

BAYER DIGESTER OPTIMIZATION STUDIES
USING COMPUTER TECHNIQUES

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Theoretically required heat transfer performance by the multistaged flash heat reclaim system of a high pressure Bayer digester unit is determined for various conditions of discharge temperature, excess flash vapor and indirect steam addition. Solution of simultaneous heat balances around the digester vessels and the heat reclaim system yields the magnitude of available heat for representation of each case on a temperature-enthalpy diagram, where graphical fit of the number of flash stages fixes the heater requirements. Both the heat balances and the trial-and-error graphical solution are adapted to solution by digital computer techniques.

Introduction

Representation of a multistaged flash heat reclaim system can be made on a temperature-total enthalpy diagram similar to Figure 1, if the amount of heat to be transferred and the available temperature driving force are known. In the design of a system these limits are either known or specified, but certain simplifying assumptions can usually be applied to obtain algebraic solutions for external heating demand and the number of flash stages required by specified reclaim heater performance (1).

Analysis of an existing Bayer digester unit, conversely, is usually complicated by such factors as external addition or removal of heat in one or more flash stages, direct steam injection, pre-heating of both caustic liquor and bauxite slurry, or dilution in a flash stage. These cases are best handled by trial-and-error graphical fit of the number of stages in the unit on a temperature-total enthalpy diagram, which defines the required heater approach temperature differences for the flow rates used in constructing the diagram.

Procedure for Graphical Solution of an Example Case

Consider a high pressure digester unit with direct steam injection heating in the digester vessels, flash heat recovery in eight indirect heating stages (one for liquor feed and seven for liquor plus slurry feed), indirect external steam addition to the first flash stage, and a dilution stream into the last flash stage. The unit, arranged as shown in Figure 2, is assumed to carry flow rates such that available heat exceeds transfer capacity of the heaters, resulting in excess flash vapor to a barometric condenser.

If temperatures are set for the liquor and slurry feed (T_L and T_S), the digester (T_R), and the blowoff and dilution streams (T_B and T_W), then the required direct steam injection rate, say S_D , is related to the heater discharge temperature from the first flash stage, D_1 , by heat balances around both the flash reclaim system and

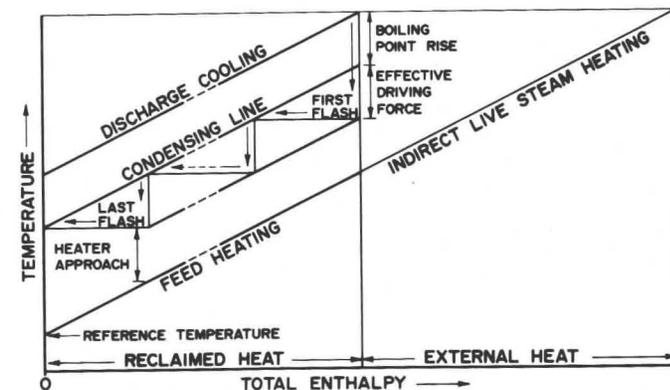


Fig. 1 - Diagrammatic representation of a multistaged flash heat reclaim system.

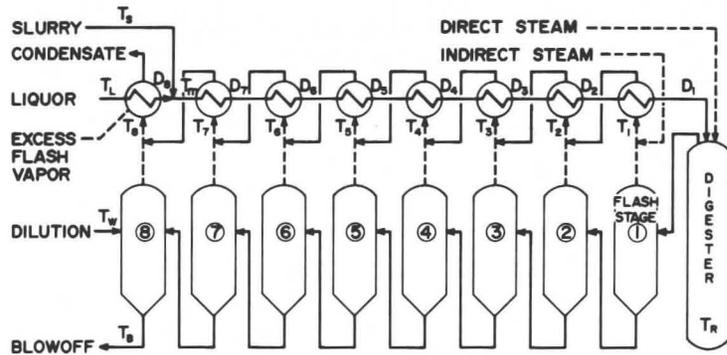


Fig. 2 - Schematic arrangement of example digester unit

the digester vessels, the terms of which are listed positively in Table I.

Data for Simultaneous Heat Balances

Average specific heats for liquor and slurry streams are best determined from empirical correlations of laboratory heat capacity versus temperature data for the system being studied. The authors' experience has shown that considering all combined slurry streams the same as liquor at the maximum attained alumina concentration gives values for specific heat which differ by less than two percent from weighted averages calculated from estimated heat capacities of the separate phases.

The mix temperature, T_m , of combined liquor and slurry can be readily calculated for specified feed rates, if the liquor heater discharge temperature is assumed and an estimated value for the mix temperature is used to determine the average liquor and slurry specific heats. The estimated T_m can then be adjusted until the calculated T_m closes to within desired agreement, while the assumed liquor heater discharge temperature, D_B in the example case, can later be changed as required to make the liquor heater approach temperature difference consistent with the average approach determined for the slurry heaters.

Feed rates of alumina trihydrate and alumina monohydrate are calculated from ABEA and MEA analyses of the bauxite (see nomenclature) and percent solids in the feed slurry. Specific heats of solution, $(\Delta h_{soln})_{tri}$ and $(\Delta h_{soln})_{mono}$, are taken as 170 kcal/kg Al_2O_3 (or 306 Btu/lb Al_2O_3) for trihydrate and 70 kcal/kg Al_2O_3 (or 126 Btu/lb Al_2O_3) for monohydrate, respectively (2), the former being assumed all dissolved in the heaters, and the latter in the digester vessels.

All specific enthalpies for steam and condensate streams are taken from a standard steam table, using the specified temperature (for saturated vapor or condensate) or temperature and pressure (for superheated steam) to enter the table for each stream. Condensate leaving each stage is assumed to be at saturation temperature

Table I. Conservation of Heat in a Digester Unit

Flash Heat Reclaim System	
Consumed Heat	Available Heat
Liquor heating, ΔH_L $Q_L(C_P)_L(D_B - T_L)$	Combined slurry cooling $(Q_L + Q_S)(C_P)_T(T_R - T_B)$
Combined slurry heating, ΔH_T $(Q_L + Q_S)(C_P)_T(D_1 - T_m)$	Condensed steam cooling $S_D(h_{fR} - h_{fB})$
Heat of solution, ΔH_{ABEA} $Q_{ABEA}(\Delta h_{soln})_{tri}$	Total condensate cooling ^a $Q_C(h_{fB} - h_{fC})$
Dilution heating, ΔH_W $Q_W(C_P)_W(T_B - T_W)$	Indirect steam heat $S_I(h_{gI} - h_{fB})$
System loss and venting $\Delta H_{loss} = \Delta H_{htr} + \Delta H_{f1}$ (est.)	
Excess flash heat, H_{XS} $S_{XS}(h_{gC} - h_{fC})$	
Digester Vessels	
Consumed Heat	Available Heat
Combined slurry heating $(Q_L + Q_S)(C_P)_T(T_R - D_1)$	Injection steam condensing $S_D(h_{gD} - h_{fR})$
Heat of solution $Q_{MEA-ABEA}(\Delta h_{soln})_{mono}$	
System loss ΔH_{loss} (estimated)	

^a Q_C , the total condensate produced, also depends on S_D .

for the pressure in the heater shell, which means that the temperature is set by boiling point rise in the flash tank, if vapor line pressure drop is neglected. Boiling point rise (BPR), like specific heat, is best estimated from laboratory data for liquors in the system being analyzed. Experience with a high pressure Bayer digester unit similar to the example has shown that BPR of the cooling slurry decreases slightly faster with decreasing temperature than it increases with increasing soda concentration, typically ranging from about 16°F (9°C) at the digester vessels to 14°F (8°C) at the low pressure end, less an additional decrease occurring in the last flash tank when the dilution stream enters.

For each case of assumed indirect steam addition and excess flash vapor rates, the amount of total condensate produced can be expressed in terms of an estimated condensate economy for the unit as follows:

$$Q_C = \text{Economy}(S_D + S_I) \quad (1)$$

Using this quantity with the rest of the discussed data plus estimated rates of heat loss by radiation, convection or venting from heaters, flash tanks and digester vessels, the heat balance equations of Table I can be solved simultaneously to obtain the heater discharge temperature, D_1 , and the resulting direct steam requirement, S_D , for the specified conditions and flow rates.

Temperature-Total Enthalpy Diagram for System

Data from the heat balance for the flash heat reclaim system can now be used to draw the heating and cooling lines of a temperature-total enthalpy (T-TE) diagram, as illustrated in Figure 3. Note that the only loss term affecting the heating line is loss from the heaters, and that the cooling line has a constant temperature displacement by the rate of heat given up by indirect steam condensing at saturation temperature of the stage where it is added. (This condensing temperature can be satisfactorily estimated, as its effect is small.) If flash vapor were being vented from a stage and removed from the system, the cooling line would be displaced at constant enthalpy an amount equivalent to the temperature drop of the slurry and condensate required to vaporize it, and that rate of heat loss would also appear in the system heat balance.

Comparison of Figure 3 with the heat balance terms of Table I reveals that the slope of a heating or cooling line for stream i is given by the reciprocal of $Q_i(C_p)_i$; thus, the diagram is affected by both the flow rate and specific heat (or apparent specific heat due to losses) of every stream in the system. If the condensate from each stage is cooled by flash evaporation in the next lower pressure stage, cascading in parallel with the slurry, then the effective mass flow of slurry plus condensate remains constant throughout the system, yielding a cooling line of constant slope, even though the slurry becomes more concentrated in each successive flash tank. This concentration change tends to minimize variation in boiling point rise through the system, though BPR need not be constant, as previously discussed. Whatever the variation, the condensing line is located below the cooling line by the best estimate of BPR at each point through the system.

With the diagram thus defined, all that remains is to adjust the slurry heater approach temperature differences, usually taken as the same for all heaters, until the exact graphical fit of the number of available stages is found. This fixes the saturation temperature and liquor or slurry heater discharge temperature for each stage, allowing the heat transferred in each stage to be read graphically (See Figure 4) or calculated by the generalized equation.

$$\Delta H_n = H_n - H_{n+1} = (Q_L + Q_S)(\bar{C}_P)_T(D_n - D_{n+1})^* \quad (2)$$

Q_S being omitted for the liquor heating stage(s).

*Using the nomenclature of Figure 4, D_{n+1} is replaced by T_m for the first slurry heating stage and by T_r for the last flash stage (first liquor heating stage) where H_{n+1} is zero.

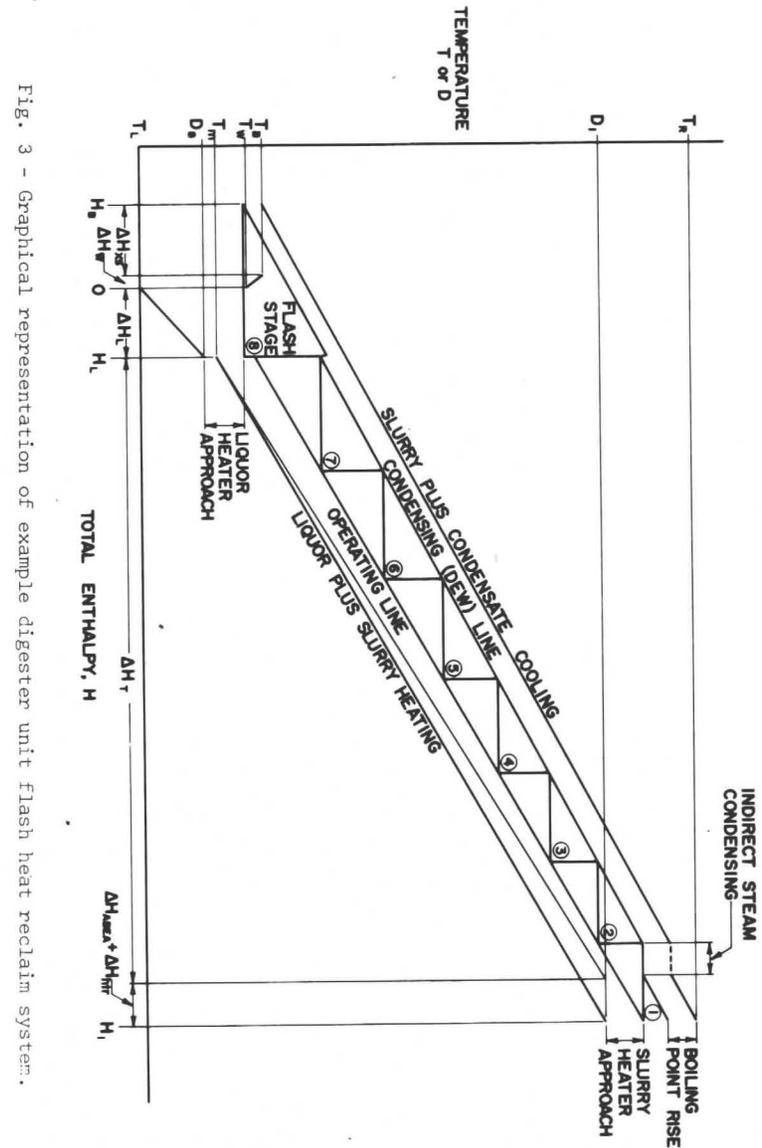


Fig. 3 - Graphical representation of example digester unit flash heat reclaim system.

Then the condensate production by evaporation in each stage is given by

$$E_n = \frac{(H_n - H_{n+1}) - Q_C^T (T_{n-1} - T_n)}{(h_g - h_f)_n} \quad (3)$$

where Q_C^T is the sum of $Q_C^1 + Q_C^2 + \dots + Q_C^{n-1}$. Exceptions are the first flash stage, where there is no cascading condensate Q_C^T to flash cool, and stage(s) with indirect steam addition, where E_n includes the indirect steam, equal to S_I if it is assumed saturated at T_n . Total condensate production by the unit is the sum of E_n values for all stages, and division of this total by the steam consumption, $S_D + S_I$, gives the unit condensate economy. This calculated economy should be compared with the estimated economy used in Equation 1, substitution being made if the difference between estimated and calculated condensate production is greater than desired.

Each such substitution or change, including adjustments to the assumed liquor and slurry heater discharge temperatures used in estimating specific heats, requires a new solution of the simultaneous heat balances in Table I plus a new graphical solution on a revised T-TE diagram. In addition, the entire process must be repeated for each case of assumed indirect steam addition, blowoff temperature and excess flash rate, to obtain a model of the digester unit over a typical range of operating conditions. Such an approach is obviously possible, but the time consuming nature of trial-and-error graphical solutions makes it impractical for analysis of an operating unit.

Application of Digital Computer Techniques

Fortunately, the assume, try and adjust pattern of defining a system T-TE diagram can be easily adapted to digital computer iteration techniques, and a method of modeling the graphical fit procedure to the computer has been developed.

Computer Solution of Heat Balances

Starting with assumed values of D_8 and T_m , the adiabatic mix temperature of the example feed liquor and slurry streams is

$$T_m = \frac{Q_L (C_P)_L D_8 + Q_S (C_P)_S T_S}{Q_L (C_P)_L + Q_S (C_P)_S} \quad (4)$$

which can be written as a FORTRAN statement, as can the required empirical specific heat calculations and the rest of the expressions to be developed. Using an IF statement, the calculated value of T_m can be substituted for the assumed T_m , repeating the operation as required to close within desired agreement between the two, as before.

Specified feed rates and temperatures, digester and blowoff temperatures, indirect steam rate and excess flash vapor rate can next be combined with the assumed values of D_1 , corresponding specific heats, and estimated condensate economy, to obtain expres-

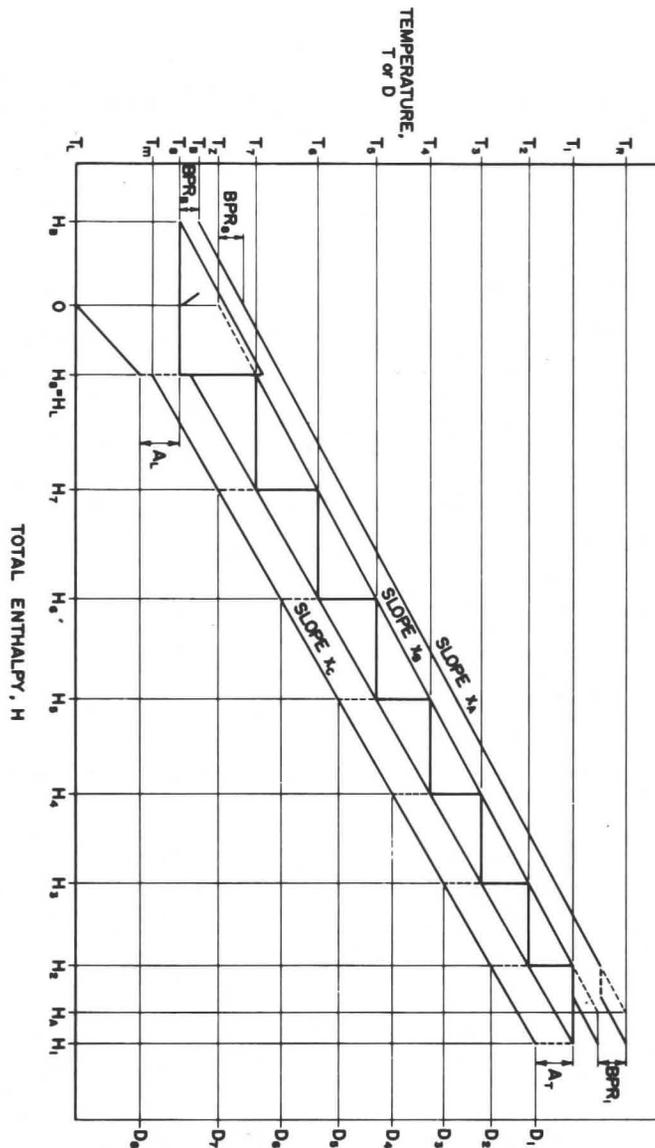


Fig. 4 - Nomenclature for stagewise calculations in computer model of example digester unit graphical fit.

sions of the form $a_{i1}D_1 + a_{i2}S_D = b_i$ for each of the heat balances in Table I. Applying determinant theory, the simultaneous solutions are

$$D_1 = \frac{b_1 a_{22} - b_2 a_{12}}{a_{11} a_{22} - a_{21} a_{12}} \quad (5)$$

and

$$S_D = \frac{a_{11} b_2 - a_{12} b_1}{a_{11} a_{22} - a_{21} a_{12}} \quad (6)$$

where

$$a_{11} = (Q_L + Q_S)(\bar{C}_P)_T \text{ from } T_m \text{ to } D_1 \quad (7)$$

$$a_{12} = -(h_{fR} - h_{fB}) - \text{Economy}(h_{fB} - h_{fC}) \quad (8)$$

$$b_1 = (Q_L + Q_S)(\bar{C}_P)_T(T_R - T_B) + \text{Economy}(S_I)(h_{fB} - h_{fC}) + S_I(h_{gI} - h_{fB}) + (Q_L + Q_S)(\bar{C}_P)_T T_m - Q_L(\bar{C}_P)_L(D_1 - T_L) - Q_{ABEA}(\Delta h_{\text{soln}})_{\text{tri}} - Q_W(\bar{C}_P)_W(T_B - T_W) - S_{XS}(h_{gC} - h_{fC}) - \Delta H_{\text{htr}} - \Delta H_{f1} \quad (9)$$

$$a_{21} = (Q_L + Q_S)(\bar{C}_P)_T \text{ from } D_1 \text{ to } T_R \quad (10)$$

$$a_{22} = h_{gD} - h_{fR} \quad (11)$$

$$b_2 = (Q_L + Q_S)(\bar{C}_P)_T T_R + Q_{MEA-ABEA}(\Delta h_{\text{soln}})_{\text{mono}} + \Delta H_{\text{loss}} \quad (12)$$

All specific heats can be calculated from empirical expressions for the specified temperatures, steam tables can be entered through sub-routines to obtain required specific enthalpies, and the term $(h_{fB} - h_{fC})$ in a_{12} and b_1 is numerically equal to boiling point rise in the last flash stage after dilution, BPR_B . For studies over a limited temperature range with constant assumed feed slurry composition, it is also practical to use typical specific enthalpy differences with empirical steam table corrections for small changes in one of the temperatures, and to correct base heats of solution by ratio of the feed slurry rate to a base rate. Finally, the calculated value of D_1 should be substituted for the assumed D_1 , repeating the simultaneous solution to close the difference, as in the calculation of T_m .

Computer Fit of System Model

With the calculation of D_1 , the limits of the temperature-total enthalpy diagram are again set for the specified temperature and flow conditions. The heating lines are located by calculating the total enthalpy limits,

$$H_1 = \Delta H_L + \Delta H_T + \Delta H_{ABEA} + \Delta H_{\text{htr}} = Q_L(\bar{C}_P)_L(D_1 - T_L) + (Q_L + Q_S)(\bar{C}_P)_T(D_1 - T_m) + Q_{ABEA}(\Delta h_{\text{soln}})_{\text{tri}} + \Delta H_{\text{htr}} \quad (13)$$

and

$$H_L = \Delta H_L = (Q_L(\bar{C}_P)_L)(D_1 - T_L) \quad (14)$$

while the lower limit of the cooling line is set by

$$H_B = -(\Delta H_{XS} + \Delta H_W) = -S_{XS}(h_{gC} - h_{fC}) - Q_W(\bar{C}_P)_W(T_B - T_W) \quad (15)$$

as shown in Figure 3. If H_A is defined as $H_1 - S_I(h_{gI} - h_{f1})$, then the cooling line slope is

$$x_A = \frac{T_R - T_B}{H_A + (\Delta H_{XS} + \Delta H_W)} = \frac{T_R - T_B}{H_A - H_B} \quad (16)$$

and the temperature on the condensing line where total enthalpy is defined as zero is given by

$$T_Z = T_B - x_A H_B - BPR_8 \quad (17)$$

where BPR_8 is the estimated point rise of slurry entering the eighth flash stage, as shown in Figure 4. The condensing line slope can then be defined as

$$x_B = \frac{T_R - BPR_1 - T_Z}{H_A} \quad (18)$$

and the slurry heating line slope as

$$x_C = \frac{D_1 - T_m}{H_1 - H_L} \quad (19)$$

which is also the operating line slope, if equal slurry heater approach temperature differences are assumed.

For a given trial value of this slurry heater approach, say A_T , the temperature of saturated condensing vapor from flash stage n , T_n , must be located on the operating line at H_n , the total enthalpy of slurry discharging the heater condensing this vapor, or

$$T_n = T_m + A_T + x_C(H_n - H_L) \quad (20)$$

Then the total enthalpy of slurry discharging the next lower temperature stage, H_{n+1} , is found on the condensing line at T_n , or

$$H_{n+1} = \frac{T_n - T_Z}{x_B} \quad (21)$$

which locates T_{n+1} on the operating line, and so on. Since H_1 is

already known (equation 13), a DO statement can be set up to solve equations 20 and 21 for T_n and H_{n+1} through H_8 (for n up to 7, the number of slurry heating stages). An IF statement can then be applied to compare H_8 with H_L (equation 14), making appropriate adjustments to A_T and repeating the DO loop calculations of T_n and H_{n+1} until the value of H_8 coincides with H_L , at which point A_T is the slurry heater approach temperature difference required to fit the example T-TE model constructed from the initial assumptions.

For the example case, the saturation temperature of vapor to the liquor heater is simply

$$T_8 = T_B - BPR_B \quad (22)$$

and the liquor heater approach temperature difference is set by the initial assumptions at

$$A_L = T_8 - D_8 \quad (23)$$

Cases with more than one liquor heating stage require construction of a liquor heater operating line of slope

$$x_D = \frac{D_{n+1} - T_L}{H_L} \quad (24)$$

where D_{n+1} is the assumed discharge temperature of liquor from the highest temperature liquor heating stage. Then another DO statement must be set up to calculate

$$T_n = T_L + A_L + x_D H_n \quad (25)$$

and

$$H_{n+1} = \frac{T_n - T_Z}{x_B} \quad (\text{same as for slurry heaters}) \quad (21)$$

starting with H_L and adjusting A_L until the calculated saturation temperature of vapor from the last flash stage (to the first liquor heater) coincides with $(T_B - BPR_B)$.

The generalized equations for a system of n slurry heating stages are obtained if substitutions are made of D_{n+1} for D_8 in equations 4, 9, 13 and 14, and BPR_{n+1} for BPR_8 in equation 17; solutions of equations 20 and 21 being made through H_{n+1} , which is forced to converge on H_L by adjustments to A_T . In all cases the heater discharge temperatures are given by

$$D_n = T_n - A_T \quad (26)$$

for slurry heating stages, or

$$D_n = T_n - A_L \quad (27)$$

for liquor heating stages. Where several liquor heating stages are used, the venting losses from the liquor heaters may be significant

enough to require adding an additional heat loss term to the calculation of H_L in equation 14.

With the calculation of heater discharge temperatures, the final corrections on the model limits can be made by calculating the unit condensate production and economy, using equations 2 and 3 as discussed in the graphical method section. The calculated economy value should then be substituted in equations 8 and 9, the entire process of solving the simultaneous heat balances and fitting the model to the number of available stages being repeated until desired agreement is reached between the condensate economy calculated from the results of equation 3 and the value used in equations 8 and 9. The resulting values of A_L and A_T are then the required liquor and slurry heater approach temperature differences theoretically required for the heat transfer rates needed to meet the specified temperature and flow conditions. A simplified logic diagram for the described computer program is given in Figure 5, which shows that results can be quickly obtained for a wide range of operating conditions by inputting values of Q_L , Q_S , Q_W , T_L , T_S , T_W , T_R , T_B , S_{XS} , S_I and D_{n+1} for each condition, if the number of slurry heating stages (n) and liquor heating stages, starting values for the T_m , D_1 , A_T , A_L and economy iterations, specific heat equations and BPR values are included in the program. The input value of D_{n+1} can be adjusted for each case as required to obtain a final value of A_L that compares with A_T as desired or as experienced in actual operation of the unit.

Analysis of Results

Care must always be taken when comparing the results of a series of computer model solutions to the performance of an operating digester unit, to avoid application of specified conditions that are physically impossible.

Effects of Blowoff Temperature

It should be kept in mind, for instance, that specification of both the blowoff temperature, T_B , and the excess flash vapor rate, S_{XS} , requires that sufficient control of pressure drop in the excess vapor line is available to maintain T_B in equilibrium with the barometric condenser discharge temperature resulting from condensation of vapor at rate S_{XS} . Should specified conditions exceed the operating range of this control, T_B and S_{XS} would actually reach the equilibrium values required by heat transfer rates in both the digester heaters and the barometric condenser of the unit, rather than the values specified in the model.

Such a system could be modeled by writing a third simultaneous heat balance around the barometric condenser, making T_B a variable in that equation and also in the heat balance equation for the flash heat reclaim system, but calculation of the approach between blowoff temperature and condenser discharge temperature would add considerable complication to the program. For the common case where unthrottled excess vapor is condensed in a cooling stream of relatively constant capacity, however, a simpler method of analysis is applicable. The approach temperature difference in

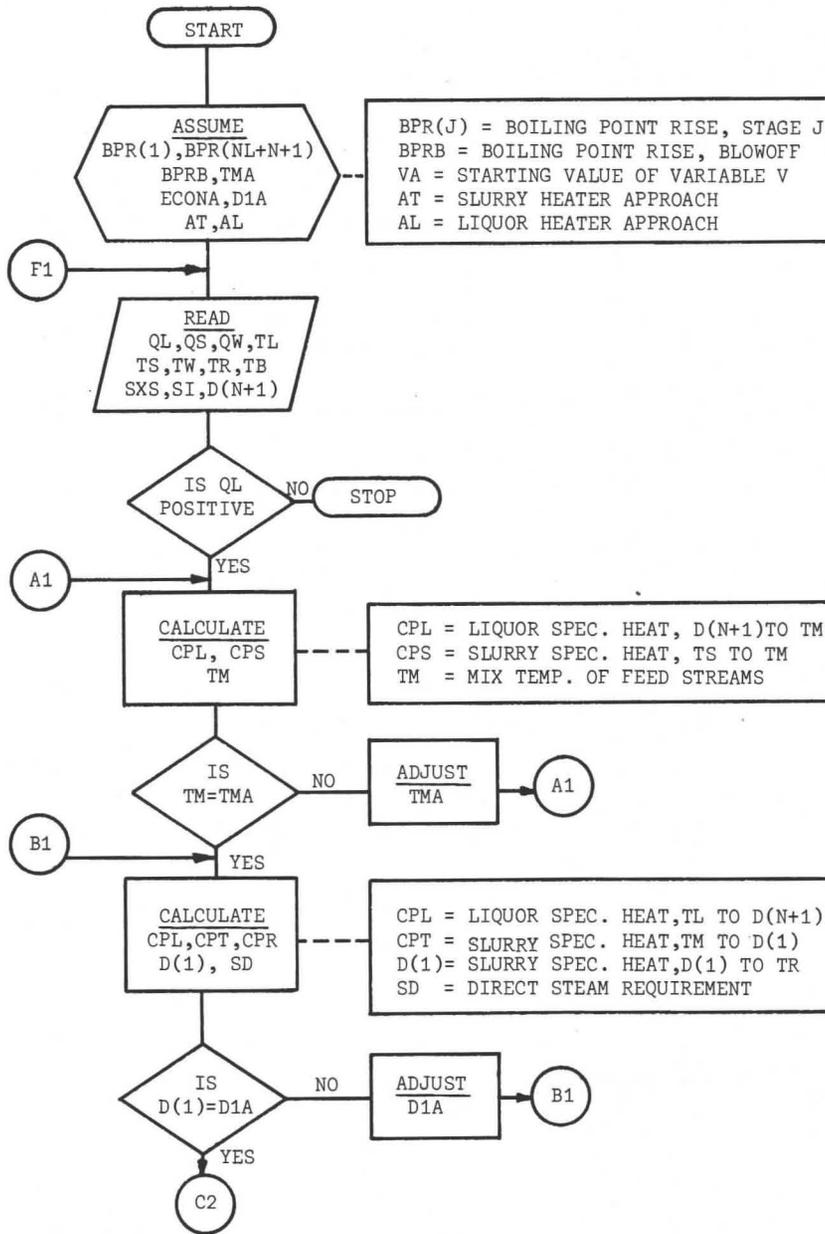


Fig. 5 - (First of two pages.)

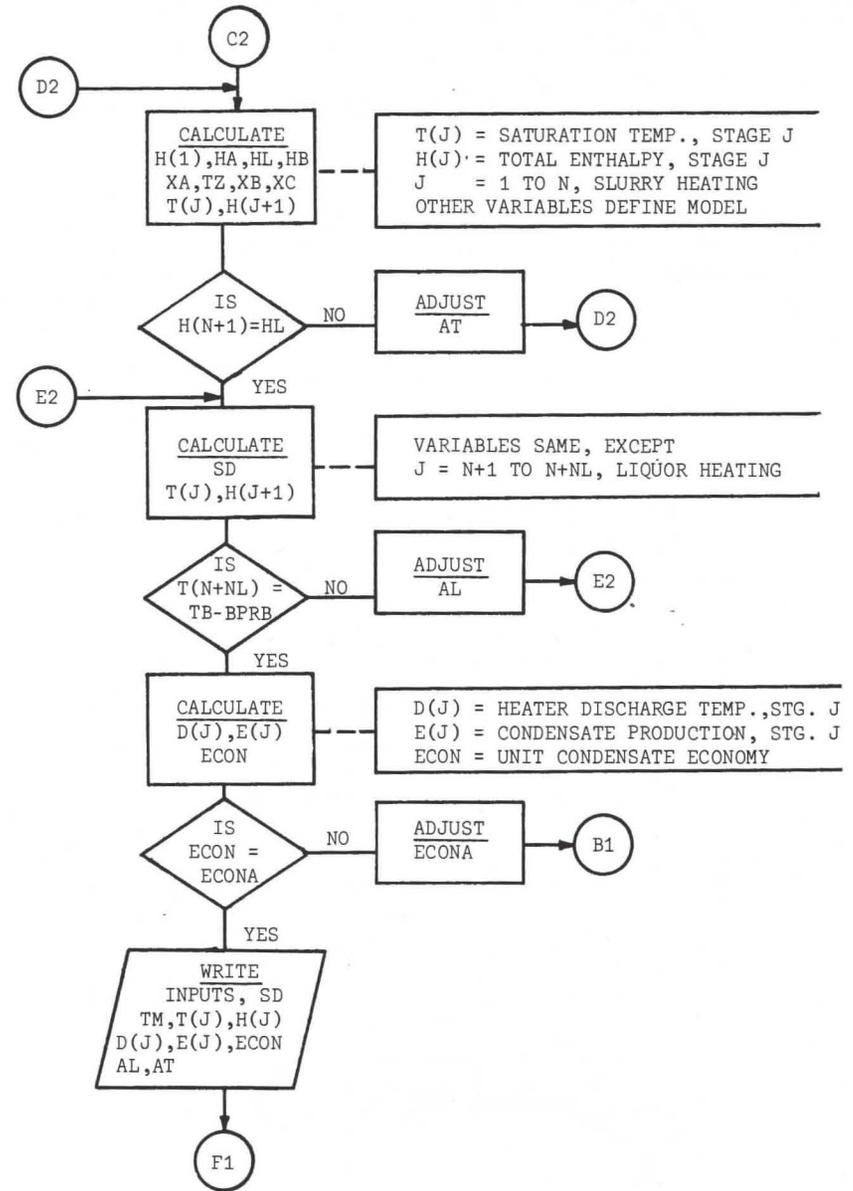


Fig. 5 - Logic diagram of computer program for construction and stagewise fit of temperature-total enthalpy model of flash heat reclaim in NL liquor heating stages and N slurry heating stages of a digester unit.

the condenser can then be assumed to change little, so that a change in condenser discharge temperature results in a nearly equal change in blowoff temperature, thus relating changes in T_B to changes in excess flash heat by the total heat capacity of the condenser cooling stream, or

$$\frac{(\Delta H_{XS})_2 - (\Delta H_{XS})_1}{(T_B)_2 - (T_B)_1} = (Q\bar{C}_P)_{\text{condenser}} \quad (27)$$

Blowoff temperature can also be related to excess flash heat for any system by fitting the computer model of the specified system at two cases of T_B , varying ΔH_{XS} until the same approximate heater approach temperature difference is calculated for each case. The corresponding two values of ΔH_{XS} are related to T_B by

$$\frac{(\Delta H_{XS})_1 - (\Delta H_{XS})_2}{(T_B)_2 - (T_B)_1} = \text{Constant} \quad (28)$$

which shows that if T_B can be lowered by reducing pressure drop in the excess vapor line, then the resulting increase in available heat is proportional to ΔT_B and will all be flashed to the barometric condenser, since no additional heat will be transferred with the same heater approaches. For systems fitting the assumptions of equation 27, the heat transfer rate in the reclaim heaters must be increased by the sum of the total heat capacity ratio of that equation plus the constant of equation 28, to lower the blowoff temperature by one degree, the equations being expressed in common units.

Effects of Indirect Steam Addition

When indirect steam is added to a flash stage, slightly more heat will be transferred at the same attained heater approaches than without this addition, due to the greater entering temperature differences in all stages. Since overall heat transfer coefficients, which can be calculated for each heating stage from the heat transfer rate of equation 2, are measures of heater performance for a particular set of flow and temperature conditions, it has been found useful to compare cases with and without indirect steam addition on the basis of equal overall coefficients, rather than equal heater approach temperature differences.

At first inspection, such comparisons for cases similar to the example have indicated that reductions in direct steam injection are less than the rate of indirect steam addition, and that the increased total steam requirement more than offsets potential benefits of the increased rate of net evaporation (defined as total evaporation minus direct steam injection) from the digester unit. Again, however, physical characteristics of the system must be considered. If steam heating of the barometric condenser discharge in addition to that from excess digester flash vapor is required, then reductions in this heating steam due to increased excess flash could make indirect steam addition desirable, if blowoff temperature were kept in control. The combined direct steam savings would then offset the cost of indirect steam, leaving any benefits of increased net evaporation as an advantage. Once

the excess vapor rate reaches the limit of throttling, however, further indirect steam heat will raise the blowoff temperature, progressively decreasing the available heat from cooling and reducing the evaporation advantage until it is overcome by the rising total steam cost.

Effects of Heat Transfer Rate

Finally, it should be recognized that improvements in heat transfer through heater cleaning, design or operation will benefit any digester unit flashing excess vapor to a condenser. Maximum benefits are realized when the excess vapor is unthrottled, while the least potential is present in systems where reduced flash vapor must be made up by external steam and the blowoff temperature cannot be allowed to drop further. All of these benefits are independent of and in addition to effects of indirect steam addition.

Conclusions

Although analysis of every flash heat reclaim system is unique in some respect, a number of characteristics have been observed with sufficient regularity to suggest general conclusions about optimal operation of a high pressure Bayer digester unit.

1. For maximum evaporation the blowoff temperature should be controlled as low as possible without upsetting other operations in the process.
2. The above condition is especially beneficial in systems where the excess flash vapor is augmented by external heating steam, in which cases it is also advantageous to add indirect steam to the flash system.
3. Computer models of a flash heat reclaim system can be quickly solved for heat transfer requirements at various conditions, to determine feasible heater performance goals at any specified feed rates.
4. Heater performance improvements are most beneficial to systems with unthrottled excess flash vapor, a situation where indirect steam addition will raise the equilibrium blowoff temperature while increased heat transfer will lower it.

Nomenclature

A_i	Average heater approach temperature difference for pre-heaters of feed stream i
BPR_n	Boiling point rise of slurry entering flash stage n
$(\bar{C}_P)_i$	Average specific heat (at constant average pressure) of stream i in the temperature range of the heat term containing it
D_n	Discharge temperature of liquor or slurry from heater condensing vapor from flash stage n

E_n	Evaporation rate calculated from temperature gain in heater condensing vapor from flash stage n
h_{fi}	Specific enthalpy of saturated condensate at temperature of stream i (from Steam Table)
h_{gi}	Specific enthalpy of steam at pressure and temperature of stream i (from Steam Table)
H_n	Total enthalpy of liquor or slurry discharging heater condensing vapor from flash stage n, above enthalpy of feed liquor equal zero
ΔH_{f1}	Estimated heat loss by radiation and convection from flash tanks and vapor lines
ΔH_{htr}	Estimated heat loss by radiation, convection and venting from feed preheaters
ΔH_{loss}	Estimated heat loss by radiation and convection from digester vessels
Q_i	Mass flow rate of liquid, slurry or alumina stream i
S_i	Mass flow rate of steam or vapor stream i
T_i	Temperature of liquid or slurry stream i
T_m	Mix temperature of feed slurry and discharge from liquor heating stage(s)
T_n	Temperature of saturated condensing vapor from flash stage n

Subscripts (i)

ABEA	refers to American Bayer extractable alumina, the Gibbsite, or alumina trihydrate, fraction of dry bauxite in feed slurry
B	refers to slurry blowoff from last stage flash tank of digester unit
C	refers to total condensate discharging heater condensing vapor from last flash stage
D	refers to steam directly injected into digester vessels
I	refers to indirect steam added to flash stage(s)
L	refers to caustic liquor feed to digester unit
MEA	refers to maximum extractable alumina, the sum of Gibbsite and Boehmite, or alumina trihydrate plus alumina monohydrate, fractions of dry bauxite in feed slurry

R	refers to reacted slurry and condensed steam discharging digester vessels
S	refers to bauxite slurry feed to digester unit
T	refers to combined liquor and slurry feed to digester vessels
W	refers to dilution stream added to last stage flash tank
XS	refers to flash vapor in excess of capacity of heaters to condense it, discharging last flash stage to barometric condenser

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2. See ref. 1, p. 57.