

A COMPUTER PROGRAM FOR SIMULATING AND EVALUATING MULTIPLE EFFECT EVAPORATORS IN THE SUGAR INDUSTRY

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Abstract

A computerised model of multiple effect evaporators in the sugar industry is described. It can be used to predict performance under given conditions, calculate the heat transfer areas required to perform a given duty, or calculate the heat transfer coefficients of the individual vessels from plant performance measurements of an operating installation. The program allows for any arrangement of vessels, whether series or parallel, co-current or counter-current, and enables any special features of an installation to be incorporated. Such flexibility is achieved by having a short program master segment for each evaporator installation, and a large number of common sub-routines which can be called by any one of the master segments. The program is iterative and operates by solving sets of simultaneous linear equations which describe the process relationships. After each iteration, the coefficients of these equations are updated.

1 Introduction

In the past, manual calculations for multiple effect evaporators were an extremely tedious chore. Even "short-cut" methods are not short in calculating time, as they usually involve an iterative procedure, and because of their underlying simplifications, they are not particularly accurate, whereas more rigorous methods would be prohibitively long to perform by hand.

For this reason it was decided to develop a full-scale computer program to perform such calculations in a reasonably rigorous manner.

This program should be able to operate in any of the following calculation modes:

- (1) Calculate the overall heat transfer coefficients for the individual vessels from measurements of vapour pressures, juice brines and clear juice flow rate. This mode is used to evaluate an existing installation.
- (2) Predict the final syrup brix as well as the process variable values (i.e. vapour flows, pressures and temperatures, and juice flows, brines and temperatures) of the intermediate streams, given the clear juice flow rate and brix, the vapour bleed rates and the calandria areas and heat transfer coefficients. This mode is used to plan plant modifications.
- (3) Calculate the clear juice flow rate and intermediate process variable values, given the clear juice and final syrup brines, the vapour bleed rates and the calandria areas and heat transfer coefficients. This mode is also used to evaluate the effect of plant modifications.
- (4) Calculate the required areas of one or more vessels and the remainder of the intermediate process variable values, given the clear juice flow rate and brix, the final syrup brix, the vapour bleed flow rates, heat transfer coefficients and the remaining areas and intermediate vapour pressures. This mode is used to plan new or expanded existing installations.

A further requirement of the program was that it should be adaptable to any desired vessel configuration, and to take into account any special features which the factory in question might have. It should be a static model, i.e. it was not required to simulate the stability or control of the process.

Thus the Program for Evaporation Simulation and Testing (PEST) was conceived.

2 General Principles

2.1 Individual evaporator

The lower part of Fig. 1 diagrammatically illustrates an evaporator of 2 effects. Incoming quantities or properties will have their names ending in I, and likewise outgoing quantities will have their names ending in O. V represents vapour flow-rate, J juice flow rate, T temperature, P pressure and BX degrees brix. TVI thus means "Temperature of Vapour at Inlet", and represents a single variable, and *not* a product of $T \times V \times I$. Furthermore, to prevent confusion between the letter X and the multiplication sign, all multiplications shall be represented by "*", as in FORTRAN notation.

A quantity VI of vapour at temperature TVI and pressure PI enters the calandria of an evaporator. The inside of the evaporator is supplied by juice at a rate JI, a temperature TJI and a brix content BXI. The incoming vapour condenses on the calandria tubes of area AC and heat transfer coefficient U, giving off its latent heat to the juice, and in so doing evaporates part of the juice water content into a quantity of vapour VO which leaves the evaporator at a temperature TVO

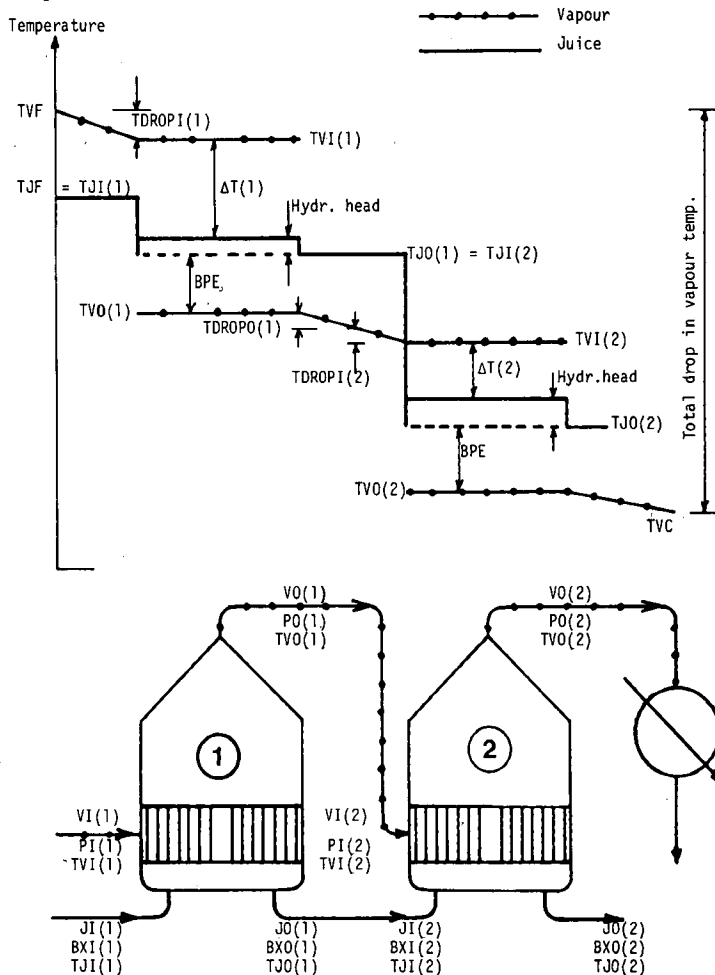


FIGURE 1 Process variables and temperature changes associated with multiple effect evaporators.

and pressure PO. The now more concentrated juice leaves the bottom of the evaporator at a rate JO, a temperature TJO and an (increased) solids concentration BXO. The amount of condensate formed in the calandria must obviously also be VI and, if no sub-cooling of the condensate is assumed, at a temperature TVI.

If heat losses, heats of solution, changes in the sensible heat of the juice and in latent heat of vapour, etc. are ignored, the amount of vapour VO released from an evaporator will be the same as the amount of vapour VI entering it, by virtue of the latent heat of condensation being approximately equal to the latent heat of evaporation of the juice solution.

2.2 Evaporators in Series

The juice JO from one evaporator can enter the next evaporator as its JI, which can be supplied by the outlet vapour VO from the previous evaporator, becoming its inlet vapour VI, as shown in Fig. 1.

The variables relating to the respective evaporators numbers 1 and 2 have accordingly been indexed, e.g. temperature of juice entering evaporator 1 is given by TJI(1).

In operation, the intermediate vapour temperatures (TVO(1) and TVI(2), in this case), will settle at that level which will let the respective ΔT 's and hence heat transfers of the 2 effects be such that vapour VO(1) = VI(2), which must of course hold.

In this manner a number of stages or effects of evaporators can be arranged in series, thus multiplying the amount of evaporative work that can be obtained from a given amount of steam supplied to the first evaporator.

For given respective temperatures of supply vapour and condenser vapour, there is however a limit to the number of evaporators that can be in series. The reasons are illustrated in Fig. 1, and are as follows:

- A temperature difference ΔT between vapour and juice is required to transfer heat across the walls of the calandria (which includes the heat resistance offered by the condensate film, scale and the boundary layer on the juice side). The more effects there are, the smaller will be the share of the total available temperature difference between supply vapour and condenser to each evaporator. This could be compensated for by larger calandria areas, but at increased capital cost.
- Boiling point elevation (BPE), whereby a solution such as sugar juice boils at a higher temperature than pure water at the same pressure, represents a reduction in the available temperature difference between supply vapour and condenser. Furthermore, the more concentrated the solution, the higher is the boiling point elevation.
- The hydrostatic head at the average depth of boiling juice raises the effective boiling temperature of the bulk of the juice by a further amount above the previously mentioned boiling point elevation.
- There are likely to be temperature and pressure losses in the inter-connecting vapour lines.

2.3 Fundamental relationships across an evaporator

From the principles of conservation of mass and energy, the following relationships hold over an evaporator:

- (1) Total heat in = Total heat out.

This means: Heat content of the incoming vapour + Heat content of the incoming juice = Heat content of the outgoing vapour + Heat content of condensate from the calandria + Heat content of the outgoing syrup + Heat losses through radiation to the atmosphere.

- (2) Latent heat of incoming vapour = Heat transfer over calandria tube walls.

This means: Latent heat of the incoming vapour = Heat transfer coefficient * Heat transfer area * (Temperature of vapour in calandria — Temperature of the boiling juice).

- (3) Juice entering evaporator = Juice leaving evaporator + Vapour leaving evaporator.
- (4) Solids in juice entering evaporator = Solids in juice leaving evaporator.

These four relationships are the basis of all the alternative methods which can be used for performing calculations on multiple effect evaporators.

Referring to Fig. 1, the following are assumed to be given: Vapour supply temperature and pressure; condenser operating temperature; inlet juice temperature, brix and flow rate; and the heat transfer coefficients and calandria areas of the respective vessels. The typical problem is to calculate the temperatures, pressures, brixes and mass flow rates of vapour and juice at entry and exit of each of the vessels. Of particular interest would be the concentration achieved in the product syrup and the amount of vapour to the 1st effect.

All these unknown values can be calculated by applying the above four fundamental relationships to the vessels and the way they are inter-connected.

Methods by which these calculations can be performed by computer are described in the next section.

3 Computer Methods Available

3.1 Iterative simulation of successive evaporator vessels

The inlet conditions of the first vessel are either known or, when not, initial estimates are made, particularly of the inlet vapour VI(1). The aforementioned 4 basic relationships, applied to that evaporator, enable the calculation of its exit vapour quantity, pressure and temperature, and its exit juice quantity, temperature and brix. These values form the input to the next vessel. In this way one can proceed up to the last, and if the value of its exit vapour pressure as per the calculations does not correspond to the condenser vapour pressure, the assumed inlet conditions at the start of the evaporator string have to be modified and the procedure repeated. This procedure has been used by several writers such as Bator and Urbaniec,¹ Bolmstedt and Jernquist,⁵ Chame and Saez⁸ and Joseph and Mellows.⁹

This method has certain drawbacks:

- The perfect, simple situation of *n* vessels, all in series as *n* effects, with the juice going in the normal co-current flow, is relatively easy to program but this situation is rarely encountered in South African mills. Frequently 2 or more vessels are used in parallel as one effect, or there could be 2 parallel strings of vessels of several effects each, all of which would make the programming extremely awkward. The amount of vapour leaving an evaporator is already dependent on the assumed vapour supply to the 1st effect, and if it splits into 2 parallel evaporators for the next effect, a further assumption on the proportions of this split will be necessary, which will in turn require additional sub-iterations.
- Where the unknowns are evaporator areas or heat transfer coefficients, a complicated program would be required.
- Although it has not been proved by the author, it is likely that such iterative methods of solution will be heavier on computer time than more direct methods.

3.2 Solution of simultaneous equations

The equations for each vessel, based on the relationship described in Section 2.3, can be set up for each evaporator, as well as additional equations to link the outputs of one evaporator with the inputs of the next.

These equations have the appearance of being linear, but many of them are not, in that the "coefficients" associated with some of the variables are themselves functions of variables. Examples: Boiling point elevation is very much a function of the brix of the solution and, to a lesser extent, of the pressure in the vessel. The heat transfer coefficients depend on the conditions in the evaporator. In some of the equations the product of mass flow rate and temperature occurs, which means a non-linearity if both are unknowns.

Because of the structure of such non-linear equations, there is no way of solving them directly in a first-time analytical method, and iterative numeric methods are necessary. The most common is the Newton-Raphson method, but other methods such as Wegstein's are also available.

The solution of simultaneous equations has the advantage that the set of unknowns to be solved can easily be varied according to the requirements of the particular problem under review.

The simultaneous equations method has been used by Radovic et al¹⁰ and Bayes, McWhinney and Murry.³ In addition, dynamic simulations by Urbaniec and Szczeniowski,¹¹ Bolmstedt⁶ and Behar⁴ also solve simultaneous equations, but these are differential equations in that they contain derivative terms.

4 The Method Chosen

Because of its mathematical tidiness and greater suitability for our requirements, PEST was based on the method of solving simultaneous equations. The equations are treated as linear by pre-calculating, where applicable, their "coefficients" from initial estimates of the variables. A neat, precise method of solving simultaneous linear equations by computer is available.

Solution of equations is done twice within the iteration: First all the equations pertaining to vapour flows between and heat balances within evaporators (relationships 1 and 2 in Section 2.3) are stated and solved, and then the juice and

solids flow between and within evaporators (relationships 3 and 4 in Section 2.3) are stated and solved. The storage space for the matrix of equations to be solved and the solution time are each proportional to the square of the number of unknowns, and such a split makes the program less unwieldy.

Fortunately the juice and solids unknowns occur only infrequently in the vapour and heat equations, and where they do appear, they do not exert too large an influence.

After passing through the two equation-solving stages, the coefficients of the linear equations are updated from the outcomes of the solutions, and the procedure is repeated.

Rather than constructing some mathematical criterion of convergence to bring the iterations to a halt, the results sheet discussed further on is printed out after each iteration, so that the user can see for himself how good the convergence is. Experience soon shows how many iterations are required to come sufficiently close to the answer. Depending on the mode of operation of the program (See Section 1), this could be from 2 to 6 iterations.

5 General Structure of the Program

This program has to cater for many different installations in the Hulett's Group. In practice, one rarely comes across the simple text book type of multiple effect evaporator installation where each effect consists of 1 vessel, but more likely there will be a large number of vessels in all kinds of parallel and series configurations. The need for counter-current vapour juice flow might arise. The condensate flash vapour return circuits usually show no consistency or symmetry. To illustrate what is meant, Fig. 2 shows the situation for the Amatikulu mill, where the quadruple effect evaporator consists of 12 vessels, with 2 parallel juice streams. To keep the drawing simple, the condensate vapour flash returns are not shown.

It is obvious that one cannot use a single program for all installations by merely changing the process variable values, calandria areas and number of effects in the data.

The other extreme would be to write a separate, complete program for each mill, catering for any peculiarities which it might have, but this would require much programming effort and be wasteful, because many of the calculating procedures

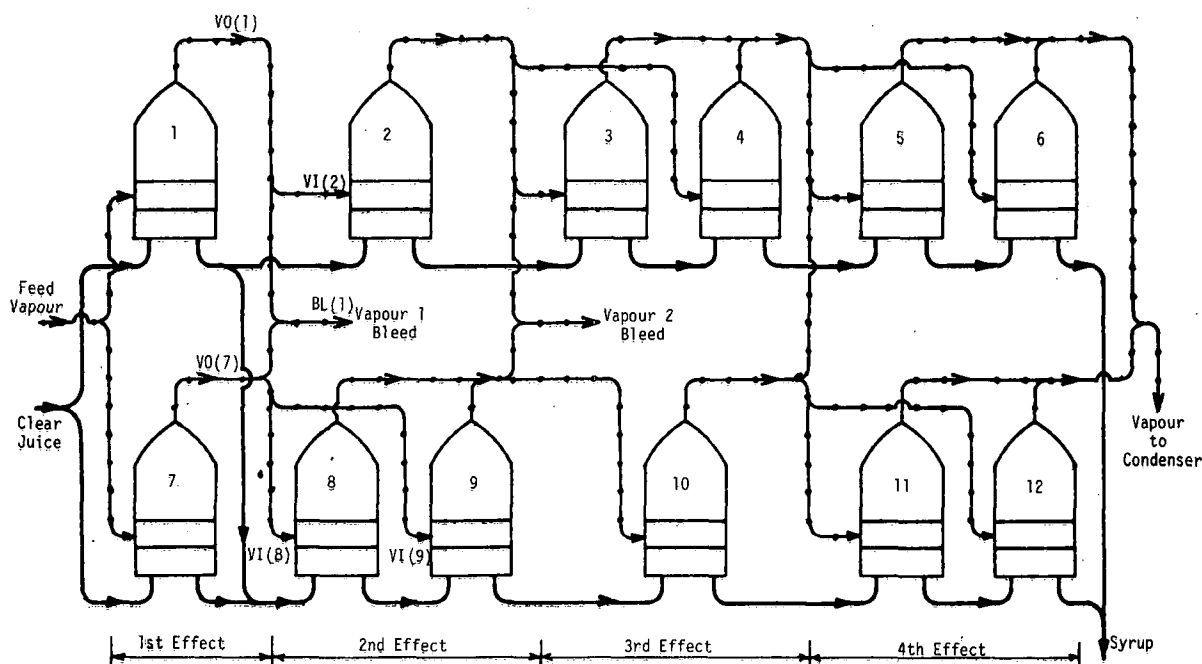


FIGURE 2 Main Vapour and juice lines of Amatikulu multiple effect evaporator.

within one program could also be used in programs for other installations.

It therefore makes sense to write a number of common sub-routines which can be accessed or called as required for the calculations of any evaporator installation.

There are 2 ways in which the calling of these sub-routines can be organised.

- A single master program (also called an executive program) can be written. In operation it will call the appropriate sub-routines for a given installation in accordance with numeric instruction data supplied to the program together with the other data on quantities, pressures, etc. This is the method followed by Bolmstedt and Jernquist⁵ and Bayes, McWhinney and Murry.³ It has the advantage that, *provided* all possibilities have been anticipated, the program, once written, can be applied to any situation without additional programming being required.
- A separate master segment can be written for each evaporator installation. This has the drawback of requiring a knowledge of computer programming and of the sub-routines, but has the advantage that each individual master segment can be much shorter and simpler than the aforementioned executive program, which has to try being all things to all men.

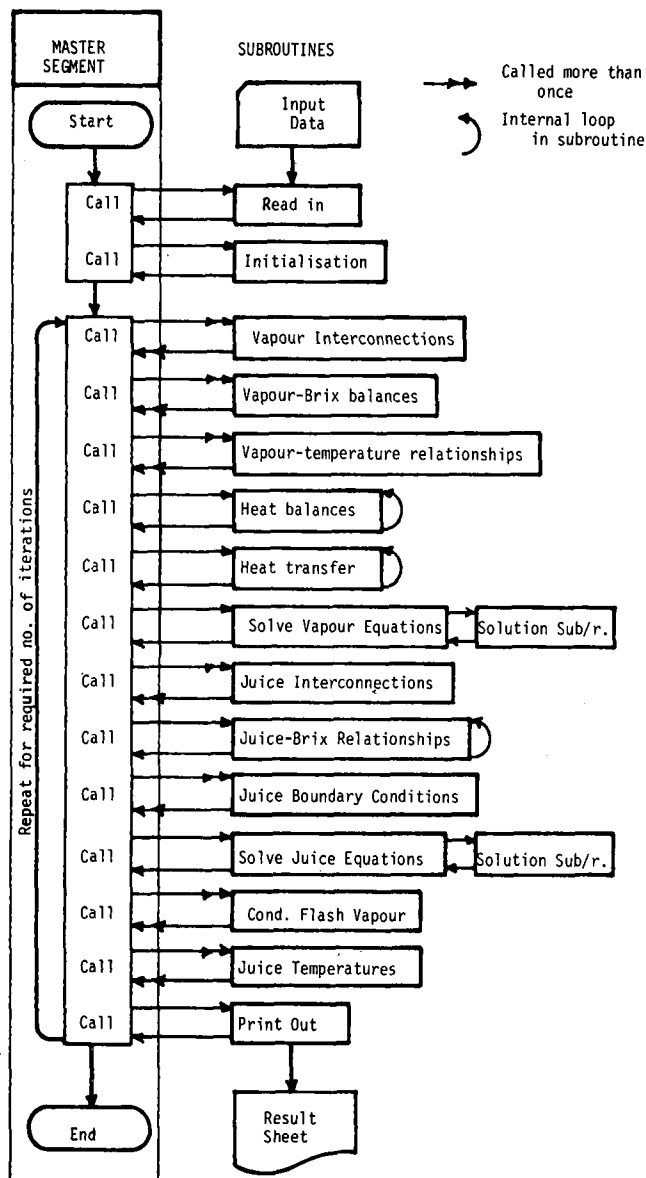


FIGURE 3 Logic flow chart of master segment and subroutines.

Experience has shown that, with the diversity of evaporator installations in the South African sugar industry, each having its own, usually unanticipated peculiarities, it is better to have the flexibility of an individual master segment for each installation, so that any special features not contained in the block of sub-routines can be dealt with in the master segment. A new master segment can easily be written by analogy with a previous one, making changes where necessary. In these times of high speed computers, the compilation time required for each new master segment is of little consequence.

For the above reasons, it was decided that PEST should be based on having an individual master segment for each installation.

A complicated installation such as Amatikulu in Fig. 2 requires only 150 executable statements in its master segment, which includes catering for condensate vapour flash returns and the peculiar partial juice cross-over after the 1st effect.

Fig. 3 shows a schematic flow chart of the operation of the program. Further details will be provided in the next section.

6 Stages in the Program Calculation Procedure

In its execution, the program passes through various stages, mainly by calling the appropriate sub-routines.

Sub-routines associated with processes inside the vessels, such as the relationships in Section 2.3, are called from the master segment only once during an iteration, and deal with the vessels in turn by going through a repetitive loop within the sub-routine. Other sub-routines, mainly those associated with interconnection of the equipment, deal with only one inter-connection situation at a time, and are called several times during an iteration.

The calculation stages are the following:

Reading in data (called only at start of program)

There is a set procedure for reading in the process data, including the initial estimates for some of the unknowns. The code no. of the desired mode of calculation is also read in. If additional data is required for a special situation, the reading-in instructions for it can follow as part of the master segment immediately after this sub-routine.

Initialisation (called only at start of program)

This consists mainly of zeroising, converting the factory tons/hour flows to kg/second and converting vapour pressures to their equivalent saturated temperatures, using a semi-empirically developed relationship.

Vapour Interconnections (called several times per iteration)

Where the vapour from one vessel passes into the next, i.e. say from i to $(i + 1)$, many writers use only one variable to express the flow rate, say $V(i)$. Although this keeps the number of variables to a minimum, it cannot cater for condensate flash vapour returns, vapour bleeds or splits and mergers in vapour streams. For this reason, separate variables were used for the respective vapour inlet and vapour outlet streams of each evaporator i , namely $VI(i)$ and $VO(i)$. For a situation such as the vapour flow from the 1st to the 2nd effect in Fig. 2, the sub-routine would be called to set up the equation:

$$VO(1) + VO(7) = VI(2) + VI(8) + VI(9) + BL(1).$$

Even in cases of a simple flow from one vessel to the next, the same policy of separate inlet and outlet quantities is used, e.g.

$VO(i) = VI(i + 1)$. Although such additional equations and variables might appear superfluous, this notation provides the

program with a degree of symmetry and consistency which lends itself far better to repetitive, vessel-by-vessel calculations.

Vapour-Brix balances (called once or more times per iteration)

In Modes 1 and 4, the clear juice flow rate and the brixes of the clear juice and the final syrup are given, and thus the difference between clear juice feed and product syrup will be equal to the total outlet vapours $\sum_i \text{VO}(i)$ from all vessels.

Additionally, for Mode 1, where inlet and outlet brix values for the juice of each vessel *i* are provided, a similar relationship can also be expressed for the outlet vapour $\text{VO}(i)$ of each individual vessel.

Vapour-Temperature relationships (called several times per iteration)

For the purposes of this program, all vapours are assumed to be saturated.

PEST allows one to take into account pressure drops (and hence temperature drops) along the inlet and the outlet vapour lines of each vessel, which is why separate variables for the outlet temperature of one vessel and inlet temperature of the next are used.

Fig. 1 illustrates how the vapour temperatures and temperature drops can be related to each other in the following equation:

$$\text{TVO}(1) - \text{TDROPO}(1) = \text{TVI}(2) + \text{TDROPI}(2).$$

There also are the boundary conditions, in that the inlet vapour temperature to the 1st effect vessel(s) will be equal to that of the vapour supply less the temperature drop, and similarly for the condenser vapour temperature and the last effects, e.g.

$$\text{TVF} = \text{TVI}(1) + \text{TDROPI}(1)$$

where TVF = temperature of the vapour feed.

Heat balances (called once per iteration, with internal loop)

By virtue of the conservation of energy, we have the following relationships across each vessel:

Heat content of entering vapour + Heat content of entering juice = Heat content of exit vapour + Heat content of exit juice + Heat content of exit condensate + Heat losses to atmosphere.

- The unknowns to be solved are the vapour flow rates and vapour temperatures at inlets and outlets, i.e. $\text{VI}(i)$, $\text{VO}(i)$, $\text{TVI}(i)$ and $\text{TVO}(i)$. Another variable in the equation is the inlet juice flow rate $\text{JI}(i)$ because of its large sensible heat content. In the solution of the set of vapour equations, $\text{JI}(i)$ is taken as a constant value, obtained from initial estimates or from the previous iteration's solution for the juice equations.
- Only the latent heat of the vapour is taken into account. The dependence of latent heat on vapour temperature is taken care of by an empirical relationship.
- The change in heat content of the juice is calculated by means of specific heats and heat of dilution, both of which can empirically be expressed in terms of degrees brix.
- Account can be taken of heat losses to the atmosphere, in that the input data contains information on the insulation heat transfer coefficients of the vessels as well as the ambient temperature.

Heat transfer (called once per iteration, with internal loop)

The relationship used is:

Heat transfer across calandria walls = Latent heat of incoming vapour.

- The driving force behind the heat transfer, i.e. $\Delta T(i)$, is the difference between the temperature of the incoming vapour $\text{TI}(i)$ and the bulk of the boiling juice. The temperature of the boiling juice $\text{TJO}(i)$ at outlet of the vessel is calculated by adding the boiling point elevation to the outlet vapour temperature $\text{TVO}(i)$. An empirical relationship for boiling point elevation in terms of vapour temperature $\text{TVO}(i)$, juice outlet brix $\text{BXO}(i)$ and juice purity has been developed.

- Because most heat transfer takes place some distance below the surface of the boiling juice, the influence of hydrostatic head on boiling point, particularly in the later effects with low absolute pressures, is important. With juice density as a function of $\text{BXO}(i)$ and the estimated average height below the surface, the additional pressure can be calculated and hence the effect on temperature by using the semi-empirical pressure temperature relationship for vapour.

- The overall heat transfer coefficient $\text{U}(i)$ is not a fixed value, but can be affected by fouling as well as process conditions. Where the HTC needs to be given for running in Modes 2, 3 or 4, we have used the relationship of Batstone and Prince,² which states that:

$$\text{Heat transfer coefficient} \propto \text{TVO}^2 * \sqrt{100 - \text{BXO}} * \sqrt{\Delta T}$$

The method is to obtain an HTC under known process conditions (usually from previous Mode 1 runs), and let the program adjust it according to the actual process conditions at the time. In most applications of PEST however, a fixed estimated HTC for each effect is submitted to the program, without adjusting for process conditions.

Solve vapour equations (called once per iteration)

After arranging the equations in a suitable form, a standard computer sub-routine is called to solve the set of linear equations.

The unknowns to be solved for will depend on the mode in which PEST is operating:

- In Modes 2 and 3, it will be the vapour flow rates and temperatures, VI , VO , TVI and TVO .
- In Mode 1, it will be the vapour flow rates and the heat transfer coefficients, VI , VO and U .
- In Mode 4, it will be the vapour flow rates and temperatures and the calandria areas, VI , VO , TVI , TVO and AC .

Juice Interconnections (called several times per iteration)

The same logic applies as for vapour interconnections discussed above, except that, at each interconnection, there will be 2 relationships: one for juice, and one for the solids content (brix mass) in the juice. For going from vessel *i* to (*i* + 1) we have:

$$\begin{aligned} \text{JO}(i) &= \text{JI}(i + 1) \\ \text{S}(i) &= \text{S}(i + 1) \end{aligned}$$

There is no need to differentiate between the respective solids mass flow rates at inlet and outlet of a vessel, because they must obviously be equal.

Juice-Brix relationships (called once per iteration, with internal loop)

For each vessel *i* there is the relationship:

Reduction in mass of juice = Amount of outlet vapour, i.e.

$$\text{JI}(i) - \text{JO}(i) = \text{VO}(i)$$

In this sub-routine the $\text{VO}(i)$ term is not treated as an unknown, but is obtained from the solution of the previous vapour equations.

Juice boundary conditions (called several times per iteration)

The degrees brix of the inlet juice to the first vessels is always given and, depending on the mode of operation, the quantity

of clear juice and/or the degrees brix of the product syrup will be given, all of which form additional relationships between the juice and its solids content.

Solve juice equations (called once per iteration)

Again the standard computer sub-routine of solving simultaneous linear equations is used, the unknowns being juice and solids mass flows JI, JO and S.

Condensate flash vapour return (called several times per iteration)

Where condensate comes from an evaporator calandria, its quantity must equal the corresponding quantity of inlet vapour VI(i), and where it comes from juice heaters, vacuum pans, etc. which have been supplied by one of the vapour bleed streams j, its quantity will equal the corresponding amount of bleed vapour BL(j). In each case, the condensate is calculated to flash from the saturated temperature of the vapour from which it was formed to that of the vapour line into which it discharges. Cascading of condensate from one flash pot to the next can also be dealt with.

Juice temperatures (called several times per iteration)

The juice outlet temperature TJO(i) from each vessel i is calculated from its outlet vapour temperature TVO(i), to which is added the calculated boiling point elevation.

The inlet temperature TJI(i) of each vessel is calculated from the specific heats and temperatures of the juice outlet streams supplying it, weighted by their respective mass flow rates.

Print-out (called once per iteration)

All the results from each iteration are printed out as shown in the example of Figure 4.

7 Special Considerations for the Individual Modes of Operation

Mode 1: Determine heat transfer coefficients

The HTC's U(i) together with the vapour inlet and outlet flows VI(i) and VO(i) are the unknowns to be solved. For each vessel the juice inlet and outlet brixes BXI(i) and BXO(i) and the vapour inlet and outlet pressures PI(i) and PO(i) are measured on the plant, and this forms part of the input data.

MATERIAL AND HEAT BALANCE FOR EVAPORATOR TRAIN AT AMATIKULU												DATE: 31/03/81	TIME: 13/29/11		
RUN NO: 0	OBJECT OF RUN: DEMONSTRATION RUN FOR AMATIKULU MILL														
TYPE OF CALCULATION: 3:															
CALC. CLEAR JUICE FLOW AND VAPOUR FLOWS AND TEMPS, GIVEN SYRUP BRIX AND BLEED FLOWS.															
EVAPORATOR REF. NO :	1	2	3	4	5	6	7	8	9	10	11	12			
EVAP. NAME IN FACTORY	KA	1A	2A	3A	4A	5A	6A	7A	8A	9A	10A	11A	12A		
EVAPORATOR EFFECT NO.	1	2	3	3	4	4	1	2	2	3	4	4			
MATERIAL FLOWS:															
VAPOUR IN:															
FLOW RATE, KG/HR	78247.	20978.	18200.	16142.	16057.	11681.	48987.	21065.	19567.	19006.	16585.	13512.			
TEMPERATURE, DEG.C.	123.1	114.5	103.6	103.6	89.4	89.4	123.1	114.8	114.8	105.6	89.4	89.4			
PRESSURE, KPA ABS	219.	168.	115.	115.	69.	69.	219.	168.	168.	115.	69.	69.			
VAPOUR OUT:															
FLOW RATE, KG/HR	74822.	22777.	19899.	15778.	17702.	11098.	46788.	25242.	19222.	21622.	18941.	12909.			
TEMPERATURE, DEG.C.	116.2	105.0	90.9	90.9	61.1	61.1	116.2	105.0	105.0	90.9	61.1	60.1			
PRESSURE, KPA ABS	176.	141.	75.	75.	41.	41.	176.	141.	141.	75.	41.	41.			
JUICE IN:															
FLOW RATE, TONS/HR	224.5	113.0	91.0	71.1	55.5	57.6	144.5	135.6	110.3	91.1	70.5	51.5			
TEMP., DEG.C.	110.0	116.4	105.4	91.5	91.9	63.7	110.0	116.4	105.3	105.5	91.7	62.4			
DEGREES BRIX	11.1	16.0	20.6	26.6	34.2	50.5	11.1	16.5	19.9	24.2	31.2	42.7			
JUICE OUT:															
FLOW RATE, TONS/HR	147.6	91.0	71.1	55.5	37.6	26.5	97.7	110.3	91.1	70.5	51.5	38.4			
TEMP., DEG.C.	116.4	105.4	91.5	91.9	63.7	63.4	116.4	105.3	105.5	91.7	62.4	62.4			
DEGREES BRIX	16.6	20.8	20.6	34.2	50.5	71.4	16.4	19.9	24.2	31.2	42.7	57.1			
EVAPORATOR DATA:															
HEAT TRANSFER:															
CALANDRIA AREA, SQ.M	2270.	743.	743.	743.	743.	743.	1857.	743.	743.	743.	743.	743.			
HYDROSTATIC HEAD, M	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3			
HYDROSTATIC TEMP. RISE, °C	0.6	0.8	1.2	1.2	3.5	3.9	0.6	0.8	0.8	1.2	3.4	3.8			
BOILING POINT ELEV., DEG.C	0.2	0.3	0.6	1.0	1.9	4.3	0.2	0.3	0.3	0.5	1.3	2.5			
O/ALL HTC, KW/SQ.M/DEGC	2.00	2.00	1.40	1.30	0.60	0.20	2.60	2.00	2.00	1.40	1.50	1.60			
O/ALL TEMP. DIFF.	6.1	8.6	10.9	10.4	22.6	20.0	6.1	8.7	8.5	10.6	23.6	23.0			
HEAT TRANSFER RATE, KW	47399.	12850.	11315.	10038.	10169.	7437.	29674.	12904.	11987.	11817.	10499.	8540.			
HEAT LOSSES:															
SHELL AREA, SQ.M	260.	131.	131.	131.	131.	146.	186.	131.	131.	131.	131.	146.			
O/ALL HTC, KW/SQ.M/DEGC	0.0034	0.0034	0.0034	0.0034	0.0034	0.0034	0.0034	0.0034	0.0034	0.0034	0.0034	0.0034			
HEAT LOSS RATE, KW	81.	30.	50.	50.	10.	18.	58.	36.	36.	30.	10.	17.			
VAPOUR BLEEDS AND CONDENSATE FLASH FEEDS:															
BLEED OR FLASH REF. NO.															
VAPOUR BLEED FLOW, KG/HR	60000.	22912.													
VAPOUR BLEED TEMP., DEG.C	110.0	98.4													
VAPOUR BLEED PRESS. KPA AB	140.	90.													
CONDENSATE FLASH, KG/HR	11013.	843.	466.												
TOTAL VAPOUR FLOWS, KG/HR :	TOTAL EVAPORATION = 305852.						TOTAL BLEEDS = 82912.			TOTAL VAPOUR FEED = 127433.					

FIGURE 4 Example of PEST print-out for Amatikulu mill evaporator installation. Values underlined represent input data. All other values shown were obtained by calculation.

Referring back to Fig. 1, it should be clear from material and energy balance considerations that specification of the brixes of the intermediate juice streams in the input data is superfluous, and will result in more relationships than unknowns. One could discard sufficient data to retain the same number of relationships as unknowns, but with more complicated arrangements, such as in Fig. 2, one could easily discard non-redundant data by mistake. Besides, the more values that can be input, the higher the accuracy of the output is likely to be.

The method decided on was to use all the brix data from the juice streams, and let the program perform a least-squares fit to solve for the unknowns, similar to multiple linear regression analysis techniques.

Mode 2: Output brix unknown

The clear juice brix and flow rate are given, and the brix of the final syrup is among the variables which have to be calculated. This mode gives rise to instability in the calculation procedure, because of the dependence of boiling point elevation and heat transfer coefficient on brix values, particularly in the later effects. The problem of divergence was overcome by using damping factors when updating the values of outlet juice brix, heat content of the entering juice and the heat transfer temperature difference, as per the following example:

$$\text{Updated BXO}(i) = \alpha(i) * \text{BXO}(i) \text{ from latest iteration} + (1 - \alpha(i)) * \text{BXO}(i) \text{ from previous iteration,}$$

where $0 < \alpha(i) \leq 1$.

There will be a separate $\alpha(i)$ value for each vessel i , determined by experience, and generally ranging from 1,0 for the first effect down to approximately 0,6 for the last.

If Mode 2 operations are not used, $\alpha = 1$ can be used throughout.

Mode 3: Clear juice flow unknown

The brixes of the clear juice and the final syrup are given, and the clear juice flow rate is among the unknowns to be determined. This mode of operation does not require any special procedures.

Mode 4: Areas unknown

In addition to the vapour flows and intermediate temperatures, one or more calandria areas are among the unknowns. Some of the intermediate temperatures also have to be specified so as to keep the total number of unknowns equal to the total number of relationships. The clear juice flow rate as well as the clear juice and final syrup brixes have to be specified.

8 Optimisation of Areas

PEST does not attempt to minimise the combined calandria areas of all the vessels in the sense of performing some kind of a search. However, the following method is used in conjunction with PEST:

From the work of Buczolic and Zadori,⁷ the following relationship for any effect i in an n -effect evaporator installation can be deduced:

$$\Delta T(i) = k * \sqrt{Q(i)/U(i)}, \text{ for } i = 1, 2, \dots, n \quad (1)$$

$$\text{where } k = \frac{\Delta T(1) + \Delta T(2) + \dots + \Delta T(n)}{\sqrt{Q(1)/U(1)} + \dots + \sqrt{Q(n)/U(n)}} \quad (2)$$

$\Delta T(i)$ = overall temperature difference across calandria walls of effect i , °C.

$Q(i)$ = total heat flux through walls of effect i , kW

$U(i)$ = overall heat transfer coefficient for effect i , kW/(°Cm²).

Note that here the index i refers to *effect* no., which is not necessarily the same as *vessel* no.

The summation of the unknown $\Delta T(i)$'s in equation (2) can be determined from:

$$\text{TVF} = \text{TVC} + \Delta T(1) + \dots + \Delta T(n) + E(1) + \dots + E(n)$$

where TVF = temperature of feed vapour, °C.

TVC = temperature of condenser vapour, °C.

$E(i)$ = boiling point elevation + hydrostatic temperature rise + temperature drops over vapour inlet and outlet lines, for $i = 1, 2, \dots, n$.

Then k can be written as

$$k = \frac{\text{TVF} - \text{TVC} - E(1) - \dots - E(n)}{\sqrt{Q(1)/U(1)} + \dots + \sqrt{Q(n)/U(n)}} \quad (3)$$

The procedure is to perform a 1st run in Mode 4, using any reasonable estimate of intermediate pressures (or temperatures) between successive effects, thus determining a 1st estimate of the values $Q(i)$, $E(i)$ and $U(i)$. Using equation (3) and then equation (1), new values of $\Delta T(i)$ and hence more exact values of intermediate vapour temperatures can be calculated. This data is entered for a 2nd run in Mode 4, the output of which will provide near-optimum values of the areas $AC(i)$.

9 Running of the Program

The program was written in ICL Extended FORTRAN, and run on an ICL 2904 computer with 64 k words. The time per iteration depends on the size of the evaporator installation, but typically requires about 20 seconds per iteration.

In case things go wrong, particularly when the master segment for a new installation is run for the first time, a facility is provided for the computer to print out the complete matrix of linear equations which is submitted to the solution routine. From this it usually is easy to trace the causes of incorrect results or of equations which cannot produce a solution.

10 Special Uses

Special applications of PEST, involving modifications to the existing master segments of the installations, included the following:

- *Provision of vapour 2 bleed*

At one of Huletts mills, which was only supplying vapour 1 bleed, runs were done to test the effect of a vapour 2 bleed on general evaporator capacity and on total steam consumption.

- *Throttling between effects*

As part of a large mill expansion project, throttling between the 2nd and 3rd effects to raise the vapour 1 and vapour 2 pressures was investigated. This was achieved by fixing the clear juice flow rate and the clear juice and final syrup brixes, but leaving out the vapour temperature relationship between the 2nd and 3rd effects.

- *Vapour recompression*

In the design of a new diffuser-equipped mill, the master segment was written to allow for vapour recompression by steam ejector or alternatively by mechanical means. The source and delivery positions of the vapour could be varied by reading in additional data cards.

- *Varying the number of effects*

In designing some new or expanded multiple effect evaporator installations, the relevant master segments were written such that the number of effects could be changed by means of a data card, without having to re-compile. This

was only practicable for the straight-forward one-vessel-per-effect situation.

11 Future Refinements

- The Newton-Raphson method of solving the simultaneous equations might be introduced. Although a more complicated program will result, it should require fewer iterations, and little or no need for the aforementioned damping factors when running in Mode 2. When running in Modes 2, 3 or 4, the given vapour bleed rates need not be fixed, but can partly be determined by the clear juice flow rate and the brix mass flow rate.
- With the recent purchase of a Data General Eclipse S/140 computer with video terminals, it is intended to let data be input in an interactive manner, with the computer prompting the user on the next data item to enter.
- A longer-term aim is to let this program be part of a larger overall factory material and heat balance.

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