STEAM BALANCE FOR THE NEW FELIXTON II MILL

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Abstract

Various aspects affecting the steam economy of a raw sugar mill are discussed. Process and prime mover steam requirements are evaluated for a new 600 ton cane/hour factory currently under construction at Felixton, and options for achieving a level of steam economy where fibre can be supplied to a downstream paper mill with the appropriate fuel economy are evaluated. The necessity to allow for flexibility should conditions affecting the bagasse supply in the future change is highlighted.

Introduction

A new mill to crush 600 tons cane/hour is presently under construction at Felixton. The choice of the site and milling capacity of the new mill, to be known as Felixton II, was motivated by a number of important factors including the existence of an established infrastructure and the proximity of a paper mill which uses bagasse fibre as a raw material for paper manufacture.

The need to provide an end user with a large portion of the bagasse which is normally used as boiler fuel makes it necessary to increase the thermal efficiency of the sugar mill to a higher than usual level. The adjacent Paper Mill presently requires up to 100 000 tons of bone-dry fibre per annum, which on a conservative basis requires Felixton II to make available up to 30% of its total bagasse when running at 600 tch. When this fibre is "de-pithed", the mill will be left with 67% of its normal fuel, which will consist mainly of pith.

This represents the major constraint in the design of the steam balance for the new mill. The thermal economy of the mill should be such that the consumption of coal is kept to a minimum under normal conditions. In addition, the design of the mill should be such that it is consistent with a plan to be able to take steam economy to much higher levels should the demand for bagasse or fibre by downstream users increase.

The options for achieving these objectives are discussed below.

Steam Requirements

It seemed likely at the outset that the mill would be exhaust steam dependent rather than high pressure steam dependent for the following reasons:

(a) There is no requirement to produce power for any outside or ancillary use (e.g. back end refinery, irrigation etc)

(b) Extraction by diffusion not milling was planned, which requires far less power but more low quality vapour for heating purposes.

Conventional South African practice involves the use of vapour bled from multiple effect evaporators for heating purposes and for use on the pan floor. It is necessary to establish these quantities initially before the requirements of the evaporator station can be determined.

Quantities of process streams

Since a high extraction is being aimed for in the diffuser, it appeared to be optimal to use a higher than normal imbibition rate. An average imbibition % fibre figure of 350 was assumed, although industry average imbibition levels are now up to the same level.

Once crushing rates and cane quality data are finalized, juice quantities can be calculated. Average data for Felixton II are as follows:

Crushing rate	600 tch
Pol % cane	12,6
Fibre % cane	17,5
Pol extraction	98%
Imbibition % fibre	350

Average mixed juice and clear juice quantities were calculated to be 860 and 780 tons/hour respectively. However these quantities vary significantly during the season as cane quality varies. Average rates were assumed to be 15% higher during the peak of the season.

Effect of diffuser requirements

Diffusers are large users of vapour for heating purposes. Generally temperatures of over 80°C are aimed for throughout a diffuser to promote extraction and eliminate losses due to microbiological action. This results in significant heat losses in the diffuser itself, and in the dewatering mills. In addition there are losses in evaporation, and by comparison with milling, losses in the higher temperature bagasse leaving the diffuser.

The major part of the heat is not lost and results in a draft juice temperature of roughly 65°C, so that the subsequent juice heating requirements are reduced somewhat. Part of the heat is added in the form of direct injection vapour into the diffuser trays to make up for heat losses. The bulk of the heating should be done at the feed end of the diffuser by indirect means, and in order to calculate the size of the scalding juice heaters the split between direct and indirect heating needs to be established.

Actual measurements of vapour heating requirements are sparse. Measurements made at Amatikulu in 1977 recorded heating vapour requirements of 13% on cane. In addition measurements taken at Gledhow indicated a requirement of 12.4% vapour on cane.[1]

In order to estimate scalding juice heating requirements, a computer programme was written to establish heat and mass balance relationships at the feed end of the diffuser, taking into account varying percolation rates. Scalding juice heaters on 1st and 2nd stage flows were assumed. The primary scalding juice heater is assumed to run at full capacity whilst vapour to the secondary heater is throttled to achieve the required bed temperature in the diffuser (85°C). From this programme, an estimated vapour requirement at the scalding juice heaters of 36 ton/h was obtained, and the draft juice temperature calculated to be 65°C. The direct injection vapour requirements were estimated from a heat balance to be 30 ton/h. Thus the total vapour heating requirement is 66 ton/h or 11% on cane.

Juice heating

Energy requirements for juice heating are easily calculated by heat balance. Juice specific heats for pure sucrose solutions are assumed to hold for mixed and clear juice, and energy requirements are calculated in MW.

In this case mixed juice and clear juice heating requirements were calculated to be 36.7 and 15.6 MW respectively. For interest the heating requirements for mixed juice assuming Felixton II to be a milling mill (i.e. mixed juice temperature = 35°C) are estimated to be 65.2 MW. This is considerably less than the combined duties for diffuser juice heating of 36.7 + 41.0 = 77.7 MW, indicating a 19% increased heating requirement for diffuser mills. In terms of total vapour usage, this represents an increase in vapour required of 9.4%.

Pan floor requirements

The major vapour requirement on the pan floor is for use in the calandrias of the vacuum pans. Smaller quantities are required for use in blow-up tanks, centrifugals, melter and calorifier, while a not insignificant amount is also required for jigger steam in low grade pans.

It was assumed that a conventional 3 boiling system with partial remelt of B and C sugars would be used. Use was made of a computer programme BOOB (Boiling house Operations Overall Balance) to calculate steam requirements.[2] This requires full specification of water and steam usage on centrifugals, movement water used on pans, and the brixes of remelt and molasses (after the blow-up tanks) to calculate steam usage. The input data required for the program is shown in Table 1.

The printout for the base case (data given in Table 1) is shown in Figure 1. This indicates a total vapour requirement of 93.2 tons/h.

TABLE 1

Input data to boiling house balance for estimating vapour usage - base case data

Syrup feed: tons total solids/h		88.8	
purity		85.5	
% total solids		68.0	
B magma: total solids %		89.0	
Final molasses: true purity		38.0	
total solids %		77.0	
	A	B	C
Massecuite total solids %	93.0	92.5	92.5
Massecuite exhaustion	66.0	58.0	—
Sugar purity	99.4	93.0	83.0
total solids %	99.3	97.5	96.0
Total solids % molasses	70.0	70.0	77.0
Movement water % solids	13.0	13.0	12.5
Centrifugal wash water % solids	1.0	5.50	—

Seasonal effects result in considerable variations in massecuite quantities in the pan house, due to changes in brix loading and syrup purities. In addition, variations in syrup brix and massecuite exhaustion also affect massecuite quantities. The overall effect of these changes is shown in Table 2.

BOILINGHOUSE OPERATIONS OVERALL BALANCE. DATE:05APR83 TIME: 12:59:51

RUN NO: FX14 MILL: FELIXTON II OBJECT: BASE CASE CONDITIONS.

RESULTS OF RUN. ITERATION 3 OF 3 84 UNKNOWNS, 84 RELATIONSHIPS.

DATA ON PROCESS STREAMS.

STREAM NO.	DESCRIPTION	TONNAGES: TOTAL MASS	TOTAL SOLIDS	TOTAL SUCROSE	TOTAL NON-SUC	WATER IN SOLN	XTAL SUCROSE	QUALITY: PURITY (%)	SOLIDS% TOTMASS	XTAL % SOLIDS	NUTSCH PURITY	NON-SUC :WATER
1	SYRUP FROM EVAPORATORS	130.59	88.80	75.92	12.88	41.79		85.50	68.00			.308
2	SYRUP + RECYCLE FEED TO A-PAN	166.46	113.19	97.27	15.92	53.27		85.93	68.00			.299
3	A-MASSECUITE FROM PAN TO XTALLISER	127.33	118.42	102.13	16.29	8.91	66.59	86.24	93.00	56.24	68.57	1.828
4	A-MASSECUITE FROM XTALLISER TO CENTR.	127.33	118.42	102.13	16.29	8.91	71.11	86.24	93.00	60.05	65.57	1.828
5	A-SUGAR PRODUCT	69.18	68.70	68.28	.41	.48	67.41	99.40	99.30	98.12	68.07	.851
6	A-MOL.FROM A-CENTR.TO A-BLOW-UP TANK	59.34	49.72	33.85	15.88	9.61		68.07	83.80			1.652
7	A-MOLASSES FROM A-BLOW-UP TO B-PAN	50.90	35.63	24.25	11.38	15.27		68.07	70.00			.745
8	A-MOLASSES FROM A-BLOW-UP TO C-PAN	20.13	14.09	9.59	4.50	6.04		68.07	70.00			.745
9	B-MASSECUITE FROM PAN TO XTALLISER	38.52	35.63	24.25	11.38	2.89	13.44	68.07	92.50	37.72	48.73	3.938
10	B-MASSECUITE FROM XTALLISER TO CENTR.	38.52	35.63	24.25	11.38	2.89	15.04	68.07	92.50	42.22	44.74	3.938
11	B-SUGAR TO MINGLER	16.63	16.22	15.08	1.14	.42	14.07	93.00	97.50	86.73	47.24	2.730
12	B-MOL.FROM B-CENTR.TO B-BLOW-UP TANK	23.85	19.41	9.17	10.24	4.43		47.24	81.41			2.311
13	B-MOLASSES FROM B-BLOW-UP TO C-PAN	27.73	19.41	9.17	10.24	8.32		47.24	70.00			1.231
14	C-MASSECUITE FROM PAN TO XTALLISER	36.22	33.50	18.76	14.74	2.72	8.30	56.00	92.50	24.79	41.50	5.427
15	C-MASSECUITE FROM XTALLISER TO CENTR.	36.22	33.50	18.76	14.74	2.72	10.65	56.00	92.50	31.78	35.50	5.427
16	C-SUGAR TO REMELTER	13.96	13.40	11.12	2.28	.56	9.73	83.00	96.00	72.58	38.00	4.080
17	C-MOLASSES PRODUCT	26.11	20.10	7.64	12.46	6.00		38.00	77.00			2.076
18	REMELT TO BLENDING WITH SYRUP	35.87	24.39	21.34	3.05	11.48		87.51	68.00			.266
19	MINGLER TO REMELTER	12.35	10.99	10.22	.77	1.36		93.00	89.00			.566
20	FOOTING FROM MINGLER TO A-PAN	5.88	5.23	4.86	.37	.65		93.00	89.00			.566
21	WATER OR CJ DILUTION TO MINGLER	1.59				1.59						

DATA ON PROCESSING UNITS.

UNIT NO.	DESCRIPTION	SUCROSE RECOV %	M/CUITE EXH.%	XTAL GR FACTOR	PURITY RISE %	VAP.USE %SOLIDS	WASH % SOLIDS	VAPOUR (TONS)	WASH (TONS)	RATIO/ SPLIT
1	A-PAN				-17.68	51.00	13.00	60.39	15.39	.050
2	A-XTALLISER			1.068	-3.00					
3	A-CENTRIFUGAL		66.00	.948	2.50		1.00		1.18	
4	A-MOLASSES BLOW-UP TANK						23.52		11.70	2.529
5	B-PAN				-19.34	47.75	13.00	17.01	4.63	
6	B-XTALLISER			1.119	-3.99					
7	B-CENTRIFUGAL		58.00	.935	2.50		5.50		1.96	
8	B-MOLASSES BLOW-UP TANK						20.02		3.89	
9	C-PAN				-14.50	47.25	12.50	15.83	4.19	.726
10	C-XTALLISER			1.282	-6.00					
11	C-CENTRIFUGAL		51.84	.913	2.50		11.48		3.85	
12	MINGLER									2.101
13	REMELTER						39.20		9.56	.820
14	MERGE OF REMELT & SYRUP TO A-PAN									.275
15	OVERALL SYSTEM	89.94	89.61							

TOTAL WASH WATER CONSUMPTION = 56.35 TONS/HR TOTAL VAPOUR CONSUMPTION = 93.24 TONS/HR

FIGURE 1 Computer printout of boiling house balance programme for average conditions

TABLE 2
Effect of changes in boiling house operations on vapour required

	Vapour usage (ton/h)	A massecuite quantity (ton/h)	Total massecuite boiled (ton/h)
Base case (fig 1)	93.2	127.3	202.1
Lower syrup brix = (65.0)	99.3	127.3	202.1
High A exhaustion (=70)	84.6	120.8	182.8
Low A exhaustion (=60)	107.9	138.8	234.6
High syrup purity (=87.5)	95.2	129.8	206.4
Low syrup purity (=83.0)	89.9	124.2	194.8
Peak season conditions (high purity, high brix load)	108.6	147.9	235.4
Early season conditions (low purity, low brix load, low molasses purity)	78.0	107.3	169.0

Total vapour required is summarised in Table 3 below. It includes estimates for jigger steam used in low grade pans, and for ancillary operations in the boiling house e.g. centrifugals, remelter, etc.

TABLE 3
Summary of vapour requirements

	Duty (MW)	Vapour Rate (tons/h)	Vapour % Cane
Diffuser heating	41.0	66	11.0
Mixed juice heating	36.7	59	9.9
Clear juice heating (1st stage)	8.8	14	2.4
Pan requirements	58.4	93	15.5
Pan jigger steam	5.6	9	1.5
Pan house ancillaries	7.1	11	1.9
TOTAL	157.6	252	42.2

Evaporator requirements

The evaporators are the major users of exhaust steam, and determine the overall steam requirements in an exhaust steam dependent factory. It was recognized therefore that the options available in evaporator configuration would largely determine the overall steam requirement.

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             MATERIAL  AND  HEAT  BALANCE  FOR  EVAPORATOR  TRAIN  AT  FELIXTON 2              .DATE:  5/ 4/83
             -------------------------------------------------------------------------        -------------

RUN NO: 55                 OBJECT OF RUN: 600 TCH. FIXED AREAS V2 TO PANS AND DIFFUSER - BASE CASE.
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TYPE OF CALCULATION:   6:
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     CALC. VAPOUR AND RECOMPR. FLOWS AND TEMPS., GIVEN SYRUP AND CLEAR  JUICE  FLOWS  AND  BRIXES.

EVAPORATOR REF. NO :         1        2        3        4        5
EVAP. NAME IN FACTORY        1        2        3        4        5
EVAPORATOR EFFECT NO.        1        2        3        4        5
MATERIAL FLOWS;
---------------
VAPOUR IN:
FLOW RATE, KG/HR          276500.  238010.  26424.   40269.   58603.
TEMPERATURE, DEG.C.         120.8    114.0   101.0     94.5     82.5
PRESSURE, KPA ABS           203.     164.    105.      83.      52.

VAPOUR OUT:
  FLOW RATE, KG/HR          275010.  241024.  31214.   43379.   62744.
  TEMPERATURE, DEG.C.         114.0    107.1    94.5     82.5     55.3
  PRESSURE, KPA ABS           164.     130.     83.      52.      16.

JUICE IN:
FLOW RATE, TONS/HR          784.0    509.0   268.0    236.8    193.4
TEMP.,DEG.C.                115.0    114.3   108.2     95.7     84.3
DEGREES BRIX                 11.3     17.5    33.1     37.5     45.9

JUICE OUT:
  FLOW RATE, TONS/HR          509.0    268.0   236.8    193.4    130.6
  TEMP.,DEG.C                 114.3    108.2    95.7     84.3     58.9
  DEGREES BRIX                 17.5     33.1    37.5     45.9     68.0

EVAPORATOR DATA:
----------------
HEAT TRANSFER:
 CALANDRIA AREA, SQ.M       11000.   11000.   2000.    2000.    2860.
 HYDROSTATIC HEAD,      M       .3       .3      .3       .3       .3
HYDROSTATIC TEMP.RISE,  C       .6       .8     1.1      1.7      4.7
BOILING POINT ELEV, DEG.C       .2      1.0     1.3      1.8      3.6
 O/ALL  HTC, KW/SQ.M/DEGC     2.60     2.60    2.00     1.50      .70
 O/ALL TEMP.DIFF.              5.9      5.1     4.1      8.5     18.8
 HEAT TRANSFER RATE,    KW 168036.  146083.   16514.   25353.   37444.

HEAT LOSSES:
 SHELL AREA,   SQ.M            0.       0.      0.       0.       0.
 O/ALL  HTC, KW/SQ.M/DEGC    .0000    .0000   .0000    .0000    .0000
 HEAT LOSS RATE,        KW     0.       0.      0.       0.       0.

VAPOUR BLEEDS AND
  CONDENSATE FLASH FEEDS:
--------------------------
BLEED OR FLASH REF.NO.       1        2        3        4
VAPOUR BLEED FLOW, KG/HR  37000.  214600.       0.
VAPOUR BLEED TEMP,  DEG.C  114.0    107.1     94.5
VAPOUR BLEED PRESS,KPA AB  164.     130.      83.
CONDENSATE FLASH,  KG/HR     0.       0.    8934.   15266.

VAPOUR FLOWS, KG/HR : TOTAL EVAPORATION  = 653372.        TOTAL BLEEDS = 251600.        TOTAL VAPOUR FEED = 276500.
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FIGURE 2 Computer printout of evaporator station calcution.

The evaporators are to concentrate 784 tons/h on average at 11 brix up to 68 brix syrup, and incorporate vapour bleeding to fulfil process steam requirements. A maximum exhaust pressure of 200 kPa (absolute) was specified, and a final effect pressure of 16 kPa was chosen since this appears to give optimum performance.[3]

A computer programme PEST (Programme for Evaporator Simulation and Testing) was utilized to compute exhaust steam requirements for any proposed configuration.[4] It incorporates provision for condensate flash, vapour bleeding and mechanical or thermo-compression of vapour. It makes provision for boiling point elevation, hydrostatic head and heat loss from the vessels, although the latter item was ignored. A sample of the data specified is given in Table 4, and a sample output is shown as Figure 2.

TABLE 4

Input data to evaporator simulation programme

No. of effects	5
1st effect calandria pressure	200 kPa
Last effect vapour space pressure	16 kPa
Juice flow to evaporator	784 ton/h
Juice purity	85
Juice brix	11.3
Syrup brix total dissolved solids	68
Juice inlet temperature	115°C
Heat transfer areas: 1st effect	11 000m^2
2nd effect	11 000m^2
3rd effect	2 000m^2
4th effect	2 000m^2
5th effect	2 860m^2
Heat transfer coefficients: 1st effect	2.6 kW/m^2°C
2nd effect	2.6 kW/m^2°C
3rd effect	2.0 kW/m^2°C
4th effect	1.5 kW/m^2°C
5th effect	0.7 kW/m^2°C
Vapour bleeds: V1	37.0 ton/h
V2	215.0 ton/h
Vapour 2 pressure minimum	130 kPa

Heat transfer coefficients were chosen based on previous measurements on evaporators, most of which has been reported elsewhere.[3] The results shown in the printout assume all exhaust condensate is taken to boiler feed under pressure. Other condensate to the boilers is flashed to vapour 3, and all reject condensate is flashed to vapour 4.

Prime Mover and Boiler Requirements

The total power requirement of the mill and factory is one of the most important variables required to complete a steam and fuel balance for a new factory.

For Felixton II, it was necessary at an early stage to draw up a preliminary electric motor list and to make early decisions on the type of prime mover to be used for large power consumers. Some of these consumers, with their prime movers are listed below:

Shredders: It was decided to use electric motors because the shredders, which are located some distance away from the main buildings, would require lengthy and expensive pipework if turbine driven. A negative aspect of an electric shredder drive is that the motor must be large enough to cope with power surges. Thus 3 000 kW motors were chosen for each shredder, i.e. 57 kW per ton fibre per hour. A further reason for choosing electric drives is that they are, in overall terms, more efficient than the type of turbine drives normally used for this purpose.

Mills: In this case a prime requirement of the mills is that the speed of the drives can be varied through a very wide range. The decision to use steam turbines was based on this need and on the fact that the steam supply is conveniently close to the mill house. The powers are sufficiently small to reduce the effect of the lower efficiency of the turbine drives.

I D fans: Electric drives were chosen for these units, based on the need for higher efficiency. Also, it was decided at an early stage to control boiler draft using dampers, and a fixed speed motor is generally more economical than a turbine of the same power.

The total installed power of electric motors for Felixton II is 34 MW. Allowing for the usual diversity factors and based on existing installations, it has been estimated that the total electrical load will be approximately 17 MW, or 28 kW per ton of cane, a figure which is fairly close to that for existing mills in the group.

The process steam demand is found in all cases to be higher than the exhaust generated by all the turbines, running at their average loads. The shortfall in process steam must be made up from high pressure steam through a reducing valve, and must subsequently be desuperheated. Therefore any change in steam turbine steam consumption, either through load or efficiency changes, does little to the level of fuel consumption.

It thus makes sense, if attempting to save fuel, to use some of the high pressure steam going to the reducing valve to power various devices which could reduce the process steam demand. Some of these devices will be discussed in the next section.

It should be noted that the desuperheating station, which uses the evaporaion of condensate to reduce temperature, increases the quantity of steam flowing to process users. This increase for the pressures and temperatures used at Felixton II is 20% of the high pressure steam let down to exhaust – a quantity which is quite significant in the overall steam balance.

The boilers were selected according to the following policy: three boilers are required, of a sufficient size that if one boiler is off range for any reason, the other two at full load can keep the factory going at rated throughput. The pressure and temperature chosen conform to what has become common for all new boilers in the Natal industry i.e. 32MPa and 400°C. In order to achieve the necessary fuel economy, the steam load will be close to but not greater than 300 tons per hour, or 50% on cane. The three boilers must therefore have a maximum continuous rating (MCR) of 150 tons of steam per hour each.

In order to provide the bagasse surplus required even during periods of low fibre % cane, the boiler house efficiency should be at least 78%. It has been conservatively assumed that in order to achieve this overall boiler house efficiency continuously the boilers must be individually designed for an efficiency of at least 82%.

The boilers have been arranged in such a way that additional economiser heating surface can easily be added in the future to increase the efficiency to 85%.

An increase in boiler pressure and superheat temperature will generally reduce the steam consumption of turbines. The effect of doubling the pressure and increasing the temperature to 450°C was investigated and was found to make an improvement of only 1,3% to fuel consumption. It was thus decided to stay with the traditional pressure and temperature – which has the added attraction that boiler maintenance and water treatment are much simplified.

Methods of Reducing Steam Consumption in Exhaust Dependent Factory

For the Felixton II situation, the required steam efficiency can be achieved by using a quintuple effect evaporator, and bleeding second vapour for use on pans and diffuser. A sketch of the steam and vapour flows for average Felixton II conditions is shown in Figure 3.

A reduction in steam consumption would be desirable in the event that a market is found for all the bagasse – which is considered to be quite likely sometime in the future. It was thus necessary to ensure that the design and layout of the steam

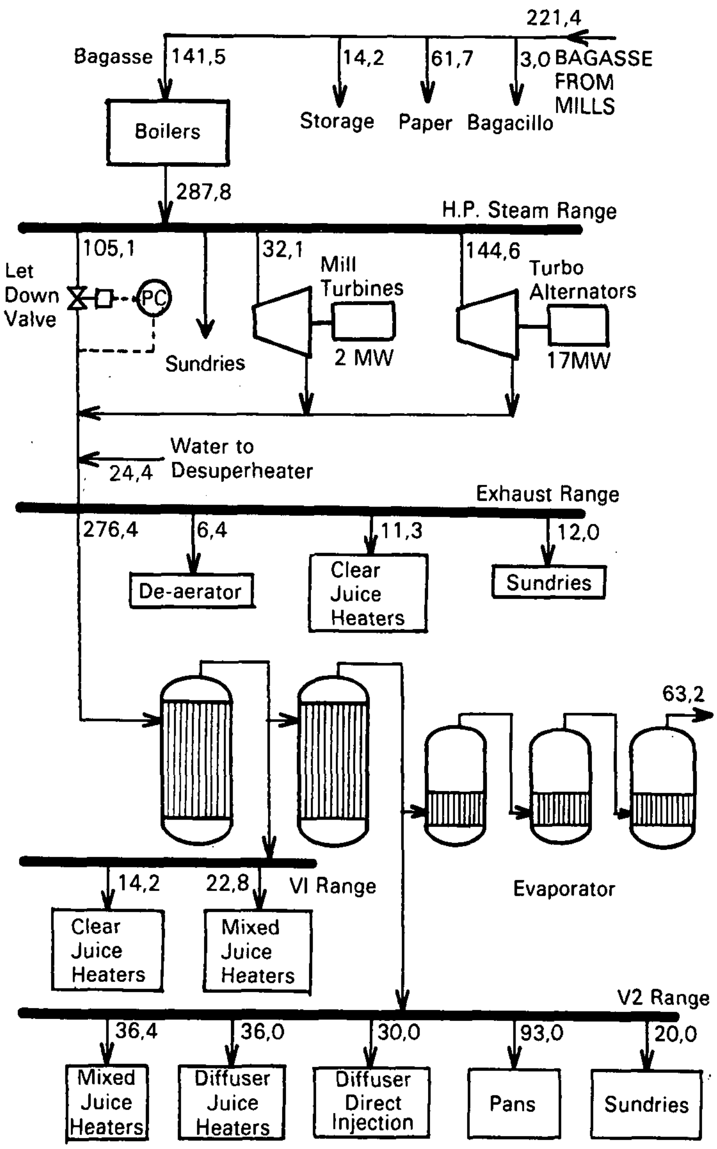

FIGURE 3 Felixton II steam balance for season average conditions

system should be such that future alterations and additions to improve the thermal efficiency, could be carried out in an economical manner.

For this purpose, the effect on thermal efficiency of a number of different evaporator and other process plant configurations was investigated. A summary of the various cases of interest is given in Table 5, and the effect of various options discussed in more detail below:

Vapour bleeding

The effect of using a low grade vapour from the evaporator for process heating, instead of using exhaust steam, is already so well known in the process industries that it is widely practised, especially in the sugar industry. It is also commonly known that the lower the effect in the evaporator train from which the vapour can be used the greater will be the thermal efficiency.

Thus at Felixton II, the evaporator heating surface has been chosen so that the pressure of second vapour is high enough to be used for vacuum pans, diffuser heating and most other factory requirements. Using V1 bleed instead of V2 bleed for this purpose increases the high pressure steam requirement by 12%. The effect of this large bleed from V2 is to increase the second effect area, reduce the heating surface requirement of the last three effects in the evaporator train, and reduce the condenser duty. It is apparent from Table 5 that better steam economy generally goes together with a lower final vapour condensing load, and a greater evaporator heating surface requirement.

If V3 bleed is used in a primary mixed juice heater, the HP steam consumption can be reduced a further 0,5%. This quantity is small and does not offset the additional heater costs. Lower effects than V3 do not provide adequate temperatures for diffuser juice heating.

Increase in number of effects

The fact that the "tail" of the evaporator is much reduced in size reduces the effect of increasing the number of vessels in an evaporator.

The difference between a quintuple and a quadruple effect when both are providing V2 to pans and diffusers is 3,6% based on HP steam consumption. On the other hand, if both quadruple and quintuple effect are providing V1 bleed, the difference in HP steam consumption is 5,1%.

A similar effect is noticed when comparing quintuple with sextuple effect evaporation. When V2 is bled, the difference is 2,2%, whereas with V1 bleed, this difference increases to 3,9%.

Thermocompression

Over a century ago, engineers were beginning to apply vapour compression to evaporative processes in order to economize on steam. Thermocompression is one method of compressing this vapour, using as an energy source the surplus high pressure steam mentioned above. A thermocompressor is simply a steam ejector, drawing vapour from the top of an evaporator vessel

TABLE 5
Summary of steam balance options

Pest Run No.	Evap Type	Vap to Pans & Diffuser	Exh Press (kPa)	V2 Press (kPa)	Exhaust to Evaps (t/h)	H P Steam (t/h)			Total (incl. sundry)	HP Steam % Cane	Total HS in Evap. (m^2)	Vapour to cond. (t/h)	HP Steam % Base Case
						Let-down	Prime movers	TC					
55	Quin (base case)	V2	203	130	276,4	105,0	176,9	–	287,8	48,0	28860	63,1	100,0
56	Quin	V1	200	130	319,0	140,1	176,0	–	322,1	53,1	30400	106,1	111,9
57	Quin, TC, 3-1	V2	200	136	160,0	11,1	176,0	85,9	279,0	46,5	28860	64,1	96,9
58	Quin, TC, 2-1	V2	200	125	160,0	11,1	176,0	92,0	285,1	47,5	28860	58,5	99,1
59	Quin, MVR, 2-1	V2	215	130	256,3	72,7	196,5	–	275,1	45,9	28860	47,3	95,6
60	Quin, MVR, 3-1	V2	228	146	245,6	45,8	218,9	–	270,7	45,1	28860	55,3	94,1
61	Quin, MVR, 2-1	V2	202	130	252,7	74,7	190,4	–	271,1	45,2	34360	45,0	94,2
62	Quin, MVR, 3-1	V2	217	148	243,4	48,0	213,9	–	267,9	44,7	34360	38,6	93,1
63	Sext	V2	200	139	269,1	99,6	176,0	–	281,6	46,9	35108	58,9	97,8
64	Sext	V1	200	139	304,0	127,9	176,0	–	310,0	51,7	33914	93,5	107,7
65	Quad	V1	200	108	339,3	156,5	176,0	–	338,6	56,4	20731	120,7	117,7
66	Quad	V2	200	130	289,7	116,3	176,0	–	298,3	49,7	28191	67,3	103,6
67	Quin, V3 bleed	V2	200	130	274,9	104,3	176,0	–	286,3	47,7	28303	62,0	99,5

TC: Thermocompressor MVR: Mechanical Vapour Compressor and compressing it through the conversion of pressure energy into kinetic energy for mixing and back into pressure energy of a lower order for use in the calandria of the vessel.

When thermocompression was first considered, it was realised that the entrainment ratio (i.e. ratio of entrained vapour/ motive steam) was a crucial variable that had to be known in advance. A literature survey was carried out to try to find a method of calculating this entrainment ratio, but with little success. Eventually it was decided to build an experimental thermocompressor which was installed first at Amatikulu and subsequently at Felixton. After a series of exhaustive tests, it was found that the entrainment ratio was somewhat lower than that predicted by various empirical formulae.[5] This information was incorporated in the PEST programme, and it was found that the addition of a thermocompressor could indeed improve the thermal efficiency of a factory, but only by about 1% to 3%, when applied in its most suitable form to a quintuple effect evaporator.

A thermocompressor has two disadvantages when applied to an evaporator. Firstly, the mixing efficiency is extremely sensitive to changes in pressure ratio, which limits the turn down to less than 10%. Adequate pressure control can therefore only be achieved by using multiple thermocompressors of different sizes, arranged in parallel, and by means of a complicated valve control system, switching them in or out as required.

Secondly, the vapour resulting from a thermocompressor is a mixture of high pressure steam and a lower grade vapour, which can be contaminated by liquid entrainment in the evaporator. Thus the condensate from the mixed vapour contaminates that from the high pressure steam which is then less suitable as boiler feedwater.

Mechanical vapour recompression (MVR)

The MVR appears to be the most promising means of improving thermal efficiency. In its most suitable form, a compressor with a throughput of 110 tons per hour, compressing V2 to V1 and requiring about 1,6 MW to drive it, would reduce high pressure steam consumption by 4,5%. This figure can be increased to nearly 6% by adding 3 000 m^2 additional heating surface to the second effect of each evaporator.

The MVR can be driven by an electric motor, or by a steam turbine. In the latter case, pressure control can be achieved by varying the speed of the turbine, whereas in the former case, the most effective means of pressure control is through adjustable inlet vanes.

A further positive aspect of the MVR is that it has been successfully used in process industries, including the beet sugar industry and sugar refining, for many years. The successful trials of an MVR at Pongola are further proof of its suitability.[6]

If the Felixton II mill is at some future date able to find a market for all of its bagasse, the boilers will have to burn coal only, and it would be feasible to spend considerable money and effort to reduce steam consumption, and thereby reduce fuel costs. The addition of an MVR of the maximum feasible size and the bleeding of third vessel vapour to its maximum extent will reduce HP steam consumption to a figure approaching 45% on cane.

Pan vapour recompression

A further reduction, down to about 35% steam on cane, can be achieved by employing an MVR with a high pressure ratio on some of the continuous pans. As the process steam consumption is reduced, a point is reached at which all the steam turbine exhaust is being used. For Felixton II, this figure works out to about 37% on cane.

Thus any further load on the power station will only serve to increase the exhaust steam to the point at which it blows off to atmosphere which is extremely wasteful.

Pan vapour recompression would therefore probably be powered by electric motors, the power for which would be purchased from ESCOM. Because of the high pressure ratio of a pan MVR, i.e. 7 to 1, the specific power consumption is nearly 10 times that for the MVR used on the evaporator. The power required would in fact be over 1 MW per continuous pan. At present day prices, the cost of power used on these pan MVR's is slightly less than the saving in coal consumption thus generated.

The recompression of pan vapour is therefore a technique which could be born in mind for the future when it may become an economically attractive proposition.

Exhaust range pressure

The effect of exhaust range pressure is often overlooked when considering the thermal efficiency of a factory.

A high exhaust range pressure may be good for evaporator capacity because of the increased temperature difference in the first vessel. But it is detrimental to the steam rate of turbines. If the quantity of exhaust generated is close to that consumed, it is important to keep the exhaust range pressure as low as possible.

Also, a lower exhaust range pressure leads to a greater quantity of water evaporated in the desuperheater, and hence a marginal reduction in the required steaming rate. In addition, higher exhaust steam pressures lead to higher temperatures in the first effect vessels, which can promote losses of sucrose through degradation. Thus it was decided to design the evaporator and the rest of the factory for as low an exhaust pressure as possible. This results in an evaporator with a large heating surface.

At Felixton II, the total heating surface of the evaporator is 28 860 square metres in total, of which 22 000 square metres (or 76%) is in the 1st and 2nd effect, consisting of four vessels each of 5 500 m^2 heating surface.

Conclusion

The Felixton II sugar mill is a mixture of modern and traditional designs and ideas. With regard to thermal efficiency, the emphasis has been on versatility. At commissioning date, the thermal efficiency will be good enough to provide sufficient fibre for the paper mill while burning a minimum of supplementary fuel. Adequate steam economy is achieved through the use of a quintuple effect evaporator, using vapour 2 for diffuser heating and in the pan house.

Allowance has been made for improving the thermal efficiency by various means, so that if a greater quantity of the bagasse is required for by-product uses, the consumption of coal, as a replacement fuel, would be kept to a minimum.

It will also be possible by modifying the evaporator to run the factory inefficiently, should the need ever arise. For example, if the paper mill should ever shut down, the evaporator would be converted to quadruple effect operation and vapour 1 would be taken off for pans and diffuser heating. In this way all of the bagasse could be used as fuel, without a surplus.

The mill has also been laid out in such a way that it can be extended to a nominal capacity of 900 tons cane per hour, which because of economies of scale, will provide up to 200 000 tons of bone dry fibre per annum without the need to burn coal.

The uncertainties inherent in future planning have been catered for as well as possible in the design of Felixton II. It has also become clear that there are many ways in which the thermal efficiency of a sugar mill can be increased, but that there is a limit beyond which any further improvement is not possible.

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