A THIELE-GEDDES
EQUILIBRIUM STAGE
CALCULATION

BY

RONALD CSUHA

A THESIS
SUBMITTED TO THE FACULTY
IN PARTIAL FULFILLMENT OF THE
REQUIREMENTS FOR THE DEGREE OF
MASTER OF SCIENCE

Houston, Texas
Dec., 1964
"Shut not your doors to me proud libraries,
For that which was lacking on all your well-fill'd shelves,
yet needed most, I bring,
Forth from the war emerging, a book I have made,
The words of my book nothing, the drift of it every thing,
A book separate, not link'd with the rest nor felt by the intellect,
But you ye untold latencies will thrill to every page."

Walt Whitman
# TABLE OF CONTENTS

<table>
<thead>
<tr>
<th>Section</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>Abstract</td>
<td>1</td>
</tr>
<tr>
<td>Introduction</td>
<td>3</td>
</tr>
<tr>
<td>The Problem</td>
<td>6</td>
</tr>
<tr>
<td>The Equations</td>
<td>10</td>
</tr>
<tr>
<td>The Solution of the Material Balance Equations</td>
<td>12</td>
</tr>
<tr>
<td>The Solution of the Enthalpy Balance Equations</td>
<td>17</td>
</tr>
<tr>
<td>Results</td>
<td>21</td>
</tr>
<tr>
<td>Round-off error in the absorber-stripper calculations</td>
<td>21</td>
</tr>
<tr>
<td>Absorber and Stripper Problems and Results</td>
<td>30</td>
</tr>
<tr>
<td>Distillation Calculations</td>
<td>37</td>
</tr>
<tr>
<td>Conclusions</td>
<td>39</td>
</tr>
<tr>
<td>Nomenclature</td>
<td>41</td>
</tr>
<tr>
<td>Bibliography</td>
<td>43</td>
</tr>
<tr>
<td>Appendix</td>
<td></td>
</tr>
<tr>
<td>Equilibrium and Enthalpy Data</td>
<td>1</td>
</tr>
<tr>
<td>Problem Data</td>
<td>v</td>
</tr>
<tr>
<td>Problem Solutions</td>
<td>vii</td>
</tr>
</tbody>
</table>
ABSTRACT

The Thiele-Geddes method as developed by Holland (ref. 3) uses dew and bubble point calculations to determine new stage temperatures for each succeeding iteration. The method to be considered here involves the use of enthalpy balances to determine these temperatures. This way of determining stage temperatures is more applicable for wide boiling range mixtures, while that used by Holland is more applicable for a narrow boiling range mixture.

The proposed method was utilized for a distillation column (employing a relatively narrow boiling range feed) and for both absorbers and strippers (employing relatively wide boiling range feeds). The absorber and stripper problems yielded good results; however, in a problem with wide boiling mixtures, the round-off error which arises in the solution of the material balance equations becomes significant. This error was effectively eliminated by performing a top-down calculation for the light components, and a bottom-up calculation for the heavy components. The decision as to when a component is heavy enough to warrant a bottom-up calculation is to be based on the magnitude of the absorption factor \( A_{ij} = \frac{L_j}{V_j K_{ij}} \).
For an $A_{ij}$ (averaged over all the equilibrium stages) less than or equal to unity, the top-down calculation is to be used; and for an $A_{ij}$ avg. greater than or equal to unity the bottom-up calculation is to be used.

Convergence could not be attained for the distillation calculation though several modifications were tried. The use of a wide boiling range feed for the distillation calculation was not used since a major change in the basic procedure would be required to eliminate the round-off error.
INTRODUCTION

There are many approaches to multicomponent separation problems. However, the rigorous approaches may, in general, be classified into three main groups -- the classification being based on the method of handling the material balance equations. For example, among the three classifications:

1. the Relaxation method (ref. 8),
2. the Lewis-Matheson method (ref. 4), and
3. the Thiele-Geddes method (ref. 11);

the Relaxation method is characterized by the fact that it utilizes a time dependent material balance; the Lewis-Matheson method — a straight difference equation without any substantial alteration in the form; and the Thiele-Geddes — a modified form of the difference equations to incorporate ratios of flow rates instead of the flow rates themselves.

With each method is associated advantages and disadvantages. Thus, while the Relaxation method follows the actual path of a distillation column from start-up, it uses too much time to effect a steady state solution. The Lewis-Matheson method, while using fewer iterations, needs initial
assumptions of column variables and, in addition, the build-up of round-off error may become significant. The Thiele-Geddes method minimizes the round-off error and uses fewer initial assumptions than the Lewis-Matheson method, but is more complex for columns with side streams. Within each method there are also modifications which may prove more advantageous than others. These modifications differ in that some affect the way in which the material balance equations are solved (for example, a matrix solution or a plate to plate calculation), and some affect the way in which the remaining restrictions are placed on the system.

The total number of restrictions to be placed on the separation system consist of the equations of the mathematical model plus the number of independent specifications. The equations of a steady state mathematical model consist of four types:

1. material balance equations for each component,
2. phase equilibrium relationships for each component,
3. enthalpy balance equations, and
4. identity relationships for the mole fractions.

A transient model would also contain a rate equation, but the rate of approach to equilibrium usually is not considered.
For the present paper, it is sufficient only to work with the steady state equations of the mathematical model. The material balance equations are normally combined with the equilibrium relationships to yield values for the individual component flow rates (or compositions). Once these are determined there remain two choices:

1. to use the compositions in a dew or bubble point calculation to determine new stage temperatures; and then to determine new values of overall flow rates by means of the enthalpy balances; or

2. to use the individual component flow rates in an enthalpy balance to determine new temperatures; and then to use the identity relationships to evaluate the new flow rates.

The use of each procedure to reevaluate initial assumptions has been discussed by Friday (ref. 2). For the case of a narrow boiling range mixture, the temperature changes that can take place are limited by the nature of the mixture; and the enthalpy of a phase of such a mixture becomes a function of the amount of that phase. The enthalpy balances would then be insensitive towards the determination of stage temperatures; and should be used to determine the flow rates.
Procedure (1) is therefore more applicable for a narrow boiling range mixture. For a wide boiling range mixture, the phase flow rates are essentially fixed. The enthalpy of a phase is then dependent on the temperature of the phase rather than the amount. The enthalpy balances then become insensitive towards the determination of the flow rates, and should be used to determine temperatures. Procedure (2) would therefore be used for wide boiling range mixtures.

The procedure (1) combined with the Thiele-Geddes method has been developed by Holland (ref. 3), and has the disadvantage of giving poor convergence for wide boiling range mixtures. The purpose of this paper is to propose the use of the procedure (2) with the Thiele-Geddes method. The proposed method, while having the advantage of converging for a wide boiling range mixture, has the disadvantage of giving poor convergence for a narrow boiling mixture.
THE PROBLEM

The problem considered in this paper and defined above is to be applied to the distillation column of figure 1 and the absorber (or stripper) of figure 2. In order to solve the mathematical model, a certain number of independent specifications must be applied. These arise out of the consideration of the number of equations and the number of unknowns, the excess of unknowns over the equations being the number of independent specifications to be placed on the system. An analysis of this kind has been made for a number of systems by Smith (ref. 9), and another analysis based on the phase rule has been utilized by Maddox (ref. 5) to achieve the same results. The results of such an analysis do not provide for the exact specifications which must be used, but only yield the number of such specifications. The choice of which specifications to utilize is open.

In the work presented here, the following specifications were used:

1. Composition, flow rate, and thermal condition of the feed,
2. Heat loss or gain from each plate,
3. Pressure on each plate,
4. Number of equilibrium stages in the column, and
5. The feed plate location.

In addition, the distillation column requires the extra specifications:

6. Reflux ratio \((L/Q/D)\),
7. Rate of flow of one product stream (or a specified number of components in that stream),
8. The condenser or reboiler heat load (for a total condenser, the temperature of the distillate will essentially fix the condenser heat load, and this together with specifications 1 and 2 will fix the reboiler heat load), and
9. Pressure of the reboiler and condenser.

Also, for the absorber or stripper there are two feeds (the oil and the gas input), and specification 1 is to apply to both. Other specifications have been used by various authors, but only the ones mentioned above will be considered here.

Closely allied with the specifications are the assumptions used, since they are used to eliminate a certain number of the specifications. For example, in almost all the rigorous separation calculations, there is assumed perfect equilibrium on each plate. If there is no equilibrium on a plate,
additional specifications for the deviation from equilibrium are needed for each plate. Similarly, the assumption of an adiabatic column (used in this paper) corresponds to specification 2, and the specification 3 has been handled by assuming the pressure on each plate equal to the total column pressure.

Equations:

The material balance equations as used by Holland (ref. 3), and those that will be used for this paper, are associated with the Thiele-Geddes method for distillation calculations. The equations are obtained by combining the equilibrium relationships:

$$A_{ij} = \frac{L_j}{V_j} \frac{K_{ij}}{i_{ij}} = \frac{i_{ij}}{v_{ij}} \text{ and } S_{ij} = \frac{1}{A_{ij}} \quad (1)$$

with the material balances. They are then manipulated to yield:

A. For distillation (ref. 3a):

1. rectifying section

A material balance about section I, of figure 1, yields:

-10-

[Image 0x0 to 612x792]
\[ \frac{v_{11}}{d_1} = A_{10} + 1 \quad \text{(2a)} \]

\[ \frac{v_{1,j+1}}{d_1} = A_{1j}(\frac{v_{1j}}{d_1}) + 1 \quad (j = 1, 2, \ldots, f-2) \quad \text{(2b)} \]

where: \[ A_{10} = \frac{L_0}{DK_{10}} \quad \text{(for a partial condenser)} \]
\[ = \frac{L_0}{D} \quad \text{(for a total condenser)} \]

2. stripping section

A material balance about section II, of figure 1, yields:

\[ \frac{l_{1,M-1}}{b_1} = S_{1M} + 1 \quad \text{(3a)} \]

\[ \frac{l_{1,k-1}}{b_1} = S_{1k}(\frac{l_{1k}}{b_1}) + 1 \quad (k = M, M-1, \ldots, f+1) \quad \text{(3b)} \]

B. For absorbing and stripping (ref. 3b)

1. top-down calculation

A material balance about section I, of figure 2, yields:

\[ \frac{v_{11}}{v_{11}} = 1 \quad \text{(4a)} \]

\[ \frac{v_{1,j+1}}{v_{11}} = A_{1j}(\frac{v_{1j}}{v_{11}}) + (1 - (\frac{l_{10}}{v_{11}})) \]
\[ (j = 1, 2, \ldots, M) \quad \text{(4b)} \]
2. bottom-up calculation

A material balance about section II, of figure 2, yields:

\[ \frac{l_{1M}}{l_{1M}} = 1 \]  \hspace{1cm} (5a)

\[ \frac{l_{i,j-1}}{l_{1M}} = S_{ij}(\frac{l_{ij}}{l_{1M}}) + (1 - (\frac{l_{i,M+1}}{l_{1M}})) \]

\( (j = M,M-1, \ldots ,1) \)  \hspace{1cm} (5b)

Solution of the material balance equations:

The problem is to solve the (b) equations with the terminal conditions (a). The actual procedure for solving the equations for distillation, case (A), differs from that for absorbing and stripping, case (B); in that both sets of equations (1 and 2) are to be used for the distillation, and either set (1 or 2) may be used for absorbing and stripping. The general procedure to be used is an iteration technique which involves:

1. the assumption of a set of solution variables \((T_j,V_j)\),

2. the calculation of \(L_j\) from an overall material balance for each plate, and \(K_{ij}\) from its functional relationship in terms of \(T_j\),
iii the calculation of the $A_{ij}$'s or $S_{ij}$'s and the solution of the material balance relationships to determine the individual component vapor and liquid flow rates,

iv the application of the enthalpy balances for each plate to determine the temperature on each plate,

v the application of the equality relationships:

$$L_j = \sum_{i=1}^{N} l_{ij}, \quad \text{and} \quad V_j = \sum_{i=1}^{N} v_{ij} \quad (6)$$

to determine new values for the total flow rates on each plate,

vi the recalculation of the $A_{ij}$'s or $S_{ij}$'s, and the iteration of steps iii through vi until convergence is attained.

For the distillation procedure, equations (2) are applied from the top plate to the plate above the feed plate, and equations (3) are applied from the bottom plate to the feed plate. This yields values for:

$$l_{i,f-1/d_i} \quad \text{(from the enriching section), and}$$

$$v_{if/b_i} \quad \text{(from the stripping section).}$$

From these and a material balance about the feed plate, a
ratio (ref. 3a):

\[
\frac{b_i}{d_i} = \frac{(l_{i,f-1}/d_i) + (l_{i,F}/F_{i,F})}{(v_{i,F}/b_i) + (v_{i,F}/F_{i,F})}
\]

(7)

may be determined. Then a component material balance about the column yields values for the distillate component flow rates:

\[
d_i = F_{i,F}/(1 + (b_i/d_i))
\]

(8)

However, since the ratios of \((b_i/d_i)\) have been determined from assumed values of \(A_{ij}'s\) and \(S_{ij}'s\), the ratios are not correct, and the product specification (specification 9) cannot be met by merely summing the distillate flow rates. This is corrected for by applying the theta method of convergence (ref. 3a):

\[
\sum_{i=1}^{s} d_i = D = \sum_{i=1}^{s} \left[ F_{i,F}/(1 + \Theta(b_i/d_i)) \right]
\]

\[
1 \leq s \leq N
\]

(9)

where \(\Theta\) is merely a correction factor to force the distillate components (8) to meet the specification of equation (9). The procedure is then to determine a value of \(\Theta\) to satisfy (9), and then to determine all the individual distillate component
flow rates:

\[ d_i = \frac{FX_{iF}}{(1 + \Theta(b_i/d_i))} \quad (i=1,2, \ldots, N) \quad (10) \]

and bottoms flow rates:

\[ b_i = FX_{iF} - d_i \quad (11) \]

The component flow rates on each plate are now determined from the previously determined ratios, \((v_{ij}/d_i)\)'s and \((l_{ij}/b_i)\)'s, by multiplying through by the corrected \(d_i\)'s and \(b_i\)'s respectively. The equilibrium relationships can then be applied to evaluate the remaining flow rates.

The procedure for the absorber is the same as that for the stripper. The discussion here, however, will not include the provision for non-distributed components such as steam in a steam stripper. Either of the equations (4) or (5) may be used in the plate to plate calculations, and the only restriction that may be laid on their use arises out of the consideration of the round-off error. This subject of round-off error will be considered in a later section. For the present consider only the theoretical aspects of solving the equations.

In order to use equations (4) or (5), it is necessary to evaluate difference terms:
This may be done by repeated substitutions and manipulation with the material balances to yield (ref. 3b):

\[
(1 - \frac{l_{10}}{V_{11}}) = \frac{v_{i,M+1} - \omega_1}{v_{i,M+1} + \Omega_1 l_{10}} \quad (12a)
\]

where

\[
\omega_1 = A_{iM} A_{i,M-1} \ldots A_{i2} A_{i1} \quad (12b)
\]

\[
\Omega_1 = 1 + A_{iM} + A_{iM} A_{i,M-1} + \ldots \ldots \ldots
\]

\[
\ldots + A_{iM} A_{i,M-1} \ldots A_{13} A_{12} \quad (12c)
\]

and

\[
(1 - \frac{v_{i,M+1}}{l_{1M}}) = \frac{l_{10} - \phi_1 v_{i,M+1}}{l_{10} + \phi_1 v_{i,M+1}} \quad (13a)
\]

where

\[
\phi_1 = S_{i1} S_{i2} \ldots \ldots S_{i,M-1} S_{i M} \quad (13b)
\]

\[
\Phi_1 = 1 + S_{i1} + S_{i1} S_{i2} + \ldots \ldots \ldots
\]

\[
\ldots + S_{i1} S_{i2} \ldots \ldots S_{i,M-2} S_{i,M-1} \quad (13c)
\]

Equations (4) may be applied with (12) to yield values of \(v_{ij}/V_{11}\), and these may be converted to values of \(l_{ij}/V_{11}\)
by applying the equilibrium relationships. Similarly, equations (5) may be applied with (13) to yield values of \( \frac{l_{1j}}{l_{1M}} \), and these may be converted to values of \( \frac{v_{1j}}{l_{1M}} \). Applying a material balance about the column:

\[
v_{11} = \frac{v_{1,M+1} - l_{1o}}{1 + (l_{1M}/v_{11})} \quad (14)
\]

\[
l_{1M} = \frac{v_{1,M+1} - l_{1o}}{1 + (v_{11}/l_{1M})} \quad (15)
\]

Therefore, in order to obtain a complete set of vapor and liquid flow rates for all components, the set of equations (4), (12), and (14) may be used for a top-down calculation; or, likewise, the set (5), (13), and (15) may be used for a bottom-up calculation.

Solution of the enthalpy balance equations:

If the heat of solution is neglected, the enthalpy balance for the plate \( j \) may be written:

\[
\sum_{i=1}^{N} v_{i,j+1} h_{i,j+1} - \sum_{i=1}^{N} l_{1j} h_{1j} + Q_j = \sum_{i=1}^{N} v_{i,j} h_{i,j} - \sum_{i=1}^{N} l_{1,j-1} h_{i,j-1} \quad (16)
\]
where \( Q_j \) is the external heat load applied to the plate \( j \) (Btu/hr),

\( h_{ij} \) is a component vapor enthalpy on the plate \( j \) (Btu/lb-mole), and

\( h_{ij} \) is a component liquid enthalpy on the plate \( j \) (Btu/lb-mole).

It is also desirable to represent the enthalpy dependency as a linear function of the temperature. Thus:

\[
H_{ij} = a_{ij}^v + b_{ij}^v T_j
\]

(17)

\[
h_{ij} = a_{ij}^l + b_{ij}^l T_j
\]

Other schemes may be used for representing the enthalpy, however (17) is the most convenient. Equation (16) may now be rearranged to give:

\[
T_j = \left[ \sum_{i=1}^{N} \frac{l_i}{(b_{ij}^v v_{ij} + b_{ij}^l l_{ij})} \right] \left[ \sum_{i=1}^{N} (v_{ij,j+1} h_{ij,j+1} + l_{ij,j-1} h_{ij,j-1}) \right]
- \sum_{i=1}^{N} (v_{ij} a_{ij}^v + l_{ij} a_{ij}^l)
\]

(18)

Since the \( v_{ij} \)'s and the \( l_{ij} \)'s have been determined from the material balance equations and since the enthalpy polynomial coefficients are known for each component, the only
remaining variables in (18) are the enthalpies $H_{i,j+1}$ and $h_{i,j-1}$; and these are known functions of $T_{j+1}$ and $T_{j-1}$ respectively. The procedure would then be to start with a known temperature for either $T_{j+1}$ or $T_{j-1}$ and to proceed step by step through the column. Thus, if $T_1$ is known, $T_2$ can be approximated from the previously assumed value of $T_3$. With the new value of $T_2$ and the old value of $T_4$, a new value of $T_3$ may be calculated. The process is then repeated step by step down the column. If a terminal temperature $T_M$ is known at the bottom of the column; $T_{M-1}$, $T_{M-2}$, etc., may be determined in a similar manner by proceeding up the column.

There is also the possibility of iterating on the enthalpy balances before proceeding to the next iteration on the material balances. Thus, once a set of $T_j$'s have been calculated from the old set of $T_j$'s, a third set of $T_j$'s may be calculated in the same manner from this second set of $T_j$'s. Likewise, a fourth set may be calculated, etc. There is, however, some question as to the worth of this type of iteration since the iteration may not be successful; and if it is successful, the final iterated values will have about the same order of error associated with them as that of the error of the non-iterated values. Iteration would probably be most useful towards the end on the calculation, when convergence is almost attained. In the calculations presented, the non-iterated temperatures were used.
The actual scheme of the calculation may vary. For the distillation column, \( T_0 \) is known in the case of the total condenser or may be calculated from an enthalpy balance around a partial condenser. This, therefore, represents the starting point of the calculation. \( T_M \) may be calculated from a balance around the column (excluding the reboiler). There are now several choices which may be made in selecting the direction of the calculation.

1. Calculate from the top plate to the feed plate and then from the bottom plate to the feed plate, or vice-versa.
2. Calculate from the top plate to the bottom plate.
3. Calculate from the bottom plate to the top plate.

For the absorber calculations \( T_0 \) and \( T_{M+1} \) are known and since there are no side feed streams, only the last two choices are available.

All three methods are numerically inequivalent. However, if iteration is carried out, and if the iteration is successful, all three methods would then become equivalent. Another method that would be equal to the iterated case involves the use of a matrix equation. Sujata (ref. 10) obtains an enthalpy function from (16) for each plate, and uses Newton's method together with matrix manipulations to obtain all the temper-
ature changes simultaneously. Another matrix method has been developed by Newman (ref. 7). Neither of the refined methods were used here. The absorber enthalpy balances were applied from the top of the column to the bottom and all three orders of calculation were applied to the distillation column.

RESULTS

Absorber and Stripper Calculations:

The calculations were performed with both sets of equations (4) and (5). Each set is mathematically correct, but the numerical results of their solution sometime differ. This can only be the effect of round-off error, of which an excellent illustration is given by Holland (réf. 3c). For the absorber calculation, there are three possible sources of this error. One such source is the initial starting ratio (4a) or (5a). However, this ratio is 1.000; and since the exact number is used by the computer, no error is introduced. This leaves two remaining sources, the multiplication by the $A_{ij}$'s or $S_{ij}$'s and the addition of the term (12a) in the case
of (4) or the term (13a) in the case of (5). The accuracy
of (12a) or (13a) is good, but the terms may cause a serious
build-up of error. The repeated process of multiplication
and addition may also produce an error which grows with
successive calculations. These two sources of error will
be discussed under the classification of round-off error,
even though the first case of error propagation is not a
true round-off error. It was found that, in the top-down
calculations, equations (4) gave negligible round-off error
for the light components, but gave unrealistic flow rates
for the heavy components. The opposite was true for equations
(5).

If the equations for the calculation are examined, it
will be found that the only parameters that affect the cal¬
culation are the $A_{ij}$'s on each plate, the number of plates,
and the ratio:

$$R_i = \frac{1_{10}}{v_{i,M+1}}$$

The round-off error must then be a function of these same
parameters. In order to examine this error more closely,
the $A_{ij}$'s for a particular component were considered inde¬
dependent of the stage number. That is:

$$A_{ij} = A_{i,j+1} = A_{i} \quad (j = 1, 2, \ldots, M-1)$$

The calculation (4) was then performed for values of $A_i$ and
$R_i$; and
the calculated ratios, \( v_{1M}/v_{1l} \), thus obtained were compared with those obtained from a formula derived by Holland (ref. 3b):

\[
\frac{l_{1M}}{v_{1l}} = \frac{\omega_{1l_{10}} + (\Omega_1 + \omega_1 - 1)v_{1,M+1}}{\Omega_{1l_{10}} + v_{1,M+1}} = \frac{\omega_1 R_1 + (\Omega_1 + \omega_1 - 1)}{\Omega_1 R_1 + 1} \tag{19a}
\]

\[
\frac{v_{1M}}{v_{1l}} = S_{1M} \frac{l_{1M}}{v_{1l}} \tag{19b}
\]

The comparison was effected by determining a relative error:

\[
e = \frac{|(v_{1M}/v_{1l}) \text{ from eq. (4b)} - (v_{1M}/v_{1l}) \text{ from eq. (19b)}|}{(v_{1M}/v_{1l}) \text{ from eq. (19b)}}
\]

for fixed values of \( A_1 \) and \( R_1 \), and for a total of 10 stages. The results are presented in figure 3, where lines of constant error are plotted in the \( A_1-R_1 \) plane.

For values of \( A_1 \) less than unity, the round-off error occurred in the term calculated from (11) and the ratios (19). The plate to plate calculation was accurate, and the effect of round-off error was taken to be negligible -- being of the order of \( 10^{-6} \).
The bottom-up calculation (5) is, by symmetry, essentially the same as the top-down calculation (4). This may be easily shown by considering the conditions under which the two calculations become numerically equal:

\[
\frac{v_{11}/v_{ll}}{(4)} = \frac{l_{1M}/l_{1M}}{(5)} \quad (20a)
\]

\[
A_i^{(4)} = S_i^{(5)} \quad (20b)
\]

\[
(1-(l_{10}/v_{ll}))^{(4)} = (1-(v_{i,M+1}/l_{1M}))^{(5)} \quad (20c)
\]

Here the superscript refers to the set of equations to be used. Thus for (20a) the value of \(v_{11}/v_{ll}\) to be used with (4) must be numerically equal to the value of \(l_{1M}/l_{1M}\) to be used in (5). This condition is obviously met since both ratios are equal to unity. Condition (20b) can be set, and for condition (20c) to hold:

\[
R_i^{(4)} = (1/R_i)^{(5)}
\]

\[
= R_i' \quad (5)
\]

Thus, if the top-down calculation has been performed for values of \(A_i\) and \(R_i\), there is no need to perform the bottom-up calculation since that calculation performed with the reciprocals of \(A_i\) and \(R_i\) (that is, \(S_i\) and \(R_i'\)) would yield
the same results. A plot of $S_1$ versus $R_1$ for lines of constant error would look exactly like the plot of figure 3.

It is interesting to note that the effect of the round-off error is not random, but is a complex function of the input variables and the starting conditions. With the exception of a certain number of irregularities (regions or points of low or high error), it is quite predictable. The irregularities themselves seem to repeat for a given $A_1$. Thus, there are four irregularities (within the given range) at an $A_1$ of about 2.5, and three at an $A_1$ of about 8.0.

The primary purpose of the examination of the round-off error was to try to develop some criteria for the decision of which calculation to use (the top-down or the bottom-up calculation). From the preceding results, it can be decided to perform the calculation (4) for values of $A_1$ less than or equal to unity, and to perform the calculation (5) for values of $A_1$ greater than or equal to unity. The $A_{ij}$'s for a particular component, however, will not be the same on all plates. But since they will be of the same order of magnitude, and average $A_1$:

$$A_1 \text{ avg} = \left( A_{i1} A_{i2} \cdots \cdots A_{iM} \right)^{(1/M)}$$

or

$$A_1 \text{ avg} = \frac{A_{i1} + A_{i2} + \cdots + A_{iM}}{M}$$
should yield satisfactory results when applying the criteria.

The procedure for the absorber or stripper would then be modified in the following manner:

1. calculate all the $A_{ij}$'s and apply the criteria to choose the order of the calculation,
2. for the light components (4) will be used and for the heavy components (5) will be used,
3. the values of $l_{ij}/l_{iM}$ can be converted to values of $v_{ij}/v_{i1}$ by applying (19a) and the equilibrium relations,
4. with the complete set of $v_{ij}/v_{i1}$'s, (14) may be applied to determine values for $v_{i1}$, and
5. the remaining calculation will then be the same as discussed previously.

The effect of the round-off error on the absorber and stripper calculations is shown in figures 4 and 5. The absorber results are shown for absorber feed 2 (AF2) and for values of $L_0/V_{M+1}$ of 0.05 and 0.50. The stripper results are shown for stripper feed 1 (SFI) and for values of $L_0/V_{M+1}$ of 0.50 and 5.00. Absorber results for $L_0/V_{M+1}$ of 5.00 and stripper results for $L_0/V_{M+1}$ of 0.05 yielded very little round-off error effect.
FIG. 4

THE EFFECT OF ROUND-OFF ERROR ON THE ABSORBER SOLUTIONS FOR AF2

- soln. with round-off error
- soln. without round-off error
FIG. 5

THE EFFECT OF ROUND-OFF ERROR ON THE STRIPPER SOLUTIONS FOR SF1

- soln. with round-off error
- soln. without round-off error
Absorber Problems:

The absorber problems were made up of 4 gas feeds. The numerical data on the feeds are included in the appendix together with the inlet temperatures used. Here, it will be sufficient to say that:

AFla is a rich gas
AFlb is a rich gas with a lower concentration of heavy components than AFla
AF2 is a medium between a rich and a lean gas
AF3 is a lean gas

The lean oil was taken to be pure nC₈ with all four gas feeds. In addition, a C₁ saturated oil was used with AF2. Three different rates of $L_o/V_{M+1}$ were used, these being 5.00, 0.50, and 0.05. The pressure of the column was taken to be 264.7 psia.

Stripper Problems:

The same procedure was followed for the stripper problem as for the absorber problem. Here the various feeds are:
SF1 is a rich oil
SF2 is a medium between a rich and a lean oil
SF3a is a lean oil
SF3b is a lean oil with a higher concentration of light components than SF3a

The numerical data on these feeds together with the inlet temperatures used are given in the appendix. The stripping gas was taken to be steam at 600°F and 50 psia. The column pressure was also taken to be 50 psia. The same three values of $L_0/V_{M+1}$ were used for the stripper as for the absorber.

The problem was then solved for three types of widely differing feeds and three values of widely differing $L_0/V_{M+1}$. The convergence of the different problems is shown in figures 6, 7, and 8, where the methane vapor rate for the top plate is plotted against the trial number. The absorber solutions for the two lower values of $L_0/V_{M+1}$ were quite stable while those for the $L_0/V_{M+1}$ of 5.00 exhibited some instability. The instability increased with the richness of the feed gas, so that while the lean gas exhibited the same type of oscillations as the other feeds, these same oscillations decreased in amplitude. The oscillations for the richer gases increased in amplitude. In all the absorber problems the oscillations could be damped by averaging the successive trial values of
FIG. 6

CONVERGENCE OF ABSORBER SOLUTIONS

$\frac{L_0}{V_{M+1}} = 0.05$

$\frac{L_0}{V_{M+1}} = 0.50$

$V_{IL}$ (methane vapor rate on top plate)

TRIAL NO.
FIG. 7

CONVERGENCE OF STRIPPER SOLUTIONS

$V_{11}$ (methane vapor rate on top plate)

TRIAL NO.

$L_0/V_{M+1} = 5.0$

$L_0/V_{M+1} = 0.5$

SF2
SF3a
SF1

SF2
SF3a
SF1
FIG. 8

CONVERGENCE OF THE ABSORBER PROBLEMS WITH $L_0/V_{M+1} = 5.00$ AND FOR THE STRIPPER PROBLEMS WITH $L_0/V_{M+1} = 0.05$
L, V, and T. The procedure, involving the averaging, constitutes a separate iteration procedure that is not the same as the non-averaged procedure. When an iteration produces a "calculated" set of L, V, and T for each plate; these values are averaged with the previously assumed values of L, V, and T. These averaged values of L, V, and T would then be used to calculate another set of L, V, and T for each plate. These, in turn, are averaged with the previously assumed values (previously averaged values) of L, V, and T. The iteration is thus carried out with a resulting dampening of any oscillations that occur in the calculation. The dotted lines of figure 8 represent the results of this averaging procedure for the absorber problems noted.

The same situation was encountered for the stripper solutions. Here the instability was encountered for the lowest value of $\frac{L_0}{V_{M+1}}$, and was about the same order of magnitude for all three ranges of feeds. In all the stripper problems, the oscillations decreased in amplitude. In order to illustrate the reciprocal relationship between the stripper and absorber problems, these results are plotted in the same figure 8 that the corresponding absorber results are plotted.

The problem solutions are presented in figures 9 through 18. T and V were selected as the variables to be plotted as
functions of the stage number. All the solutions had the same
general shapes for the profiles. Thus, for the absorber solu-
tions, the temperature profiles for:

\[ \frac{L_0}{V_{M+1}} = 0.05 \] had a generally negative slope,
\[ = 0.50 \] had a generally positive slope, and
\[ = 5.00 \] had a generally zero slope;

and for the stripper solutions, the temperature profiles for:

\[ \frac{L_0}{V_{M+1}} = 0.05 \] had a generally negative slope with a
positive second derivative,
\[ = 0.50 \] had a generally negative slope with an
inflection point, and
\[ = 5.00 \] had a generally zero slope with an inflec-
tion point.

Figure 11 represents the comparison between the solution
for AF2 with a pure oil inlet and that for AF2 with a C1 sat-
urated oil inlet. Figures 9 and 10 present the comparison
between AF1a and AF1b solutions; and figure 15 compares the
difference between the solutions for SF3a and SF3b. Selected
component flow rates, for AF2, are presented in figures 16,17,
and 18. The components selected were methane (an extremely
light component), propane (a moderately light component), and
normal octane (an extremely heavy component).
Distillation Calculations:

The distillation results are not as satisfactory as those for the absorber and stripper. Reasons for this behavior were discussed in the introduction. The primary result of this section is that convergence was never attained. However, in the course of the investigation, several modifications to the basic method were considered.

The first modification to be considered here was suggested by the behavior of the iteration process. If an overall material balance (11) is satisfied, the value of $b_i/d_i$ which is consistent with that balance is the corrected value:

$$(b_i/d_i)_{\text{corr.}} = \Theta (b_i/d_i)_{\text{calc.}}$$

However, the value of $b_i/d_i$ consistent with the balance about the feed plate is the calculated value, $(b_i/d_i)_{\text{calc.}}$. Thus, there are two choices: to satisfy the balance about the feed plate and try to converge to a balance about the column, or to satisfy the column balance and try to converge to the balance about the feed plate. Neither case was successful in that the iteration process produced successively greater mismatches at the bottom plate or at the feed plate.
A second modification was a consideration of varying the convergence method. This consisted in forming three distillate flow rate expressions:

\[ d_1 = \frac{Fx_1}{l + \theta \left(\frac{b_1}{d_1}\right)} \]

\[ d_1 = \frac{Fx_1}{\theta + \left(\frac{b_1}{d_1}\right)} \]

\[ d_1 = \frac{\theta Fx_1}{l + \left(\frac{b_1}{d_1}\right)} \]

to be used for the specification of equation (9). The case "a" is the original method discussed previously. The cases "b" and "c" necessitate the satisfying of the match at the feed plate since negative flow rates can be obtained for the individual bottoms flow rates if an overall balance is used. Cases "b" and "c" gave convergence, but yielded results that would not satisfy the material balances (values of \( \theta \) obtained were not equal to unity).

The third modification was in the direction of satisfying the enthalpy balances. All three directions (page 20) were considered, and it was found that the direction had a significant influence on the stability of the solution. Still, convergence could not be attained. The best results of the distillation calculation were produced with the distillate rate "a" and a match at the bottom of the column (mismatch at the feed) in conjunction with the enthalpy balance "c".
While this procedure failed to converge, it was the most stable. Convergence results are presented in figure 19.

The use of a wide boiling range feed was not used for the distillation calculation since a major change in the basic procedure would be required to eliminate the round-off error that arises in the solution of the material balance equations.

CONCLUSIONS

1. The proposed method for a rigorous separation calculation is more applicable for wide boiling range mixtures, and therefore yields good results for absorbers and strippers. However, for such mixtures the round-off error that arises in the solution of the material balance equations becomes significant; and the basic procedure for the solution of these equations must be altered to eliminate the effects of this error.

2. The modification used to eliminate the effect of this error was to perform a top-down calculation for the light components (components that have an average $A_{ij}$ less than
or equal to unity), and to perform a bottom-up calculation for the heavy components (components that have an average $A_{ij}$ greater than or equal to unity).

3. Convergence could not be attained for the distillation calculation, though several modifications were tried. The major problem seemed to be that of converging to a match at the feed (or that of converging to an overall column balance in the case where the match at the feed was forced).

4. The use of a wide boiling range feed for the distillation column was not used since a major change in the basic procedure would be required to eliminate the effects of round-off error.
NOMENCLATURE

$A_{ij}$ is the absorption factor $(L_j/V_jK_{ij})$.

$b_i$ is the component bottoms flow rate (moles/hr).

$d_i$ is the component distillate flow rate (moles/hr).

$D$ is the total distillate flow rate (moles/hr) or a partial distillate flow rate.

$e$ is a relative error defined on page 23.

$F$ is the feed flow rate (moles/hr).

$h_{ij}$ is the enthalpy of the pure liquid component "i" on the stage "j" (Btu/mole).

$H_{ij}$ is the enthalpy of the pure vapor component "i" on the stage "j" (Btu/mole).

$K_{ij}$ is the distribution ratio $(y_{ij}/x_{ij})$ for the component "i" on the stage "j".

$l_{ij}$ is the individual component liquid rate for the plate "j" (moles/hr).

$L_j$ is the total liquid rate for the stage "j" (moles/hr).

$M$ is the total number of stages including the reboiler.

$N$ is the total number of components.

$Q_j$ is the external heat load applied to the stage "j" (Btu/hr).

$R_i$ is a ratio of the inlet feeds for the absorber $(l_{10}/v_iM+1)$.

$R_i'$ is the reciprocal of $R_i$.

$S_{ij}$ is the stripping factor $(1/A_{ij})$.

$T_j$ is the stage temperature ($^\circ$F).

$v_{ij}$ is the individual component vapor rate for the plate "j"
\( V_j \) is the total vapor rate for the stage "j" (moles/hr).

\( x_{ij} \) is the liquid composition of the component "i" on the stage "j".

\( y_{ij} \) is the vapor composition of the component "i" on the stage "j".

**Subscripts**

- \( f \) refers to the feed plate.
- \( F \) refers to the feed.
- \( i \) refers to the component.
- \( j \) refers to the stage.

**Greek letters**

- \( \Theta \) is a correction factor.
- \( \omega \) is defined by equation (12b).
- \( \Pi \) is defined by equation (12c).
- \( \phi \) is defined by equation (13b).
- \( \phi \) is defined by equation (13c).
BIBLIOGRAPHY


   a. Chapter 3.
   b. Chapter 8.
   c. p. 145.
   d. pp. 494 and 496.


EQUILIBRIUM AND ENTHALPY DATA

Polynomial approximations were used to express the equilibrium and enthalpy data input. Relative volatilities were given the form:

\[ \alpha_i = \frac{K_i}{K_{\text{ref}}} = a_i + b_i T + c_i T^2 \]

with the reference K value:

\[ \ln K_{\text{ref}} = C_1 + C_2/(T + 459.69) . \]

The enthalpy functions were:

\[ h_1 = a_1^1 + b_1^1 T \]

and

\[ H_1 = a_1^V + b_1^V T . \]

All temperature are expressed in degrees Fahrenheit, and the polynomial coefficients are given in tables 1 through 6.
Table 1: Equilibrium Data for the Distillation Problem
(ref. 3d). Pressure = 120 psia.

<table>
<thead>
<tr>
<th>Component</th>
<th>$a_1$</th>
<th>$b_1 \times 10^2$</th>
<th>$c_1 \times 10^5$</th>
</tr>
</thead>
<tbody>
<tr>
<td>$C_3H_8$</td>
<td>+ 11.061</td>
<td>- 5.2007</td>
<td>+ 9.2489</td>
</tr>
<tr>
<td>1-$C_4$</td>
<td>+ 4.6929</td>
<td>- 1.8243</td>
<td>+ 3.1755</td>
</tr>
<tr>
<td>n-$C_4$</td>
<td>+ 3.0703</td>
<td>- 0.83565</td>
<td>+ 1.2144</td>
</tr>
<tr>
<td>1-$C_5$</td>
<td>+ 1.0000</td>
<td>- 0.00000</td>
<td>+ 0.0000</td>
</tr>
<tr>
<td>n-$C_5$</td>
<td>+ 0.73827</td>
<td>- 0.05246</td>
<td>+ 0.0189</td>
</tr>
</tbody>
</table>

$ln K_{1C_5} = +5.98 - 4100.0/(T + 469.69)$

Table 2: Enthalpy Data for the Distillation Problem (ref. 3d).
Pressure = 120 psia.

<table>
<thead>
<tr>
<th>Component</th>
<th>$a^v_1$</th>
<th>$b^v_1$</th>
<th>$a^l_1$</th>
<th>$b^l_1$</th>
</tr>
</thead>
<tbody>
<tr>
<td>$C_3H_8$</td>
<td>+ 7488.6</td>
<td>+ 17.505</td>
<td>+ 0.0000</td>
<td>+ 25.219</td>
</tr>
<tr>
<td>1-$C_4$</td>
<td>+ 9592.0</td>
<td>+ 18.435</td>
<td>+ 0.0000</td>
<td>+ 33.450</td>
</tr>
<tr>
<td>n-$C_4$</td>
<td>+ 8002.9</td>
<td>+ 43.827</td>
<td>+ 0.0000</td>
<td>+ 29.600</td>
</tr>
<tr>
<td>1-$C_5$</td>
<td>+11646.0</td>
<td>+ 27.706</td>
<td>+ 0.0000</td>
<td>+ 36.813</td>
</tr>
<tr>
<td>n-$C_5$</td>
<td>+12004.0</td>
<td>+ 31.687</td>
<td>+ 0.0000</td>
<td>+ 38.450</td>
</tr>
</tbody>
</table>
Table 3: Equilibrium Data for the Absorber Problem (ref. 3d).

Pressure = 264.7 psia

<table>
<thead>
<tr>
<th>Component</th>
<th>$a_i$</th>
<th>$b_i \times 10^2$</th>
<th>$c_i \times 10^5$</th>
</tr>
</thead>
<tbody>
<tr>
<td>CH$_4$</td>
<td>+ 63.765</td>
<td>- 29.519</td>
<td>+ 42.579</td>
</tr>
<tr>
<td>C$_2$H$_6$</td>
<td>+ 10.937</td>
<td>- 3.8843</td>
<td>+ 5.558</td>
</tr>
<tr>
<td>C$_3$H$_8$</td>
<td>+ 3.648</td>
<td>- 1.066</td>
<td>+ 1.6179</td>
</tr>
<tr>
<td>n-C$_4$</td>
<td>+ 1.000</td>
<td>- 0.0000</td>
<td>+ 0.0000</td>
</tr>
<tr>
<td>n-C$_5$</td>
<td>+ 0.17185</td>
<td>+ 0.22019</td>
<td>- 0.2826</td>
</tr>
<tr>
<td>n-C$_8$</td>
<td>- 0.01184</td>
<td>+ 0.02524</td>
<td>+ 0.0593</td>
</tr>
</tbody>
</table>

$$\ln K_{nC_4} = + 6.09 - 4085.0/(T + 459.69)$$

Table 4: Enthalpy Data for the Absorber Problems (ref. 3d).

Pressure = 264.7 psia.

<table>
<thead>
<tr>
<th>Component</th>
<th>$a^V_i$</th>
<th>$b^V_i$</th>
<th>$a^l_i$</th>
<th>$b^l_i$</th>
</tr>
</thead>
<tbody>
<tr>
<td>CH$_4$</td>
<td>+ 4864.2</td>
<td>+ 7.7362</td>
<td>+ 1840.3</td>
<td>+ 13.453</td>
</tr>
<tr>
<td>C$_2$H$_6$</td>
<td>+ 9394.5</td>
<td>+ 12.807</td>
<td>+ 3147.1</td>
<td>+ 22.225</td>
</tr>
<tr>
<td>C$_3$H$_8$</td>
<td>+11806.0</td>
<td>+ 17.974</td>
<td>+ 4425.2</td>
<td>+ 29.686</td>
</tr>
<tr>
<td>n-C$_4$</td>
<td>+14851.0</td>
<td>+ 23.120</td>
<td>+ 5553.2</td>
<td>+ 36.431</td>
</tr>
<tr>
<td>n-C$_5$</td>
<td>+18757.0</td>
<td>+ 29.035</td>
<td>+ 6663.2</td>
<td>+ 41.618</td>
</tr>
<tr>
<td>n-C$_8$</td>
<td>+25539.0</td>
<td>+41.192</td>
<td>+9845.0</td>
<td>+53.538</td>
</tr>
</tbody>
</table>
Table 5: Equilibrium Data for the Stripper Problems (ref. 1).
Pressure = 50 psia.

<table>
<thead>
<tr>
<th>Component</th>
<th>$a_1$</th>
<th>$b_1 \times 10^2$</th>
<th>$c_1 \times 10^5$</th>
</tr>
</thead>
<tbody>
<tr>
<td>CH$_4$</td>
<td>+ 100.5</td>
<td>- 55.34</td>
<td>+ 94.63</td>
</tr>
<tr>
<td>C$_2$H$_6$</td>
<td>+ 15.39</td>
<td>- 6.681</td>
<td>+ 11.22</td>
</tr>
<tr>
<td>C$_3$H$_8$</td>
<td>+ 4.244</td>
<td>- 1.193</td>
<td>+ 1.429</td>
</tr>
<tr>
<td>n-C$_4$</td>
<td>+ 1.000</td>
<td>+ 0.000</td>
<td>+ 0.000</td>
</tr>
<tr>
<td>n-C$_5$</td>
<td>+ 0.2164</td>
<td>+ 0.1052</td>
<td>- 0.000785</td>
</tr>
<tr>
<td>heavy</td>
<td>+ 0.0001759</td>
<td>+ 0.006691</td>
<td>+ 0.007222</td>
</tr>
</tbody>
</table>

\[ \ln K_{nC_4} = + 7.2666 - 4040.0/(T+459.69) \]

Table 6: Enthalpy Data for the Stripper Problems (ref. 6).
Pressure = 50 psia.

<table>
<thead>
<tr>
<th>Component</th>
<th>$a_v^1$</th>
<th>$b_v^1$</th>
<th>$a_l^1$</th>
<th>$b_l^1$</th>
</tr>
</thead>
<tbody>
<tr>
<td>CH$_4$</td>
<td>+ 4748.0</td>
<td>+ 9.472</td>
<td>+ 2553.0</td>
<td>+ 11.00</td>
</tr>
<tr>
<td>C$_2$H$_6$</td>
<td>+ 8556.0</td>
<td>+ 14.88</td>
<td>+ 3516.0</td>
<td>+ 18.90</td>
</tr>
<tr>
<td>C$_3$H$_8$</td>
<td>+ 11910.0</td>
<td>+ 20.50</td>
<td>+ 4936.0</td>
<td>+ 25.52</td>
</tr>
<tr>
<td>n-C$_4$</td>
<td>+ 15240.0</td>
<td>+ 27.72</td>
<td>+ 5440.0</td>
<td>+ 38.04</td>
</tr>
<tr>
<td>n-C$_5$</td>
<td>+ 18590.0</td>
<td>+ 34.41</td>
<td>+ 6307.0</td>
<td>+ 46.08</td>
</tr>
<tr>
<td>heavy</td>
<td>+ 20720.0</td>
<td>+ 40.07</td>
<td>+ 5985.0</td>
<td>+ 56.58</td>
</tr>
</tbody>
</table>
PROBLEM DATA

Absorber Feeds:

Table 7: Compositions of the Absorber Feeds.

<table>
<thead>
<tr>
<th>Component</th>
<th>AFla</th>
<th>AFlb</th>
<th>AF2</th>
<th>AF3</th>
</tr>
</thead>
<tbody>
<tr>
<td>CH₄</td>
<td>0.60</td>
<td>0.60</td>
<td>0.70</td>
<td>0.900</td>
</tr>
<tr>
<td>C₂H₆</td>
<td>0.05</td>
<td>0.20</td>
<td>0.15</td>
<td>0.080</td>
</tr>
<tr>
<td>C₃H₈</td>
<td>0.05</td>
<td>0.10</td>
<td>0.10</td>
<td>0.010</td>
</tr>
<tr>
<td>n-C₄</td>
<td>0.10</td>
<td>0.05</td>
<td>0.04</td>
<td>0.005</td>
</tr>
<tr>
<td>n-C₅</td>
<td>0.20</td>
<td>0.05</td>
<td>0.01</td>
<td>0.005</td>
</tr>
<tr>
<td>n-C₈</td>
<td>0.00</td>
<td>0.00</td>
<td>0.00</td>
<td>0.000</td>
</tr>
</tbody>
</table>

Table 8: Thermal Data on the Absorber Feeds.

<table>
<thead>
<tr>
<th>Temperature (°F)</th>
<th>AFla</th>
<th>AFlb</th>
<th>AF2</th>
<th>AF3</th>
</tr>
</thead>
<tbody>
<tr>
<td>bubble point</td>
<td>- 50.0</td>
<td>- 52.0</td>
<td>- 58.0</td>
<td>-----</td>
</tr>
<tr>
<td>dew point</td>
<td>+ 57.0</td>
<td>+ 88.0</td>
<td>+ 42.0</td>
<td>-----</td>
</tr>
<tr>
<td>gas inlet T</td>
<td>+ 160.0</td>
<td>[ + 160.0 ]</td>
<td>+ 75.0</td>
<td>+ 90.0</td>
</tr>
<tr>
<td>oil inlet T</td>
<td>+ 90.0</td>
<td>+ 90.0</td>
<td>+ 90.0</td>
<td>+ 90.0</td>
</tr>
</tbody>
</table>
Table 9: Compositions of the Stripper Feeds.

<table>
<thead>
<tr>
<th>Component</th>
<th>SF1</th>
<th>SF2</th>
<th>SF3a</th>
<th>SF3b</th>
</tr>
</thead>
<tbody>
<tr>
<td>CH₄</td>
<td>0.0001</td>
<td>0.0001</td>
<td>0.0001</td>
<td>0.0001</td>
</tr>
<tr>
<td>C₂H₆</td>
<td>0.0006</td>
<td>0.0049</td>
<td>0.0049</td>
<td>0.0099</td>
</tr>
<tr>
<td>C₃H₈</td>
<td>0.0006</td>
<td>0.0150</td>
<td>0.0050</td>
<td>0.0100</td>
</tr>
<tr>
<td>n-C₄</td>
<td>0.1000</td>
<td>0.0300</td>
<td>0.0100</td>
<td>0.0200</td>
</tr>
<tr>
<td>n-C₅</td>
<td>0.0600</td>
<td>0.0500</td>
<td>0.0300</td>
<td>0.0100</td>
</tr>
<tr>
<td>heavy</td>
<td>0.8387</td>
<td>0.9000</td>
<td>0.9500</td>
<td>0.9500</td>
</tr>
</tbody>
</table>

Table 10: Thermal Data on the Stripper Feeds.

<table>
<thead>
<tr>
<th>Temperature (°F)</th>
<th>SF1</th>
<th>SF2</th>
<th>SF3a</th>
<th>SF3b</th>
</tr>
</thead>
<tbody>
<tr>
<td>bubble point</td>
<td>+ 306.0</td>
<td>+ 342.0</td>
<td>+ 426.0</td>
<td>+ 377.0</td>
</tr>
<tr>
<td>dew point</td>
<td>+ 584.0</td>
<td>+ 594.0</td>
<td>+ 601.0</td>
<td>+ 601.0</td>
</tr>
<tr>
<td>steam inlet T</td>
<td>+ 600.0</td>
<td>+ 600.0</td>
<td>+ 600.0</td>
<td>+ 600.0</td>
</tr>
<tr>
<td>oil inlet T</td>
<td>+ 280.0</td>
<td>+ 340.0</td>
<td>+ 340.0</td>
<td>+ 340.0</td>
</tr>
</tbody>
</table>
FIG. 9: TEMPERATURE PROFILES FOR AF1

- AF1a - $T_i = 160 \, ^\circ$F
- AF1b - $T_i = 160 \, ^\circ$F
- AF1b - $T_i = 90 \, ^\circ$F
FIG. 10
VAPOR PROFILES FOR AFI

AFI_a \quad T_u = 160 \degree F
AFI_b \quad T_u = 160 \degree F
AFI_c \quad T_u = 90 \degree F
FIG. 11

ABSORBER SOLUTIONS FOR AF2

- PURE $L_o$
- C. SATURATED $L_o$
FIG. 12

ABSORBER SOLUTIONS FOR AF3

$L_0/N_m = 0.50$
$= 5.00$

$L_0/N_{mT_1} = 0.05$
$= 0.50$
$= 5.00$
FIG. 13

STRIPPER SOLUTIONS FOR SF1

\[ \frac{L_0}{V_{M_{l1}}} = 5.00 \]
\[ = 0.50 \]
\[ = 0.05 \]
FIG. 14

STRIPPER SOLUTIONS FOR SF2
FIG. 15

STRIPPER SOLUTIONS FOR SF3

- SF3 a
- SF3 b

$\frac{L_0}{V_m} = 5.00$
$= 0.50$
$= 0.05$

TEMPERATURE (°F)

V

STAGE NO.

TOP

BOTTOM
**FIG. 17**

$C_3$ VAPOR PROFILES FOR AF2

- **PURE OIL**
- **$C_1$ SATURATED OIL**

$L_0/V_{M,1} = 0.05$

- 0.50
- 5.00

**Legend**

- TOP
- BOTTOM

**Ysis (Propane Vapor Rate)**

**Stage No.**

0, 1, 2, 3, 4, 5, 6, 7, 8, 9, 10, 11
FIG. 18

n-C_8 VAPOR PROFILES FOR AF2

- - - - C. SATURATED OIL

- - - - PURE OIL

\( \frac{L_0}{V_{R0}} = 0.50 \)
\( = 0.05 \)
\( = 5.00 \)
FIG. 19

THE EFFECT OF THE INITIAL TEMPERATURE PROFILES ON THE CONVERGENCE OF THE DISTILLATION CALCULATION

LINEAR TEMPERATURE PROFILES

$T_i = T_m$

- 160 - 250
- 155 - 245
- 147 - 234
- 130 - 240

$\Theta = 1.0$

TRIAL NO.