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onderwerp:

Raw Juice Extraction from Sugar Cane

and Sugar Beet

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LIST OF SYMBOLS

\( c_0 \)  sugar concentration in raw juice  
\( Q \)  quantity of raw juice relative to 1 unit of cane  
\( \varepsilon \)  diffuser efficiency  
\( z_L \)  sugar concentration in bagasse at diffuser output  
\( y_0 \)  sugar concentration in juice at diffuser output  
\( J \)  quantity of juice in cane relative to 1 unit of cane  
\( L \)  length of the diffuser  
\( S \)  sugar concentration in cane  
\( S_1 \)  sugar concentration in juice at 1st mill output  
\( \sigma \)  coefficient taking into account differences between \( S \) and \( S_1 \)  
\( z_0 \)  sugar concentration in bagasse at diffuser input  
\( G \)  quantity of bagasse relative to 1 unit of cane at input  
\( s \)  sugar concentration in bagasse at diffuser input  
\( \Delta \)  diffusion coefficient  
\( \text{GSM} \)  mean specific heat  
\( P_i \)  mass of gas in flue gases relative to 1 kg of fuel  
\( C_i \)  mean specific heat of gas \( i \)  
\( c \)  mean specific heat of water or steam

(The significance of remaining symbols are obvious from the text)
ABSTRACT

This study contains two mains parts:

a) Raw juice extraction from sugar cane,

b) Raw juice extraction from sugar beet.

The first part is divided in:

(i) Energy requirement of the plant
(ii) Energy production for the plant
(iii) By-products manufacture.

The second part is divided in:

(i) Energy requirement of the plant
(ii) Energy production for the plant
(iii) Pressing and drying of wet pulp.

From sugar cane, raw juice can be obtained either by milling or diffusion. Energy considerations and total investment studies have given diffusion to be the most economical process. From a plant capacity of 100 t.c.h., we obtain 116 tons of raw juice per hour with a percentage by weight of 12.5%. Starting from cane with a fibre content of 0.125, 25 tons of bagasse are produced per hour from which 10 tons only are necessary for the total energy supply of the plant. The remaining 15 tons per hour are used for manufacturing briquettes.

From sugar beet, raw juice can only be obtained by diffusion. For a plant capacity of 100 t.c.h. we obtain 119 tons per hour with the same percentage as for sugar cane. Unfortunately, beet pulp cannot be burnt as such in the boiler furnaces to supply the necessary energy for the process. Therefore we have to rely on fuel. An economical study has pointed out that the lowest production cost for dried pulp is obtained by combining high pressing efficiency and drying (hence by using three presses), instead of a lower pressing efficiency (two presses) followed by drying.
INTRODUCTION

In November 1975, Brazil's National Alcohol Programme (PHA), under decree 76.593, was legislated for the production of anhydrous alcohol (anidrol) initially as a blending component in gasolines, and at a later stage as a feedstock for chemical manufacture. The present aim of the programme is to save U.S. $500 million per year in foreign exchange credits on the importation of petroleum. This expresses the response of the Brazilian government to the 1973 oil crisis and shows the will of people to get out of the energy dependence problematic.

At present, serious consideration is being given to sugar cane and sugar beet for use in distilleries. However other crops are also being considered: manioc, a root crop, babassu, a nut, Jerusalem artichoke in the French Carburol programme, and sorgho, the sweet steem, similar to sugar cane but with a thrice as large growth rate and a much more resistant crop.

However, at the present moment, in the optic of the "Fabrieksvoortentwerp" programme at the Delft Institute of Technology, we are exclusively interested in alcohol production from sugar cane and beet. Two other fellow students, Mr van Baalen and Hogendoorn were assigned the task of calculating the distillation process starting from raw juice (the extraction product of cane and beet). Ours is to supply this particular starting material and the necessary energy for the entire plant. An economic study follows, which is determinant in the choice of the adopted extraction process and the possibility of utilizing the by-products is also punctiliously considered.
PART ONE:

THE SUGAR CANE EXTRACTING UNIT
COMBINATION OF CANE PREPARATORS

There are various types of equipment which are placed ahead of the mills with the object of preparing the cane so that the pressure applied at the mills will yield a maximum extraction:

a) The knives
b) The crushers
c) The shredders

Hugot (1, p. 74) recommends for a capacity of 100 tch:
Two sets of knives, the first placed at the bottom of the inclined portion of the carrier, the second just above, with minimum clearance.
Then, a hammer-type shredder which achieve a complete preparation.
After the shredder, a mill-crusher.
Then, four other mills.

This gives us a tandem of 15 rollers. From (1, fig. 21.3), by extrapolation, we realise that the same extraction will be obtained from a 12 rollers battery with an imbibition of \( \lambda = 3.2 \) than from a 15 rollers battery with an imbibition of \( \lambda = ? \) (a common value for imbibition).
In a normal sugar factory, the imbibition water must be calculated to minimise the losses during crystallisation;
In our plant it is necessary to dilute the raw juice to 12.5 % in sugar.
So, we are free to use a high quantity of imbibition water.
Then, by using a 12 rollers battery we will minimise the power consumption and capital cost of the milling unit.

Fig. 21.3 — Variation du jus perdu % de ligneux en fonction de l'imbibition % de ligneux \( \lambda \) (imbibition composée : \( \beta = 0.6 \)).
EXTRACTION EFFICIENCY OF MILLING

In this chapter we will calculate the extraction efficiency which can be obtained with a 12 rollers battery and a high imbibition water.

By extrapolation from (1, fig. 21.3), we find that for a 12 rollers battery with an imbibition $\lambda = 3.2$, the losses in the bagasse will be 33 % undiluted juice % fibre.

If we assume a quantity of cane equal to 1,
The fiber is 0.125
The total sugar content in cane 0.15
The water content of cane 0.725
Hence, we loose in bagasse:
$0.33 \times 0.125 = 0.041$ non diluted juice.
The concentration of non diluted juice is:

$$\frac{\text{sugar in cane}}{\text{sugar in cane} + \text{water in cane}} = \frac{0.15}{0.15 + 0.725} = 0.1714$$

So, the amount of sugar lost in bagasse will be:
$0.1714 \times 0.041 = 0.0070$
and in % bagasse:
$$\frac{0.0070 \times 100}{0.25} = 2.8 \% \text{ bagasse}$$

Hence, the extraction efficiency will be:
$$e = \frac{\text{sugar in cane} - \text{sugar lost in bagasse}}{\text{sugar in cane}} \times 100 = \frac{0.15 - 0.0070 \times 100}{0.15} = 95 \%$$

Dilution of raw juice.
The definition of imbibition is:
$$\lambda = \frac{w}{f}$$

$w$ : quantity of imbibition water
$f$ : fiber content of cane
So, $w = \lambda \times f = 3.2 \times 0.125 = 0.40$

The amount of sugar obtained in cane is:
cane sugar content - sugar lost in bagasse =
0.15 - 0.0070 = 0.1430

The weight of raw juice obtained will be:
water in cane - water in bagasse + imbibition water + sugar in raw juice:
0.725 - 0.25 \times 0.48 + 0.40 + 0.143 = 1.15

So, the concentration of raw juice will be:
\[
\frac{0.143}{1.15} = 12.5\% \text{ by weight;}
\]

**Conclusion**

With a 12 rollers battery and a high imbibition factor (\( \lambda = 3.2 \)) we can obtain a good extraction (95\%) and a raw juice at the required concentration (12.5\% by weight).
PRE-MILLING OPERATIONS

Various types of equipment are placed ahead of the mills with the object of preparing the cane. These are:

(i) The cane carrier
(ii) The knives
(iii) The shredder

The cane carrier

The cane carrier is the moving apron which conveys the cane into the factory, and which assures the feed to the mills by transporting the cane from the yard to the crusher.

Roughly, we may reckon, as a first approximation for the power consumed by cane carriers:

\[ p = \frac{3z + A}{20} = 12.5 \text{ h.p.} \quad (1, \text{p.25}) \]

\[ z = \text{total length of the carrier, in meters} \]
\[ A = \text{crushing rate of the mills, in t.c.h.} \]

The installed power should be appreciably higher, say:

\[ P_i = \frac{3z + A}{10} = \frac{150 + 100}{10} = 25 \text{ h.p.} = 18.4 \text{ kW} \]

Intermediate carriers

The intermediate carriers are the conveyors which move the bagasse from one mill to the feed of the next. The mean power consumption of the intermediate carrier is generally scarcely a matter of concern, since this power is furnished by the mill itself, and in a way forms an integral part of the power required to operate the mill (1, p.80).

Bagasse conveyor

Bagasse leaving the last mill should be elevated for distribution to
the boiler furnaces. It is picked up by a bagasse elevator, which drops it into a horizontal conveyor; this distributes it along the length of the boiler station to the furnaces.

Roughly speaking, we may take 1.10 kW for every 10m total length of the bagasse conveyor (i.e., about twice the actual length of the carrier = upper run + lower run). If we choose 40m, we obtain:

\[ P = 4 \times 1.10 = 4.40 \text{ kW} \quad (1, \text{ p. 96}) \]

Knives

As one set of knives is insufficient to assure satisfactory slicing up of the cane (1, p. 74), we shall commence with two sets, the first placed at the bottom of the inclined portion of the carrier, the second just above, with minimum clearance.

The average power absorbed by a set of knives depends on:

1) The tonnage of cane
2) The fibre in cane
3) The nature of the fibre, whether more or less resistant.
4) The proportion of cane actually cut
5) The number of blades
6) The speed of rotation
7) The radius of the cutting circle
8) Diverse variable factors: friction, lubrication, knives more or less worn.

The following figures are quoted by Hugot (1, p. 42)

**TABLE I**

<table>
<thead>
<tr>
<th></th>
<th>Java</th>
<th>Australia</th>
<th>Queensland</th>
<th>South Africa</th>
</tr>
</thead>
<tbody>
<tr>
<td>1st set</td>
<td></td>
<td></td>
<td></td>
<td>1.5-2</td>
</tr>
<tr>
<td>2nd set</td>
<td></td>
<td></td>
<td></td>
<td>2.5-3</td>
</tr>
<tr>
<td>both</td>
<td>4</td>
<td>2</td>
<td>2</td>
<td></td>
</tr>
</tbody>
</table>

values in h.p./t.c.h.
For mean power consumption, we may estimate in general, for knives, approximately 0.74 - 1.47 kW per t.c.h. On account of the large variations experienced in the feeding of the knives, the power to be installed for the drive motor should be substantially higher than the mean power estimated.

From Hugot (1, p.44) it is advisable to install a motor of 25 h.p. per t.f.h. for factories working at more than 6 t.f.h. Thus, for our case with 12.5 t.f.h. we then install 230 kW.

The shredder

The object of the shredder is to complete the preparation and desintegration of the cane, so as to facilitate the extraction of the juice by the mills.

It is generally assumed that shredders of the Searby type consume on an average about 1.47 kW per t.c.h. However, in order to allow for momentary peak loads, the motors installed have a nominal power of 1.84-2.21 kW per t.e.h. We choose 210 kW per t.e.f.t.

POWER REQUIREMENTS OF MILLS

The determination of the power required by a mill is rather complex because a number of factors enter into it. To begin with, this power may be split up into 6 different terms:

a) Power consumption by compression of bagasse

b) Power consumed in friction between the bearings of the rollers and the shafts

c) Power consumed by friction between bagasse and trash plate

d) Power consumed by friction of scrapers and toe of the trash plate against the rollers, to which should be added the work of dislodging the bagasse at these points

e) Power consumed in driving the intermediate carriers

f) Power absorbed in the gearing

Furthermore, these components of the power themselves depend on certain factors rather difficult to measure or estimate, such as: variety of cane, state of the friction surfaces, quality and regularity of lubrication, adjustment of the setting and of the trash plate, etc... Owing
to the impossibility of taking into account all these factors, the values found in practice may differ appreciably from the mean power figures which we shall derive. This difference may be as much as 20 or even 25% of the normal value furnished by the formula.

Summarizing Hugot's work (1, ch.14), and assembling the mean terms for the powers quoted above, we can write for the total power consumption of one mill:

\[ P = \frac{nD}{\rho} \left[ F \left( 0.4 \frac{6r - 5}{\sqrt{r(1 + \sqrt{r})}} \cdot \sqrt{r} + 0.08 \right) + 4L \right] \]  

(1, eq. 14.27)

P = total power consumed by a three cylinder mill  
\( n \) = rotation velocity of the cylinders, r.p.m.  
\( D \) = cylinder diameter, m  
\( L \) = cylinder width, m  
\( \rho \) = efficiency of the gearings  
\( F \) = total hydraulic pressure exerted on upper cylinder, tons  
\( r \) = reabsorption factor  
\( C_a \) = specific opening between cylinders

N.B. In the case of the first mill, the power required for breaking up the structure of the cane, even when prepared by knives, is substantially higher than that absorbed by compression of bagasse. For this reason, it is recommended (1, p.232) to replace for the first mill the coefficient 0.40 by 0.45.

In order to calculate the power consumed in a given mill, we must first dimensionate that mill, that is, find the values of the different terms in formula [3].

Dimensioning of the mills

From table(II), (1,p. 181) we can derive the number of mills necessary for a capacity of 100 t.e.h., the cylinders diameter and length, the number of rollers. (For more details on choice of number of mills, see discussion on p. 3).
### TABLEAU II

CAPACITÉ DES MOULINS

Pour toute combinaison d’engins correspondant à un coefficient \( c' \neq c \), multiplier par \( \frac{c'}{c} \).

#### TABLEAU 12.3

**Capacité des batteries de moulins, en t c h**

<table>
<thead>
<tr>
<th>Composition de la batterie</th>
<th>D = 3M</th>
<th>4M</th>
<th>D = 5M</th>
<th>6M</th>
<th>D = 6M</th>
<th>7M</th>
</tr>
</thead>
<tbody>
<tr>
<td>Nombre de cylindres ( N )</td>
<td>11</td>
<td>12</td>
<td>14</td>
<td>15</td>
<td>17</td>
<td>18</td>
</tr>
<tr>
<td>( \frac{V}{N} )</td>
<td>3,32</td>
<td>3,46</td>
<td>3,74</td>
<td>3,87</td>
<td>4,12</td>
<td>4,24</td>
</tr>
<tr>
<td>( D \times L ) mm²</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>42 ( \times ) 84</td>
<td>1'065</td>
<td>2'134</td>
<td>2,420</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>40 ( \times ) 84</td>
<td>1'016</td>
<td>2'134</td>
<td>2,199</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>38 ( \times ) 84</td>
<td>965</td>
<td>2'134</td>
<td>1,987</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>36 ( \times ) 84</td>
<td>915</td>
<td>2'134</td>
<td>1,787</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>36 ( \times ) 78</td>
<td>915</td>
<td>1'980</td>
<td>1,658</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>34 ( \times ) 78</td>
<td>864</td>
<td>1'980</td>
<td>1,478</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>34 ( \times ) 72</td>
<td>864</td>
<td>1'830</td>
<td>1,366</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>32 ( \times ) 72</td>
<td>813</td>
<td>1'830</td>
<td>1,210</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>32 ( \times ) 66</td>
<td>813</td>
<td>1'675</td>
<td>1,107</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>30 ( \times ) 66</td>
<td>760</td>
<td>1'675</td>
<td>0,967</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>30 ( \times ) 60</td>
<td>760</td>
<td>1'525</td>
<td>0,881</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>28 ( \times ) 60</td>
<td>710</td>
<td>1'525</td>
<td>0,769</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>28 ( \times ) 54</td>
<td>710</td>
<td>1'370</td>
<td>0,691</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>26 ( \times ) 54</td>
<td>660</td>
<td>1'370</td>
<td>0,597</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>26 ( \times ) 48</td>
<td>660</td>
<td>1'220</td>
<td>0,531</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>24 ( \times ) 48</td>
<td>610</td>
<td>1'220</td>
<td>0,454</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>24 ( \times ) 42</td>
<td>610</td>
<td>1'065</td>
<td>0,396</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>22 ( \times ) 42</td>
<td>560</td>
<td>1'065</td>
<td>0,334</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>20 ( \times ) 36</td>
<td>510</td>
<td>915</td>
<td>0,238</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

\[ D = 760 \text{ mm} \]
\[ L = 1'675 \text{ mm} \]
\[ N = 12 \]

Hence,

\[ \frac{V}{N} = \frac{1065 \times 2134}{2420} \]

\[ D \times L = 1.065 \times 2.134 \text{ mm}^2 \]

\[ n = 5 \text{ tours/minute} \]
The roller speed is then given by formula \( 4 \), (1, p.180)

\[
A = 0.8 \frac{cn(1 - 0.06uD)LD^2\sqrt{N}}{f}
\]

where:
- \( A \) = battery capacity (100t.c.h.)
- \( f \) = percentage bone dry fibre (0.125)
- \( c \) = coefficient relative to preparing machine (1.25) see (1. p.184)
- \( n \) = rotation velocity of cylinders, in r.p.m.

By replacing letters by ciphers, we find \( n = 4.77 \) r.p.m.

Calculation of cylinder opening

Rearranging formula (1, p. 191) we get:

\[
\varepsilon_a = \frac{A f}{330nD^2 Lf_i^1}
\]

values for \( \varepsilon_a \) are tabulated on p. 12
values for \( f_i^1 \) are found in (1, p.196)

Calculation of the reabsorption factor \( r \)

\[
r_i = 0.75 + 0.017v + 0.65 \varphi_i
\]

(1, p.153)

\[
v = \pi nD = 11.39 \text{m/mm}
\]

\[
\varphi_i = \frac{A f}{60WnD\varepsilon ai}
\]

(1, p. 131)

Knowing \( \varepsilon ai \) for each mill, we can calculate \( r_i \). Values for \( r_i \) are tabulated on p. 12

Calculation of the hydraulic pressure \( F \)

\[
F = 1300LD\sqrt{\varepsilon_a \cdot d_{b_1}^2(1 + \sqrt{r} - 1)}
\]

(1, eqn. 10.90)

The only unknown here is \( d_{Bi} \)

\[
d_{Bi} = \frac{1}{V.V.E.} \quad \text{and} \quad V.V.E. = \frac{1}{d_{jc} \times 1.01} - \left( \frac{1.20}{d_{jc} \times 1.01} - 0.86 \right) f'
\]
We have already fixed $f'$ for each mill. Hence

$$d_{jo} = \begin{array}{cccc}
1^{st} \text{mill} & 2^{nd} \text{mill} & 3^{rd} \text{mill} & 4^{th} \text{mill} \\
1.04 & 1.03 & 1.02 & 1.01 \\
\end{array}$$

From this we can calculate $(V.V.E.)$, which gives $d_{Bl}$ and consequently $F_i$.

Table III concentrates the results of the preceding section and gives besides the calculated powers of the four mills from formula 14.27.

### Table III

<table>
<thead>
<tr>
<th>Mill</th>
<th>$f_i$</th>
<th>$e_{ai}$</th>
<th>$\varepsilon_{ai}$</th>
<th>$r_i$</th>
<th>$d_{jo}$</th>
<th>$F_i$</th>
<th>$P_i$</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>mm</td>
<td>$10^2$</td>
<td>kg/m$^3$</td>
<td>kg/dm$^3$</td>
<td>tons</td>
<td>kW</td>
<td></td>
</tr>
<tr>
<td>1</td>
<td>0.33</td>
<td>15.91</td>
<td>2.49</td>
<td>0.648</td>
<td>1.04</td>
<td>258</td>
<td>173</td>
</tr>
<tr>
<td>2</td>
<td>0.42</td>
<td>14.85</td>
<td>1.96</td>
<td>0.824</td>
<td>1.03</td>
<td>307</td>
<td>187</td>
</tr>
<tr>
<td>3</td>
<td>0.47</td>
<td>13.28</td>
<td>1.75</td>
<td>0.922</td>
<td>1.02</td>
<td>370</td>
<td>219</td>
</tr>
<tr>
<td>4</td>
<td>0.50</td>
<td>12.47</td>
<td>1.64</td>
<td>0.982</td>
<td>1.01</td>
<td>390</td>
<td>229</td>
</tr>
</tbody>
</table>

**TOTAL POWER NECESSARY FOR THE MILLING PROCESS**

Now, we can calculate the total power necessary for the milling battery, that is, for the conveying of the cane from the yard through the knives, shredder, mills, and the final distribution of bagasse to the boiler furnaces.

- Cane carrier: 9.2 kW
- Knives: 176.4 kW
- Shredder: 147.0 kW
- First mill: 173.0 kW
- Second mill: 187.0 kW
- Third mill: 219.0 kW
- Fourth mill: 229.0 kW
- Bagasse conveyor: 4.4 kW

**Total**: 1145 kW
In this chapter, we will summarize the advantages and disadvantages of electric drive and steam drive of mills.

A. Advantages of electric drive

a. Neatness and cleanliness
An electric mill house is much neater than one where the mills are driven by steam engines, and even somewhat neater than one with turbine drive: there are no steam joints leaking or dripping, no oil splashing from the crank oiler or dripping from the lubricator, no cumbersome steam pipes.

b. More complete and definite control
Each mill is driven by a separate motor, since electric drive lends itself much better to individual drive than the steam engine and at least as well as the steam turbine. The power consumed by each mill is ascertained at any moment by the simple reading of an ammeter; This is an important point in favour of electric drive, and is quickly translated into improved extraction.

c. Ready general regulation of speed.
The speed of the whole mill tandem is controlled, very conveniently, from the power house.

d. Ease of starting and stopping.
The mills are started by a push-button control. Stopping the mills is equally simple, also their reversal, as required in the case of a choke.

e. Lower operating and maintenance cost.
Costs of lubrication for electric motors are much lower than those involved for steam engines. In the same way costs of maintenance are much lower.

f. Accidents fewer.
No fear of water-hammer, or of fracture of a crank-pin.

g. Safeguard against passage of large pieces of tramp iron.
When a large piece of iron reaches the mill, the electric motor stops...
immediately; with a steam turbine the inertia of the flywheel compels the foreign body to pass, at the expense of the roller grooving.

B. Disadvantages of electric drive

a. Higher first cost of installation
The combination: high pressure boiler, turbo-alternator, switchboards, motors and cables, and supplementary stage of gearing, costs more than the combination: low pressure boiler, steam pipes and steam engines;

b. Additional double transformation of energy.
With an electrical installation, in addition to the analogous transformation effected in the turbo set, the energy must also undergo:
the transformation of movement into electric energy in the alternator;
the transformation of electric energy into movement in the mills motors.
Each of these transformations involves a loss of efficiency.

c. Extra stage of reduction gearing.
The speed of electric motors necessitates the interposition of an extra stage of speed reduction between motor and mill, taking up additional space and involving further loss of power.

d. Less complete speed control.
Turbines and particularly steam engines maintain their power better at low speeds, and are more flexible. The steam range obtainable with a steam engine is much greater and more complete.

e. Accidents more serious.
The accidents liable to occur with electric drive are more serious, and necessitate a judicious provision of spare parts.

f. More specialised personnel.
The electric installation requires more specialised personnel.
D.L. Hughes, (1, p.240) gave figures for the relative performance of a turbine and that of an a.c.-d.c. cascade system (see table IV). He considered that these two systems are those which, with modern technique, would give the most complete answer to the requirement of mill drive.

TABLE IV

Comparaison entre turbine à vapeur et cascade alternatif-continu
faite à égalité de puissance nominale

<table>
<thead>
<tr>
<th>Couple disponible à :</th>
<th>Turbine à vapeur</th>
<th>Cascade altern.-continu</th>
</tr>
</thead>
<tbody>
<tr>
<td>Vitesse 100 %</td>
<td>100 %</td>
<td>100 %</td>
</tr>
<tr>
<td></td>
<td>118 %</td>
<td>125 %</td>
</tr>
<tr>
<td></td>
<td>135 %</td>
<td>166 %</td>
</tr>
<tr>
<td>Consommation de vapeur pour couple 100 % :</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Vitesse 100 %</td>
<td>100 %</td>
<td>100 %</td>
</tr>
<tr>
<td></td>
<td>87 %</td>
<td>80 %</td>
</tr>
<tr>
<td></td>
<td>74 %</td>
<td>60 %</td>
</tr>
<tr>
<td>Couple de surcharge sur 2,5 % du couple de pleine charge :</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Vitesse 100 %</td>
<td>100 %</td>
<td>125 %</td>
</tr>
<tr>
<td></td>
<td>118 %</td>
<td>156 %</td>
</tr>
<tr>
<td></td>
<td>135 %</td>
<td>207 %</td>
</tr>
<tr>
<td>Puissance à installer à capacité égale :</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>100 %</td>
<td>77 %</td>
</tr>
</tbody>
</table>

Since we know now that Hughes was speaking in favour of electric drive, we must admit that, if this table was given in absolute values, instead of in values relative to those corresponding to full speed as a basis, there would be several items where the advantage would lie with the turbine.

However, having balanced the various advantages and desavantages of both techniques, the final choice is more a matter of economics and personal taste. If we agree that the company may afford a higher investment cost by purchasing mills for substantial added returns due to the much lower operating and maintenance costs, then electric drive may be a very good paying proposition.
DIFFUSION

Description

There are two ways of extracting sucrose from cane by diffusion. The first is pure diffusion or diffusion of cane. It consists of operating in manner completely analogous to that of beet; the cane is prepared and broken into small pieces by means of knives, shredder or desintegrator, then is sent to the diffuser. The prepared cane retains its full weight and still contains all its sugar. If the factory is treating 100 t.c.h. at 15% sugar content, the 100 tons of cane and the 15% sugar pass through the diffuser.

The second type of diffusion is called diffusion after mills or bagasse diffusion. The cane, prepared as for milling, goes first into a mill which extracts as much juice as possible, say 65 - 70% of the sugar in the cane. It is the bagasse from this mill which is sent to the diffuser; the diffuser thus receives only 30 - 35% of the sugar in the cane, and the weight of bagasse is perhaps 40 tons for 100 tons of cane. This method of bagasse diffusion is based on the idea that, while it is very expensive to employ 4 or 5 mills to extract 94 - 96% of the sugar in the cane, it is a much more payable proposition to employ one mill to extract 70%; the extraction of 70% by one mill is three or four times superior to that of 16 - 23% for each mill of the train.

For the following reason we prefer the bagasse diffusion process:

a) Passage of the cane through a first mill with coarse grooving conveniently completes the preparation of the cane for diffusion

b) Losses by inversion and fermentation during diffusion no longer act on the whole of the sugar content of the cane but only on 30% of it

c) The diffuser can be greatly shortened, since the material treated no longer contains 14% of sugar but only 4%, that is, 30% of the original sugar within the cane.

DIFFUSION PROCESSES

The diffusion processes which can claim at present to have a place in the world sugar industry are the following:
THE D.d.S. PROCESS

D.d.S. are the initial of "De Danske Sukkerfabrikker", the organisation which produces most of the sugar made in Denmark. The D.d.S. diffusion is the only continuous diffusion process in the strict use of the term. It is simple, not expensive, and the space required is relatively small. For the description of the diffuser, we refer to (1, P.338). The following scheme will give an idea of the process.

Steam consumption: about 25 kg/t.c., say 2500 kg/h for a capacity of 100t.c.h.

Power consumption:

- Cane preparation: The best cane preparation for the D.d.S. process is that obtained by knives, and it is desirable to have two sets. The preparation may also be completed
by a shredder, but the bagasse from a shredder is less permeable than
that from knives, and there is a risk that the gain in extraction ob­
tained by this better preparation may be lost or offset due to the con­
sequent difficulties in regulation and operation. Power of the set of
knives: 176.4 kW (see p. 8).

Mills: The diffuser is preceded by one mill which achieves a complete
preparation of the cane entering the diffuser. In any diffusion process
the megasse cannot be left as such; the cane sugar factory requires fuel
and it is quite impossible to burn megasse, on account of its high wa­
ter content. It is therefore necessary to press this megasse in order
to remove the excess juice. Two mills are generally used to reduce the
moisture content from 85% to 48%, in other words to convert megasse
into ordinary bagasse which can be burnt in the boiler furnaces. Hence,
one mill upstream from the diffuser, Two mills downstream, with an approx­
imate power of 191.1 kW.

Diffuser: Power requirement is relatively low. Each helix is driven by
a d.c. motor through reduction gearing and cardan shafts. The direct
current is given by a set consisting of an asynchronous motor, driving
an independently exited generator. Including the discharge wheel and the
conveyors, the total power required for diffusion is approximately 1.5 kW
per t.c., that is 150 kW for a capacity of 100 t.c.h.

From there we can estimate the total power required by the d.D.S. process

| 2 sets of knives | 176.4 kW |
| 3 mills         | 573.3 kW |
| 1 diffuser      | 147 kW   |
| **Total**       | **896.7 kW** |

THE de SMET PROCESS

The de Smet diffusion process differs from the d.D.S. diffusion process
as well as in energy consumption as in steam consumption. It offers also
some advantages:
a) Good visibility of what is taking place inside the diffuser
b) Possibility of regulating the height of the megasse layer as well as its speed of movement
c) Opportunity of sampling the juice from each tray, over the full length of the diffuser.

Fig. 24.16 show the essential of the process.

Fig. 24.16. — Diffusion du Smet. Bilan-matières.

Power consumption:

The consumed power is about 1.03 kW per t.c.h.; the installed power is around 1.2 kW per t.c.h., say then 120 kW for a capacity of 100 t.c.h. For the preparation of the cane, two ordinary sets of knives may be used, plus one mill. The megasse compression to bagasse requires two mills downstream to the diffuser. From there the total power requirement can be calculated:

2 sets of knives 176.4 kW
3 mills 573.3 kW
1 diffuser (1, p. 352) 120 kW

Total 869.7 kW
Steam consumption:

The quantity of steam consumed varies between 80 and 85 kg per t.c.h.; say 8500 kg/h for a capacity of 100 t.c.h.

B.M.A. DIFFUSION PROCESS

The de Smet diffuser is the standard type of equipment and process from which several other diffusers have been developed, in particular the B.M.A. in Egypt and the Silver Process in the U.S.A. The B.M.A. diffuser is very similar to the de Smet diffuser and is distinguished from it only by details. One of the main differences lies in the method of moving the megasse, which obtained by scrapers pushing the material on fixed perforated plates, with conical holes of one cm in diameter. On account of the fine bagasse passing through these holes with the juice the juice is handled by an unchokeable pump which returns it to the preceding compartment. B.M.A. advises against a shredder and recommend preparation by two sets of knives and one mill. Shredded cane gives lower speeds of percolation (see fig. 24.21 B.M.A. diffuser).

![Diagram](image-url)
Power consumption:

We may reckon approximately 1.00 kW per t.c.h. for the B.M.A. diffuser. For the preparation of the cane, power calculation is similar to previous diffusion processes (1, p. 366)

- 2 sets of knives: 176.4 kW
- 2 mills + 1 roller: 573.3 kW
- 1 diffuser: 100 kW

Total: 849.7 kW

The steam consumption is approximately 120 kg/t.c.h., hence 12000 kg/h for a capacity of 100 t.c.h.

THE SILVER DIFFUSION PROCESS

The Silver diffusion process is also derived from the de Smet type. This process, however, displays its originality not so much in the diffuser as in the efforts made to break as completely as possible with the classical methods of extraction, namely knives and mills.

The power consumption can be calculated as such (1, p. 372)

- Knives: 0.51 kW / t.c.h.
- Cane buster: 3.45 "
- Fiberizer: 3.01 "
- Diffuser: rotation: 0.04 "
- Screws: 0.51 "
- Horizontal bagasse conveyor: 0.04 "
- French press: 4.48 "

Total: 12.04 kW / t.c.h.

For a capacity of 100 t.c.h., this becomes 1204 kW.

The steam consumption for the heaters and steam injectors is of the order of 4 to 5% on cane. That is approximately 4500 kg / h.
CHOICE OF A DIFFUSION PROCESS

Now we have given the characteristics and power requirements of the available diffusion processes, we have to make a choice among them. Therefore, it is necessary to estimate the total investment cost for each process and deduce the production cost of raw juice in the case a diffuser only would be sufficient. Knowing the price of a B.M.A. diffuser, 3 000 000 f (delivered but not installed), we shall assign it a coefficient 1 and make a rough estimate of the other diffusers' costs by comparing their volumes (masses), material used for construction, power consumption, degree of automatisatation, complexity of construction, space required by the unit, etc...

Consequently, the value found for these complexity factors are labelled on p. 23.

For the price used for electricity (0.07 f/kWh), see bagasse manufacturers.

For the price of steam, we used the following calculation:

Energy obtainable from L.P. steam:

\[
\left(143 \degree C, 3 \text{ bar} \right) - \left(90 \degree C, 1 \text{ bar} \right) \times 4.18 = 2236 \text{ kJ/kg}
\]

Corresponding power:

\[
\frac{(625 - 90) \times 4.18}{3600} = 0.6212 \text{ kW/kg/h}
\]

Let us apply a conversion factor of 0.3 to obtain equivalent electric power:

The conversion factor becomes: 0.6212 \times 0.3 = 0.1864

### TABLE V

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>D.d.S.</td>
<td>1</td>
<td>0.76</td>
<td>2 280 000</td>
<td>3 275 400</td>
</tr>
<tr>
<td>de Smet</td>
<td>1.4</td>
<td>1.08</td>
<td>3 240 000</td>
<td>11 761 200</td>
</tr>
<tr>
<td>B.M.A.</td>
<td>1.3</td>
<td>1.00</td>
<td>3 000 000</td>
<td>10 390 000</td>
</tr>
<tr>
<td>Silver</td>
<td>1.5</td>
<td>1.15</td>
<td>3 450 000</td>
<td>12 523 500</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Eq. Pow. of Steam (kW)</th>
<th>Steam Cost</th>
<th>PRODUCTION COST</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Incl. ST &amp; C.F.</td>
<td>Excl. St</td>
</tr>
<tr>
<td>466</td>
<td>156 576</td>
<td>2 195 878</td>
</tr>
<tr>
<td>1584.4</td>
<td>532 258</td>
<td>3 288 617</td>
</tr>
<tr>
<td>2838.8</td>
<td>751 565</td>
<td>3 323 964</td>
</tr>
<tr>
<td>281 837</td>
<td>3 191 081</td>
<td>2 909 244</td>
</tr>
</tbody>
</table>
D.d.S. DIFFUSION. MASS BALANCE.

Our choice goes then to the D.d.S. diffuser. For this type of process, we are now going to calculate the extraction efficiency and the amount of sugar left in the bagasse. For this calculation, we will make use of the formulae developed by Hugot, chapter 24. These equations are:

\[ (1, 24.24; 24.26; 24.27; 24.28; 20.4) \]

Since the simultaneous resolution of these equations gives rise to 3rd and 4th order equation systems, we will proceed by iteration.

We start from 1 of cane. Let suppose a bagasse with 2% sugar. Hence the quantity of sugar in the bagasse will be:

\[ 0.02 \times 0.25 = 0.005 \]

Gazing at figure 24.11 we deduce that:

\[ c_0 \cdot Q + 0.005 = 0.15 \]

since the sugar content in cane is 15%.

From there:

\[ Q = \frac{0.15 - 0.005}{0.125} = 1.16 \]

\[ c_0 = 0.125 \] is our desired final concentration of raw juice.

From formula (24.24) we can now come to know

\[ \varepsilon = \frac{z \cdot L \cdot \gamma_0}{y_0} \left( \frac{Q - J}{Q} \right) \exp \left( \frac{\Delta L (1 - J)}{Q} \right) + J - 1 \]

\[ Q = 1.16 \]

\[ J = 0.15 + 0.725 = 0.875 \]

\[ \Delta L = p \cdot t = 0.13 \times 30 = 3.90 \]

\[ t = \text{residence time of juice in the diffuser} \]

\[ p = \text{factor given by Hugot (1, p. 348)} \]

Hence, \( \varepsilon = 0.4183 \)
From eq. (24.26) we can calculate \( y_o \):

\[
y_o = \frac{JS - (J - S) S_1}{Q - J + G + (1 - J) \varepsilon}
\]

(1. eqn. 24.26)

\( S = \) Sugar \% of cane = \[
\frac{0.15}{0.15 + 0.725} = 0.1714
\]

\( S_1 = \sigma S = 1.03 \times 0.1714 = 0.1843 \) (see eq. 20.4)

\[
f' = \frac{f}{f + G} \quad \Rightarrow \quad G = \frac{f(1 - f')}{f'} = 0.2538
\]

since \( f' = 0.33 \) for the first mill

\( f = 0.125 \) for cane

Knowing all terms in eq. (24.26) we can find \( y_o \)

\( y_o = 0.0600 \)

Out of eq. (24.27) flows \( z_o \)

\[
z_o = \frac{G, g + (J - G) S_1}{J}
\]

and

\[
g = \frac{0.15 - (J - G) S_1}{G} = 0.1399
\]

Hence,

\( z_o = 0.0832 \)

From:

\[
\varepsilon = \frac{z_L}{y_o} \quad \Rightarrow \quad z_L = 0.0251
\]

\[
c_o = \frac{(J - G) S_1 + Q - (J - G) y_o}{Q} = 0.1239
\]

The quantity of sugar in the bagasse becomes:

\( (1 - J)z_L = 0.125 \times 0.0251 = 0.00341 \)

The percentage of sugar in the cane is then:

\[
c_o \cdot Q + 0.000341 = 0.1472
\]

which is almost 15%.

The extraction efficiency is:

\[
e = \frac{\text{sugar in cane} - \text{sugar in bagasse}}{\text{sugar in cane}} = \frac{0.15 - 0.00341}{0.15} = 97.7\%
\]
Fig. 24.11. — Diffusion D.d.S. Efficacité.
CHOICE BETWEEN DIFFUSION AND MILLING

In the great majority of cane sugar factories throughout the world, extraction of sugar from cane is effected by means of mills. Today, many manufacturers of cane sugar return to the other method of extraction, namely, diffusion. There are sound reasons which have prompted them:

The mills consume considerable power, out of proportion to the result obtained.

The mills are very heavy and very expensive, both the purchase price and in cost of operation and maintenance.

Whatever the power expended, it is acknowledged that a certain proportion of the juice contained in the cane cannot be extracted by pressure.

Diffusion permits gains in extraction, but is more critical to control than milling, on account of the risks of inversion and deterioration which are involved. The initial cost is slightly higher for diffusion: it is appropriate to compare the normal tandem with its cane preparators and its four mills, with the same tandem with the diffuser, only three mills and the preparatory plant. Now, the cost of the diffuser is close to that of 1½ mill. The initial cost is thus roughly equivalent, or slightly higher for the diffusion plant.

However, apart from the gain in extraction, diffusion presents an other advantage:

The maintenance of the diffuser is less costly, on account of the cost of rollers, trash plates and scrapers.

Taking into account on one hand the gain in recovery and these economies, on the other hand the slightly higher investment cost, our final preference will go to diffusion.
STEAM PRODUCTION AND USAGE

The fibre in the cane is generally sufficient to enable the bagasse produced by the mills to supply all the steam necessary for power production and for manufacture, when utilized as fuel in the boiler furnace. The excess bagasse can be used in the cane sugar by-products industry, and we will see some applications in a later study.

We shall study successively:

a) the composition of bagasse
b) the combustion of bagasse
c) the boiler and accessories
d) the steam turbine

THE COMPOSITION OF BAGASSE

Final bagasse or simply bagasse is the solid material which leaves the delivery opening of the last mill of the tandem, after extraction of the juice. It is the residue from the milling of the cane.

Physical composition

In spite of the diversity of milling plants and machines employed, the physical composition of bagasse varies between rather narrow limits. Its most important property, from the point of view of steam production, is its moisture content. The most frequent values goes from \( w = 45\% \) to 50\%. (\( w \) is the moisture content of the bagasse.) Generally we shall not involve any great error in adopting for practically all cases the standard value: \( w = 48\% \).

In addition to water bagasse contains:

a) insoluble matter consisting mainly of cellulose, and comprising the fibre content of bagasse
b) substances in solution in the water, consisting of sugar and impurities. They represent between 2 to 4\% of the weight of bagasse.

It remains for the fibres:

\[
f' = 100 - 48 - 2 = 50\%
\]

\( f' = \) fibre \% bagasse
Quantity of bagasse

We have fixed for the fibre content of the cane, \( f = 0.125 \) (mean value given by Hugot (1)). We obtain the quantity of bagasse produced from 100 tons of cane per hour by equating the weight of fibre entering the mills to that leaving:

\[ 100 f = B \cdot f' \]

hence

\[ B = \frac{100 \times 0.125}{0.5} = 25 \text{ t/h} \]

Calorific value of bagasse

The calorific value (or C.V.) is the quantity of heat which will be released by combustion of unit weight of the fuel under consideration. We have two different calorific values:

a) The gross calorific value (G.C.V.): this is the heat liberated by the combustion of 1 kg of the fuel, taken at 0°C and 760 mm Hg, all the products of combustion being reduced to the same conditions. The water present in the fuel, as well as the water formed by combustion of the hydrogen entering into its composition, is consequently condensed.

b) The Nett Calorific Value (N.C.V.), which assumes on the contrary that the water formed by combustion, as also the water of constitution of the fuel, remains in the vapour state.

But since in industrial practice, it has not yet been found practicable to reduce the temperature of the combustion products below the dew point, the N.C.V. gives a more accurate indication of the heat practically obtainable.

The N.C.V. of a fuel is therefore given by the formula:

\[ \text{N.C.V.} = \text{G.C.V.} - 600E \]

\( E = \) weight of vapour present in the gases produced by the combustion of 1 kg fuel, expressed in kg

600 = vaporisation enthalpie of water in kcal/kg
Cross Calorific Value of dry bagasse

In spite of considerable differences in appearance between different varieties of cane, the G.C.V. of dry bagasse is remarkably constant in all countries and for all varieties of cane.

\[ \text{G.C.V.} = 4600 \text{ kcal/kg} \]

Calorific Value of wet bagasse

The G.C.V. and N.C.V. of wet bagasse is given by the formula (1, p. 813)

\[ \text{G.C.V.} = 4600(1 - w) - 1200s \]

\[ \text{N.C.V.} = 4250 - 4850w - 1200s \]

\( s \) = sugar % of bagasse
\( w \) = moisture % of bagasse

We have found (see p.28): \( s = 0.02 \) and \( w = 0.48 \), hence

\[ \text{G.C.V.} = 2368 \text{ kcal/kg} \]

\[ \text{N.C.V.} = 1898 \text{ kcal/kg} \]

Quantity of steam obtainable

We may now calculate the quantity of steam which we can obtain from 1 kg of bagasse. The losses of heat in the furnace and at the boiler consist of the following:

a) Latent heat of the water formed by combustion of hydrogen in the bagasse
b) Latent heat of the water contained in the bagasse
c) Sensible heat of the flue gases leaving the boiler
d) Losses in unburnt solids
e) Losses by radiation from the furnace and especially from the boiler
f) Losses due to bad combustion of carbon giving CO instead of CO₂.

Now, the use of the N.C.V. has already taken into account losses a) and b). The loss c) is given by the formula (1, p. 824)

\[ q = [(1 - w)(1.4m - 0.13) + 0.5]t \]

- \( q \) = sensible heat loss in flue gases, in kcal/kg
- \( t \) = temperature of flue gas, in °C
- \( w \) = moisture content of bagasse
- \( m \) = ratio of air weight used for combustion to that theoretically necessary

For \( m = 1.4, w = 0.48 \) and \( t = 180°C \), we find

\[ q = 260 \text{ kcal/kg bagasse} \]

The three other losses are taken into account by means of coefficients applied to the total quantity of heat which is still available after the first three losses:

\( \alpha \) = coefficient taking into account losses due to unburnt solids; \( \alpha \) is of the order of 0.98 for ordinary furnaces

\( \beta \) = coefficient taking into account losses due to radiation; for a well-lagged boiler a value of 0.975 may be used

\( \eta \) = coefficient taking into account losses due to incomplete combustion

For a well-conducted combustion, we may use a value of 0.96

the quantity of heat remaining to be transferred to the steam is therefore given by the expression:

\[ M_v = (4250 - 1200s - 4850w - q)\alpha \beta \eta \]

Hence,

\[ M_v = 1503 \text{ kcal/kg} \]
The enthalpy of high pressure steam at 400 °C and 40 kg/cm² is 767 kcal/kg.
The enthalpy of feed water at 90 °C to the economiser is 90 kcal/kg. Hence, the energy necessary to supply 1 kg high pressure steam will be:

\[767 - 90 = 677 \text{ kcal/kg} \]

So, the amount of high pressure steam produced by one kg bagasse will be:

\[
\frac{1503}{677} = 2.22 \text{ kg}
\]

Composition of flue gases

We know the total weight of flue gases per kg bagasse burnt:

\[
P_g = 5.76 \left(1 - w\right) m + 1 (1, \text{ eqn. 42.29})
\]

\[
w = 0.48
\]

\[
m = 1.4
\]

Hence, \(P_g = 5.193 \text{ kg}\)

The weight of each individual gas is (cf. 1, p. 817):

a. Nitrogen \(N_2\):
\[
N_2 = 4.43 \left(1 - w\right) m = 3.225 \text{ kg} \quad 62.3 \%
\]

b. Oxygen \(O_2\):
\[
O_2 = 1.33 \left(1 - w\right) (m - 1) = 0.277 \quad 5.3 \%
\]

c. Water \(H_2O\):
\[
H_2O = 0.585 \left(1 - w\right) + w = 0.784 \quad 15.1 \%
\]

d. Carbon dioxide \(CO_2\):
\[
CO_2 = 1.72 \left(1 - w\right) = 0.894 \quad 17.3 \%
\]

Specific heat of flue gases

Mean specific heat of flue gases between 0 (or 30 °C) and \(T\) (1, p. 822)

\[
\begin{align*}
N_2 : \quad & CSM = 0.246 + 0.000020 T \\
O_2 : \quad & CSM = 0.214 + 0.000018 T \\
H_2O : \quad & CSM = 0.468 + 0.000156 T \\
CO_2 : \quad & CSM = 0.199 + 0.000082 T
\end{align*}
\]

The mean specific heat of flue gases will be:
CSM = \sum X_i CSM_i

X_i = weight of gas i in flue gases
CSM_i = specific heat of gas i.

Hence,
CSM = 0.270 + 0.000051 \cdot T

Calculation of combustion temperature

The combustion temperature prevailing in the bagasse furnace is readily calculated from the fact that the heat produced during the combustion is received in the gases passing from the furnace to the boiler:

\[ t = t_0 + \alpha \beta \frac{\sum N_i}{\sum P \cdot C} \cdot CSM \]

\[ t_0 = 103 \text{ °C} \]
\[ \alpha = 0.98 \]
\[ \beta = 0.975 \]
\[ N_i = 1898 \text{ kcal/kg} \]
\[ P = 5.193 \text{ kg} \]
\[ CSM = 0.270 + 0.000051 \cdot T \]

We have to calculate \( T \) by trial and error:
\[ T = 1100 \text{ °C} \quad \Rightarrow \]
\[ CSM = 0.270 + 0.000051 \times 1100 = 0.325 \text{ kcal/kg °C} \]

Hence, \( t = 1182 \text{ °C} \)
STEAM TURBINE

The steam turbine will produce electricity from high pressure steam coming from the boiler.

For complete details about the calculation of the steam turbine, refer to (1, p. 987, 1007).

Data

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure at entry</td>
<td>$p_a = 38 \text{ kg/cm}^2$ \text{ eff.}</td>
</tr>
<tr>
<td>Temperature of superheated steam at entry</td>
<td>$t_a = 390 \text{ °C}$</td>
</tr>
<tr>
<td>Back-pressure at exhaust</td>
<td>$p_b = 3 \text{ kg/cm}^2$ \text{ eff.}</td>
</tr>
<tr>
<td>Power required at the alternator terminals</td>
<td>1991 kW</td>
</tr>
<tr>
<td>Rotational speed of the turbine</td>
<td>$n = 9000 \text{ rpm}$</td>
</tr>
</tbody>
</table>

Calculations

If we assume an alternator of 1500 rpm, we shall need a speed reducer of 6 to 1.

Heat drop. The Mollier diagram gives:

$\Lambda_a = 760 \text{ kcal/kg}$

$\Lambda_b = 625 \text{ kcal/kg}$

$\Lambda_a - \Lambda_b = 135 \text{ kcal/kg}$

Diameter

Since $U = \frac{D \cdot n}{60}$

Let us take $200 = \frac{3.14 \cdot D \cdot 9000}{60}$

Hence, $D = 0.425 \text{ m}$

Number of wheels.

$x \cdot 0.04 = \frac{8205 \cdot 135 \cdot \frac{3^2}{200^2}} = 27.7 \cdot \frac{3^2}{200^2}$
Let us look for the best solution among the different corresponding values.

For $x = 3$  
$\gamma = 0.331$

For $x = 4$  
$\gamma = 0.382$

For $x = 5$  
$\gamma = 0.427$

For $x = 6$  
$\gamma = 0.470$

$x = 6$ gives us a value very close to 0.45 and so the best efficiency.

If on the contrary we required simple and cheap turbine, we could come down to $x = 3$.

But we shall reject these two extreme solutions and take an intermediate solution with $x = 4$ and with $\gamma = 0.382$

Scale of pressure.

We have then $\gamma = 0.382$. Hence:

$$V = \frac{U}{\gamma} = \frac{660}{0.382} = 524 \text{ m/s}.$$  

On the other hand:

$$V = 91.5 \cdot 0.94 \cdot q_1$$

First wheel. Hence:

$$q_1 = \frac{524^2}{91.5^2 \cdot 0.94} = 34.8 \text{ kcal/kg}.$$  

On the Mollier diagram, we look for point 1 on the vertical of A, such that:

$$\Lambda_a - \Lambda_1 = q_1$$

So $\Lambda_1 = \Lambda_a - q_1 = 760 - 34.8 = 725.2 \text{ kcal/kg}.$

Point 1 corresponds to: 29 kg/cm² abs. and 315 °C.

Other wheels.

We have just found the heat drop for the first wheel, $q_1$. We have now to release:
\[ \Lambda_1 - \Lambda_b = 725.2 - 625 = 100.2 \text{ kcal/kg.} \]

and to use three wheels. Consequently each of them will have to work under a heat drop of:

\[ q_n = \frac{100.2}{3} = 33.4 \text{ kcal/kg.} \]

And we can check that we have in fact:

\[ \xi = \frac{U}{V} = \frac{200}{91.5 \times 0.98 \times 33.4} = 0.382. \]

With the help of the Mollier diagram, we can now establish the scale of pressure stages: (see p. 37)

<table>
<thead>
<tr>
<th>Total heat (kcal/kg)</th>
<th>Temperature (°C)</th>
<th>Abs. Pres. (kg/cm²)</th>
<th>Eff. Pres. (kg/cm²)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boiler</td>
<td>767</td>
<td>400</td>
<td>40</td>
</tr>
<tr>
<td>Admission</td>
<td>760</td>
<td>390</td>
<td>39</td>
</tr>
<tr>
<td>1st wheel</td>
<td>725.2</td>
<td>315</td>
<td>29</td>
</tr>
<tr>
<td>2nd wheel</td>
<td>691.8</td>
<td>240</td>
<td>16</td>
</tr>
<tr>
<td>3rd wheel</td>
<td>658.4</td>
<td>170</td>
<td>8</td>
</tr>
<tr>
<td>4th wheel</td>
<td>625</td>
<td>143</td>
<td>4</td>
</tr>
</tbody>
</table>

Steam consumption.

The turbine will drive the alternator. Then, the steam consumption per kWh at the terminals of the switch board, is given by:

\[ Q = \frac{860}{(\Lambda_a - \Lambda_b) \cdot \eta \cdot \eta' \cdot \eta} \quad (1, \text{eqn. 44-42}) \]

\[ \eta = 0.72 \quad (1, \text{table 44-1}) \]
\[ \rho_r = 0.98 \quad (1, \text{p. 919}) \]
\[ \rho_s = 0.93 \quad (1, \text{p. 919}) \]

Hence, \[ Q = \frac{860}{135 \times 0.72 \times 0.98 \times 0.93} = 9.71 \text{ kg/kWh}. \]

In order to obtain the actual steam consumption, we shall have to add to the above value \( Q \):

For losses through condensation 3% (1, p. 919)

For losses by leaks 2% (1, p. 919)

So, \[ Q = 9.71 \times 1.05 = 10.2 \text{ kg/kWh}. \]

The quantity of steam necessary to provide 1991 kW will be:

1991 \times 10.2 = 20306 \text{ kg/h}. 

---

FIG. 41: Diagram de Mollier.
Bagasse furnace

There are four types of bagasse furnace:

a. The step-grate furnace
b. The cook or horsehoe furnace
c. The ward furnace
The spreader-stoker furnace.

a. The step-grate furnace
The step-grate furnace, the oldest one is now no longer produced.

b. The horsehoe furnace, more recent permits of very high combustion rates and gives excellent efficiency.

c. The ward furnace, of american origin, closely ressembles the horsehoe furnace. It permits a great saving in the total floor space necessary for furnace and boiler, and a substantial saving in refractories. It has given excellent result and high efficiency.

d. The spreader stoker furnace
This is the most recent type of furnace. The quantity of unburnt solids remaining in the ash pit is greater (c=0.975). But it is considered that the spreader stoker furnace permits of reducing the normal excess air to 40 % and consequently of improving the efficiency consequently. Moreover, this type of furnace permits a combustion rate very superior to that of other types (35 to 40 kg of steam per hour and m² of heating surface) Finally, the spreader stoker furnace makes ash removal easy, is easy to clean and economical in brickwork. Because of all these advantages, a spreader stoker furnace will be chosen.

Combustion chamber volume

The volume of combustion chamber should be proportioned to the volume of gases necessary for combustion. The volume of the combustion chamber is then given by:

\[ V = \frac{B \cdot N_i}{200000} \]

\[ V = \text{combustion chamber volume in m}^3 \]
\[ B = \text{weight of bagasse's wt. in kg/h: 12500 kg/h} \]
\[ N_i = NCV \text{ of bagasse} = 1503 \text{ kcal/kg} \]

Hence, \( V = 101 \text{ m}^3 \)

**Calculation of the superheater**

We have two principal equations:

\[ M = \alpha PC (T_1 - T_2) = p ((1 - X) r + c (T - t)) \]  \( (1, \text{ eqn. 42.60}) \)

- \( M \): quantity of heat transmitted to the steam, in kcal/h.
- \( \alpha \): coefficient of efficiency, generally 0.90
- \( P \): weight of gas passing over the superheater, in kg/h
- \( C \): specific heat of these gases, in kcal/kg °C
- \( T_1 \): temperature of gases at entry to the superheater, in °C.
- \( T_2 \): temperature of gases leaving the superheater, in °C.
- \( p \): weight of steam to be superheated, in kg/h.
- \( X \): dryness fraction of saturated steam, between \( t \) and \( T \)
- \( t \): temperature of saturated steam, at the boiler pressure.
- \( T \): temperature of superheat desired.

\[ M = kS \left( \frac{T_1 + T_2}{2} - \frac{T + t}{2} \right) \]  \( (1, \text{ eqn. 42.61}) \)

- \( k \): coefficient of heat transfer, in kcal/m²/h/°C
- \( S \): heating surface of the superheater, in m²

Heat available per kg bagasse burnt : 1503 kcal

Weight of steam supplied per kg bagasse burnt : 2.22 kg

Hence, total weight of bagasse to be burnt to produce 29967 kg/h steam

\[ B = \frac{29967}{2.22} = 13498 \text{ kg/h} \]

Weight of gases produced. We have \((1, \text{ eqn. 42.29})\):

\[ P_g = 5.76 (1 - w) m + 1 = 5.193 \text{ kg} \]

Or, in total: \( P = 5.193 \times 13498 = 70055 \text{ kg/h} \).
Temperature of gases leaving the superheater.

\[ T_1 - T_2 = \frac{p}{\alpha \cdot PC} ((1 - x) \cdot r + c \cdot (T - t)) \]  

(l, eqn. 42-60)

\[ \alpha = 0.90 \]  

(l, p.844)

\[ p = 29967 \text{ kg/h} \]

\[ T_1 = 1182 \text{ °C} \]

\[ P = 70055 \text{ kg/h} \]

\[ C = 0.333 \text{ kcal/kg °C} \]  

(between \( T_1 \) and \( T_2 \))

\[ x = 0.98 \]  

(l, p.845)

\[ r = 411 \text{ kcal/kg} \]  

(l, table 42-1)

\[ c = 0.614 \text{ kcal/kg °C} \]  

(l, eqn. 42-64)

\[ T = 400 \text{ °C} \]

\[ t = 249 \text{ °C} \]  

(l, table 42-1)

Hence, \( T_2 = 1182 - \frac{29967}{0.90 \cdot 70055 \cdot 0.333} \cdot (0.02 \cdot 411 + 0.614 \cdot (400-249)) \)

\[ = 1038 \text{ °C} \]

The heating surface of the superheater installed is thus:

\[ S = \frac{p \cdot (1-x) \cdot r + pc(T - t)}{k \cdot (\frac{T_1 + T_2}{2} - \frac{T + t}{2})} \]

\[ k = 55 \]  

(l, p. 844)

Hence \( S = \frac{29967 \cdot (1 - 0.98) \cdot 411 + 29967 \cdot 0.614 \cdot (400-249)}{55 \cdot (\frac{1182 + 1038}{2} - \frac{400 + 249}{2})} \)

\[ = 70 \text{ m}^2. \]
BOILER

Quantity of heat transmitted to steam or water:

\[ M = \alpha \cdot P \cdot C \cdot (T_o - T_1) = P \cdot (X \cdot r + c \cdot (T - t_o)) \]

\[ \alpha = 0.90 \ (1, \ p. \ 844) \]

\[ P = 70055 \ \text{kg/h} \]

\[ T_o = 1038 \ ^\circ C \]

\[ C = 0.339 \ \text{kcal/kg} \ ^\circ C \ (\text{between} \ T_o \ \text{and} \ T_1) \]

\[ p = 29967 \ \text{kg/h} \]

\[ X = 0.98 \ (\text{dryness fraction of steam at boiler output}, \ 1 \ p. \ 845) \]

\[ r = 411 \ \text{kcal/kg} \ (1, \ \text{table 42.1}) \]

\[ c = 1 \ \text{kcal/kg} \]

\[ T = 249 \ ^\circ C \ (1, \ \text{table 42.1}) \]

\[ t_o = 161 \ ^\circ C \ (\text{water temperature at economiser output}) \]

Hence, \( T_1 = 350 \ ^\circ C \)
ECONOMISER

The boiler feed water is approximately 90°C.

Now, the saturation temperature at which the water should be fed into the boiler to be transformed into steam, according to the pressure used, is 249°C.

There is then a large margin of temperature to make up in the boiler. This margin means that a substantial proportion of the total heat has to be supplied to the water before evaporation proper commences.

Total heat content of liquid water at 249°C and 40 kg/cm²: 258 kcal/kg.

Total heat content of liquid water at 90°C: 90 kcal/kg.

Sensible heat to be supplied: 168 kcal/kg.

Total heat content of steam at 400°C and 40 kg/cm²: 767 kcal/kg.

So, the total heat supply to obtain steam at 400°C and 40 kg/cm², is 767 kcal/cm².

Hence, the sensible heat represents 25% of the total heat supply.

Now, the combustion gases leave the boiler at a temperature which is still relatively high, and generally above the saturation temperature. This sensible heat content of the gases would be lost in the stack. Hence, the idea of a heat exchanger to raise the temperature of the incoming water.
Air heater

In addition to the economiser, there exist another type of equipment permitting of the partial recovery of the sensible heat of the combustion gases, which are still hot as they pass to the chimney. Instead of absorbing this heat in the boiler feed water, it is absorbed by the air which is to be used for combustion in the furnace. The equipment is then called an "air heater" or "pre-heater".

We shall use in this paper an economiser as well as an air-heater. The combination of these two heat exchangers presents some advantages:

a. A large safety margin for the economiser, since only part of the possible temperature rise is being sought.

b. No risk of deterioration of refractories due to the high air temperature for the same reason.

c. In the case of a breakdown of the first heat-exchanger, there is a possibility that the second exchanger will compensate for it to a certain extent, consequently on the increase of temperature in the gases entering it.

Calculation of heat exchange surface of the economiser and air-heater

The basic equations are:

\[ M = \alpha \cdot P \cdot C \cdot (T_1 - T_2) = P \cdot C \cdot (t - t_0) \]

\[ M' = \alpha' \cdot P \cdot C \cdot (T_1 - T) = P' \cdot C' \cdot (t' - t'_0) \] (1, eqn. 42.67)

We will choose \( M = 2 \cdot M' \) because the heat exchange between air and water is more efficient than air-air exchange, and because the flue gases temperature is higher in the first exchanger. So, we will obtain an economiser and a air heater with about the same surface.

Hence:

\[ \alpha \cdot P \cdot C \cdot (T_0 - T_1) = 2 \cdot \alpha' \cdot P \cdot C \cdot (T_1 - T) \]

\[ \alpha \cdot (t_0 - T_1) = 2 \cdot \alpha' \cdot (T_1 - T) \]

\[ T_1 = \frac{\alpha \cdot T_0 + 2 \cdot \alpha' \cdot T}{2 \cdot \alpha' + \alpha} \]

\[ \alpha = 0.95 \quad (1, \ p. \ 849) \]

\[ \alpha' = 0.935 \quad (1, \ p. \ 855) \]

\[ T_0 = 350 \ ^\circ C \quad (1, \ p. \ 850) \]

\[ T = 180 \ ^\circ C \]

Hence, \( T_1 = 237 \ ^\circ C \)
So, \( r = \frac{\alpha \cdot PC}{pc} = \frac{t - t_o}{T_o - T_1} \)  

(1, eqn.42.69)

\[ \alpha = 0.95 \]

\[ P = 70055 \text{ kg/h.} \]

\[ C = 0.284 \text{ kcal/kg °C} \]

\[ p = 29967 \text{ kg/h} \]

\[ c = 1 \text{ kcal/kg °C} \]

So, \( r = \frac{0.95 \cdot 70055 \cdot 0.284}{29967 \cdot 1} = 0.623 \)

\[ t = t_o + r \cdot (T_o - T_1) \]  

(1, eqn.42.73)

Hence \( t = 90 + 0.628(350 - 237) = 161 \text{ °C} \).

Heat exchange surface of the economiser

\[ S = \frac{\alpha \cdot PC}{k \cdot (1 - r)} \cdot \frac{\ln \frac{T_o - t}{T_1 - T_o}}{1} \]  

(1, eqn.42.70)

\[ k = 25 \]  

(1, p. 850)

Hence, \( S = \frac{0.95 \cdot 70055 \cdot 0.284}{25 \cdot (1 - 0.628)} \cdot \ln \frac{350 - 161}{237 - 90} = 509 \text{ m}^2 \).

Calculation of the air-heater exchange surface

\[ r' = \frac{\alpha' \cdot PC}{p'c'} = \frac{t' - t_o}{T_1 - T} \]

\[ \alpha' = 0.935 \]  

(1, p.355)

\[ c' = 0.24 \text{ kcal/kg °C} \]

Calculation of \( p' \)

We know that \( p_a = 5.76 \cdot (1 - w) \cdot m \)  

(1, eqn. 42-28)

\[ w = 0.48 \]

\[ m = 1.4 \]
Hence \( p_a = 4.19 \) kg air/kg bagasse.

Quantity of bagasse burnt : 13498 kg/h.

So \( p' = 4.19 \times 13498 = 56557 \) kg/h

Hence \( r' = \frac{0.935 \times 70055 \times 0.284}{56557 \times 0.24} = 1.37 \)

So \( t' = t_o + r(T_f - T) \)

\( t'_o = 25 \) °C (room air)

\( T = 180 \) °C (temperature of the flue gases at the stack)

Hence \( t'_f = 25 + 1.37(237 - 180) \)

\( = 103 \) °C.

So, the air heater heat exchange surface is:

\[
S' = \frac{\delta \cdot P C \cdot \ln(T_f - t'_f)}{k'(1 - r') \cdot T - t'_o}
\]

\( k' = 18 \) (1, p.857)

Hence, \( S' = \frac{0.935 \times 70055 \times 0.284 \cdot \ln(237 - 103)}{18 \cdot (1 - 1.37) \cdot 180 - 25} \)

\( = 406 \text{ m}^2 \)
DRAUGHT FAN POWER REQUIREMENTS.

In order to maintain the temperature and the rate of combustion, it is necessary to pass through the furnace and over the fuel bed, the required quantity of air. Since the path of the gases is complex, with many resistances to overcome, maintaining this flow of gases demands a certain expenditure of energy which is normally supplied in the form of a motive pressure by a fan.

a. Forced draught. The aim is to establish in the furnace a pressure of:

To this must be added pressure drops in:

- air duct: 10 mm
- changes of direction: 30 mm
- across air heater (air side): 20 mm

90 mm W.G.

b. Induced draught. At the top of the furnace there is a required suction of:

Pressure drops are:

- across the boiler tubes: 25 mm
- across the economiser: 10 mm
- across the air heater (gas side): 15 mm
- across the stack: 7 mm

60 mm W.G.

Power requirements of the fans.

The approximate formula for power used in sugar factories is:

\[ P_f = \frac{B \cdot d \cdot (273 + t)}{8400} \]  
(1, eqn. 42-89)
\[ P_f = \text{power requirement of the fan, in h.p.} \]
\[ B = \text{quantity of bagasse consumed by the furnace served by the fan in t/h.} \]
\[ d = \text{draught at suction of fans, in mm of water.} \]
\[ t = \text{gas temperature at suction of the fan, in } ^\circ\text{C.} \]

a. **Forced draught**

\[ B = 13 \text{ t/h} \]
\[ d = 90 \text{ mm} \]
\[ t = 25 ^\circ\text{C} \]

Hence, \[ P_{f1} = \frac{13}{8400} \times 90 \times (25 + 273) = 41.5 \text{ h.p.} = 30 \text{ kW} \]

b. **Induced draught.**

\[ B = 13 \text{ t/h} \]
\[ d = 60 \text{ mm} \]
\[ t = 180 ^\circ\text{C} \]

So, \[ P_{f2} = \frac{13}{8400} \times 60 \times (180 + 273) = 42 \text{ h.p.} = 31 \text{ kW} \]

**BOILER FEED WATER PUMP.**

The pump has to work under the following pressure (1, p.871):

a. **Certified pressure**  
\[ 40 \text{ kg/cm}^2 \]

b. **Excess pressure to free safety valve**  
(about 3% of working pressure):  
\[ 1.2 \text{ kg/cm}^2 \]

c. **Head losses in the piping about:**  
Gravity head corresponding to the height of the boiler above the pump:  
\[ 0.1 \text{ kg/cm}^2 \]

e. **Head losses in the economiser, about:**  
\[ 0.5 \text{ kg/cm}^2 \]

f. **Head losses in the regulated valve:**  
\[ 1 \text{ kg/cm}^2 \]

g. **A safety margin (about 10% of the total preceeding):**  
\[ 4.2 \text{ kg/cm}^2 \]

\[ 48 \text{ kg/cm}^2 \]
Power requirement.

Power requirement of centrifugal pumps is given by:

\[ P = \frac{Q \cdot \Delta H}{75 \cdot \rho_m} \]  
(1, ecm. 46-9)

\[ \rho_m = 0.65 \]  
(1, p.870)

\[ \Delta H = 480 \text{ m} \cdot \text{bar} \]  

So, \[ P = \frac{29000 \cdot 480}{75 \cdot 0.65 \cdot 3600} = 79 \text{ h.p.} = 58 \text{ kW} \]

STEAM REDUCING VALVE AND DE-SUPERHEATER

The quantity of low pressure steam produced by the turbine is: 20306 kg/h.

The steam required for the diffusion process is: 2500 kg/h.

The steam required by the manufacture of by-products is: 8800 kg/h.

So we have an excess of low pressure steam of: 9006 kg/h.

Then, the process requires 19400 kg/h of superheated steam at 3 kg/cm² and 180 °C.

This steam will be obtained by deriving a part of the steam produced by the boiler directly into the low pressure steam system, with a pressure regulator.

The steam obtained is then superheated. Hence it is necessary to follow the pressure regulator by a de-superheater.

Calculations.

Total heat content of high pressure steam at 390 °C and 39 kg/cm²: 760 kcal/kg

Total heat content of low pressure steam at 180 °C and 3 kg/cm²: 672 kcal/kg

Total heat content of low pressure steam at 143 °C and 3 kg/cm²: 625 kcal/kg

Total heat content of liquid water at 90 °C: 90 kcal/kg.

So we will obtain the quantity of high pressure steam and liquid water to add to the
low pressure steam by a mass balance and a heat balance:

\[ x = \text{quantity of high pressure steam} \]

\[ y = \text{quantity of liquid water} \]

\[
\begin{align*}
19400 &= 9006 + x + y \\
19400 \cdot 672 &= 9006 \cdot 625 + x \cdot 760 + y \cdot 90
\end{align*}
\]

\[
\begin{align*}
10394 &= x + y \\
7408050 &= 760 \cdot x + 90 \cdot y
\end{align*}
\]

Hence \[
\begin{align*}
x &= 9661 \text{ kg/h} \\
y &= 733 \text{ kg/h}
\end{align*}
\]
VI ENERGY REQUIREMENTS OF THE PLANT

We have made a choice between the different possibilities, milling or diffusion, electrical or steam drive of the machines, and we have calculated the power requirements of the electric motors of the plant. So we can now calculate the total power requirement of the whole plant.

<table>
<thead>
<tr>
<th>Component</th>
<th>Power Requirement</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cane carrier</td>
<td>12.5 h.p. = 9.2 kW</td>
</tr>
<tr>
<td>Two sets of knives</td>
<td>240 h.p. = 176.4 kW</td>
</tr>
<tr>
<td>Three mills</td>
<td>780 h.p. = 573.3 kW</td>
</tr>
<tr>
<td>D.d.S. Diffuser</td>
<td>200 h.p. = 147 kW</td>
</tr>
<tr>
<td>Bagasse distributor</td>
<td>6 h.p. = 4.4 kW</td>
</tr>
<tr>
<td>Forced draught Fan</td>
<td>41 h.p. = 30 kW</td>
</tr>
<tr>
<td>Induced draught Fan</td>
<td>42 h.p. = 31 kW</td>
</tr>
<tr>
<td>Boiler feed water pump</td>
<td>79 h.p. = 58 kW</td>
</tr>
<tr>
<td>Manufacture of briquettes</td>
<td>682 kW</td>
</tr>
<tr>
<td>Manufacture of alcohol</td>
<td>280 kW</td>
</tr>
<tr>
<td>Total</td>
<td>1991 kW</td>
</tr>
</tbody>
</table>
BY-PRODUCTS MANUFACTURE
UTILIZATION OF BAGASSE AS A BY-PRODUCT

The fibre in the cane is generally sufficient to enable the bagasse produced by the mills to supply all the steam necessary for power production and for manufacture, when utilized as fuel in the boiler furnaces. With a normal fibre content (12.5%) and a well-balanced and well-designed factory, there remains in addition an excess of bagasse (or of steam) which may be used for other purposes; irrigation pumping, steam for the distillery, supply of electric energy to the regional network, manufacture of by-products. It is this last aspect that we are going to take a close look at.

As we have seen in the preceding section, an alcohol plant with a total capacity of 100 t.c.h., supplies every hour 25 tons of bagasse. The quantity of bagasse used for the plant needs, from the conveying of the cane until final distillation is 10 tons per hour. There remains thus an excess of 15 t/h surplus bagasse.

WHAT CAN WE DO NOW WITH THIS SURPLUS BAGASSE

At present, bagasse is used for many purposes, some of which have given more or less satisfying results. Briefly, we may quote:

1) Direct or indirect utilization as fuel
   a) steam production
   b) electricity generation
   c) manufacture of briquettes
   d) bagasse charcoal
   e) methane production

2) Utilization of fibrous products of bagasse
   a) pulp and paper manufacture
   b) paper board, corrugating board, box board
   c) fibre board
   d) particle board
3) **Chemicals from bagasse**

a) production of furfural  
b) production of α-cellulose  
c) production of plastics  

4) **Miscellaneous**

a) poultry litter  
b) mulch  
c) soil conditioner  
d) bagasse concrete

Let us follow this list and immediately eliminate the by-products which for one reason or another, are not a paying proposition for our manufacture.

1-a,b) We have already seen how bagasse could be used for electricity and steam production. Although we will be concerned with this aspect for the surplus bagasse, only the economic side will be treated.

1-d) The production of bagasse charcoal is a by-product of the sugar factory. For this, bagasse is mixed with a certain quantity of molasse, and further carbonized. In our plant, we do not dispose of any molasse.

1-e) The production of methane by anaerobic fermentation of organic wastes is rather incompetently mastered. Theoretically, it is possible to get from one ton of bagasse, 200 m$^3$ biogas, which is a mixture of 65% methane and 35% CO$_2$. But results varies so much from plant to plant that a strict calculation method cannot be given.

2-a,b) Paper production is a very competitive business, and only the relative shortage of softwood pulps in certain areas can create a rendable market for bagasse pulp. To achieve economy of scale, the factory paper production should have a minimum production of 500 tons a day. Perhaps that for smaller capacities the production of paper from bagasse is possible in a protec-
ted market; however, we must emphasize that in a protected mar-
ket uneconomic products can temporarily be sold; but very soon
the acid test of free competition plays havoc with such produc-
tions.

3-a) Furfural has actually a declining market in the world. It was
used before in the manufacture of adiponitrile from which ny-
on 66 is made. But it has to compete now against the much cheaper
cyclohexane and butadiene.

3-b,c) The manufacture of dissolving pulp from bagasse is not encour-
raging. To be competitive it is necessary to produce about 400
tons a day of plastics, which is not possible in our case.

4-a,b,c) Here, we simply have to consider the profit obtainable by sel-
ling mill run bagasse. Therefore a cost price for mill run ba-
gasse has to be fixed. We do not expect much profit from these
points for, the bagasse having been submitted to very lit-
tle transformation cannot yield much.

PRODUCTION OF ELECTRICITY FROM SURPLUS BAGASSE

The burning of bagasse for steam generation and the utilization of this
high or medium pressure steam for electricity generation is a standard
practice as we have seen in the preceding section. We are here concerned
with the burning of surplus bagasse for the production of surplus elec-
tricity to be sold to the grid.

We may consider here three possibilities:

a) Electricity generation at the plant
b) Selling bagasse to a central power station
c) Being the central power station

a) Electricity generation at the plant

Our factory, with a total capacity of 100 t.c.h. has a steam requirement
of 22 tons per hour. It produces 25 tons per hour bagasse. From 1 kg of
bagasse, it is possible to produce 2.22 kg of high pressure steam (see
p.32). Hence, with a suitable boiler we could generate in total:

\[ 25000 \times 2.22 = 55.6 \text{ tons steam per hour} \]
After distillation, we are left with 33.6 t. Since 10.2 kg high pressure steam generate 1 kWh electricity, from the 33.6 we may recover 3294 kW. To produce our surplus electricity we would require a turboalternator with an installed power of 3788 kW (3294 + 15%). The price of such a turboalternator, inclusive boiler, control commands, building, is about 1000f/kVA. The cost price of our unit will then be: 3 788 000f.

The price at which the electrical unit should be sold to make generation at the factory a paying proposition, can be roughly evaluated as follow:

Let us consider the 3788 kW turbo-alternator producing on average 3300 kW for 20 hours a day for a 240 days-crop, i.e., 15 840 000 kWh. The capital cost, that is, the installed cost of the turbo-alternator and accessories is f-3 780 000.

Assume 10% for depreciation, 5% for maintenance and man power, and 10% interest on 40% of the invested capital. Hence, the minimum price at which unit electrical energy can be sold, can be estimated:

\[
X = \frac{3 780 000 \times (0.10 + 0.05 + 0.04)}{15 840 000} = 0.05 \text{ f/kWh}
\]

Since the official price for electricity produced by enterprises is 0.07 f/kWh, we can expect a net profit of:

\[
15 840 000(0.07 - 0.05) = 316 800 \text{ f/year}
\]

which is 8% of the capital cost, and 42% on the production cost.

Cost of bagasse as a raw material

Note that in the above we have reckoned no value for the bagasse burnt. As we can imagine, bagasse, although often regarded as a waste product, if it is to be further utilized outside the cane distillation steam generating plant, must bear the costs of handling and storage, to which must be added, very often, the cost of depithing and transport. Consequently, unless some added return is indicated, over and above the fuel oil replacement value and the other handling and storage charges, there can be little incentive for the distilling factories to embark on capital expenditure to convert there furnaces from bagasse to fuel oil.
However, even in cases where a surplus is available, again a definite price incentive should exist, commensurate with the storage and handling charges.

Taking this into account, we may adopt 2 prices for bagasse, whether it is to replace fuel oil in the factory, or to be sold just as surplus.

If $X$ is the cost of fuel oil in f/tons
$Y$ the cost of labour in f/man-hour
$Z$ the cost of depithing

(i) Then, for a factory with surplus bagasse, the cost of this bagasse would be approximately:

<table>
<thead>
<tr>
<th>Cost Item</th>
<th>Cost (f)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Baling wire</td>
<td>1.20</td>
</tr>
<tr>
<td>Baling station</td>
<td></td>
</tr>
<tr>
<td>Storage site</td>
<td>0.24</td>
</tr>
<tr>
<td>Equipment</td>
<td>2.00</td>
</tr>
<tr>
<td>Labour</td>
<td>2.50Y</td>
</tr>
<tr>
<td>Depithing</td>
<td>$Z$</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>$3.64 + 2.50Y + Z$</td>
</tr>
</tbody>
</table>

(ii) For a factory with bagasse replaced by fuel oil, the cost of this bagasse could be approximately:

<table>
<thead>
<tr>
<th>Cost Item</th>
<th>Cost (f)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Baling wire</td>
<td>1.20 f/ton</td>
</tr>
<tr>
<td>Baling station</td>
<td>0.20 &quot;</td>
</tr>
<tr>
<td>Storage site</td>
<td>0.24 &quot;</td>
</tr>
<tr>
<td>Equipment</td>
<td>2.00 &quot;</td>
</tr>
<tr>
<td>Boiler conversion cost</td>
<td>0.48 &quot;</td>
</tr>
<tr>
<td>Labour</td>
<td>2.50Y &quot;</td>
</tr>
<tr>
<td>Fuel replacement value</td>
<td>0.123X &quot;</td>
</tr>
<tr>
<td>Depithing</td>
<td>$Z$</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>$4.12 + 2.50Y + 0.123X + Z$</td>
</tr>
</tbody>
</table>

The fuel oil replacement index can be calculated as follow:

$$\frac{(\text{Calorific value of bagasse}) \cdot (\text{boiler efficiency for bagasse})}{(\text{Calorific value of fuel oil}) \cdot (\text{boiler efficiency for fuel oil})} = 1$$
\[
\frac{1503 \times 4.18}{4000 \times 0.75} = 0.123
\]

From the above we can calculate (i) and (ii).

\( Y = 4 \) f/man-hour for involved country

\( Z = 0 \)

\( X = 420 \) f/ton heavy oil

Hence,

(i) = 13.64 f/ton mill run bagasse

(ii) = 65.78 f/ton

From this we can estimate the total entries for surplus bagasse over the whole year:

\[ 13.64 \times 15 \times 20 \times 240 = 982,000 \text{ f/year.} \]

Since there exists no market price for bagasse, we must rely on offer and demand.

The factory willing to replace bagasse by fuel oil in order to use his bagasse for higher added returns, can only do it if these added returns exceed:

\[ 65.78 \times 10 \times 20 \times 240 = 3,157,440 \text{ f/year} \]

The actual price of fuel makes this proposition of little incentive for the factory, unless the bagasse saved can further be converted into high value products.

Central power station

Let us consider now the case where the factory is willing to functionate as a central power station in order to produce additional electricity for intercrop utilization. Two additional problems present themselves in this case:

(i) The transport of baled bagasse from the sugar estates to the central station during the crop;

(ii) The storage and handling of the bagasse at the central station for its intercrop utilization.
The transport of baled bagasse, which has generally a density of 550 kg /m$^3$, is a costly item if the factories supplying the bagasse are relatively far from the central station. Now, we are going to make a purely speculative calculation. Say, the average distance from the central station of the three supplying factories is 50 km. The price of transport per km and per ton of bagasse can be calculated:

\[
x = \text{price of transport per km and per ton bagasse} \\
y = \text{rent price of truck per day} \\
z = \text{truck capacity in tons} \\
n = \text{number of kilometers covered in one day} \\
g = \text{price of diesel oil in guilders per liter.} \\
c = \text{truck oil consumption in liters per km}
\]

Hence, the price of transport per ton and per km will be:

\[
x = \frac{y}{z \cdot n} + \frac{g \cdot c}{z} = \frac{850}{20.450} + \frac{3.33}{20} = 0.26 \text{ f/km/ton}
\]

From there we can estimate the production cost per ton of bagasse:

<table>
<thead>
<tr>
<th>Cost</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cost of transport (0.26x50)</td>
<td>13.00 f/ton</td>
</tr>
<tr>
<td>Cost of baling</td>
<td>0.20 &quot;</td>
</tr>
<tr>
<td>Cost of storage</td>
<td>0.24 &quot;</td>
</tr>
<tr>
<td>Cost of handling (2.50x4)</td>
<td>2.00 &quot;</td>
</tr>
<tr>
<td>Total</td>
<td>15.44 &quot;</td>
</tr>
<tr>
<td>+ 15% loss in storage and transport</td>
<td>2.32 &quot;</td>
</tr>
<tr>
<td></td>
<td>17.76 &quot;</td>
</tr>
</tbody>
</table>

Now, knowing that 1 kg mill run bagasse produces 2.22 kg steam and that the steam consumption of the turbo-alternator will be 10.2 kg steam per kWh, then the fuel cost per kWh will be:

\[
\frac{17.76 \times 10.2}{1000 \times 2.22} = 0.08 \text{ f/kWh}
\]
For a 5000 kW power station we can estimate the fixed annual charges as 25% of the invested capital. The price of the turbo-alternator inclusive boiler, commands, building, is 1000 f/kVA. For a 5000 kW power station, it comes to 5 000 000 f. The fixed annual charges amount to 1 250 000 f, it comprises: maintenance, depreciation, interest on capital, labour, overheads. Then, the fixed charges per kWh are:

\[
\frac{1 250 000}{5000 \times 24 \times 365} = 0.03 \text{ f/kWh}
\]

From the above, it apparent that the central station generating cost would be at least: 0.03 + 0.08 = 0.11 f/kWh. This unfortunately does not compare favorably with a station running diesel engines on heavy fuel oil, and therefore generally speaking we can say that electricity generation using bagasse is more economic when performed at individual factories rather than at a central generating plant, except in special favourable conditions.

Manufacture of briquettes

Especially in developing countries where open fires are still used for cooking at home, briquettes are a good substitute for charcoal and wood. In some particular countries, like Haiti, where the government has to face serious problems due to erosion caused by non-systematic use of fresh wood for the peasants energy needs, the briquette alternative is a welcome one.

The briquetting plant consists of three main sections:

a) A fibre preparation section, where dry depithing is performed and the pith separated generally returns to the factory furnaces for burning

b) A drying section, where the back pressure steam of the turbine is made use of

c) A briquetting section, where the depithed bagasse is compressed by a 1000 kg/cm^2 press into briquettes.

The surplus bagasse is 15 tons/h. The number of tons of briquettes we may produce (since moisture content of bagasse goes from 50% to 20% after drying and since bagasse contents 35% pith) is:

\[
15 \times 0.50 \times 0.65 = 6.09 \text{ tons/h}
\]

\[
\frac{0.80}{0.80}
\]
The commercial presses designed by the Swiss Precision Machinery and Co, have a capacity of 2 tons of briquettes per hour; we then need 3 presses. The capital cost of a plant producing some 30000 kg / day of briquettes was in 1952 about $ 100000, inclusive building, services, land. If we assume a constant inflation of 8% during the elapsed time and a conversion factor of 2.48 for the rate of exchange, the actual value of such a plant will be:

\[ 100000 \times (1.08)^{30} \times 2.48 = 2500000 \, \text{f} \]

The cost of production per ton of briquettes produced is given by the manufacturer as follow:

(i) Wear parts
   a) Fibre preparation section 1.00 f/ton
   b) Drying section 0.60 "
   c) Briquetting section 2.40 "

(ii) Lubricating
   0.40 "

(iii) Labour (0.5 man-hour)
   2.00 "

(iv) Power consumption
   a) Fibre preparation section 60 kWh
   b) Drying section 22 kWh
   c) Briquetting section 30 kWh
   \[ \text{Total kWh} = 112 \, \text{kWh} \]
   \[ \frac{112 \, \text{kWh}}{\text{Ton}} = 7.84 \, \text{f/ton} \]

(v) Heat required for drying (800000 kcal/Ton) (assumed free)

(vi) Depreciation (10% of capital cost)
   \[ \frac{2500000 \times 0.10}{29232} = 8.55 \, " \]

(vii) Interest on capital cost (10% of 40%)
   \[ 0.10 \times 0.40 \times \frac{2500000}{29232} = 3.42 \, " \]

(viii) Cost of bagasse (assumed free)

The total cost of production per ton of briquettes produced is:

26.21 f/ton

It means we can't sell our ton of briquettes for less than 26.21 f/ton. Since there exists any known price on the international market, we can speculate on two prices:
a) a price based on the calorific value of charcoal
b) a price based on the calorific value of wood

The price of charcoal is 200 f/ton and its calorific value 6500 kcal/kg.
That of wood is $585 f/m^3$ and the calorific value 2800 kcal/kg.

The corresponding conversion indices are then 23% and 54%.

The equivalent prices of the briquettes will be:

a) charcoal equivalent: 46 f/ton
b) Wood equivalent: 221 f/ton

Fibreboard

Fibreboard, used as building material, is a broad generic name which encompasses sheet materials of widely varying densities manufactured from refined or partially refined wood fibre or other vegetable fibres. There are two main types of fibreboard:

a) Insulating board
b) Hardboard

There are some fibreboard plants around the world utilizing almost exclusively bagasse which have operated successfully for a number of years. The better known are: The Celotex Corporation (U.S.A), The Colonial Sugar Refining Company (Sydney), The Changhwa Fibreboard (Taiwan), The Canec Wall Board Plant (Hawaii).

Economics of the fibre plant

The capacity of such a plant could be:

hardboard: $\frac{6.09 \times 20}{1.15} = 106$ tons/day (see p. 61)

Insulating board $\frac{6.09 \times 20}{1.25} = 97$ tons/day (see p. 61)

From figure 34 (3, p. 83) we can derive the capital cost:

<table>
<thead>
<tr>
<th>Year</th>
<th>Hardboard</th>
<th>Insulating board</th>
</tr>
</thead>
<tbody>
<tr>
<td>1968</td>
<td>$3,500,000</td>
<td>$4,000,000</td>
</tr>
<tr>
<td>1980 GLD</td>
<td>$18,700,000</td>
<td>$21,400,000</td>
</tr>
</tbody>
</table>


ITEMS QUANTITY AND % COST PER TON OF PRODUCT

<table>
<thead>
<tr>
<th>ITEMS</th>
<th>INSULATING BOARD</th>
<th>HARDBOARD</th>
</tr>
</thead>
<tbody>
<tr>
<td>A) RAW MATERIAL</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Depithed screened bagasse (tons)</td>
<td>1.15</td>
<td>15.68</td>
</tr>
<tr>
<td>Rosin (kg)</td>
<td>10</td>
<td>10.0</td>
</tr>
<tr>
<td>Alum. or ferric sulphate (kg)</td>
<td>30</td>
<td>12</td>
</tr>
<tr>
<td>Miscellaneous: felts, wires, lubricants etc...</td>
<td>--</td>
<td>3.24</td>
</tr>
<tr>
<td>Labour (man-hours)</td>
<td>6</td>
<td>24</td>
</tr>
<tr>
<td>B) SERVICES</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Steam H.P. (tons)</td>
<td>3.9</td>
<td>26.76</td>
</tr>
<tr>
<td>Electricity (kWh)</td>
<td>550</td>
<td>38.5</td>
</tr>
<tr>
<td>Water (m³)</td>
<td>50</td>
<td>2.60</td>
</tr>
<tr>
<td>C) FIXED CHARGES</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Repairs and maintenance</td>
<td>18.30</td>
<td></td>
</tr>
<tr>
<td>(2% of capital cost)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Depreciation</td>
<td>110.30</td>
<td></td>
</tr>
<tr>
<td>(10% of Capital cost)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Overhead and administration</td>
<td>50.47</td>
<td></td>
</tr>
<tr>
<td>(6% of capital cost)</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

From this table we can now calculate the total production cost:

- Hardboard: 283.01 f/ton
- Insulating board: 311.85 f/ton

Price of hardboard on market: \(2.86 \text{ f/m}^2\)

- thickness: 35 mm  
  - Hence: 814.17 f/ton
- density: 1.00

Price of Insulating board: \(3.50 \text{ f/m}^2\)

- thickness: 10 mm  
  - Hence: 1400.01 f/ton
- Density: 0.25
COST EVALUATION

INVESTMENT COST

As far as investment is concerned we have made use of the Lang Method for estimating installed prices of equipment. We have sought first the delivery prices of the various machines. On the one hand, machines prices are almost exclusively coming from the WEBCI 1980 booklet. Some other prices have been given to us by Mr Mouris. On the other hand we approximate prices of machines dating from a few years ago by introducing a conversion factor calculated as follow:

\[
\text{Conversion Factor} = (1 + \text{inflation rate})^{\text{number of years}} \times \text{rate exchange}
\]

inflation rate = 8%
rate of exchange = (\$ 1 = 2.48 Gld) February 1980

Delivery prices of machines

<table>
<thead>
<tr>
<th>Machines</th>
<th>Number</th>
<th>Price</th>
</tr>
</thead>
<tbody>
<tr>
<td>Belt Conveyor</td>
<td>M2</td>
<td>107 500 Gld</td>
</tr>
<tr>
<td>BAGASSE elevator</td>
<td>M34</td>
<td>75 000 &quot;</td>
</tr>
<tr>
<td>Knives</td>
<td>M3 + M5</td>
<td>M11 + M24 + M27 21 000 000 &quot;</td>
</tr>
<tr>
<td>Shredder</td>
<td>M7</td>
<td>2 000 000 &quot;</td>
</tr>
<tr>
<td>3 Mills (installed)</td>
<td>M14</td>
<td>3 000 000 &quot;</td>
</tr>
<tr>
<td>Diffuser</td>
<td>V9 + V15</td>
<td>52 000 &quot;</td>
</tr>
<tr>
<td>Hoppers (20 m³)</td>
<td>V43</td>
<td>48 000 &quot;</td>
</tr>
<tr>
<td>Tank (20 m³)</td>
<td>P36</td>
<td>25 000 &quot;</td>
</tr>
<tr>
<td>Compressor</td>
<td>P19 + P33 + P40</td>
<td>19 500 &quot;</td>
</tr>
<tr>
<td>3 Pumps (10 kW, 150 m)</td>
<td>P44 + P42</td>
<td>26 000 &quot;</td>
</tr>
<tr>
<td>2 fans (30 kW)</td>
<td>P21 + P22</td>
<td>35 000 &quot;</td>
</tr>
<tr>
<td>BAGASSE MANUFACTURE</td>
<td></td>
<td>2 500 000 &quot;</td>
</tr>
<tr>
<td></td>
<td></td>
<td>29 888 000 &quot;</td>
</tr>
</tbody>
</table>
This gives a total investment on delivery of 29,880,000 gld. All those prices are exclusive A.V.T.(B.T.W.). If we include those 18%, The delivery price will be: 35,267,840 Gld. Starting from this value we make an estimate of the real investment in the following manner: The price corresponding to the total investment is taken into account by multiplying the delivery price of the various type of equipment by a factor F which is found as follow:

Price of foundations, installation \( f_1 = 1.43 \)

Price of piping

<table>
<thead>
<tr>
<th>Type</th>
<th>( f_1 )</th>
</tr>
</thead>
<tbody>
<tr>
<td>Solid-liquid</td>
<td>1.25</td>
</tr>
<tr>
<td>Gas-liquid</td>
<td>1.60</td>
</tr>
<tr>
<td>Solids only</td>
<td>1.10</td>
</tr>
</tbody>
</table>

Price of electrical installation, utilities, \( f_1 = 1.50 \)

Indirect costs such as construction, overhead, engineering \( f_1 = 1.38 \)

The total Lang-Factor is now the product of the more specific \( f_1 \) factors.

<table>
<thead>
<tr>
<th>Factor</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>F (solids only)</td>
<td>3.10</td>
</tr>
<tr>
<td>F (liquid-solid)</td>
<td>3.63</td>
</tr>
<tr>
<td>F (liquid-gas)</td>
<td>4.74</td>
</tr>
</tbody>
</table>

From there the total investment cost can be calculated (inclusive turbine and boiler, furnace etc...) (see later)

Total investment: 68,368,020 Gld

Process dependant costs

Steam and electricity expenses.

The energy required for the entire process is supplied by the factory itself. Hence, steam and electricity expenses will be accounted for in the investment part of the costs concerned with energy production. That is:

- Turbine + alt. (2250 kW), 2,000,000 gld
- Furnace + boiler + Heat exch. + Chimney, 3,600,000 "

These are naturally delivery prices.
Water expenses

The total water consumption of the plant is 41 tons per hour. That is yearly equal to 196,800 tons. Since the price of water is approximately 0.05 f/m³, the total expenses for water will be: 9840 f.

Raw material expenses

Sugar cane prices fluctuate very strongly. According to Mr. Mouris the price nowadays could be taken as 24f/ton in South and Middle America. The yearly consumption of our plant is: 100 x 20 x 240 = 480,000 tons. The total cost for cane is then: 11,520,000 f/year.

Labour cost

The labour cost is calculated by using the Wessel relation. For continu processes it works as follow:

\[
\frac{\text{man-hours}}{\text{tons product}} = \frac{\text{number of sections} \times 10}{(\text{capacity per day})^{0.76}}
\]

The number of sections is taken to be 3. The capacity per day is: 116 x 20 = 2320 tons/day. This gives 192.68 man-hours/day. This number divided by 8 gives 24 men/day in shift. To this number we add 15 persons as high qualified staff. The mean annual salary is 60,000 f. Hence the total labour expenses per year is: 2,340,000 f.

ECONOMIC EVALUATION

The total investment cost amounts to 68,368,020 f. It is agreed that the plant will depreciate in 10 years. The depreciation is linear. It means that the yearly depreciation amounts to 6,836,802 f. The rate of interest is equal to 10%. Interest has to be paid on 60% of the invested capital. The annual expenses as a result of interest, depreciation and maintenance (5% of capital cost) are: 14,357,284 f.

PRODUCTION COST

The total production cost can be calculated from there:
depreciation 6 836 802 gld
interest 4 102 081 "
maintenance 3 418 401 "
water expenses 9 840 "
raw material 11 520 000 "
labour costs 2 340 000 "

Total 28 227 124 "

The production cost of raw juice per ton is:

\[
\frac{28 227 124}{116 \times 20 \times 240} = 50.70 \text{ gld/ton} \approx 400 \text{ kg/ton}
\]

The production cost of briquettes, if subtracted from that of raw juice will lower even more the total cost of production:

\[
28 227 124 - 766 170 = 27 460 953 \text{ f}
\]

The resulting price of production for raw juice is:

\[
\frac{27 460 953}{116 \times 20 \times 240} = 49.32 \text{ f/ton.}
\]
CONCLUSION

Although most sugar cane factories still use milling as extraction mode for raw juice, an economical study has showed that diffusion is cheaper and less energy consumer.

Among the various types of diffusion equipments available, a production cost study has given the D.d.S. as the cheapest extraction process.

The D.d.S. diffusion use three mills. A balance of advantages and disadvantages of electric or steam drive, has given electric drive as the best paying proposition.

General calculation of calorific value of bagasse and detailed calculation of energy requirements of the various types of equipments have showed that only 40% of the total amount of bagasse have to be burnt in order to supply the entire energy necessary for the plant, while the resting 60% could be used in a by-product manufacture.

Although the production of fiberboard from bagasse seems to be a profitable proposal, we have nevertheless concentrated on the production of briquettes because it is actually realized in developing countries, and also because briquettes can be considered as the necessary raw material for making fiberboard, paper pulp, and all other products based on bagasse.

Since there is no assigned price for bagasse briquettes the sale price of briquettes has been taken equal to the production cost. This gives us a production cost of raw juice from sugar cane of 49.32 Gld per ton.
### Massa-en Warmtebalans

**Sugar Cane**

<table>
<thead>
<tr>
<th>IN</th>
<th>Voorwaarts</th>
<th>Retour</th>
<th>UIT</th>
</tr>
</thead>
<tbody>
<tr>
<td>M</td>
<td>Q</td>
<td></td>
<td></td>
</tr>
<tr>
<td>27.78</td>
<td>29.03</td>
<td></td>
<td></td>
</tr>
<tr>
<td>11.39</td>
<td>11.90</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Diagram**:
- **1**: Flow to M11
- **2**: Flow to M24
- **3**: Flow to M14
- **4**: 90°C Water to M14
- **5**: Flow to V15
- **6**: Flow from M14
- **7**: Flow from M2A
- **8**: Flow from M2B
- **9**: Flow from M25
- **10**: Flow from V15

**Table**:

<table>
<thead>
<tr>
<th>M</th>
<th>Q</th>
</tr>
</thead>
<tbody>
<tr>
<td>40.53</td>
<td>11.00</td>
</tr>
<tr>
<td>0.6944</td>
<td>18.14</td>
</tr>
<tr>
<td>20.83</td>
<td>43.53</td>
</tr>
<tr>
<td>20.33</td>
<td>43.53</td>
</tr>
<tr>
<td>6.95</td>
<td>44.32</td>
</tr>
<tr>
<td>4.92</td>
<td>10.23</td>
</tr>
<tr>
<td>3.75</td>
<td>7.83</td>
</tr>
<tr>
<td>19.46</td>
<td>30.63</td>
</tr>
<tr>
<td>15.74</td>
<td>16.23</td>
</tr>
<tr>
<td>1.72</td>
<td>3.59</td>
</tr>
<tr>
<td>14.97</td>
<td>23.90</td>
</tr>
<tr>
<td>17.25</td>
<td>18.03</td>
</tr>
<tr>
<td>1466</td>
<td></td>
</tr>
</tbody>
</table>
Massa in kg/s
Warmte in kW

Fabrieksvoorontwerp
No.
PART TWO:

THE BEET SUGAR EXTRACTING UNIT
I Characteristics of beet, wet pulp, dry pulp.
(Physico-chemical composition, amount of material treated, calorific value)

Composition of beet
We will start with the following standard values:
-Sugar content: 15%
-Dry material (cellulose) 5%
-Water 80%

Element analysis of dry pulp.
As we don't know the element analysis of beet pulp, we will use the same values than for cane fiber, because these two materials are mostly cellulose.

<table>
<thead>
<tr>
<th>Element</th>
<th>Percentage</th>
</tr>
</thead>
<tbody>
<tr>
<td>C</td>
<td>47%</td>
</tr>
<tr>
<td>H</td>
<td>6.5%</td>
</tr>
<tr>
<td>O</td>
<td>44%</td>
</tr>
<tr>
<td>Ashes</td>
<td>2.5%</td>
</tr>
</tbody>
</table>

100%

Composition of wet pulp
After being pressed, the wet pulp has the following mean composition:
Dry material (cellulose) 22%
Water 78%

Composition of dried pulp
The beet pulp is dried and sold as cattle food.
The dried pulp has the following composition:
Dry material (mostly cellulose) 90%
Water 10%

A Mass balance for the hydrogen gives us the possibility to calculate the percentage of Hydrogen in the dried pulp and in the wet pulp.

Hydrogen percentage in dry pulp 6.5%
Hydrogen percentage in dried pulp 5.85%
Hydrogen percentage in wet pulp 1.43%

These percentages are calculated on a wet basis. The hydrogen of water is not included.
Calorific values

It is possible to calculate two different calorific values: The gross calorific value (GVC) and the Nett calorific value (NVC) (cf. p. 869). For the dried pulp, NVC = 18000 kj/kg.

So, calorific value of dry pulp:

\[
\text{GVC} = 5202 \text{ kcal/kg} \\
\text{NVC} = 5202 - 5400 \times 0.065 = 4851 \text{ kcal/kg}
\]

Calorific value of dried pulp (10% moisture):

\[
\text{GVC} = 5202 \times 0.9 = 4682 \text{ kcal/kg} \\
\text{NVC} = 4682 - 5400 \times 0.0585 - 600 \times 0.10 = 4306 \text{ kcal/kg} (= 18000 \text{ kj/kg})
\]

Calorific value of wet pulp (78% moisture):

\[
\text{GVC} = 5202 \times 0.22 = 1144 \text{ kcal/kg} \\
\text{NVC} = 1144 - 5400 \times 0.0143 - 600 \times 0.78 = 599 \text{ kcal/kg}
\]
DIFFUSER . HEAT AND MASS BALANCES

The calculations of diffuser, presses and drier are based upon the results of the " Raffinerie Notre-Dame à Oreye " (Reference 5 ).

Mass balance

We will refer to a quantity of beet of 1.

Quantity of sugar lost in pulp : 2.6 % dry substance (5, p, 6)
Hence, : 0.026 x 0.05 = 0.0013
So, quantity of sugar obtained in raw juice :
0.15 - 0.0013 = 0.1487
We want a concentration of sugar in raw juice of 12.5 % by weight.
So, quantity of raw juice :
\[ \frac{0.1487}{Q} = 0.125 \]
Hence, \( Q = 1.190 \)

Moisture content of wet pulp : 72 % (5, p. 2)
So, quantity of wet pulp at diffuser output :
0.05 = M x 0.72
Hence, \( M = 0.6944 \)

Moisture content of pressed pulp at presses output : 22 %
So, quantity of pressed pulp at presses output :
0.05 = M' x 0.22
Hence, \( M' = 0.2273 \)

So, quantity of juice obtained at presses :
0.6944 - 0.2273 = 0.4671

The mass balance for diffuser is :
Beet + Fresh water + Juice from presses =
Wet pulp + Juice at diffuser output
Hence :
1 + 0.4671 + X = 0.6944 + 1.190
Hence, \( X = 0.4173 \)

So, the amount of fresh water to supply to the diffuser for a capacity of 100 t beet per h, will be :
0.4173 x 100 = 41.73 t/h
(5, p. 3 ) reports : 40 to 41 t/h
Heat balance

We will assume a specific heat of 1 kcal/kg °C for water, juice and pulp. The juice obtained at presses, will be heated in a heat exchanger before returning to the diffuser.

The quantity of steam used for the diffuser is : 2200 kg/h
The characteristics of the steam are :
\[ p = 3 \text{ kg/cm}^2 ; T = 143 °C ; \Lambda = 625 \text{ kcal/kg} \]
Hence, the heat supplied by steam will be :
\[ M = 2200 \left( 625 - 90 \right) = 1.177 \times 10^6 \text{ kcal/h} \]

The temperature of presses juice at heat exchanger output is given by :
\[ M \alpha = p \left( T - T_o \right) \]
\[ p = 46710 \text{ kg/h} \]
\[ \alpha = 0.90 \text{ (rendement of heat exchange)} \]
\[ t_o = 45 °C \text{ (juice temperature at presses output)} \]
Hence, \[ T = 66.5 °C \]

Temperature of beet at diffuser input : 25 °C
Temperature of pulp at diffuser output : 50 °C
Temperature of fresh water : 25 °C
Hence, the raw juice temperature at diffuser output will be calculated from a heat balance for the diffuser :
\[ 25 \times 1 + 0.4671 \times 66.5 + 0.4173 \times 25 = 50 \times 0.6944 + 1.190 T_j \]
Hence, \[ T_j = 27 °C \]
II Calorific value of fuel-oil

We will assume a fuel-oil n° 2 (table 9.3)
Hence \( CCV = 140000 \text{ Btu:gal} = 10830 \text{ kcal:kg} \)

Element analysis % (table 9.11)

<table>
<thead>
<tr>
<th>Element</th>
<th>%</th>
</tr>
</thead>
<tbody>
<tr>
<td>C</td>
<td>87.3</td>
</tr>
<tr>
<td>H</td>
<td>12.6</td>
</tr>
<tr>
<td>O</td>
<td>0.04</td>
</tr>
<tr>
<td>N</td>
<td>0.006</td>
</tr>
<tr>
<td>S</td>
<td>0.22</td>
</tr>
<tr>
<td>Ash</td>
<td>0.01</td>
</tr>
</tbody>
</table>

\[ NVC = GVC - 5400 \cdot H \]
\[ = 10830 - 5400 \cdot 0.126 = 10150 \text{ kcal/kg} \]

Combustion of Fuel-oil

Composition of air (table 42-9)
Oxygen 23.15%
Nitrogen 76.85%

Reactions of combustion
\[ C + O_2 \rightarrow CO_2 \]
By weight: 12 + 32 \rightarrow 44
Or \[ : 1 + 2.67 \rightarrow 3.67 \]
\[ H_2 + O_2 \rightarrow H_2O \]
\[ 1 + 8 \rightarrow 9 \]
\[ S + O_2 \rightarrow SO_2 \]
\[ 1 + 1 \rightarrow 2 \]
Properties of gaseous products of combustion

<table>
<thead>
<tr>
<th></th>
<th>Mol. wt.</th>
<th>Density (kg/m³) at 0°C and 760 mm Hg</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO₂</td>
<td>44</td>
<td>1.977</td>
</tr>
<tr>
<td>H₂O</td>
<td>18</td>
<td>0.804</td>
</tr>
<tr>
<td>O₂</td>
<td>32</td>
<td>1.429</td>
</tr>
<tr>
<td>N₂</td>
<td>28</td>
<td>1.256</td>
</tr>
<tr>
<td>SO₂</td>
<td>64</td>
<td>1.293</td>
</tr>
<tr>
<td>air</td>
<td>29</td>
<td>2.857</td>
</tr>
</tbody>
</table>

A. Combustion of fuel-oil without excess air.
   a. Oxygen. In order to burn 1 kg of fuel-oil, we require:

   \[
   \begin{align*}
   &C: 0.873 \times 2.67 = 2.33 \\
   &H: 0.126 \times 8 = 1.01 \\
   &S: 0.022 \times 1 = 0.022
   \end{align*}
   \]

   But the fuel-oil already contains O = 0.04, hence the air must supply 3.30.

   b. Nitrogen. This oxygen brings with it:

   \[
   \frac{76.85}{3.30} = 10.76 \text{ kg of } N₂
   \]

   Hence, the total weight of air required is 14.26 kg.

B. Combustion of fuel-oil with excess air
   Weight of air required
   \[
   P_a = 14.26 \times 1.3 = 18.54 \text{ kg}
   \]
   Weight of gaseous products of combustion:
   \[
   P_g = 14.26 \times 1 + 1 = 19.54 \text{ kg}
   \]
   \[
   \begin{align*}
   P_{gs} &= \text{weight of dry air. We shall obtain this quantity by deducing from } \ P_g \\
   &\text{the water formed by combustion of the hydrogen of the fuel-oil:}
   \\
   &P_{gs} = 19.54 - 0.126 \times 9 = 18.40 \text{ kg}
   \end{align*}
   \]

Composition of flue-gases
   We know the total weight of gases: \( P_g = 19.54 \text{ kg} \)
   The weight of individual gases is given by:
   a. Nitrogen
      \[
      N₂ = 10.76 \times 1.3 = 13.99
      \]
      72.4
b. Oxygen

\[ O_2 \text{ derived from the air} = 3.30 \times 1.3 = 4.29 \]
\[ + O_2 \text{ derived from the fuel-oil} = 0.04 \]
\[ - O_2 \text{ used to form water} = 1.01 \]
\[ - O_2 \text{ used to form CO}_2 = 2.33 \]
\[ - O_2 \text{ used to form SO}_2 = 0.02 \]

or

\[ 0.97 \quad 5\% \]

c. Water

Water of constitution

\[ 1.13 \quad 5.8\% \]

d. Carbon dioxide

\[ 3.20 \quad 16.6\% \]

e. Sulfur dioxide

\[ 0.044 \]

Total

\[ 19.33 \]

The slight difference between the total and \( P \), is due to ash content, as the sum of components \( C + H + O + S \) of fuel-oil which we have assumed does not amount to \( 1 \) kg.

Calculation of combustion temperature.

The combustion temperature \( T \) prevailing in the fuel-oil furnace, is readily calculated from the fact that the heat developed in the combustion is received in the gases passing from the furnace to the boiler.

Taking into account losses:

\[
t = t_0 + \frac{\alpha \beta_0 N_i}{\left( \sum P_i c_i T \right) t_0} \quad (1, \text{ eqn. 42.43})
\]

Mean specific heat.

Mean specific heat of combustion gases between \( 0^\circ C \) and \( 1700^\circ C \)

\[
N_2 : \text{CSM} = 0.246 + 0.000020 \times T
\]
\[
O_2 : \text{CSM} = 0.214 + 0.000018 \times T
\]
\[
H_2O : \text{CSM} = 0.468 + 0.000156 \times T
\]
\[
CO_2 : \text{CSM} = 0.199 + 0.000082 \times T
\]
CSM = \Sigma X_i CSM_i

Hence,
CSM = 0.249 + 0.000038 \times T

Combustion temperature:
\[ t_0 = 87 \, ^\circ C \]
\[ \alpha = 0.98 \]
\[ \beta = 0.975 \]
\[ N_1 = 10150 \, \text{kcal/kg} \]
\[ p_g = 19.33 \, \text{kg} \]

Hence,
\[ t = 87 + 0.98 \times 0.975 \times 10150 \times \frac{1}{19.33 \times 0.317} = 1669 \, ^\circ C. \]

Quantity of steam obtainable

We may now calculate the quantity of steam which we can obtain from 1 kg fuel oil.

\[ M_v = (NCV - q) \alpha \beta \eta \]

\[ q = \text{sensible heat of combustion gases leaving the economiser;} \]
\[ \alpha = 0.98 \]
\[ \beta = 0.975 \]
\[ \eta = 0.96 \]

Calculation of \( q \)

\[ q = P_g \times c_{p,g} \times T_g \]
\[ P_g = 19.33 \, \text{kg} \]
\[ T_g = 180 \, ^\circ C \]
\[ c_{p,g} = 0.249 + 0.000038 \times T_g \]

Hence, \[ c_{p,g} = 0.256 \, \text{kcal/kg} \times ^\circ C \]

So, \[ q = 890 \, \text{kcal/kg} \]

Hence:
\[ M_v = 8848 \, \text{kcal/kg} \]
The enthalpy of high pressure steam at 400 °C and 40 kg/cm² is:
767 kcal/kg (1, table 42.3)
The enthalpy of feed water at 90 °C to the economiser is 90 kcal/kg
Hence, the energy necessary to produce 1 kg high pressure steam will be
767 - 90 = 677 kcal/kg
So, the amount of high pressure steam produced from one kg fuel oil:
\[ \frac{M_v}{677} = 13.23 \text{ kg} \]
STEAM TURBINE

The details about the calculations of the steam turbine can be found in the chapter "Steam turbine" in the first part: "Cane distillation plant".

Data
Pressure at entry: \( p_a = 38 \text{ kg/cm}^2 \) eff.
Temperature of superheated steam at entry: \( t_a = 390 \degree C \)
Back pressure at exhaust: \( p_b = 3 \text{ kg/cm}^2 \) eff.
Power required at the alternator terminals: 846 kW

Calculations
The Mollier diagram gives:
\( \Lambda_a = 760 \text{ kcal/kg} \)
\( \Lambda_b = 625 \text{ kcal/kg} \)
\( \Lambda_a - \Lambda_b = 135 \text{ kcal/kg} \)
Consequently, the scale of pressures is:

<table>
<thead>
<tr>
<th>Total heat</th>
<th>temperature</th>
<th>abs. pres.</th>
<th>eff. pres.</th>
</tr>
</thead>
<tbody>
<tr>
<td>(kcal/kg)</td>
<td>t (\degree C)</td>
<td>p (kg/cm(^2))</td>
<td>p' (kg/cm(^2))</td>
</tr>
<tr>
<td>Boiler</td>
<td>767</td>
<td>400</td>
<td>40</td>
</tr>
<tr>
<td>Admission</td>
<td>760</td>
<td>390</td>
<td>39</td>
</tr>
<tr>
<td>1st wheel</td>
<td>725.2</td>
<td>315</td>
<td>29</td>
</tr>
<tr>
<td>2nd wheel</td>
<td>691.8</td>
<td>240</td>
<td>16</td>
</tr>
<tr>
<td>3rd wheel</td>
<td>658.4</td>
<td>170</td>
<td>8</td>
</tr>
<tr>
<td>4th wheel</td>
<td>625</td>
<td>143</td>
<td>4</td>
</tr>
</tbody>
</table>

Steam consumption

\[
Q = \frac{860}{(\Lambda_a - \Lambda_b)\eta f_t f_s} \quad (1, \text{eqn. 44-42})
\]
\( \eta = 0.72 \quad (1, \text{table 44-1}) \)
\( f_t = 0.98 \quad (1, \text{p. 919}) \)
\( f_s = 0.93 \quad (1, \text{p. 919}) \)

Hence \( Q = 9.71 \text{ kg/kWh} \)
Losses by condensation 3 
Losses by leaks 2 
Hence Q = 9.71 \times 1.05 = 10.2 \text{ kg/kWh} 
So the quantity of steam required to produce 846 kW is: 
846 \times 10.2 = 8630 \text{ kg/h}.
CALCULATION OF THE SUPERHEATER

We need 21606 kg/h of high pressure steam, so we are going to calculate a steam boiler with a capacity of 22000 kg/h to maintain a safety margin.

So, weight of fuel oil to be burnt:

\[ B = \frac{22000}{13.23} = 1918 \text{ kg/h (Cf. 'Calorific value of Fuel oil')} \]

Weight of combustion gases

We have calculated \( P = 19.33 \text{ kg/kg fuel oil} \)

So, in total:

\[ P = 19.33 \times 1918 = 37083 \text{ kg/h} \]

Temperature of gases leaving the superheater:

\[ T_2 = T_1 - \frac{P}{a \cdot c} \left( (1 - X) + c (T - t) \right) \quad (1, \text{ eqn. 42.60}) \]

\[ T_1 = 1669 \degree C \]
\[ p = 22000 \text{ kg/h} \]
\[ a = 0.90 \quad (1, \text{ p. 844}) \]
\[ P = 37083 \text{ kg/h} \]
\[ X = 0.98 \quad (1, \text{ p. 845}) \]
\[ r = 411 \text{ kcal/kg} \quad (1, \text{ table 42.1}) \]
\[ t = 249 \degree C \quad (1, \text{ table 42.1}) \]
\[ T = 400 \degree C \quad \text{(superheat required)} \]
\[ C = 0.304 \text{ KCal/kg \degree C (between} T_1 \text{ and} T_2 \text{)} \]

Hence,

\[ T_2 = 1669 - \frac{22000}{0.90 \times 37083 \times 0.304} \left( 0.02 \times 411 + 0.650 \times (400 - 249) \right) \]

\[ = 1439 \degree C \]

The heating surface of the superheater installed is thus:

\[ S = \frac{p \left( 1 - X \right) + p c \left( T - t \right)}{k \left( \frac{T_1 + T_2}{2} - \frac{T + t}{2} \right)} \]

\[ k = 55 \quad (1, \text{ p. 844}) \]

Hence, \( S = 70 \text{ m}^2 \).
BOILER

Quantity of heat transmitted to steam or water:

\[ M = \alpha \cdot P \cdot C \cdot (T_o - T_1) = P \cdot (X \cdot r + c \cdot (T - t_o)) \]

\[ \alpha = 0.90 \quad (1, \text{ p. 844}) \]
\[ P = 37083 \, \text{kg/h} \]
\[ C = 0.316 \, \text{kcal/kg \textdegree C} \quad \text{(between T_o and T_1)} \]
\[ T_o = 1439 \, \text{\textdegree C} \]
\[ X = 0.98 \quad (\text{dryness fraction of steam at boiler output, 1, p. 845}) \]
\[ r = 411 \, \text{kcal/kg} \quad (1, \text{ table 42.1}) \]
\[ c = 1 \, \text{kcal/kg \textdegree C} \]
\[ T = 249 \, \text{\textdegree C} \quad (1, \text{table 42.1}) \]
\[ t_o = 127 \, \text{\textdegree C} \quad \text{(temperature of water at economiser output)} \]

Hence, \( T_1 = 350 \, \text{\textdegree C} \).
Calculation of the economiser

The details about theses calculations are given in the first part "Raw Juice extraction from sugar cane".

\[ r = \frac{\alpha \cdot P_c}{C} = \frac{t - t_0}{T_0 - T_1} \quad (1, \text{eq. 42.69}) \]

\[ \alpha = 0.95 \]
\[ P = 37083 \, \text{kg/h} \]
\[ C = 0.204 \, \text{kcal/kg/°C} \]
\[ p = 22000 \, \text{kg/h} \]
\[ c = 1 \, \text{kcal/kg/°C} \]

Hence,

\[ r = \frac{0.95 \times 37083 \times 0.204}{22000} = 0.3207 \]

Since,

\[ t = t_0 + r(T_0 - T_1) \]

and

\[ t_0 = 90^\circ\text{C} \quad (\text{feed water temperature}) \]
\[ T_0 = 350^\circ\text{C} \]
\[ T_1 = 237^\circ\text{C} \]

It follows that:

\[ t = 90 + 0.3267(350 - 237) = 127^\circ\text{C} \]

Heat exchanging surface of the economiser:

\[ S = \frac{\alpha \cdot P_c}{k(1 - a)} \ln \frac{T_0 - t}{T_1 - T_0} \quad (1, \text{eq. 42.70}) \]

\[ k = 25 \quad (1, \text{p. 850}) \]

Hence,

\[ S = 179 \, \text{m}^2 \]

Calculation of the air heater

\[ r' = \frac{\alpha' \cdot P_c}{p' \cdot c'} = \frac{t' - t_0'}{T_1 - T} \quad (1, \text{eqn. 42.74}) \]

\[ p_a = 18.33 \, \text{kg air/ kg fuel oil burnt (cf. "Calorific value of fuel oil")} \]

So, in total \( p' = 18.33 \times 1918 = 35157 \, \text{kg/h} \)
\[ \alpha' = 0.935 \ (1, \ p. \ 855) \]
\[ c' = 0.24 \text{ kcal/kg}^\circ \text{C} \ (1, \ p. \ 857) \]
\[ P = 37087 \text{ kg/h} \]
\[ C = \text{ MCV of combustion gases between 180 and 237}^\circ \text{C} \]
\[ C = 0.3098 \text{ kcal/kg}^\circ \text{C} \]
\[ \text{So, } r' = \frac{0.935 \times 37087 \times 0.245}{35157 \times 0.24} = 1.0885 \]

Hence:
\[ t = t' + r \cdot (T_1 - T) \]
\[ = 25 + 1.0885 \cdot (237 - 180) \]
\[ = 87 \text{ }^\circ \text{C} \]

So, heat exchange surface of the air heater:
\[ S' = \frac{\alpha' \cdot PC}{k' \cdot (1 - r')} \cdot \frac{\ln \frac{T}{T'} - \ln \frac{T}{T''}}{T - t_0} \quad (1, \ \text{eqn. 42.75}) \]
\[ k' = 18 \ (1, \ p. \ 855) \]
\[ \text{So, } S' = \frac{0.935 \times 37087 \times 0.265}{18(1 - 1.0885)} \cdot \frac{\ln 237 - 87}{180 - 25} \]
\[ S' = 189 \text{ m}^2 \]

**DRAUGHT. POWER REQUIREMENTS OF FANS**

\[ P_{f1} = \frac{V_a \cdot d_1}{75 \cdot \phi} \quad (1, \ \text{eqn. 42.89}) \]
\[ d_1 = 95 \text{ mm W.G.} \ (1, \ p. \ 865) \]
\[ \phi = 0.5 \ (1, \ p. \ 866) \]
\[ V_a = \frac{P_a}{d_a} \]
\[ d_a = \text{air density} = 1.29 \text{ kg/m}^3 \]
\[ p_a = 27715 \text{ kg/h} \ (\text{cf., calculation of air heater}) \]
\[ V_a = \frac{27715}{1.29} = 21489 \text{ m}^3/\text{h} = 5.97 \text{ m}^3/\text{s} \]

Hence
\[ P_{f1} = \frac{5.97 \times 95}{75 \times 0.5} = 15.1 \text{ h.p.} = 11.1 \text{ kW} \]

\[ P_{f2} = \frac{V \cdot d_2}{75 \cdot \phi} \]
\[ V, \text{ the volume of flue gases, is almost equal to the air volume } V_a: \]
\[ d_2 = 66 \text{ mm W.G.} \ (1, \ p. \ 866) \]
\[ \text{So, } P_{f2} = \frac{5.97 \times 66}{75 \times 0.5} = 10.5 \text{ h.p.} = 7.7 \text{ kW} \]
\[ \text{So, in total } P_{f1} + P_{f2} = 18.8 \text{ kW} \]
POWER REQUIREMENTS OF THE FEED WATER BOILER PUMP

\[ P = \frac{Q \cdot \Delta H}{75 \cdot \rho_m} \quad (1, \text{ eqn. 46.9}) \]

\[ \rho_m = 0.65 \quad (1, \text{ p. 870}) \]

\[ H = 480 \text{ m} \quad (\text{cf. Part I}) \]

So,

\[ P = \frac{20000 \cdot 480}{75 \cdot 0.65 \cdot 3600} = 54.7 \text{ h.p.} = 40.2 \text{ kW} \]

POWER REQUIREMENTS OF THE PLANT

We can now calculate the total power required by the plant:

<table>
<thead>
<tr>
<th>Equipment</th>
<th>Power (h.p.)</th>
<th>Power (kW)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Carriers</td>
<td>20</td>
<td>14.7</td>
</tr>
<tr>
<td>Slicers</td>
<td>130</td>
<td>101.4</td>
</tr>
<tr>
<td>Diffuser</td>
<td>200</td>
<td>147</td>
</tr>
<tr>
<td>Draught fans for boiler</td>
<td>25.6</td>
<td>40.2</td>
</tr>
<tr>
<td>Feed water boiler pump</td>
<td>54.7</td>
<td>40.2</td>
</tr>
<tr>
<td>Three presses</td>
<td>40</td>
<td>94</td>
</tr>
<tr>
<td>Drum drier</td>
<td>40</td>
<td>29.4</td>
</tr>
<tr>
<td>Drier fans</td>
<td>40</td>
<td>121.2</td>
</tr>
<tr>
<td>Alcohol manufacture</td>
<td>200</td>
<td>280</td>
</tr>
</tbody>
</table>

Total                                    |              | 846        |
STEAM REDUCING VALVE AND DE-SUPERHEATER

With the same procedure than in part I, we can now calculate the quantity of high pressure steam and liquid water to add to low pressure steam:

The mass balance and heat balance are:

\[
\begin{align*}
19400 & = 6424 + x + y \\
19400 \times 672 & = 6424 \times 625 \times 760 + y \times 90 \\
\end{align*}
\]

\[x = \text{quantity of high pressure steam in kg/h}\]
\[y = \text{quantity of liquid water in kg/h}\]

So,

\[
\begin{align*}
12976 &= x + y \\
9021800 &= 760 \times x + 90 \times y \\
\end{align*}
\]

Hence:

\[
x = \frac{9021800 - 90 \times 12776}{760 - 90} \\
x = 11749 \text{ kg/h}
\]

So, \[y = 1227 \text{ kg/h}\]

So, the 19400 kg/h of steam at 180 °C and 3 kg/cm², will be provided by 11749 kg/h of steam at 390 °C and 38 kg/cm² plus 1227 kg/h of liquid water at 90 °C and 6424 kg/h of steam at 143 °C and 3 kg/cm².
PRESSING AND DRYING

Nowadays, most of the beet pulp produced, is dried before being used as cattle feed. The pulp is processed in a pulp drier directly linked with the diffusion unit. Thus the value of the pulp as cattle feed is maintained even in case of storage. The pulp being discharged from the diffusion unit has to be dewatered before entering the drier in pulp screw presses. Efficient pressing of wet materials prior to drying results in reduced costs and improved economy. The problem we will study in this chapter, is the optimisation -on a basis of production cost- of pressing efficiency. We will compare the cost of production of dried pulp by mean of a high pressing efficiency (22 % in dry substance), and a lower pressing efficiency (18 % dry substance).

From (g, p. 10), we find that an efficiency of 22 % can be obtained with three BS 64 S presses for a capacity of 100 t beet/h. On the other hand, an efficiency of 18 % in dry substance will be obtained with two BS 64 B presses for the same capacity. So, we have to compare a system including three presses and a drier with a system including two presses and a drier.
1 conveyor
2 feeding scroll
3 mixing scroll and molasses tank
4 drum with vapour outlet
5 drum drive and support rollers
6 discharge housing
7 discharge scroll
8 vapour fan
9 dust separation
10 cell sluice
11 elevator
12 hot-gas furnace
We will describe drum drier manufactured by the firm A B R equipped with a furnace from BALCKE and BOCHUM.

The furnace is a cylinder 2.80 m inside diameter and 7.5 m long, divided into two parts, the furnace itself and the dilution chamber.

The furnace has a capacity of 2000 kg fuel oil per hour and can evaporate 22500 kg/h water.

The combustion air is provided by a fan with a capacity of 30000 m$^3$/h and a power consumption of 40 h.p.

In the dilution chamber, the combustion gases (1500-1600 °C) are mixed with fresh air, so the gas temperature at the drier inlet is compatible with the pulp to be dried.

The drum drier is a cylinder 3 m diameter and 16.75 m long.

It is moved by an electric motor with power requirement of 40 h.p.

The mixing of combustion gases and vapors is moved by a fan with a power requirement of 125 h.p. and a flow rate of 130000 m$^3$/h, and sent to a cyclone 20 m high and 4.90 m in diameter in order to collect the dust.

Rendement of the drier.

The fuel consumption is 0.076 kg fuel oil/kg evaporated water, with a net calorific value of 9650 kcal/kg fuel oil.

So, the thermal consumption is 730 kcal/kg evaporated water, and the rendement will be:

$$\frac{600}{730} = 82\%$$

The capacity of the drying unit 25 t/h pulp 22 % dry substance or:

$$25 \times \frac{22}{5} = 110 \text{ t/h beet}.$$

The pulp produced is 90 % in dry substance.

POWER REQUIREMENTS FOR PRESSING

The power consumption of a press handling maximum of 3400 t/day is appr. 130 kW (p. 9).

So, for a press handling 800 t/day the power consumption will be appr.

$$130 \times \frac{800}{3400} = 31 \text{ kW}$$
Capacity of Presses

From fig. 6 (8, p. 10) we see that a feed of 1200 t. beet/24 h will give a pressing efficiency of appr. 18%. On the contrary a feed of only 800 t. beet/24 h, will give us a pressing efficiency of appr. 22%. Hence, a high pressing efficiency requires three presses working continuously, and a low pressing efficiency of 18% can be obtained with two presses only.

![Graph showing daily slicing capacity for one BS 64 S press](image-url)
CALCULATIONS OF THE DRIER

In this chapter we will calculate the quantity of air to supply to the drier, the temperature of gases at furnace output, at dilution chamber output, the heat and mass balances for the drier.

Furnace

Mass balance:
\[ M_g = M_a + M_f \]  

Heat balance:
\[ M_g c_p,g T_g = M_a c_p,a T_a + M_f c_p,f T_f + M_f h_{comb} \]  

\[ M_g', c_p,g T_g' : \text{Mass, specific heat and temperature of combustion gases} \]
\[ M_a, c_p,a T_a : \quad \text{fresh air} \]
\[ M_f, c_p,f T_f : \quad \text{fuel oil} \]
\[ h_{comb} : \text{calorific value of fuel oil} \]

Dilution chamber

Mass balance:
\[ M_g' = M_a + M_g \]  

\[ M_g' = \text{mass of gases at dilution chamber output} \]
\[ M_a' = \text{mass of fresh air to supply at dilution chamber} \]

Heat balance:
\[ h_g = M_g' c_p,g T_g' = M_a c_p,a T_a' + M_f c_p,f T_f + M_f h_{comb} \]  

\[ h_g : \text{enthalpy of gases at dilution chamber output} \]

Drier

Mass balance for water:
\[ M_g H_1 + Q = M_g H_2 \]  

\[ Q = \text{quantity of water to be evaporated} \]
\[ H_1 = \text{absolute humidity of gases at drier input (in kg/kg dry air)} \]
\[ H_2 = \quad \text{output} \]
Heat balance:

\[ M_{p,\text{in}} \cdot c_{p,p} \cdot T_{p,\text{in}} + h_g = M_{p,\text{out}} \cdot c_{p,p} \cdot T_{p,\text{out}} + Q \cdot h_{\text{vap}} + M'_p \cdot c_{p,g} \cdot T'_{g,\text{out}} \quad (7) \]

- \( M_{p,\text{in}}, M_{p,\text{out}} \) = mass of pulp at input and output of the drier
- \( c_{p} \) = specific heat of pulp
- \( T_{p,\text{in}}, T_{p,\text{out}} \) = temperature of pulp at input and output of the drier
- \( h_{\text{vap}} \) = heat of vaporization of water

**Approximate solution**

If we neglect the sensible heat of air and fuel oil, the equation (4) becomes:

\[ h_g = M_p \cdot h_{\text{comb}}. \]

\[ h_{\text{comb}} = 9650 \text{ kcal/kg} \quad (5, \text{p. 11}) \]

We need 0.0076 kg fuel oil per kg evaporated water (5, p. 11)

Hence,

\[ h_g = 9650 \times 0.0076 = 730 \text{ kcal} \]

If we neglect the solid, the equation (7) becomes:

\[ h_g = Q \cdot h_{\text{vap}} + M'_p \cdot c_{p,g} \cdot T'_{g,\text{out}} \]

The temperature \( T'_{g,\text{out}} \) is measured at drier output:

\[ T'_{g,\text{out}} = 143 \degree C \quad (5, \text{p. 10}) \]

\[ c_{p,g} = 0.26 \text{ kcal/kg}^\circ C \]

Hence,

\[ h_g = Q \cdot h_{\text{vap}} + M'_p \cdot c_{p,g} \cdot T'_{g,\text{out}} \]

Hence, \( M'_p = 3.50 \text{ kg dry air/kg evaporated water} \)

From equation (3):

\[ \frac{H_2 - H_1}{M'} = \frac{1}{M'_p} = \frac{1}{3.5} = 0.286 \text{ kg water/kg dry air} \]

In total,

\[ Q = 17170 \text{ kg evaporated water per h} \]

Hence,
\[ M_g = 3.50 \times 17170 = 60100 \text{ kg/h} \]
\[ M_a = (3.50 - 0.076) \times 17170 = 58700 \text{ kg/h} \]

\[ M_{g,\text{out}} = M_{g,\text{in}} + Q \]

Hence, \( M_{g,\text{out}} = 77300 \text{ kg/h} \)

Temperature of gases at dilution chamber output

\[ h_g = M'_g \cdot c_{p,g} \cdot T'_g \]

\[ M'_g = 3.50 \text{ kg} \]
\[ c_{p,g} = 0.26 \text{ kcal/kg°C} \]
\[ h_g = 730 \text{ kcal} \]

Hence, \( T'_g = 802 \text{ °C} \)

Temperature of combustion gases at furnace output

\[ h_g = M_g \cdot c_{p,g} \cdot T_g \]

\[ M_g = 19.33 \text{ kg ( Cf. 'Calorific value of fuel oil') } \]
\[ c_{p,g} = 0.316 \text{ kcal/kg°C} \]

Hence, \( T_g = 1636 \text{ °C} \)
CAPITAL COST

Price of a drier : $3 \times 10^6$ gld (from Mr. Mouris)
Price of a press : $0.465 \times 10^6$ gld (from Mr. Mouris)

System including two presses
Price before installation = $3 \times 10^6 + 2 \times 0.465 \times 10^6 = 3.93 \times 10^6$ gld

Installed price. We have to use a Lang factor of 3.10 (10, pIII-15)
So, installed price = $3.93 \times 10^6 \times 3.10 = 12.183 \times 10^6$ gld

Cost per year.
In order to calculate the cost per year, we will use:
10% for depreciation
5% for maintenance
10% out of 60% of installed price for financial expenses:
So, cost per year = $(0.10 + 0.05 + 0.10 \times 0.60) \times 12.183 \times 10^6$

= $0.21 \times 12.183 \times 10^6$

= $2.558 \times 10^6$ gld/year

Cost per hour
The factory works 240 days per year and 20 hours per day, hence 4800 hours/year.
So the capital cost per hour is:

$\frac{2.558 \times 10^6}{4800} = 914$ gld/h.

System including three presses

Price before installation
$3 \times 10^6 + 3 \times 0.465 \times 10^6 = 4.395 \times 10^6$ gld

Installed price
$4.395 \times 10^6 \times 3.10 = 13.625 \times 10^6$

Cost per year
$13.625 \times 10^6 \times 0.21 = 2.861 \times 10^6$ gld

Cost per hour
$\frac{2.861 \times 10^6}{4800} = 1022$ Gld./h.
Power consumption

System including two presses.
The power required for two presses is appr. 62 kW
The power consumption of the drier is:
- Drum electric motor: 40 h.p. = 29.4 kW
- Combustion air fan: 40 h.p. = 29.4 kW
- Draught fan: 125 h.p. = 91.8 kW
- Total: 213 kW

Power expense.
The power expense will be:
\[ 213 \times 0.07 = 15 \text{ Gld/h} \] (0.07 is the price of 1 kWh)

System including three presses.
- Power required for three presses: 94 kW
- Power required for drying unit: 151 kW
- Total: 245 kW

Power expense
The power expense will be:
\[ 151 \times 0.07 = 17 \text{ Gld/h} \]

Fuel oil consumption

System including two presses.
The quantity of wet pulp processed is:
\[ \frac{100 \times 5}{18} = 27.78 \text{ t/h including} \]
27.78 \( (1 - 0.18) = 22.78 \text{ t/h water (the presses deliver a 18 \% dry substance pulp)} \)
The quantity of dried pulp (90 \% dry substance) produced is:
\[ \frac{100 \times 5}{90} = 5.6 \text{ t/h including 0.56 t/h water.} \]
So, the quantity of water to be evaporated is:
22.78 \( - 0.56 = 22.22 \text{ t/h} \)
So we need:
\[ 22.22 \times 0.076 = 1689 \text{ kg fuel oil/h.} \]
Fuel oil price is appr. 420 Gld/t.
So the fuel oil expense will be:
1.689 \times 420 = 709 \text{ gld/h}.

System including three presses.
The quantity of wet pulp processed is:
\[
\frac{100 \times 0.5}{22} = 22.73 \text{ t/h including } 22.73(1 - 0.22) = 17.73 \text{ t/h water. (the three presses system delivers 22 % dry substance pulp)}
\]
The dried pulp 90 % DS produced is 5.6 t/h including 0.56 t/h water.
So, water to be evaporated:
17.73 - 0.56 = 17.17 t/h.
So we need 17.17 \times 0.076 = 1305 \text{ kg/h fuel oil}.
Then the fuel oil expense will be:
1305 \times 420 = 548 \text{ Gld/h}.

**Total production cost of pressing and drying.**

We can now calculate the total cost of pressing and drying for both systems:

System including two presses:

- **Capital cost:** 914 Gld/h
- **Fuel oil expenses:** 709 Gld/h
- **Power expenses:** 15 Gld/h
- **Total:** 1638 Gld/h.

System including three presses:

- **Capital cost:** 1022 Gld/h
- **Fuel oil expenses:** 548 Gld/h
- **Power expenses:** 17 Gld/h
- **Total:** 1587 Gld/h.

The production cost is less when using three presses. The difference is:
1638 - 1587 = 51 Gld/h.
So for one year:
51 \times 4800 = 244800 \text{ Gld}.
So, we will choose a high efficiency pressing unit.
THE WET PULP AS A FUEL

The wet pulp coming from the presses is 22 % DS. It is processed to the drum drier and then sold as cattle feed. The drying procedure costs a lot of fuel oil, so in this chapter we will see whether it is possible to burn the pulp immediately after pressing to produce the steam and power required by the plant. We will calculate the temperature of a furnace where would be burnt the wet pulp, and see whether this temperature is high enough to allow a good combustion of wet pulp.

Calorific value of wet pulp

Composition of air (1, table 42.9)

<table>
<thead>
<tr>
<th></th>
<th>By weigh</th>
<th>By volume</th>
</tr>
</thead>
<tbody>
<tr>
<td>Oxygen</td>
<td>23.15 %</td>
<td>20.84 %</td>
</tr>
<tr>
<td>Nitrogen and inert gases</td>
<td>76.85 %</td>
<td>79.16 %</td>
</tr>
</tbody>
</table>

Reactions of combustion

The combustible elements in pulp are carbon and hydrogen. The combustion products are:

\[
\text{C} + \text{O}_2 \rightarrow \text{CO}_2 \\
12 + 32 = 44 \text{ g (by weight)} \\
\text{Or} \ 1 + \ 2.67 = \ 3.67 \\
\text{H}_2 + \text{O}_2 \rightarrow \text{H}_2\text{O} \\
2 + 16 = 18 \text{ g (by weight)} \\
\text{Or} \ 1 + \ 8 = \ 9
\]

A. Combustion of wet pulp without excess air (cf. Characteristics of beet)

a. Oxygen. To burn 1 kg dry pulp we need:

\[
\text{C} : \ 0.470 \cdot 2.67 = 1.253 \text{ kg O}_2 = 0.877 \text{ m}^3 \text{ O}_2 \\
\text{H} : \ 0.065 \cdot 8 = 0.520 \text{ kg O}_2 = 0.308 \text{ m}^3 \text{ O}_2
\]

So, in total 1.773 kg O2 = 1.241 m³ O2

But the wet pulp already contains:

0.440 kg O2 = 0.308 m³ O2

b. Nitrogen. This oxygen brings with it:
1.333 \cdot \frac{76.85}{23.15} = 4.425 \text{ kg } \text{N}_2 = 3.522 \text{ m}^3 \text{N}_2

So, the total quantity of air necessary is: \(5.758 \text{ kg} = 4.455 \text{ m}^3\) air

The gases volumes are taken at 0°C and 760 mm Hg.

The weight of water formed is:
\[0.065 + 0.520 = 0.585 \text{ kg} = 0.728 \text{ m}^3\text{ steam}.

B. Combustion of wet pulp with excess air.

To obtain a complete combustion without unburnt material and to obtain from carbon only \(\text{CO}_2\), we need an excess air.

So we have:

Humidity of wet pulp out of \(1 : w\)

Ratio of air weight used over minimum weight necessary: \(\phi\)

\(P_a\), weight and \(V_a\), volume of air used per kg wet pulp.

\(P_g\), weight and \(V_g\), volume of combustion gases.

\(P_{gs}\), weight and \(V_{gs}\), volume of combustion gases assumed dry.

So,
\[P_a = 5.76.(1 - w).m \quad (1, \text{eqn. 42.28})\]
\[P_g = 5.76,(1 - w).m + 1 \quad (1, \text{eqn. 42.29})\]
\[P_{gs} = (1 - w).(5.76.m + 0.415) \quad (1, \text{eqn.42.31})\]
\[V_a = 4.45.(1 - w).m \quad (1, \text{eqn. 42.32})\]
\[V_g = 4.45.(1 - w).m + 0.572.w + 0.672 \quad (1, \text{eqn. 42.35})\]
\[V_{gs} = 4.45.(1 - w).m - 0.056.(1 -w) \quad (1, \text{eqn. 42.36})\]

These volumes are calculated at 0°C and 760 mm Hg. To obtain these volumes at any temperature \(t\), we can use the equation:

\[V_t = V_o \cdot \frac{273 + t}{273}\]

\(V_t\): volume at temperature \(t\)

\(V_o\): volume at 0°C

Composition of combustion gases. We know the total weight of combustion gases per kg of pulp burnt:

\[P_g = 5.76.(1 - w).m + 1\]

The weight of each individual gas is:

a. Nitrogen \(\text{N}_2 = 4.44.(1 - w).m\)
b. Oxygen \(\text{O}_2 = 1.33.(1 - w).(m - 1)\)
c. Water \(\text{H}_2\text{O} = 0.585.(1 - w) + w\)
d. Carbon dioxyde \(\text{CO}_2 = 1.72.(1 - w)\)
For wet pulp, \( w = 0.78 \)
Let us use \( m = 1.4 \)

So \( P = 5.76 \times (1 - 0.78) \times 1.4 + 1 = 2.77 \text{ kg} \)

\[
\begin{align*}
N_2 &= 4.44 \times 0.22 \times 1.4 = 1.36 \text{ kg} & (49\%) \\
O_2 &= 1.33 \times 0.22 \times 1.4 = 1.36 \text{ kg} & (4.2\%) \\
CO_2 &= 1.72 \times 0.22 = 0.378 \text{ kg} & (13.6\%) \\
H_2O &= 0.22 \times 0.78 = 0.909 \text{ kg} & (32.8\%) \\
\text{Total} &= 2.77 \text{ kg} & (100\%)
\end{align*}
\]

Calculation of combustion temperature.

The combustion temperature in the furnace is easy to calculate since the heat produced by combustion is transferred to the gases which leave to the furnace towards the boiler. So after taking into account losses:

\[
T = T_0 + \frac{\alpha \beta}{\sum \beta_i} \frac{N_i}{T_0} \quad (1, \text{ eqn. 42.43})
\]

\( \alpha \) : coefficient for unburnt solids. \( \alpha = 0.985 \) (1, p. 821)
\( \beta \) : coefficient for losses due to radiations. \( \beta = 0.985 \) (1, p. 821)

\( N_i \) : 599 kcal/kg (cf. "Characteristics of beet")

We have \( \sum \beta_i \frac{T}{T_0} = P \times \text{MCV} \)

\( \text{MCV} \) : Mean calorific value of combustion gases between \( T_0 \) and \( T \)

On the other hand, \( \text{MCV} = \sum x_i \cdot \text{MCV}_i \)

\( x_i \) : weight fraction of individual gases

\( \text{MCV}_i \) : Mean calorific value of individual gases between \( T_0 \) and \( T \).

If we assume \( T = 600 \degree \text{C} \), from (1, table 42.15) we obtain:

\[
\begin{align*}
N_2 & : \text{MCV} = 0.246 + 0.00020 \times 600 = 0.258 \text{ kcal/kg} \degree \text{C} \\
O_2 & : \text{MCV} = 0.214 + 0.00018 \times 600 = 0.225 \\
H_2O & : \text{MCV} = 0.488 + 0.000156 \times 600 = 0.562 \\
CO_2 & : \text{MCV} = 0.199 + 0.000082 \times 600 = 0.248
\end{align*}
\]

Since we know the composition of combustion gases, we can calculate:

\[
\begin{align*}
\text{MCV} &= 0.258 \times 0.49 + 0.225 \times 0.042 + 0.562 \times 0.328 + 0.248 \times 0.136 \\
&= 0.354 \text{ kcal/kg} \degree \text{C}.
\end{align*}
\]

So, \( T - T_0 = 0.91 \times 599 = 593 \degree \text{C} \)

If we have \( T_0 = 25 \degree \text{C} \), \( T = 618 \degree \text{C} \).
CONCLUSION

This temperature does not permit a correct combustion of pulp. It is estimated that the combustion is not satisfactory when the furnace temperature decreases under 900 °C. (1, p. 825).

So, the wet pulp (22 % dry substance) is not suitable as fuel in an ordinary furnace.

A number of possibilities to permit a good combustion of wet pulp could be studied:

- Drying the pulp up to a moisture content of 50 %, for instance.
- Mixing of pulp with a better fuel (wood, coal, fuel oil, etc...)
- Installation of a mixed furnace with fuel oil burners to increase the temperature etc...
THE DRIED PULP AS A FUEL. ECONOMICAL STUDY.

We will compare the cost of steam obtained from dried pulp, and the cost of steam obtained from fuel oil.

From 1 kg dried pulp, we can obtain in steam:

\[ M_v = (NCV - q) \times \alpha \rho_s \eta \]  
(1, eqn. 42.49)

\[
NCV = 4306 \text{ kcal/kg} \quad (\text{cf. "Characteristics of pulp"})
\]
\[
q = 290 \text{ kcal/kg} \quad (\text{cf. "Calorific value of pulp"})
\]
\[
\alpha = 0.975 \quad (1, \text{p. 821})
\]
\[
\rho_s = 0.975 \quad (1, \text{p. 821})
\]
\[
\eta = 0.98 \quad (1, \text{p. 821})
\]

Hence \[ M_v = (4306 - 290) \times 0.975 \times 0.975 \times 0.98 \]
\[ = 3750 \text{ kcal/kg} \]

From 1 kg fuel oil n° 2, we can obtain:

\[ M_v = 8622 \text{ kcal/kg in steam} \quad (\text{cf. "Calorific value of fuel oil"}) \]

The price of dried pulp sold as cattle feed is now: 420 Gld/t.
So, 1 kcal costs:

\[
\frac{420}{3750 \times 10^3} = 1.12 \times 10^{-4} \text{ Gld.}
\]

The price of fuel oil is appr.: 420 Gld/t.
So, 1 kcal costs:

\[
\frac{420}{8626 \times 10^3} = 0.487 \times 10^{-4} \text{ Gld.}
\]

These prices do not include the cost of furnace and boiler. But we know that a boiler working with pulp would be more expensive than a boiler using fuel oil.

Hence, using dried pulp as fuel is far from being a paying proposition. This result is not surprising, since drying the pulp is an expensive operation which costs a lot of energy.
COST EVALUATION

INVESTMENT COST

The investment cost will be calculated in the same way as in the first part 'Sugar cane plant'.

The ex. V.A.T. delivery prices of machines mostly come from the Webci 1980 booklet.

Delivery prices of machines

<table>
<thead>
<tr>
<th>Machines</th>
<th>Number</th>
<th>Specifications</th>
<th>Delivery price</th>
</tr>
</thead>
<tbody>
<tr>
<td>Belt conveyor</td>
<td>M2</td>
<td>100 tbh; 20 m x 2</td>
<td>2150 x 50 Gld</td>
</tr>
<tr>
<td>3 Screw conveyors</td>
<td>M4, M14, M16</td>
<td>100 tbh; 10 m</td>
<td>34000 x 3 &quot;</td>
</tr>
<tr>
<td>3 Slicers</td>
<td>M3</td>
<td>180 tbh; 100 kW</td>
<td>200000 x 3 &quot;</td>
</tr>
<tr>
<td>Diffuser</td>
<td>M6, M7, P9</td>
<td>100 tbh; 150 kW</td>
<td>3 000 000 &quot;</td>
</tr>
<tr>
<td>3 Presses</td>
<td>M11, M12, M13</td>
<td>3 BS 64 S; 100 kW</td>
<td>465 000 x 3 &quot;</td>
</tr>
<tr>
<td>Drier</td>
<td>F24, M23, M13</td>
<td>100 tbh; 17 twh</td>
<td>3 000 000 &quot;</td>
</tr>
<tr>
<td>Hopper</td>
<td>V29</td>
<td>20 m³</td>
<td>26 000 &quot;</td>
</tr>
<tr>
<td>Furnace + Boiler</td>
<td>H18, H20</td>
<td>2000 kg/h fuel</td>
<td>2 000 000 &quot;</td>
</tr>
<tr>
<td>+ Superheater +</td>
<td>F10</td>
<td>22 t/h steam</td>
<td>3 600 000 &quot;</td>
</tr>
<tr>
<td>Ecomiser + air-heat</td>
<td>H5, H17</td>
<td>400 °C; 40 kg/cm²</td>
<td>4 8 000 &quot;</td>
</tr>
<tr>
<td>Tank</td>
<td>V26</td>
<td>20 m³</td>
<td>4 8 000 &quot;</td>
</tr>
<tr>
<td>Compressor</td>
<td>P25</td>
<td>500 m³; 40 kW; 23 t/h</td>
<td>25 000 &quot;</td>
</tr>
<tr>
<td>Turbine</td>
<td>M15</td>
<td>975 kW</td>
<td>1 560 000 &quot;</td>
</tr>
<tr>
<td>Fans</td>
<td>P28, PX</td>
<td>40 kW; 30000 m³/h</td>
<td>15 250 &quot;</td>
</tr>
<tr>
<td>TOTAL</td>
<td></td>
<td></td>
<td>13 478 750 Gld</td>
</tr>
</tbody>
</table>

The investment cost is now the product of delivery price by the Lang factor taken as follow:

F(solid only) 3.10
F(liquid; liquid-solid) 3.63
F(gas; solid-gas; liquid-gas) 4.74
Hence, the total investment cost ex; V.A.T. will be: 55,700,000 Gld.

In order to calculate the total investment cost V.A.T. included, 18% have to be added. Hence, in total 65,725,000 Gld.

Process dependant costs

The process dependant costs per year will be calculated for 4800 working hours per year.

Fuel oil expenses

The fuel oil consumption of the plant is:
- 1918 kg/h for steam and electricity supply
- 1305 kg/h for pulp drying
The cost of fuel oil is 420 Gld/ton, hence the fuel oil expenses amount to 1353.66 Gld per hour or 6,498,000 Gld per year.

Water expenses

The water consumption of the plant is 41.73 t/h, or 200300 tons per year. The water price is approximately 0.05 Gld/m³ in Holland. Hence, the total water expenses will be 10,000 Gld per year.

Raw material expenses

Sugar beet price is a composite one: we have to use a mean value among the different prices encountered on the market. This mean value has been estimated as 80 Gld/t. The total sugar beet expense per year is then: 38,400,000 Gld.

Labour cost

The labour cost is calculated by the Wessel relation:

\[
\text{Man-hours / ton product} = \frac{\text{number of sections} \times 10}{0.76 \times \text{capacity per day}}
\]

The number of sections is taken to be 3.
The capacity is 2000 t beet per day, hence this gives 185.94 man-hours per day.
This number divided by 8 gives: 23 men/day in shift. To this number we add 15 persons as high qualified staff. The mean annual salary is 60 000 Gld. Hence, the labour expenses per year is: 2 340 000 Gld.

Economic evaluation

As in the first part, the annual investment cost as a result of depreciation, maintenance, and interest are taken to be: 21% of the total investment cost;

Production cost

The production cost can be calculated from there:

Depreciation
+ interest
+ maintenance
+ water expenses
+ fuel oil expenses
+ raw material
+ labour costs
- sale of dried pulp

TOTAL

49 760 000 Gld

Sale of dried pulp

The quantity of dried pulp produced is: 5.6 t/h, or 26 880 t per year. The dried pulp price is 420 Gld/t. Hence, in total: 11 290 000 Gld per year.

The resulting production cost per ton beet is then:

\[
\frac{49 760 000}{4800 \times 100} = 103.65 \text{ Gld}
\]

Or, the production cost per ton raw juice (12.5% sugar):
\[
\frac{49760000}{4800 \times 119} = 87.10 \text{ Gld.}
\]
CONCLUSION

The production of raw juice from sugar beet does not permit to consider numerous technical alternatives on the contrary of sugar cane. We have concentrated on the energy consumption of the plant.

A study of calorific value of fuel oil and a detailed calculation of the energy requirements of the various types of equipments has permitted to calculate the quantity of fuel oil necessary to supply electricity and steam to the plant. A production cost evaluation has showed that a highly efficient pressing of pulp will save a substantial quantity of fuel oil during the drying process.

The calorific value of wet pulp has been studied in order to envisage wet pulp as an alternative to fuel oil. The calculations have showed that this proposition was technically possible only in combination with a partial drying of pulp or in addition with a better fuel, such as wood, coal or fuel oil.

The high commercial value of dried pulp - used as cattle feed - does not make it a good paying alternative to fuel oil. After substraction of sale price of dried pulp, the production cost of raw juice is 87.10 Gld. per ton.
RAW JUICE EXTRACTION FROM SUGAR BEET AND PULP DRYING

P: PUMP
F: FURNACE
M: PRESS
H: HEAT EXCHANGER
S: SCREW CONVEYOR
M: STEAM TURBINE
M: MURPHY W. CONVEYOR
H: SUPERHEATER
B: BOILER
M: DESUPERHEATER
H: ECONOMISER
F: FAN
P: PULVERIZER
V: BUFFER TANK
D: DRIER
P: PUMP
V: VACUUM PUMP
F: COAL CRUSHER
M: MURPHY W. CONVEYOR
H: CHIMNEY
C: CYCLONE

RAW JUICE EXTRACTION FROM SUGAR BEET AND PULP DRYING

ALCOHOL PLANT
Masse in kg/s
Warmte in kW

Fabrieksvoorontwerp
No.
<table>
<thead>
<tr>
<th>Massa in kg/s</th>
<th>Warmte in kW</th>
</tr>
</thead>
<tbody>
<tr>
<td>66.36</td>
<td>42090</td>
</tr>
</tbody>
</table>
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RAW JUICE EXTRACTION FROM SUGAR CANE AND BY-PRODUCTS MANUFACTURE

J.P. BLANCHARD
C. MAILHE
FEBRUARY 1981
SUGAR BEET

RAW JUICE EXTRACTION FROM SUGAR BEET AND PULP DRYING

ALCOHOL PLANT