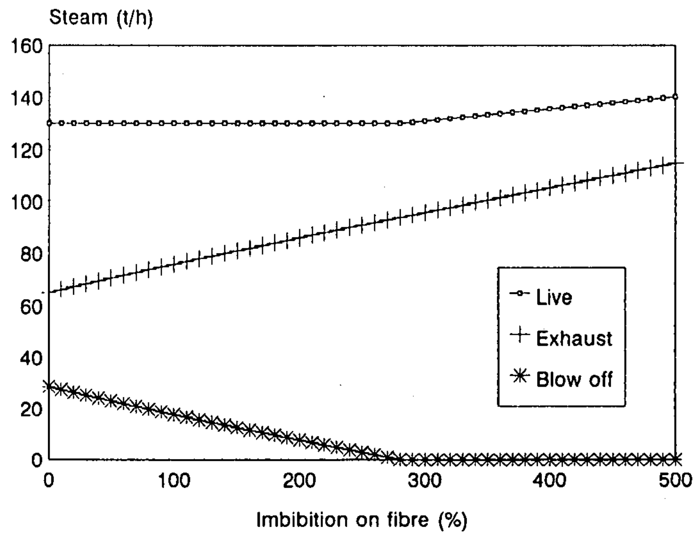

FIGURE 8 Live steam, exhaust steam and exhaust blow off.

live steam remains constant while the excess exhaust steam is blown off to atmosphere. The use of condensing turbo alternators rather than back pressure turbo alternators ensure the minimum consumption of live steam and blow off. At imbibition levels greater than 280% and a proper balance between condensing and back pressure turbines, all the exhaust steam is used, and there should be no need for the blow off or let down of steam.

An increase in fuel consumption is not only due to an increase in steam demand but also because of a decrease in the quality of bagasse with respect to its lower calorific value (LCV). This calorific value is calculated as:

$$LCV = 18\ 309 - 31{,}14 * Brix - 207{,}6 * Moisture - 196{,}05 * Ash$$

It is assumed that the ash % bagasse remains constant but the mass balances indicate a decrease in brix % bagasse and an increase in moisture % bagasse with increasing imbibition. This is shown in Figure 9.

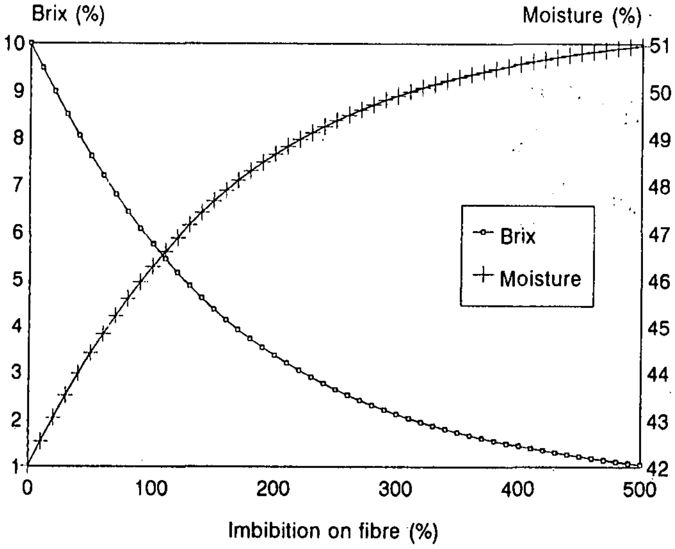

FIGURE 9 Brix and moisture % bagasse.

These combined effects decrease the calorific value and increase the fuel consumption. In addition the quantity of bagasse decreases slightly. The available bagasse quantity plus the calorific value of bagasse are shown in Figure 10.

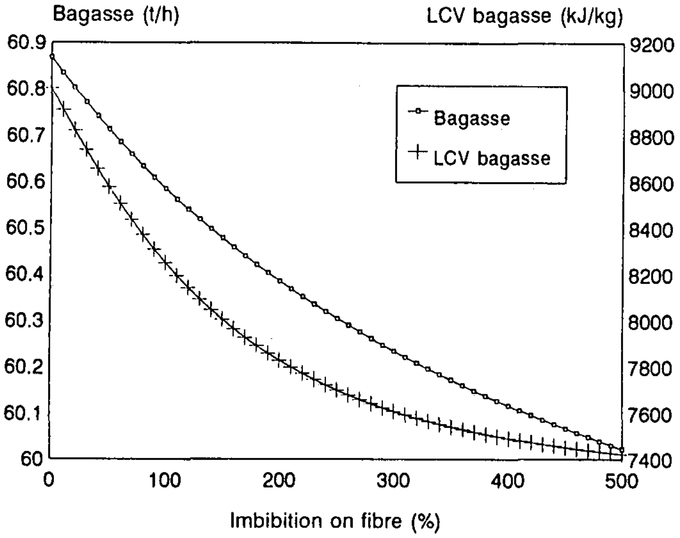

FIGURE 10 Tons bagasse and lower calorific value of bagasse.

Figure 11 shows the calculated bagasse and coal consumption. At up to 280% imbibition the live steam demand is constant and the increase in the bagasse consumption is entirely due to a decrease in its calorific value. Above 280% a steeper increase can be seen which is caused by a combination of a further decrease in the calorific value together with an increasing demand for exhaust (process) steam. A shift of the electrical load from condensing to passout turbo alternators provides for this additional exhaust steam in the most economical manner. At an imbibition of 320% the available bagasse is no longer sufficient and coal has to be added as additional fuel.

Note: These calculations assume that evaporator capacity exists to cope with these high imbibition levels. Evaporator capacity could impose an upper limit on imbibition level.

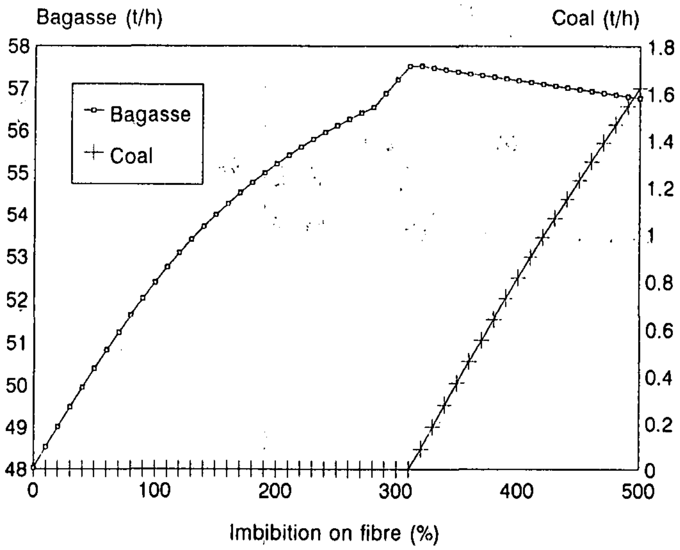

FIGURE 11 Bagasse and coal consumption.

The optimum imbibition level

From the mass balances the sugar and molasses output follows as a function of the imbibition while the steam balances provide the required fuel as a function of the imbibition. From this together with the prices of sugar (R1295/ton), molasses (R150/ton) and coal (R120/ton) the incre-

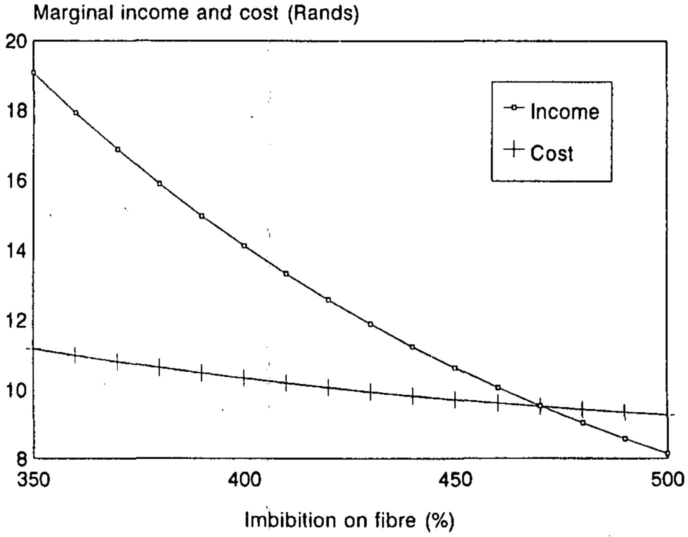

FIGURE 12 Optimum imbibition level at the present sugar price.

mental income and the incremental cost can be calculated at various imbibition levels as the imbibition is increased. The imbibition level at which the incremental cost and income are equal is the optimum level. Figure 12 shows this optimum level to be 470%. Above this level, the cost of additional coal exceeds the benefit of additional imbibition.

This optimum is dependent on the nett price of A-pool sugar. Although most of the sugar is A-pool, in a normal season there is always some B-pool sugar and any marginal sugar should therefore be considered B-pool at a much lower price. However with the abolishing of the pool system there will be only one sugar price which will be somewhere between the two but close to the A-pool price which justifies the use of the latter. The effect of any change in that price on the optimum imbibition level is shown in Figure 13.

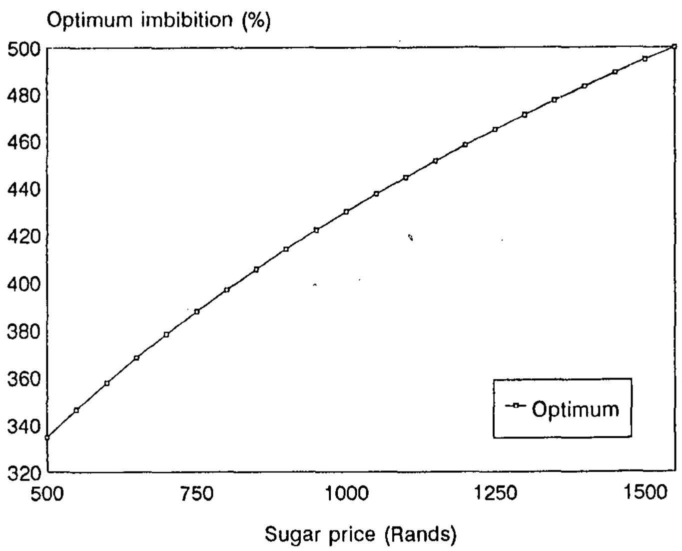

FIGURE 13 Optimum imbibition versus sugar price.

Conclusions

The optimum imbibition is calculated using a series of mass and steam balances. The most difficult part is to obtain the data required for these balances. Thereafter the method followed is quite straightforward although based on some assumptions. There is however no reason to believe that these assumptions should have a significant effect on the final result. At the present sugar price and considering coal as the only marginal cost the optimum imbibition is calculated to be 470%. With 300% Mount Edgecombe operates at well below this optimum. It is most likely that with this high optimum imbibition the evaporator capacity would be the limiting factor rather than the cost of coal.

Acknowledgements

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REFERENCES

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APPENDIX A

Mill mass balance

Overall mass balance

Description		*Description*	
Cane (t/h)	211,55	Bagasse (t/h)	60,23
Brix cane (%)	15,23	Brix bagasse (%)	2,12
Fibre cane (%)	14,43	Fibre bagasse (%)	47,99
Water cane (%)	70,34	Water bagasse (%)	49,89
Imbibition (t/h)	86,72	Mixed juice (t/h)	238,04
Imbibition eff (%)	73,54	Brix Mjuice (%)	13,00
Separation eff (%)	95,29	Fibre Mjuice (%)	0,68
Brix extraction (%)	96,05	Water Mjuice (%)	86,32

Individual mill mass balances

Description	*Mill 1*	*Mill 2*	*Mill 3*	*Mill 4*	*Mill 5*	*Mill 6*	*Mill 7*
Cane (t/h)	211,55	243,84	197,86	187,49	184,73	175,71	157,53
Brix cane (%)	15,23	7,73	4,80	3,34	2,49	1,83	1,22
Fibre cane (%)	14,43	14,12	17,37	18,62	20,16	20,54	19,95
Water cane (%)	70,34	78,15	77,83	78,05	77,35	77,63	78,83
Bagasse (t/h)	85,61	90,19	81,18	79,82	78,42	70,80	60,23
Brix bagasse (%)	10,14	7,26	5,21	4,14	3,27	2,72	2,11
Fibre bagasse (%)	31,63	35,52	38,47	40,81	42,82	44,39	47,99
Water bagasse (%)	58,23	57,22	56,32	55,05	53,91	52,89	49,89
Exprs juice (t/h)	125,94	153,65	116,68	107,67	106,31	104,91	97,29
Brix Ejuice (%)	18,69	8,00	4,52	2,74	1,91	1,23	0,67
Fibre Ejuice (%)	2,74	1,56	2,68	2,16	3,45	4,45	2,59
Water Ejuice (%)	78,57	90,44	92,80	95,10	94,64	94,32	96,74
Imbibition (t/h)	41,55	158,23	107,67	106,31	104,91	97,29	86,72
Imbibition eff (%)	107,97	90,33	79,90	68,39	63,50	55,88	45,04
Separation eff (%)	81,01	88,95	84,57	88,40	82,89	78,34	87,02
Reabsorption coeff (%)	0,95	1,07	1,14	1,18	1,19	1,19	1,12

APPENDIX B

Factory Steam Balance

Extraction

Description	*Tot/Ave*	*Description*	*Tot/Ave*
Cane (t/h)	211,55	Imbibition (t/h)	86,72
Fibre cane (%)	14,43	Bagasse (t/h)	60,23
Brix cane (%)	15,23	Fibre bagasse (t/h)	28,91
Imbibit fibre (%)	300,00	LCV bagasse (kJ/kg)	7 608,97
Brix bagasse (%)	2,12	Fibre bagasse (%)	47,99
Water bagasse (%)	49,89	Mixed juice (t/h)	238,04
Ash bagasse(%)	1,41	Brix Mjuice (t/h)	30,95
Fibre Mjuice (%)	0,68	Brix Mjuice (%)	13,00

Clarification

Description	*Tot/Ave*	*Description*	*Tot/Ave*
Bagacillo Mjuice (%)	1,10	Bagacillo (t/h)	2,62
Wash Water Mjuice (%)	7,84	Wash water (t/h)	18,66
Filtrate Mjuice (%)	15,00	Filtrate (t/h)	35,71
Brix filter cake (%)	0,84	Clear juice (t/h)	247,23
Water filter cake (%)	75,37	Brix Cjuice (t/h)	30,90
Brix Cjuice (%)	12,50	Filter cake (t/h)	12,09

Steam Properties

Description	*Pressure* kPa(abs)	*Temper* °C	*Latheat* kJ/kg	*Enthalpy* kJ/kg
High press steam	1500,00	270,00	0,00	2970,94
Exhaust steam	200,00	120,35	2202,60	2706,49
Vapour-1 steam	150,00	111,71	2228,29	2696,01
Vapour-2 steam	115,00	104,06	2249,16	2684,87
Vapour-3 steam	80,00	94,09	2274,45	2668,42
Vapour-4 steam	48,00	80,93	2305,75	2644,58
Vapour-5 steam	15,00	54,20	2367,51	2594,44

Boilers and power station

Description		*Description*	
Electrical power (kW)	8500,00	Desup.recom.vapour (%)	0,00
BFW temperature (°C)	115,00	HPress steam miscell (t/h)	0,00
Boiler efficiency (%)	75,00	HPress steam losses (%)	3,00
LCV of coal (kJ/kg)	27500,00	Exhaust steam miscell (t/h)	0,00
Export of bagasse (%)	0,00	Exhaust steam losses (%)	3,00

Juice Heaters

Description	*Vapour* source	*Juic Fl* % tch	*Tempin* °C	*Tempout* °C	*Ht Surf* m^2
Mjuice heaters-1	2,00	129,40	55,00	85,00	500,00
Mjuice heaters-2	1,00	129,40	85,00	104,00	750,00
Cjuice heaters-1	0,00	116,87	98,00	112,00	220,00

Description	*Lmtd* °C	*Juic Fl* t/h	*Heat Fl* kW	*Vap Fl* t/h	*Heat TC* kW/m^2°C
Mjuice heaters-1	31,73	273,75	8806,41	14,10	0,56
Mjuice heaters-2	15,29	273,75	5577,39	9,01	0,49
Mjuice heaters-1	14,21	247,23	3723,78	6,09	1,19

Pan Floor

Description	*Vapour* source	*BxSyr* %	*Bx Msc* %	*Massec* m^3/t bx	*Vap Fl* t/h
A-Pan boiling	1,00	66,56	91,67	1,04	18,66
B-Pan boiling	1,00	70,00	92,69	0,35	5,42
C-Pan boiling	1,00	70,00	95,88	0,23	4,12

Description	*Pan Wat* % vap	*Jig Stm* % vap	*Pan Wat* t/h	*Jig Stm* t/h	*Vap Fl* t/h
A-Pan boiling	30,00	0,00	5,60	0,00	5,60
B-Pan boiling	30,00	0,00	1,63	0,00	1,63
C-Pan boiling	30,00	0,00	1,24	0,00	1,24

Steam Turbines

Description	*Quantity* no	*Efficy* %	*Loading* kW	*Sst Dem* kg/kWh	*Sst Sup* kg/kWh
Mill steam turb-1	2,00	39,00	350,00	23,48	24,65
T/A Bpress turb-1	1,00	55,00	12000,00	16,65	17,01
T/A conden turb-1	1,00	55,00	4000,00	8,74	0,00

Description	*Lpress* kPa(abs)	*Ltemp* °C	*Loading* kW	*Tst Dem* t/h	*Tst Sup* t/h
Mill steam turb-1	200,00	174,67	700,00	16,44	17,26
T/A Bpress turb-1	200,00	143,35	4612,61	76,80	78,47
T/A conden turb-1	15,00	45,11	3887,39	33,96	0,00

Evaporators

Description	*Flashing* 0/1/2/3	*Bx Juic* %	*Tp Juic* °C	*Ret Bld* t/h	*Ht Surf* m^2
Evap effect-1	2,00	18,84	112,04	0,00	3110,00
Evap effect-2	2,00	24,55	104,55	0,00	1350,00
Evap effect-3	1,00	30,65	94,78	0,00	1100,00
Evap effect-4	1,00	41,49	82,07	0,00	930,00
Evap effect-5	0,00	66,56	57,34	0,00	1550,00

Description	*Stm Fl* t/h	*Vap Fl* t/h	*Juic Fl* t/h	*Cond Fl* t/h	*Heat TC* kW/m^2°C
Evap effect-1	86,77	83,23	164,00	86,77	2,06
Evap effect-2	37,56	38,15	125,85	124,32	2,41
Evap effect-3	24,06	25,04	100,81	23,62	1,47
Evap effect-4	25,48	26,33	74,48	24,87	1,44
Evap effect-5	26,94	28,06	46,42	26,94	0,47

Steam demand and supply

Exhaust demand	t/h	*Hpress Demand*		t/h
Juice heaters	6,09	Mill turbines		16,44
Evaporators	86,77	Miscell turbines		0,00
Pan floor	0,00	Bpress turbo-alts		76,80
Refinery	0,00	Condens turbo-alts		33,96
Exhaust miscell	0,00	Hpress miscell		0,00
Exhaust blow off	0,00	Hpress let down		0,00
Exhaust losses	2,87	Hpress losses		3,93
	——,——			——,——
	95,73			131,14
Exhaust Supply	t/h	*Hpress Supply*		t/h
Mill turbines	17,26	Bagasse	57,21 t/h	131,14
Miscell turbines	0,00	Coal	0,00 t/h	0,00
BPress turbo-alts	78,47			——,——
HPress let down	0,00			131,14
	——,——			
	95,73	Surplus	0,41 t/h	0,94