1981

**Frequency Response Analysis of Distillation Column Interaction.**

Chien Wang  
*Louisiana State University and Agricultural & Mechanical College*

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Frequency Response Analysis
of
Distillation Column Interaction

A Dissertation

Submitted to the Graduate Faculty of the
Louisiana State University and
Agricultural and Mechanical College
in partial fulfillment of the
requirement for the degree of
Doctor of Philosophy

in
The Department of Chemical Engineering

by
Chien Wang
B.S. National Taiwan University, R.O.C., 1976
M.S. Louisiana State University, 1978
August 1981
Dedicated to my mother

Mrs. Wei-Wu (HAN) WANG
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ABSTRACT

A dynamic pressure model of multicomponent distillation is developed in the frequency domain. The model consists of a set of linearized overall and component balances, enthalpy balance, pressure relationship, hydraulic equation and equilibrium relationships for each tray, the condenser and the reboiler of a general column. A numerical method consisting of total scaled pivoting Gaussian elimination with the method of residues is worked out for solving the model in the frequency domain to obtain all of the desired transfer functions. The resulting transfer functions are presented as Bode plots. The measure of interaction between the various control loops can then be calculated as a function of frequency based on the transfer functions between the chosen manipulated and controlled variables. As a result, this analysis can be useful in defining the control problem and predicting interaction difficulties.

The computer program developed in this research can be used to study the effect of different model simplifying assumptions. The models being studied are pressure model, enthalpy model and equimolar overflow model.

Also, a generalized steady state program and a mechanical design program were developed to make the
frequency domain distillation column more realistic and generally applicable to a variety of columns at a variety of operating conditions.
CHAPTER ONE

INTRODUCTION

Process control is an important consideration for the operation of a distillation column. Once the distillation column is set up, it has to be operated for years. Generally speaking, the control systems that satisfy one distillation column might not satisfy another. The stability of the distillation column control system depends on the interactions between the process manipulated variables and measured variables. The term interaction arose because industry has often found it desirable to control the multivariable process as if it were made up of isolated single variable processes. In this research, a complete package is developed in the frequency domain to identify the distillation column process transfer functions, calculate the interaction measure as a function of frequency and determine the best pairing of the controlled and manipulated variables.

In the past, some distillation columns have been simulated in the frequency domain. Lamb, Pigford and Rippin (1961) designed an analogue simulation for a binary distillation column model with variable hold up on the
tray. In this simulation, they also included a tray efficiency factor with the same set of equations. Cadmen and Carr (1966) extended Lamb's distillation column to be a multicomponent distillation column. They inverted the frequency domain distillation column equations set at each frequency to get one point for the transfer function. Wood (1967) continued Cadman and Carr's column by allowing liquid flow perturbation to be caused by a change in vapor rate. The Shunta and Luyben (1971) approach removed the equimolal overflow restriction of Lamb's model by adding the enthalpy balance. F. P. Stainthorp (1973.A,B) use a set of equimolal dynamic equations to investigate the composition responses of a binary mixture on individual plates with respect to reflux flow changes and vapor flow changes.

All the above mentioned models are not flexible enough to handle different types of columns, in other words, the work performed by each of these authors is only applied to a particular distillation column. For example, Cadman and Carr's model was developed for a total condenser, single feed stream, with distillate product stream and residue product stream. None of the above models had provisions for multiple feed and sidestream. Also, these models depend on simplified steady state calculations, therefore these models were limited in their usefulness for control
For the measure of interaction within multi-control loops, Rijnsdorp (1965) proposed an interaction index for a two variable control system for the distillation column. Bristol (1966) also proposed a so called "Relative Gain Array" (RGA) method. This RGA method only requires the steady state process gain. E. J. Davison (1969) defined a non-minimum phase index for a linear time invariant multivariable system. It is then used to give a measure of the degree of interaction which may occur in a multivariable control system. Nisenfeld and Schultz (1971) have extended the relative gain array in what they call interaction analysis. Their approach involves a transformation of the diagonal elements of the relative gain array. Witcher and McAvoy (1977) in a report showed that although the Bristol RGA method and Rijnsdorp interaction index method start from different points, they actually end up with the same result. Witcher and McAvoy (1977) indicate that the influence of dynamics can be significant, therefore they improved the RGA method to a "Relative Dynamic Array", the RDA method, with the definition of dynamic potential. This RDA method was based on open loop response and concerned both steady state and dynamic interaction. They also pointed out the best way to interpret the interaction measure is in the frequency
domain in terms of magnitude and phase angle. In a 1978 paper, Bristol (1978) generalized his original "Relative Gain Array" analysis by replacing the steady state gains with the corresponding process transfer functions. His analysis provides useful insight into the dynamic behavior. In 1979, Jean Gagnepain (1979) pointed out that there exist defects in both the RGA method and the RDA method and suggested to use what he called the "averaged Relative Gain Method".

In this work, we have developed a set of computer programs to analyze distillation column interactions in the frequency domain. These programs provide the steady state distillation column profile, the mechanical design of the distillation column, and the dynamic response of the column in the frequency domain.

The steady state distillation column and the mechanical design of the distillation column are discussed individually in chapters two and three. Chapter four presents a detailed linearized distillation column model in the frequency domain. Chapter five presents the numerical method used in solving the linearized distillation column model and the concept of the interaction measure between the multiple manipulated variables and measured variables. The results are then presented in chapter six.

A basic structure was carefully designed for the
computer program to provide maximum flexibility and minimum reprogramming requirements for handling various distillation column systems and control system configurations. The structure is shown in FIG 1. Basically, there are six sets of computer programs.

(1) Steady state distillation column
(2) Mechanical distillation column design program.
(3) Constant coefficient generation program. This program generates all the necessary constants required by the frequency response analysis for a given control configuration.
(4) Frequency response analysis program. This program sets up the coefficient matrix and constant vector array at each specific frequency. Then the equation set is solved by Gaussian elimination with the method of residues for all the measured variables at each specific frequency.
(5) Bode plot program. This program makes Bode plots for the process transfer functions.
(6) Interaction measure calculation program. This program calculates the interaction measure in the frequency domain and the results are presented as Bode plots.

With reference to FIG 1, data base 1 is the data required to run the steady state program (the details are presented in the steady state design manual, Appendix K). Two sets
of steady state profiles are generated by the steady state program and transferred to the disk as data bases 2 and 3. Data base 4 is the necessary information to run the mechanical design program (the details are presented in the mechanical design manual, Appendix L). The mechanical design results are then stored into data base 5. The constant generation program first picks up the steady state profile from data base 3 and the mechanical design results from data base 5. Then, the constants are generated corresponding to different control configurations and put in data base 6. Data base 7 contains the necessary information to run the frequency response analysis program, such as minimum and maximum frequency, desired manipulated variables, disturbance variables and measured variables. The Bode plot program makes varian Bode plots based on data base 8, the process transfer function information generated by the frequency response analysis. The interaction measure program calculates the interaction measure also based on the process transformation information generated by the frequency response analysis.
CHAPTER TWO

STEADY STATE DISTILLATION WITH THE FLASH FEED

(A) Steady State Distillation

In order to study the frequency response of a general linearized distillation column model, the steady state program is needed to facilitate the set up of the points around which the equations are linearized. The steady state distillation model must provide the design options of multiple feed streams, multiple product sidestreams and interstage heaters or coolers. There are a number of steady state distillation methods available such as Lewis-Matheson method, Thiele-Geddes method, Relaxation method and so forth. However, as pointed out by Friday and Smith (1964), some methods are numerically unstable, others converge too slow and some require a considerable amount of computer storage.

In this research, the Wang/Henke method (1966) is used to generate the steady state profile. It is assumed that each stage can have one feed stream, one vapor side stream, one liquid side stream and one interstage heater or cooler. The working equations which describe each column
stage are obtained from material balances, equilibrium relationships, summation equations and enthalpy balances. This general model column can be reduced to any simpler model by setting the appropriate terms to zero. Another advantage of this method is that the component material balances can be arranged into the tridiagonal matrix form which saves considerable computer memory.

The Wang/Henke computer program used in this research was set up originally by Kurt Kominek (1979). Kominek put in also the Holland's Theta method and partial sustititution method to accelerate the convergence. This author has added the flash feed calculation and the Murphree tray efficiency to accomplish this general purpose steady state distillation program.

This steady state program allows the following design options:

1. multi-component distillation
2. multiple feed streams
3. multiple product sidestreams
4. inter-stage heaters or coolers
5. flash feed calculation
6. partial or total condenser
7. ideal or non-ideal liquid - vapor equilibrium
8. tray efficiency
9. equimolar or non-equimolar overflow calculation
The Wang/Henke calculation procedure is summarized as follows:

(1) Read the input data.
(2) Make the flash feed calculation for each feed stream.
(3) Generate the estimated liquid flow profile, vapor flow profile and composition profile.
(4) Arrange each component material balance into the tri-diagonal matrix form. Invert each matrix to solve for a new composition profile.
(5) Correct the composition profile by using Holland's Theta method.
(6) Generate the new temperature profile by the bubble point temperature calculation.
(7) Modify the temperature profile by partial substitution.
(8) Simultaneously solve the heat and material balance for each stage for new liquid flow rate and vapor flow rate.
(9) Modify liquid flow rate profile and vapor flow rate profile by partial substitution method.
(10) Check the relative error in temperature profile, composition profile, liquid flow and vapor flow profile. If smaller than the tolerance, exit.
Otherwise, go to step (4)

The design results were verified by comparison with the results by Wang and Henke (1966).

(B) Feed To The Distillation Column

As shown in figure 2, the upstream feed passes through a restriction valve before it enters the distillation column. The downstream pressure of the restriction valve is the feed stage pressure, neglecting the pipe friction loss. The total feed rate \( F \), feed composition \( Z \), feed temperature \( T_U \) and feed pressure \( P_U \) are the given upstream feed conditions.

The flash feed calculation is essentially an isenthalpic process which can be verified from the mechanical energy balance by neglecting the kinetic energy loss through the restriction valve. Whether the feed is all liquid, all vapor or a partially vaporized feed can be determined from the comparison among the total feed enthalpy with the bubble point reference enthalpy which assumes the feed is all liquid at its bubble point temperature, or with dew point reference enthalpy which assumes that the feed is all vapor at its dew point temperature. Therefore if the total feed enthalpy is below the bubble point reference enthalpy, the feed is all liquid. The feed is all vapor whenever the
FIG 2. Feed Into The Distillation Tower.

FIG 3. Liquid Feed.
FIG 4. Vapor Feed.

FIG 5. Flash Feed.
total feed enthalpy is above the dew point reference enthalpy. Supposed the total feed enthalpy is in-between the bubble point and dew point reference enthalpy, then the feed is a flash feed which consists of both liquid and vapor stream.

An equilibrium flash feed model and the flash feed calculation procedure are shown in Appendix A.

Therefore, there exist three kinds of feed conditions into the distillation column feed stage.

(1) downstream feed is all liquid, FIG 3.
(2) downstream feed is all vapor, FIG 4.
(3) downstream feed is flash feed which consists of both liquid and vapor, FIG 5.

This chapter has outlined the steady state calculation of the distillation column with the flash feed. Based on the steady state profile generated by the steady state calculation program, the mechanical design of the distillation column is discussed in the next chapter.
CHAPTER THREE

MECHANICAL DESIGN OF THE DISTILLATION COLUMN

The many aspects of mechanical design of a distillation column are essential to the study of the dynamic behavior of the distillation tower. The mechanical design of the distillation column usually consists of the tray column, reboiler, condenser and reflux drum.

The purpose of this chapter is to present computer methods to design each part of the distillation tower based on the steady state profile generated by the steady state program outlined in chapter two. Since the design equations recommended in this chapter have practically all been used in the actual plant design, there is no attempt to present theoretical developments of the design equations. Graph relationships for hand calculation purposes are correlated by SAS (Statistical Analysis System) (1979) into equations to use in this computer design.

The distillation column mechanical design options are as follows:

(A) reboiler

(1) kettle reboiler, vertical thermosyphon reboiler or forced circulation reboiler.

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FIG 6. Vertical Thermosyphon Reboiler.

FIG 7. Kettle Reboiler.
(B) condenser
   (1) total condenser or partial condenser.
   (2) vertical condenser or horizontal condenser.
(C) horizontal reflux drum
   (1) drum for total condenser or for partial condenser.
(D) tray column
   (1) single tray design or tray column design.
   (2) Ballast tray designed by the Glitsch company or
       flexitray designed by the Koch company.

REBOILER DESIGN

The scope of the design of reboilers in this research
is limited to vertical thermosyphon reboiler, FIG 6, forced
circulation reboiler, FIG 8, and kettle reboiler FIG 7.

The vertical thermosyphon reboiler and forced
circulation reboiler are vertical heat exchangers with
"circulating" boiling inside the tubes. Flow to the tubes
is by natural or forced circulation and vaporization inside
the tubes is usually less than 100%. The kettle reboiler
is a horizontal heat exchanger with "non-circulating"
(pool) boiling outside the tubes. The advantages of
vertical thermosyphon reboiler are that it is capable of
very high heat transfer rate and requires compact, simple
piping. Forced circulation reboilers are used for viscous
and solids-containing liquids. The advantage is that the circulation rate can be controlled so that the vapor blanking phenomena can be eliminated to a large extent. The maintenance and cleaning for the kettle reboiler are easier, but heat transfer rate is usually lower compared to the others. The kettle reboiler is used for small heat duty operation.

The design procedures for the vertical thermosyphon reboiler, forced circulation reboiler and kettle reboiler are summarized in Appendices B, C and D, respectively. The design results were verified by comparison with the results by Kern (1950) and Ludwig (1977).

Total Condenser or Partial Condenser Design

For single component vapor, the dew point temperature is equal to the bubble point temperature. If the vapor is a multi-component mixture, the condensation occurs over a temperature range. The upper limit of the condensing range is the dew point temperature of the vapor entering the condenser. For a total condenser, the lower limit is the bubble point temperature of the liquid leaving the condenser. However for a partial condenser, the lower limit is set by the design and is somewhere in between the inlet dew point temperature and the pseudo bubble point temperature
which assumes total condensing of the inlet vapor.

Since the condensing of the vapor mixture may occur over a wide temperature range, the fraction of the total heat load delivered during a fractional decrease in the vapor temperature need not be uniform over the entire condensing range, and this invalidates the use of the logarithmic mean temperature difference for the condenser design.

The condensing curve is a plot of vapor heat content versus vapor temperature over the entire condensing temperature range. Since the equilibrium in the condenser is actually a differential condensing, the condensing curve can be constructed by dividing the condensing temperature range into several temperature grid points. At each grid temperature, the flash calculation method outlined in the second chapter is used to calculate the vapor composition and the vapor heat content.

The percentage rise in the temperature of the cooling medium at any cross section of the shell is taken as proportional to the percentage of the heat load removed from any condenser cross section to the exit. The weighted delta T is then the averaged temperature difference between the condensing curve and a straight line representing the cooling medium.

The design procedure for either the total condenser or
FIG 9. Cross Section of the Reflux Drum

\[ h \geq 20\% \ D \]
\[ D - h \geq 10\% \ D \]
partial condenser is summarized in Appendix E. The design results were verified by comparison with the results by Kern (1950) and Ludwig (1977).

Optimum Design Of Reflux Drum For Partial Condenser

B. Sigales (1975) pointed out that ordinarily the dimensions of a reflux drum must be fixed to allowed (1) droplet disengagement in the vapor phase, (2) enough surge time for the liquid phase and (3) settling of small amount of a second heavy liquid phase, if there is one.

For the partial condenser's reflux drum, in the vapor, the droplet separation is attained by limiting vapor velocity which can be calculated by Newton's law. Vapor flow sets the minimum vapor passage area, i.e. the section free of liquid. This area should include a minimum clearance, hH above the high liquid level, FIG 9, equal to 20% of the drum diameter.

Liquid stream requires a minimum surge time in the drum to facilitate control, absorb process upsets and fluctuations, maintain safe operation, etc. Surge time is defined as the time required to empty the vessel if inflow is stopped. Sigales suggested that surge time is dependent on the holdup between the high liquid level LH and low
liquid level $L_l$, FIG 9. This approach excludes the liquid below $L_l$, which makes for a conservative design.

TABLE 1 lists the typical values suggested by Siegales; but only the larger volume should be used.

**TABLE 1. Holdup Of The Reflux Drum.**

<table>
<thead>
<tr>
<th></th>
<th>Holdup from $L_h$ to $L_l$, MIN.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tower reflux</td>
<td>5</td>
</tr>
<tr>
<td>Product to storage</td>
<td>2</td>
</tr>
<tr>
<td>Product to heat exchanger, along with other process stream.</td>
<td>5</td>
</tr>
<tr>
<td>Product to heater</td>
<td>10</td>
</tr>
</tbody>
</table>

If no second liquid phase is to be settled, $L_l$ is situated above vessel bottom at a distance $D-hL$ equal to 10% of the drum diameter. If possible, this distance should be greater than 5 inches. This circular segment of height 10% of the diameter has an area equal to 5% of the total vessel cross sectional area.

The optimum design procedures of reflux drum for either the total condenser or partial condenser are summarized in Appendix F.
Design of Single Tray or Tray Column

The design of individual trays and tray column used in this research is based on the work published by Economopoulos (1978).

In tray column design, some decisions must be made to fix the following tray design parameters:

1) Flood factor and weep factor. Low flood factors are known to yield large tray sizes and increased tower volume. Low weep factors are known to cause increased tower height or undesirably high pressure drop.

2) Weir height, deck thickness and tray hole diameter.

3) Maximum tray pressure drop.

4) System derating factor, depends on the foaming tendency of the system.

5) Tray spacing, is used for the initial sizing of each tray in the column.

In this research, the Ballast tray of Glitsch Company and the Flexitray of the Koch Company are used for the tray column design. The working equation set and the design procedures are summarized in Appendix G. The design results were verified by comparison with the results in the "Ballast Tray Design Manual" (1970).

In order to analyze the dynamics of the distillation
column, it is also necessary to have a dynamic model besides the steady state calculations and the mechanical design of the distillation column. In the next chapter, a generalized distillation pressure model is derived both in the time domain and the Laplace domain.
CHAPTER FOUR

DYNAMIC MATHEMATICAL MODELING OF THE DISTILLATION COLUMN

In this chapter, the dynamic model of the distillation column is derived mathematically in a very general manner. Besides the condenser and the reboiler, each tray of the distillation column may have liquid and vapor feed, liquid and vapor sidestream and inter-stage heater or cooler. The condenser is defined to be the first stage and the reboiler is defined to be the last stage \( j \) of the distillation column.

The derivations are divided into the following six parts:

(1) General assumptions.

(2) Physical properties. The physical properties are linearized and Laplace transformed. These physical properties include:
   (a) Enthalpy of the vapor mixture.
   (b) Enthalpy of the liquid mixture.
   (c) Enthalpy of the feed vapor.
   (d) Enthalpy of the feed liquid.

(3) Equilibrium relationships. Each equilibrium relationship is linearized and Laplace transformed.

(4) A generalized stage of the distillation column is first

FIG 11. Generalized Nth Tray
derived mathematically in the time domain. Then, all the equations are linearized and Laplace transformed.

(5) A condenser is first derived mathematically in the time domain. Then, all the equations are linearized and Laplace transformed.

(6) A reboiler is derived mathematically first in the time domain. Then, all the equations are linearized and Laplace transformed.

A typical distillation column with total condenser and reboiler is sketched in FIG 10.

**General Assumptions For The Dynamic Distillation Column**

(1) Tray vapor holdups are negligible in comparison with the liquid holdups.

(2) Adiabatic distillation column operation, i.e., no heat is gained or lost through the column wall.

(3) Perfect mixing is assumed for the vapor and liquid on each tray, in the reflux drum, at the column base, and in the coolant and steam side of the heat exchanger.

(4) Level control was achieved in both the reflux drum and the distillation tower base. (i.e. no occurrence of overflow or running dry).

(5) For the steady state model, linear pressure drop is assumed from the bottom to the top of the distillation
column.

(6) Ideal vapor - liquid equilibrium relationship is assumed to obey Raoult's law.

(7) The convective heat transfer coefficients of the heat exchanger are assumed to be constant.

(8) The liquid and vapor density on each tray, at the column base, and in the reflux drum are assumed to be constant.

(9) Vapor in the reflux drum is assumed to be a perfect gas and has the same temperature as the vapor leaving the condenser.

(10) Dynamics of component material balance and enthalpy balance of vapor space in reflux drum are negligible.

(11) The trays, condenser and reboiler are made of the same material.

(12) For each active tray, the liquid inlet height over weir is the same as the outlet height over weir. i.e. the liquid head gradient across the active tray is negligible.

(13) The vapor in the condenser is assumed to be a perfect gas and in phase equilibrium with the liquid leaving the condenser.

(14) Liquid and vapor holdup in the condenser itself is neglected.

(15) The liquid in the downcomer is not aerated.
**Equilibrium relationships**

Vapor - liquid equilibrium relationship is assumed to obey Raoult's law.

\[
\frac{Y_{i,n}}{P_{i,n}} = \frac{S}{P_{i,n}} = f(T_n, P_n)
\]

Where:

- \( P_n \): pressure, PSI, of tray \( n \).
- \( S \): saturated vapor pressure, PSI, of component \( i \) at tray \( n \).
- \( P_{i,n} \): saturated vapor pressure, PSI, of component \( i \) at tray \( n \).

\[
S_{i,n} = 0.0193368 \times \exp \left( \frac{\text{ANTA}_{i}}{T_{n}/1.8} + 255.55 + \text{ANTC}_{i} \right)
\]

\( \text{ANTA}_i, \text{ANTB}_i, \text{ANTC}_i \): saturated vapor pressure coefficient of process component \( i \).

\( T_n \): temperature, \( F \), on tray \( n \).

In Laplace domain, equilibrium relationships are given by:

\[ K_{i,n}(s) = C_1 * T_{i,n}(s) + C_8 * P_{i,n}(s) \] (4.3)

where:

\[ C_1 = \frac{0.0107427}{\exp(A N T_A - \frac{255.55 + A N T_C}{1.8} - \frac{255.55 + A N T_C}{1.8} + \frac{255.55 + A N T_C}{1.8})} \]

\[ C_8 = \frac{-0.0193368}{\exp(A N T_A - \frac{255.55 + A N T_C}{1.8} - \frac{255.55 + A N T_C}{1.8} + \frac{255.55 + A N T_C}{1.8})} \] (4.4)

**Physical Properties**

(1) enthalpy of the vapor mixture.

The enthalpy of the vapor mixture is assumed to have the following form:

\[ H_V = \sum_{n} (\alpha_1 + \alpha_2 * T + \alpha_3 * T^2 + \alpha_4 * T^3 + \alpha_5 * T^4) Y_{i,n} \]
In Laplace transformation notation,
\[ HV (s) = \sum_{i} \left( C_{2} K(s) + C_{3} X(s) + C_{4} T(s) \right) \]

where:
\[ C_{2} = (\alpha + \alpha T + \alpha T + \alpha T + \alpha T + \alpha T) X \]
\[ C_{3} = (\alpha + \alpha T + \alpha T + \alpha T + \alpha T + \alpha T) K \]
\[ C_{4} = (\alpha + 2\alpha T + 3\alpha T + 4\alpha T + 5\alpha T) X K \]

Substitute the \( K(s) \) in (4.6) by (4.3), then,
\[ HV (s) = \sum_{i} \left( C_{2} + C_{4} \right) T(s) + \sum_{i} C_{3} X(s) \]

\[ + \sum_{i} C_{2} C_{8} P(s) \]

(2) enthalpy of the liquid mixture

The enthalpy of the liquid mixture is assumed to have the following form:
In Laplace transformation notation:

\[ H_L (S) = \sum_{i=1}^{n} C_6 * T (s) + \sum_{i=1}^{n} C_7 * X (s) \] (4.10)

where:

\[ C_6 = \sum_{i=1}^{n} (\phi + 2*\phi *T + 3*\phi *T + 4*\phi *T + 5*\phi *T) X \]

\[ C_7 = \phi + \phi *T + \phi *T + \phi *T + \phi *T \]

The enthalpy of the feed liquid mixture is assumed to have the following form:

\[ H_{FL} = \sum_{i=1}^{n} (\phi + 2*\phi *T + 3*\phi *T + 4*\phi *T + 5*\phi *T) Z_L \] (4.12)

In Laplace transformation notation:

\[ H_{FL} (S) = \sum_{i=1}^{n} C_{6F} * TF (s) + \sum_{i=1}^{n} C_{7F} * ZL (s) \] (4.13)
where:

\[
\overline{C_{6F}} = \sum (\phi_1 + 2\phi_2 * T_F + 3\phi_3 * T_F + 4\phi_4 * T_F + 5\phi_5 * T_F) + Z_L
\]

\[
\overline{C_{7F}} = \phi_1 + \phi_2 * T_F + \phi_3 * T_F + \phi_4 * T_F + \phi_5 * T_F + Z_V
\]

(4.14)

(3) enthalpy of feed vapor mixture

The enthalpy of the feed vapor mixture is assumed to have the following form:

\[
H_{FV} = \sum (\alpha_1 + \alpha_2 * T_F + \alpha_3 * T_F + \alpha_4 * T_F + \alpha_5 * T_F) * Z_V
\]

(4.15)

In Laplace transformation notation:

\[
H_{FV}(s) = \overline{C_{2F}} * T_F(s) + \sum \overline{C_{3F}} * Z_V(s)
\]

(4.16)

where:

\[
\overline{C_{2F}} = \sum (\alpha_1 + 2\alpha_2 * T_F + 3\alpha_3 * T_F + 4\alpha_4 * T_F + 5\alpha_5 * T_F) + Z_V
\]

\[
\overline{C_{3F}} = \alpha_1 + \alpha_2 * T_F + \alpha_3 * T_F + \alpha_4 * T_F + \alpha_5 * T_F + Z_V
\]

(4.17)
**Generalized Equilibrium Tray**

A generalized nth tray is shown in FIG 11. The equations which describe the dynamic behavior of the nth tray in the time domain are listed below:

(1) Total material balance

\[
\frac{dM}{dt} = L + V - (L + SSL) - (V + SSV)
\]  

(4.18)

(2) Component material balances

\[
\frac{d(M_i X_i)}{dt} = (L_i X_{i,n-1}) + (V_i Y_{i,n+1}) + (F_i L_i Z_{i,n}) + (F_i V_i Z_{i,n}) - (L_i + SSL_i) X_{i,n} - (V_i + SSV_i) Y_{i,n}
\]  

(4.19)

(3) Enthalpy balance

\[
\frac{dT}{dt} = (H_i L_i M_i) + \hat{H} \cdot \Delta C M \cdot C_i - (H_i L_i) + (H_i V_i) + (H_i F_i L_i) + (H_i F_i V_i) - H_i (L_i + SSL_i) - H_i (V_i + SSV_i) + Q
\]  

(4.20)

(4) Equilibrium relationships.

\[
Y_{i,n} = K_{i,n} X_{i,n}
\]  

(4.21)
\[ K_{i,n}^{p,n} = \ldots \] (4.22)

(5) Summation equations.
\[ \sum_{i}^{n} x_i = 1.0 \] (4.23)
\[ \sum_{i}^{n} y_i = 1.0 \] (4.24)

(6) Feed split equations.
\[ F = F_L + F_V \] (4.25)
\[ Z_{i,n} * F = Z_{L,i,n} * F_L + Z_{V,i,n} * F_V \] (4.26)
\[ H_{F,n} * F = H_{FL,n} * F_L + H_{FV,n} * F_V \] (4.27)

(7) Thermal properties
\[ H_{L,n} = f(T, X_{i,n}) \] (4.28)
\[ H_{V,n} = f(P, T, X_{i,n}) \] (4.29)

(8) Tray hydraulics.
(a) downcomer liquid level.
The material balance for the downcomer liquid is,
\[ \frac{dH_{DC,n}}{dt} = \frac{W_{n,n} - W_{DC,n}}{n-1} \] (4.30)

where:
\[ W_{DC,n} : \text{flow, LB/SEC, from downcomer of tray } n. \]
FIG 12. Distillation Tray Schematic For Flows And Liquid Elevation
W : overflow, LBM/SEC, from tray n-1 into the
downcomer.

ρDC : liquid density in downcomer, LBM/FT**3, assume
the density does not vary.

ADC : downcomer cross sectional area, FT**2, tray n.

HDC : liquid height in downcomer, FT, tray n.

(b) downcomer pressure drop and flow.

The downcomer pressure is determined by the difference
between the liquid heads and static heads.

\[ ΔPDC = \frac{g}{n} (ρDC * HDC - ρTR * HTR') - (P' - P'') \]

\[ c \]

(4.31)

where:

ΔPDC : downcomer pressure drop, LBF/FT**2, tray n.

ρTR : density of aerated liquid on tray n, LBM/FT**3.

HTR' : height, FT, of aerated liquid on tray n just
downstream of the inlet weir.

P' : pressure, LBF/FT**2, above liquid on tray n-1.

P' : pressure, LBF/FT**2, on tray n.

The liquid flow pressure drop under the downcomer apron
is given by Van Winkle (1967) as follows:

\[
h_d = \frac{0.03 \cdot Q_{DC}}{n} \quad (4.32)
\]

where:

- \( h_d \): liquid head in inches of liquid flow pressure drop through the downcomer, tray \( n \).
- \( Q_{DC} \): downcomer flow, GPM, tray \( n \).
- \( A_{DM} \): free area for liquid flow under the downcomer apron, FT\(^2\), normal tray construction practice yields approximately 0.42 of the downcomer area.

The above equation can be re-written to be:

\[
\frac{h_d}{n} = \frac{\Delta P_{DC}}{\rho_{DC} \cdot \frac{\Delta P_{DC} \cdot \rho_{DC}}{\sqrt{\frac{2}{\rho_{DC}}} \cdot A_{DM} \cdot \rho_{DC}} \cdot \frac{2}{\rho_{DC}}} \quad (4.33)
\]

or

\[
W_{DC} = 4.456 \cdot A_{DM} \cdot \sqrt{\frac{\Delta P_{DC} \cdot \rho_{DC} \cdot \rho_{DC}}{\sqrt{\frac{2}{\rho_{DC}}}}} \quad (4.34)
\]
(c) Aerated liquid holdup and gradient on tray

With reference to FIG 12, a linear gradient across the tray can be assumed,

\[
\frac{V_{TR}}{A_{TR}} = \frac{(H_{OW} + H_{W}) + H_{TR}'}{2} \quad (4.35)
\]

where:

- \( V_{TR} \): volume, FT**3, of aerated liquid on active area of tray n. (downcomer area excluded).
- \( A_{TR} \): active area of tray n, FT**2.
- \( H_{OW} \): height, FT, of crest of liquid above the outlet weir, tray n.
- \( H_{W} \): height, FT, of outlet weir.
- \( H_{TR}' \): holdup of active tray n, LBM.

(d) Inlet liquid height over weir.

Assume that the change in inlet weir height over weir is the same as the change in outlet height over weir.

\[
H_{TR}' = H_{W} + H_{OW} \quad (4.36)
\]

(e) Volume of liquid on active tray.

\[
\frac{W_{TR}}{V_{TR}} = \frac{\rho_{TR}}{\rho_{TR}} \quad (4.37)
\]

where:

- \( W_{TR} \): holdup of active tray n, LBM.
(f) active tray total material balance.

The total material balance for active tray is given by:

\[
\frac{dW_{TR}}{dt} = W_{DC} - W + W_{FL} + W_{VF} - W_{SSL} - W_{SSV} + W_{V} - W_{V}^{n+1}
\]

(4.38)

where:

- \( W \) : overflow, LBM/SEC, from tray \( n \).
- \( W_{FL} \) : feed liquid, LBM/SEC, to tray \( n \).
- \( W_{VF} \) : feed vapor, LBM/SEC, to tray \( n \).
- \( W_{SSL} \) : sidestream liquid, LBM/SEC, from tray \( n \).
- \( W_{SSV} \) : sidestream vapor, LBM/SEC, from tray \( n \).
- \( W_{V} \) : vapor rate, LBM/SEC, leaves tray \( n+1 \).

(g) aerated liquid density as a function of vapor velocity.

In the active area, tray action is characterized by the froth. The observable height of this mass is designated as \( h_{f} \). The density of the mass varies with height, being greatest near the tray floor; however, it is convenient to utilize a height average density \( \rho_{TR} \). There is also an effective hydrostatic head of the aerated mass \( h_{l} \), which could be measured by a floor manometer. If the weight of the vapor portion of the mass is neglected, \( h_{l} \) is the settled
height of the clear liquid, having a clear liquid density of $\rho_L$. A relative froth density, $\phi_F$, can be defined as follows:

$$\phi_F = \frac{h_{\text{TR}}}{h_f} \frac{\rho_{\text{TR}}}{\rho_L}$$

(4.39)

Van Winkle (1967) presents a plot which shows that froth density decreases with an increase in vapor rate. The plot is a relationship for sieve tray's froth density vs. kinetic energy parameter, $F$, and the correlation is as follows:

$$\phi_F = 0.954 - 1.238F + 0.682F^2 - 0.1272F^3$$

(4.40)

From equation (4.41),

$$\rho_{\text{TR}} = (0.954 - 1.238F + 0.682F^2 - 0.1272F^3) \rho_L$$

(4.41)

The kinetic energy parameter is defined as,

$$F = \frac{0.5}{UVA \rho_V}$$

$$= \frac{WV}{A_{\text{TR}}}$$

(4.42)
where:

\[ \text{UVA} : \text{vapor velocity, FT/SEC, corresponding to ATR} \]

\[ \text{WV} = \frac{\rho V}{\text{ATR}} \]

\[ V : \text{vapor density, LBM/FT}^3, \text{corresponds to WV}. \]

(h) liquid overflow from tray.

The Francis Weir formula can be written as

\[ W = \rho TR \times (KK \times \text{HOW})^{3/2} \]  

(4.43)

where:

\[ KK : \text{a constant for any given weir} \]

\[ = 3.03588 \times LW \times FW^{-1.5} \]

\[ LW : \text{weir length, FT.} \]

\[ FW : \text{weir correction factor.} \]

(i) dry hole pressure drop.

In the frequency analysis of distillation column in this research, the Ballast tray designed by the Glitsch company is used. The pressure drop of Glitsch's Ballast tray is a function of vapor and liquid rates, type, metal density, and thickness of the valve, weir height and weir length. At low to moderate vapor rates,
when the valves are not fully opened, the dry tray pressure drop is proportional to the valve weight and is essentially independent of the vapor rate. At vapor rates sufficiently high to open the valves fully, the dry tray pressure drop is proportional to the square of the vapor velocity through the orifice. The following equations are recommended by the Glitsch design manual 4900, (1970).

(i.1) tray valve unit part opened.

\[
\Delta \text{PD}_{RY} = (1.35 * t_m * \rho_m + K1 * U_h * \rho_v) \left(\frac{1}{n-1}\right) \left(\frac{1}{n-1}\right) \left(\frac{2}{n-1}\right) \left(\frac{1}{n}\right) \left(\frac{1}{12}\right) \left(\frac{g}{c}\right)
\]

(i.2) tray valve unit fully opened.

\[
\Delta \text{PD}_{RY} = (K2 * U_h * \rho_v) \left(\frac{1}{n-1}\right) \left(\frac{1}{n-1}\right) \left(\frac{2}{n-1}\right) \left(\frac{1}{n}\right) \left(\frac{1}{12}\right) \left(\frac{g}{c}\right)
\]

Where:

\( \Delta \text{PD}_{RY} \) : dry hole pressure drop for tray n-1, LBF/FT**2.  
\( t_m \) : Ballast tray valve thickness, IN, of tray n-1.  
\( \rho_m \) : valve metal density, LB/FT**3, of tray n-1.  
\( K1, K2 \) : pressure drop coefficients, defined by Glitsch design manual 4900, (1970).  
\( U_h \) : velocity, FT/SEC, through the holes on tray n-1.
\[
\frac{WV}{n} = \frac{\rho V \cdot AH}{n \cdot n-1} \quad (4.46)
\]

(j) **total tray pressure drop.**

(Ballast tray, V-1 type, Gage 14 thickness, carbon steel).

\[
\Delta PTR' = p' - p' \quad n-1 \quad n \quad n-1
\]

\[
\Delta PTR' = \Delta PDY' + 0.372787 \left( \frac{g}{n-1} \right) \left( \frac{W}{2/3} \right) * \rho L \quad n-1
\]

\[
+ 0.4 * \frac{HW}{n-1} * \rho L * (\frac{g}{c}) \quad n-1 \quad g \quad c
\]

where:

\[
\Delta PTR' : \text{total tray pressure drop for tray } n-1, \quad \text{LBF/FT}^2.
\]

\[
p' : \text{pressure, LBF/FT}^2, \text{of tray } n-1.
\]

\[
HW : \text{weir height, FT, of tray } n-1.
\]

\[
LW : \text{weir length, FT, of tray } n-1.
\]

(k) **steady state downcomer backup.**

The clear liquid height in the downcomer is equal to the height of liquid immediately outside the downcomer (\(HW + HOW\)) increased by the head loss of the liquid flowing under the downcomer and the total pressure drop.
\[
HDC = \frac{\rho L}{n} \left( HW + (ht + hd) \right)^n - \frac{\rho L}{n} \rho V_{n+1}
\]

\[
W_n = \frac{2}{3} + 0.372787 \times \frac{\rho L}{LW} n
\]

Where:

\(hd\) : liquid head in FT of liquid flow pressure drop through the downcomer. (4.32)

\(ht\) : total pressure drop in FT of liquid.

(1) steady state tray holdup.

\[M' = \rho DC_n \left( ADC_n \cdot HDC_n \right) + \left[ ATR_n \times \left( HW_n + HOW_n \right) \right] \cdot \rho TR_n\]

where:

\(M'\) : holdup of tray \(n\), LBM.

(9) pressure drop through the tray.

\[\Delta PTR_n = P_n - P_{n-1}\]

\[
= \Delta PD_{DRY} + 0.0025888 \times \frac{\rho L}{LW} \frac{2}{3} n-1^2 + 0.002777 \times HW \cdot \rho L \cdot \left(\frac{\rho L}{c}\right) n-1^2
\]

(4.50)
where:

\[ P_n : \text{pressure, PSI, of tray } n. \]

\[ P_{n-1} : \text{pressure, PSI, of tray } n-1. \]

\[ \Delta P_{TH} : \text{total pressure drop, PSI, through tray } n-1. \]

\[ \Delta P_{DRY} : \text{dry tray pressure drop, PSI.} \]

Laplace Domain Generalized Equilibrium Tray

If the above generalized equilibrium tray time domain equations are linearized, Laplace transformed and simplified, (as shown in Appendix H), the following equations can be reached with variables \( L(s), V(s), T(s), P(s) \) and \( N_{CIXi}(s) \) for each generalized equilibrium tray, \( n \), \( n \), \( n \), \( n \), \( i,n \)

The equations are:

1. \( N_{C-1} \) Laplace domain component material balance, (H.4).
2. Laplace domain enthalpy balance, (H.6).
3. Laplace domain summation equations, (H.8) and (H.9).
4. Laplace domain pressure drop relationship, (H.43).
5. Laplace domain tray hydraulics equation.
   a. Laplace domain generalized tray hydraulics equation, (H.35), (not including the column top tray).
   b. Laplace domain column top tray hydraulics, (H.40).
FIG 13. Reboiler.
Reboiler (Stage j) Time Domain Relationships

The time domain equations which describe the dynamic behavior of the reboiler shown in FIG 13 are listed below:

(A) Process side.

(1) Total material balance.
\[ \frac{dM_j}{dt} = L_j - V_j - L_j \quad (4.51) \]

(2) Component material balances.
\[ \frac{d}{dt} (M_{ji} X_i) = X_i L_j - V_j Y_{ji} - L_j X_i \quad (4.53) \]

(3) Enthalpy balance.
\[ \frac{d}{dt} (M_{3j} H_{3j}) = H_{3j} L_j - H_V V_j - H_L L_j + U_i A_i (T_j - T_j) \quad (4.54) \]

(4) Equilibrium relationships.
\[ Y_{ij} = K_{ij} X_{ij} \quad (4.55) \]
\[ K_{ij} = \frac{s_{ij}}{P_{ij}} \quad (4.56) \]

(5) Thermal properties.
\[ H_L = f(T_j, X_{ij}) \quad (4.57) \]
\[ H V = f \left( P, T, X \right) \] (4.58)
\[ \sum_{i,j} \alpha_{i,j} = 1 \] (4.59)
\[ \sum_{i,j} \beta_{i,j} = 1 \] (4.60)

(6) Summation equations.

(7) Controller equation.

Assume the column base level controller controls the residue flow rate.

\[ L = KR \times H \] (4.61)

(B) Enthalpy balance for the reboiler metal.

\[ \frac{dT}{dt} = \frac{M \times C}{MR} \left( \frac{U \times A \times (T - T_B)}{S} - \frac{U \times A \times (T - T_B)}{S_M} \right) \] (4.62)

(C) Reboiler steam side.

(1) Total material balance.

\[ \frac{d\rho}{dt} \]
\[ \frac{V}{S C} = \frac{F}{S C} - \frac{F}{S C} \] (4.63)
\[ \frac{U \times A \times (T - T_B)}{S} \]
\[ \frac{F = \frac{U \times A \times (T - T_B)}{S} \times \frac{\lambda}{C}}{\frac{\rho}{S}} \]
\[ \frac{F = 0.0176 \times C \times \sqrt{\frac{(P - P_S)}{S}} \times \frac{\rho}{S}}{\frac{S}{SU}} \] (4.64) (4.65)

where:

\[ V \] : volume of the reboiler steam chest, FT**3.
\[ \rho \] : steam density in reboiler steam chest, LBM/FT**3.
\[ R \]
F : reboiler steam inlet mass flow rate, LBM/SEC.

F C : reboiler condensate outlet flow rate, LBM/SEC.

U S : reboiler steam side heat transfer coefficient, BTU/SEC, FT**2, F.

A S : reboiler steam side heat transfer area, FT**2.

T S : saturated steam temperature within the reboiler steam chest, F.

T MR : reboiler heat transfer metal temperature, F.

λ V : latent heat of the saturated steam within the steam chest, BTU/LBM.

C V : reboiler steam inlet control valve size.

C M : reboiler steam inlet control valve opening, range between 0 and 1.

P SO : steam supply pressure, PSI, to the reboiler.

P S : saturated steam pressure, PSI, within the reboiler steam chest.

ρ SU : reboiler steam supplied density, LBM/FT**3.

(2) Absolute pressure calculated from saturated vapor temperature (range : 280 F - 382 F)

\[
P_{S} = -3.0071444 + (0.028981108)\cdot T_{S} + (0.31012535E-10)\cdot T_{S}^{5}
\]
(3) Latent heat of saturated vapor calculated from saturated vapor temperature. (range 280 F - 382 F)

\[ \lambda = 1098.95665806 - (0.59359378)T + (0.12328925E-4)T^2 - (0.14660088E-8)T^4 \]  

(4.67)

(4) Density of steam calculated from saturated vapor temperature. (range 280 F - 382 F)

\[ \rho = (0.87273217E-3) + (0.17042733E-10)T - (0.32985211E-14)T^5 + (0.32587406E-16)T^6 \]  

(4.68)

where:

- \( T \) : saturated steam vapor temperature, F.
- \( P \) : absolute steam vapor pressure, PSI.
- \( \lambda \) : latent heat of saturated steam, BTU/LBM.
- \( \rho \) : saturated steam density, LBM/FT**3.

(d) Pressure drop through the bottom tray.

\[ \Delta P_{j-1} = P - P_j \]
\[ \Delta P_{DRY} = 0.0025888 \frac{g}{j-1} \frac{2}{3} \frac{LW}{j-1} \frac{\rho L}{j-1} + 0.002777 \frac{HW}{j-1} \frac{\rho L}{j-1} \left(\frac{g}{c}\right) \]  

(4.69)

(e) Size of the reboiler steam control valve

The following steps were used to size the reboiler steam control valve:

1. From the steady state of steam temperature and steam pressure, calculate the steam specific gravity.
2. Calculate the steady state steam volumetric flow rate in GPM.
3. Set the normalized valve opening to be 0.5.
4. Calculate the pressure drop across the steam control valve.
5. Solve for CV, the control valve sized by the design equation.

\[ V = CV \times CM \times \sqrt{\frac{P - p}{S_0 - S}} \left(\frac{S}{\text{sp. gr.}}\right) \]  

(4.70)

where:

- **CM** : valve opening, range between 0 and 1.
- **CV** : control valve size.
- **V** : steam volumetric flow rate, GPM.
- **S** :
\( P \): upstream supply steam pressure, PSI.

\( S_0 \)

\( P_S \): saturated steam pressure in reboiler, PSI.

\( S \)

sp.gr: specific gravity of the upstream steam.

(f) Liquid holdup at the column base.

The steady state liquid holdup at the distillation column base is sized for five minutes retention time of the residue product rate.

**Reboiler Laplace Domain Relationships**

If the above reboiler time domain equations are linearized, Laplace transformed and simplified (as shown in Appendix I), the following equations can be reached with variables, \( L_j(s) \), \( V_j(s) \), \( T_j(s) \), \( P_j(s) \) and \( NC_j X_{i,j} \).

The equations are:

(1) NC-1 Laplace domain component material balance, (I.6).

(2) Laplace domain total material balance, (I.4).

(3) Laplace domain vapor phase summation equation, (I.9).

(4) Laplace domain liquid phase summation equation, (I.10).

(5) Laplace domain metal and steam side enthalpy balance, (I.31).

(6) Laplace domain pressure drop relationship through the column bottom tray, (I.40).
Condenser (Total Condenser or Partial Condenser)

The time domain equations which describe the dynamic behavior of the condenser shown in FIG 14 are listed below.

condenser time domain relationships

(A) Process side (equilibrium flash calculation)

(1) Total material balance.

\[ V = L + V \]
\[ 2 C \]
\[ C \] (4.71)

(2) Component material balance (for \( i = 1, NC \)).

\[ V * Y = L * X + V * Y \]
\[ 2 i, 2 C C, i C C, i \] (4.72)

(3) Enthalpy balance.

\[ H V * V = L * H L + V * H V + U * A * (T - T) \]
\[ 2 2 C C C C C C C MC \] (4.73)

(4) Equilibrium relationships (for \( i = 1, NC \)).

\[ Y = K * X \]
\[ C, i C, i C, i \] (4.74)

s

\[ P \]
\[ C, i \]
\[ K = \frac{\text{-------}}{\text{-------}} \]
\[ C, i P 1 \] (4.75)

(5) Summation equations

\[ \Sigma X = 1 \]
\[ i C, i \] (4.76)
\[ \sum_{i} Y_i = 1 \]  
\[ (4.77) \]

(6) Pressure relationships.

\[ P = P_1 - \Delta P \]
\[ 1 \quad 2 \quad S \]  
\[ (4.78) \]

Kern (1950) suggested the following relationship to calculate the condenser shell side pressure drop.

\[ \frac{2}{S} \frac{f * G * D * (N+1)}{S} = \frac{\Delta P}{5.22E10 * D * S} \]
\[ (4.79) \]

where:

- \( \Delta P \): pressure drop through the condenser shell side, PSI.
- \( S \)
- \( N+1 \): number of shell side crosses.
- \( f \): Fanning friction factor.
- \( G \): mass velocity, LBM/SEC, FT**2.
- \( S \)
- \( D \): shell inside diameter, FT.
- \( S \)
- \( D_e \): equivalent diameter of tubes, FT.
- \( s \): average fluid specific gravity between the inlet and outlet of the condenser.

For simplicity, two reasonable assumptions were made in this research:

1. assume the Fanning friction \( f \) is constant.
2. assume average fluid specific gravity is constant.

Then, the dynamic shell side pressure drop can be related to
the steady state pressure drop by:

\[
\frac{\Delta P}{S} = \frac{G}{S} = \frac{\bar{G}}{S}
\]

where:

- \( CP \): steady state shell side pressure drop, PSI.
- \( \bar{G} \): steady state mass flow rate, LB/SEC.
- \( G \): mass flow rate, LB/SEC.

Since,

\[
G = V \cdot (\sum Y_i \cdot MW_i)
\]

(4.81)

\[
\bar{G} = \bar{V} \cdot (\sum \bar{Y}_i \cdot MW_i)
\]

(4.82)

Therefore;

\[
\frac{\Delta P}{S} = CP \cdot \frac{\left[ \frac{V}{2} \cdot (\sum Y_i \cdot MW_i) \right]^{\frac{2}{2}}}{\left[ \frac{\bar{V}}{2} \cdot (\sum \bar{Y}_i \cdot MW_i) \right]^{\frac{2}{2}}}
\]

(4.83)

(7) Process side temperature approximation.

\[
T = T_{C\; BC}
\]

(4.84)

(8) Thermal properties.
\[ \mathbf{H}_L = f(T, X_{c,i}) \]  
(4.85)

\[ \mathbf{H}_V = f(T, Y_{c,i}, P) \]  
(4.86)

(B) Enthalpy balance for the condenser metal.

\[
\frac{dT}{dt} = \frac{M*C}{MC} \cdot \Delta T_{MC} - \frac{W*C}{MC} \cdot \Delta T_{W} - \frac{W*C}{MC} \cdot \Delta T_{IN} + \frac{W*C}{MC} \cdot \Delta T_{OUT}
\]
(4.87)

(C) Condenser coolant side enthalpy balance.

\[
\frac{dT}{dt} = \frac{V*p*C}{W} \cdot \Delta T_{W} - \frac{U*A*(T - T)}{W} - \frac{U*A*(T - T)}{W} + \frac{U*A*(T - T)}{W} + \frac{U*A*(T - T)}{W}
\]
(4.88)

(D) Reflux drum equations.

(a) Vapor phase.

(1) Total material balance.

\[
\frac{dM}{dt} = V - SSV
\]
(4.89)

\[
M = \frac{V \cdot P}{D} \cdot \frac{R \cdot T}{V}
\]
(4.90)

where \( V \) is the vapor space of the drum plus all the \( D \) vapor space in the column, \( FT^{**3} \).

(2) Component material balance. (neglect the dynamics of drum vapor phase component material balances).

\[ Y_{i,1} = Y_{i,1} \]  
(4.91)
(3) Enthalpy balance. (Neglect the dynamics of drum vapor phase enthalpy balance).

\[
HV = HV \quad 1 \quad 1 \quad C \quad (4.92)
\]

\[
T = T \quad 1 \quad 1 \quad BC \quad (4.93)
\]

(b) Liquid phase.

(1) Total material balance.

\[
\frac{dM}{dt} = L - L - SSL \quad ; \quad M = A * \rho * H \quad (4.94)
\]

(2) Component material balance, for \( i = 1, NC \).

\[
\frac{d}{dt} (M * X_i) = L * X_i - X_i * (L + SSL) \quad (4.95)
\]

(3) Enthalpy balance.

\[
\frac{d}{dt} (M * H_L) = L * H_L - (L + SSL) * H_L \quad (4.96)
\]

(4) Controller equation.

(c.ase 1) \( SSL = KCL * H \quad (4.97) \)

(c.ase 2) \( L = KCL * H \quad (4.98) \)

(5) Summation equation.

\[
\sum_{i=1}^{NC} X_i = 1.0 \quad (4.99)
\]

The drum volume for a 2 : 1 ellipsoidal heads cylindrical vessel is,
\[ V = \frac{2 \pi^*D}{4} * (L + D/3) \quad (4.100) \]

Assume the drum liquid level is operated in the middle of the drum; therefore, the cross sectional area is:

\[ A = \frac{L*D^2}{1 + \frac{D}{8}} \quad (4.101) \]

**Condenser Laplace Domain Relationships**

If the above condenser time domain equations are linearized, Laplace transformed and simplified (as shown in Appendix J), the following equations can be reached with variables, \( T(s), P(s), L(s) \) or SSL \( (s) \), \( T(s) \) and NC \( C, i \) by eliminating \( X(s), H_{L}(s), H_{V}(s), T(s), P(s) \) and \( Y(s) \) -

\( \Delta P, L(s), V(s), M(s), H_{L}(s), H_{V}(s), T(s), Y(s) \)

\( C, i \) and \( C, i \) -

The equations are \( (J.31), (J.33), (J.34), (J.35), (J.36) \) or \( (J.37), NC-1 (J.32) \).
NOMENCLATURE FOR THE DYNAMIC DISTILLATION COLUMN

\[ A = \text{reboiler process side heat transfer area, } \text{FT}^2. \]
\[ B = \text{condenser process side heat transfer area, } \text{FT}^2. \]
\[ c = \text{downcomer area, } \text{FT}^2, \text{ of tray } n. \]
\[ \text{ADM} = \text{free area, } \text{FT}^2, \text{ for liquid flow under the downcomer apron, normal tray construction practice yields approximately } 0.42*\text{ADC}, \text{ tray } n. \]
\[ \text{AH} = \text{total hole area, } \text{FT}^2, \text{ of tray } n. \]
\[ A = \text{cross sectional area, } \text{FT}^2, \text{ of the distillation tower base.} \]
\[ \text{ANTA} = \text{Antoine coefficients of component } i \text{ temperature in KELVIN and pressure in MMHG.} \]
\[ S = \text{reboiler steam side heat transfer area, } \text{FT}^2. \]
\[ \text{ATR} = \text{active tray area, } \text{FT}^2, \text{ of tray } n. \]
\[ A = \text{cross sectional area, } \text{FT}^2, \text{ of the reflux drum.} \]
\[ 1 = \text{since the liquid level is adjusted by the liquid level controller, the cross sectional area is assumed to be constant.} \]
\[ C = \text{heat capacity, BTU/LBM,F, of the metal of the condenser.} \]
\[ \text{MC} = \text{steady state condenser shell side pressure drop, PSI.} \]
\[ \text{CV} = \text{valve coefficient of the reboiler steam inlet control valve.} \]
\[ W = \text{heat capacity, BTU/LBM,F, of the coolant in the condenser.} \]
\( F_n \) = kinetic energy parameter of tray \( n \).

\( F_C \) = condensate flow rate, LBM/SEC, from the reboiler steam trap.

\( FL_n \) = feed liquid, LBMOLE/SEC, to tray \( n \).

\( FS \) = reboiler steam inlet flow rate, LBM/SEC.

\( FV_n \) = feed vapor, LBMOLE/SEC, to tray \( n \).

\( FW \) = condenser coolant inlet flow rate, LBM/SEC.

\( hdn \) = liquid head in inches of liquid flow pressure drop through the downcomer, tray \( n \).

\( HDC_n \) = liquid height in the downcomer, FT, of tray \( n \).

\( HF_n \) = total feed enthalpy, BTU/LBMOLE, to tray \( n \).

\( HPL_n \) = liquid feed enthalpy, BTU/LBMOLE, to tray \( n \).

\( HFV_n \) = vapor feed enthalpy, BTU/LBMOLE, to tray \( n \).

\( H_J \) = liquid height, FT, at the distillation tower base.

\( HL_n \) = liquid enthalpy, BTU/LBMOLE, of tray \( n \).

\( HL_C \) = liquid enthalpy, BTU/LBMOLE, leaving the condenser.

\( HM_n \) = construction metal enthalpy, BTU/LBM, of tray \( n \).

\( HOW_n \) = height, FT, of crest of liquid above the outlet weir of tray \( n \).

\( HTR_n \) = height, FT, of aerated liquid on tray \( n \) just downstream of the inlet weir.
\( H_{V}^{C} \) = vapor enthalpy, BTU/LBMOLE, leaving the condenser.

\( H_{V}^{n} \) = vapor enthalpy, BTU/LBMOLE, leaving tray \( n \).

\( H_{W}^{n} \) = outlet weir height, FT, of tray \( n \).

\( H_{L}^{1} \) = liquid level, FT, in the reflux drum.

\( K_{i,n} \) = equilibrium coefficient of component \( i \) of tray \( n \).

\( K_{C}^{i} \) = condenser's liquid level transmitter gain.

\( K_{R}^{i} \) = equilibrium coefficients of component \( i \) of the stream leaving the condenser.

\( K_{R} \) = reboiler's liquid level transmitter gain.

\( L_{C}^{n} \) = liquid flow rate, LBMOLE/SEC, leaving tray \( n \).

\( L_{C} \) = liquid flow rate, LBMOLE/SEC, leaving the condenser.

\( L_{W}^{n} \) = weir length, INCHES, tray \( n \).

\( M_{n} \) = liquid hold up, LBMOLE, of tray \( n \).

\( M_{n}^{n} \) = construction material mass, LBM, of tray \( n \).

\( M_{D} \) = reflux drum vapor phase holdup, LBMOLE.

\( M_{j} \) = liquid hold up, LBMOLE, at the distillation tower base.

\( M_{MC} \) = metal mass, LBM, of the condenser.

\( M_{W}^{i} \) = molecular weight of process component \( i \).
\[ M \]
\[ = \text{liquid holdup, LBMOL, of the reflux drum.} \]

\[ \text{NC} \]
\[ = \text{number of process components.} \]

\[ \text{P} \]
\[ n \]
\[ = \text{pressure, PSI, of tray } n. \]

\[ \text{P} \]
\[ C \]
\[ = \text{pressure, PSI, of the condenser.} \]

\[ \text{P} \]
\[ S \]
\[ = \text{reboiler steam side saturated steam pressure, PSI.} \]

\[ \text{P} \]
\[ S \]
\[ i, n \]
\[ = \text{saturated vapor pressure, PSI, of component } i \text{ of tray } n. \]

\[ \text{P} \]
\[ S \]
\[ 0 \]
\[ = \text{reboiler steam supply pressure, PSI.} \]

\[ \text{P} \]
\[ 1 \]
\[ = \text{pressure, PSI, of reflux drum.} \]

\[ \text{Q} \]
\[ n \]
\[ = \text{inter stage heater/cooler heat duty, BTU/SEC, of stage } n. \]

\[ \text{QDC} \]
\[ n \]
\[ = \text{downcomer flow, GPM, tray } n. \]

\[ \text{QQ} \]
\[ n \]
\[ = \text{overflow, FT}^3/\text{SEC, of aerated liquid.} \]

\[ \text{R} \]
\[ = \text{gas constant, } 10.73 \text{ (PSI*FT}^3)/(\text{LBMOL*R}). \]

\[ \text{SSL} \]
\[ n \]
\[ = \text{sidestream liquid, LBMOL/SEC, from tray } n. \]

\[ \text{SSV} \]
\[ n \]
\[ = \text{sidestream vapor, LBMOL/SEC, from tray } n. \]

\[ \text{T} \]
\[ n \]
\[ = \text{process temperature, F, of tray } n. \]

\[ \text{T} \]
\[ BC \]
\[ = \text{stream temperature, F, leaving the condenser.} \]

\[ \text{T} \]
\[ C \]
\[ = \text{condenser process side temperature, F.} \]
\( T_{F_n} \) = feed temperature, \( F \), of tray \( n \).

\( T_{MC} \) = condenser heat transfer metal temperature, \( F \).

\( T_{MR} \) = reboiler heat transfer metal temperature, \( F \).

\( T_{S} \) = reboiler steam side saturated steam temperature, \( F \).

\( T_{V} \) = \( TV \) equals to \( T_{BC} \).

\( T_{W} \) = average condenser coolant temperature, \( F \).

\( T_{W_{IN}} \) = condenser coolant inlet temperature, \( F \).

\( T_{W_{OUT}} \) = condenser coolant outlet temperature, \( F \).

\( U_B \) = reboiler process side convective heat transfer coefficient, \( \text{BTU}/\text{FT}^2\text{,F,SEC.} \)

\( U_C \) = condenser process side convective heat transfer coefficient, \( \text{BTU}/\text{FT}^2\text{,F,SEC.} \)

\( U_{H_n} \) = vapor velocity, \( \text{FT}/\text{SEC} \), through the holes of tray \( n \), \( \frac{WV}{(\rho V_n)^{1/2}AH_n} \).

\( U_S \) = reboiler steam side convective heat transfer coefficient, \( \text{BTU}/\text{FT}^2\text{,F,SEC.} \)

\( U_{VA_n} \) = vapor velocity, \( \text{FT}/\text{SEC} \), corresponding to \( \text{ATR} \), \( \frac{WV_n}{(\rho V_n)^{-0.5}}/\text{ATR} \).

\( U_W \) = condenser water side convective heat transfer coefficient, \( \text{BTU}/\text{FT}^2\text{,F,SEC.} \)

\( V_n \) = vapor flow rate, \( \text{LBMOLE/SEC} \), leaving tray \( n \).

\( V_C \) = vapor flow rate, \( \text{LBMOLE/SEC} \), leaving the condenser.
\( V_D \) = volume, \( \text{FT}^3 \), of the reflux drum vapor space plus all the vapor space in the column.

\( V_{TR\,n} \) = volume, \( \text{FT}^3 \), of aerated liquid on active area of tray \( n \) (downcomer area excluded).

\( V_W \) = volume, \( \text{FT}^3 \), of the condenser coolant side.
(i.e., the volume between the shell and the tubes)

\( W_n \) = overflow, \( \text{LBM/SEC} \), from tray \( n \).

\( W_{DC\,n} \) = flow, \( \text{LBM/SEC} \), from the downcomer of tray \( n \).

\( W_{FL\,n} \) = feed liquid, \( \text{LBM/SEC} \), to tray \( n \).

\( W_{FV\,n} \) = feed vapor, \( \text{LBM/SEC} \), to tray \( n \).

\( W_{L\,n} \) = weir length, \( \text{FT} \), of tray \( n \).

\( W_{SSL\,n} \) = sidestream liquid, \( \text{LBM/SEC} \), from tray \( n \).

\( W_{SSV\,n} \) = sidestream vapor, \( \text{LBM/SEC} \), from tray \( n \).

\( W_{TR\,n} \) = holdup, lbm, on the active tray

\( W_{V\,n} \) = vapor rate, \( \text{LBM/SEC} \), leaving tray \( n \).

\( X_i\,n \) = liquid mole fraction of component \( i \) of tray \( n \).

\( X_C\,i \) = mole fraction of component \( i \) of the liquid stream leaving the condenser.

\( Y_i\,n \) = vapor mole fraction of component \( i \) of tray \( n \).

\( Y_C\,i \) = vapor mole fraction of component \( i \) leaving the condenser.
$Z_{i,n}$ = total feed mole fraction of component $i$ to tray $n$.

$Z_{L_{i,n}}$ = liquid feed mole fraction of component $i$ to tray $n$.

$Z_{V_{i,n}}$ = vapor feed mole fraction of component $i$ to tray $n$.

$\rho_{V_{n}}$ = vapor density, LBM/FT$^3$, corresponding to $HV_{n}$.

$\rho_{C_{n}}$ = clear liquid density, LBM/FT$^3$, of tray $n$.

$\rho_{DC_{n}}$ = liquid density, LBM/FT$^3$, in the downcomer.

$\rho_{j}$ = liquid density, LBMOLE/FT$^3$, at the distillation tower base.

$\rho_{S}$ = reboiler steam side saturated steam density, LBM/FT$^3$.

$\rho_{TR_{n}}$ = aerated liquid density, LBM/FT$^3$, of tray $n$.

$\rho_{W}$ = condenser coolant density, LBM/FT$^3$.

$\rho_{1}$ = liquid density, LBMOLE/FT$^3$, in the reflux drum.

$\lambda_{V}$ = latent heat of the saturated steam, BTU/LBM.

$\alpha_{1,i}, \alpha_{2,i}, \alpha_{3,i}, \alpha_{4,i}, \alpha_{5,i}$

= constant coefficients for pure component $i$ vapor enthalpy calculation.

$\phi_{1,i}, \phi_{2,i}, \phi_{3,i}, \phi_{4,i}, \phi_{5,i}$

= constant coefficients for pure component $i$ liquid enthalpy calculation.

$\bar{v}$ = bar on top of a variable means the steady state value of that variable.
(S) = if (S) following a variable that means variable in LAPLACE domain transformation notation.

\( \Delta P \) = pressure drop, PSI, through the condenser shell side.

\( \Delta P_{DC}^n \) = downcomer pressure drop, LBF/FT**2, of tray n.
CHAPTER FIVE

FREQUENCY RESPONSE ANALYSIS

In chapter four, the dynamic performance of the distillation column has been derived in Laplace domain in terms of the complex frequency variable s. In this research, the method of frequency response will be used to analyze the distillation column. The frequency response of a system is defined as the steady state response of the system to a sinusoidal input signal. The sinusoid is a unique input signal, and the resulting output signal for a linear system is sinusoidal in the steady state. It differs from the input waveform only in amplitude and phase angle. The advantage of the frequency response method is that the transfer function describing the sinusoidal steady state behavior of a system can be obtained by replacing s with \( j\omega \) in the system transfer functions. The transfer function representing the sinusoidal steady state behavior of the system is a complex function in \( j\omega \) which possesses a magnitude and phase angle. The magnitude and phase angle are then readily represented by graphical plot, such as Bode plot in this research, which provide a significant insight for the analysis of process or design of the control systems.
In this chapter, all the frequency domain relationships derived in chapter four for the condenser, the generalized tray and the reboiler will be considered as a whole. The numerical method will also be discussed to show how one can solve for the process transfer functions from these relationships in the frequency domain. The resulting process transfer functions are presented as Bode plots in chapter six. The Bode plots present the two components of a transfer function, amplitude ratio and phase angle, as functions of frequency. These plots completely define the input/output relationship for either manipulated variable / measured variable or disturbance variable / measured variable. The variables in the distillation tower can be classified into three categories, the measured variables, manipulated variables and disturbance variables.

(A) Manipulated variables.

The manipulated variables are those variables that can be changed in order to control the process. For the distillation tower, the possible manipulated variables are:

(1) sidestream liquid flow rate of tray n
(2) sidestream vapor flow rate of tray n
(3) condenser coolant flow rate
(4) reboiler steam control valve opening
(5) inter-stage heater/cooler heat load
(6) liquid reflux from the reflux drum
(7) distillate vapor flow rate (partial condenser)
(8) distillate liquid flow rate

(B) Disturbance variables.

The disturbance variables are input variables to the process that can't be controlled. For the distillation tower, the possible disturbance variables are:

1. feed temperature of tray n
2. liquid feed flow rate to tray n
3. vapor feed flow rate to tray n
4. liquid feed composition i of tray n
5. vapor feed composition i of tray n
6. condenser coolant inlet temperature
7. reboiler steam supply temperature

(C) Measured variables, (controlled variables).

Measured variables are those variables which can be measured by the measuring devices. The possible measured variables for the distillation tower are:

1. liquid composition i of tray n
2. vapor composition i of tray n
3. temperature of tray n
4. liquid flow rate leaving tray n
5. vapor flow rate leaving tray n
6. pressure of tray n
Process Transfer Functions And Frequency Response Analysis

If only the process is considered, i.e. not including any control systems, process transfer functions are the transfer functions between the measured variables and the manipulated variables or between the measured variables and the disturbance variables.

Consider the two variable open loop system with a load disturbance variable $L(S)$ sketched in FIG 15. The system is described by the following equations in the Laplace domain.

\[
X_1(S) = M_1(S)G_{11}(S) + M_2(S)G_{21}(S) + L_1(S)G_{L1}(S) \tag{5.1}
\]
\[
X_2(S) = M_1(S)G_{12}(S) + M_2(S)G_{22}(S) + L_2(S)G_{L2}(S) \tag{5.2}
\]

where:

- $X_1(S), X_2(S)$: Laplace domain controlled variables (or measured variables).
- $M_1(S), M_2(S)$: Laplace domain manipulated variables
- $G_{11}(S), G_{21}(S), G_{12}(S), G_{22}(S), G_{L1}(S), G_{L2}(S)$ are the process transfer functions between either the measured variable/manipulated variable or measured variable/disturbance variable as specified below.
FIG 15. Block Diagram For A 2x2 Open Loop System With A Load Disturbance.
\[
\begin{align*}
X(s) & \quad X(s) \\
G(s) = \frac{1}{M(s)} & \quad G(s) = \frac{1}{M(s)} \\
G(s) = \frac{2}{M(s)} & \quad G(s) = \frac{2}{M(s)} \\
G(s) = \frac{1}{L(s)} & \quad G(s) = \frac{2}{L(s)}
\end{align*}
\]

L(s) : Laplace domain disturbance variable.

Refer to FIG 15 in frequency domain, by simply setting

\[M(j\omega) = 1+0i, L(j\omega) \text{ and } M(j\omega) = 0+0i\]

at each frequency, 1

then, the value of the transfer function \[G(j\omega), G(j\omega)\]

at any specific frequency can be calculated as \[X(j\omega)\]

and \[X(j\omega)\]

respectively. Similarly, the value of process transfer

function \[G(j\omega), G(j\omega)\]

at each frequency can be

\[L(j\omega) = 0+0i, M(j\omega) = 1+0i \text{ and } M(j\omega) = 1+0i\]

L(j\omega) = 0+0i. For the process transfer function \[G(j\omega), G(j\omega)\]

L1

G(j\omega), the settings are \[M(j\omega) = 0+0i \text{ and } M(j\omega) = 0+0i\]

L2

0+0i, and L(j\omega) to 1+0i.

If the Laplace operator s in all the relationships

which describe the distillation column are replaced by j\omega,

they become complex algebraic equations in the frequency

domain. The manipulated and disturbance variables are

input variables, constants as either 1+0i or 0+0i, and all

the measured variables are unknowns to be solved at each
frequency.

In summary, the condenser has $NC+4$ unknowns within $NC+4$ complex number equations.

<table>
<thead>
<tr>
<th>equation</th>
<th>no</th>
<th>unknown</th>
<th>no</th>
</tr>
</thead>
<tbody>
<tr>
<td>(J.31)</td>
<td>1</td>
<td>$T(jw)$</td>
<td>1</td>
</tr>
<tr>
<td>(J.32)</td>
<td>$NC-1$</td>
<td>$P(jw)$</td>
<td>1</td>
</tr>
<tr>
<td>(J.33)</td>
<td>1</td>
<td>$L(jw)$ or $SSL(jw)$</td>
<td>1</td>
</tr>
<tr>
<td>(J.34)</td>
<td>1</td>
<td>$X(jw)$</td>
<td>$NC$</td>
</tr>
<tr>
<td>(J.35)</td>
<td>1</td>
<td>$T(jw)$</td>
<td>1</td>
</tr>
<tr>
<td>or (J.37)</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

The column top tray has $NC+3$ unknowns within $NC+3$ complex domain equations.

<table>
<thead>
<tr>
<th>equation</th>
<th>no</th>
<th>unknown</th>
<th>no</th>
</tr>
</thead>
<tbody>
<tr>
<td>component mat. bal.</td>
<td>$NC-1$</td>
<td>$L(jw)$</td>
<td>1</td>
</tr>
<tr>
<td>enthalpy balance</td>
<td>1</td>
<td>$V(jw)$</td>
<td>2</td>
</tr>
<tr>
<td>summation equation 1</td>
<td>1</td>
<td>$T(jw)$</td>
<td>2</td>
</tr>
<tr>
<td>summation equation 2</td>
<td>1</td>
<td></td>
<td></td>
</tr>
<tr>
<td>tray hydraulic eq.</td>
<td>1</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>$NC$</td>
<td>$X(jw)$</td>
<td>$i,2$</td>
</tr>
</tbody>
</table>

Also, each generalized tray has $NC+4$ unknowns within
NC+4 complex domain equations.

<table>
<thead>
<tr>
<th>equation</th>
<th>no</th>
<th>unknown</th>
<th>no</th>
</tr>
</thead>
<tbody>
<tr>
<td>component mat. bal.</td>
<td>(H.4) NC-1</td>
<td>L (jw)</td>
<td>1</td>
</tr>
<tr>
<td>enthalpy balance</td>
<td>(H.6) 1</td>
<td>V (jw)</td>
<td>n</td>
</tr>
<tr>
<td>summation equation 1</td>
<td>(H.8) 1</td>
<td>T (jw)</td>
<td>n</td>
</tr>
<tr>
<td>summation equation 2</td>
<td>(H.9) 1</td>
<td>X (jw)</td>
<td>NC</td>
</tr>
<tr>
<td>tray hydraulic eq.</td>
<td>(H.35) 1</td>
<td>P (jw)</td>
<td>1</td>
</tr>
<tr>
<td>pressure equation.</td>
<td>(H.43) 1</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>NC+4</td>
<td></td>
<td>NC+4</td>
</tr>
</tbody>
</table>

The reboiler has NC+3 unknowns within NC+3 complex domain equations.

<table>
<thead>
<tr>
<th>equation</th>
<th>no</th>
<th>unknown</th>
<th>no</th>
</tr>
</thead>
<tbody>
<tr>
<td>component mat. bal.</td>
<td>(I.6) NC-1</td>
<td>L (jw)</td>
<td>j</td>
</tr>
<tr>
<td>total mat. balance</td>
<td>(I.4) 1</td>
<td>V (jw)</td>
<td>j</td>
</tr>
<tr>
<td>summation equation 1</td>
<td>(I.9) 1</td>
<td>T (jw)</td>
<td>j</td>
</tr>
<tr>
<td>summation equation 2</td>
<td>(I.10) 1</td>
<td>X (jw)</td>
<td>NC</td>
</tr>
<tr>
<td>metal and steam side</td>
<td></td>
<td>P (jw)</td>
<td>j</td>
</tr>
<tr>
<td>enthalpy balance</td>
<td>(I.37) 1</td>
<td></td>
<td></td>
</tr>
<tr>
<td>pressure equation</td>
<td>(I.40) 1</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>NC+4</td>
<td></td>
<td>NC+4</td>
</tr>
</tbody>
</table>

Totally for each distillation column, there are

NS*(NC+4)-1 equations and NS*(NC+4)-1 unknowns which are
the measured variables. If all these frequency domain equations are arranged in order from the first stage to the last stage, they form linear algebraic complex number equation set at each specific frequency. This algebraic equation set can be expressed by matrix notation as:

\[
A \times X = B \quad (5.3)
\]

where:

- **A**: coefficient matrix of the algebraic equation set in the complex domain.
- **X**: column vector which contains all the measured variables to be solved.
- **B**: constant column vector of the algebraic equation set in the complex domain.

If the above equation set is solved, then,

\[
X = A^{-1} \times B \quad (5.4)
\]

For illustration purposes, the following example distillation column has been chosen, FIG 16. This example distillation column consists of six stages and three process components. The first stage is a total condenser and the last stage is the reboiler. The reflux drum level controller controls the liquid reflux and the column base level controller controls the residue rate. If all frequency domain linear relationships are lined up from
FIG 16. Example Distillation Column
FIG 17. Coefficient Matrix Of The Example Distillation Column.
the first stage to the last stage, the coefficient matrix $A$
for this particular system is as shown in FIG 17.

**Guassian Elimination With The Method Of Residues**

In this research, Guassian elimination with total scale pivoting is used to minimize the effects of the round off error in solving the distillation column complex domain equation set at each specific frequency. Probably in the majority of cases, the method will be a fairly accurate approximation of the true inverse. However, if the matrix is ill conditioned, the method of residues is used to improve the accuracy.

The method of residues is an iterative procedure and is illustrated as follows:

Let $X$ be the approximate solution of $A^1 * X = B$ obtained by the Gaussian elimination with total scaled pivoting. By substituting $X$ into the the system $A^1 * X = B$ again, one can compute an error vector residue $E$ defined by:

$$A^1 * X - B = E \quad (5.5)$$

Then, a correction vector $X$ can be defined such that,

$$A^1 * \delta X = - E \quad (5.6)$$
By adding \( \delta X \) to \( X \), one can compute another error residue \( E_1 \) and another \( X_2 \).

\[
A \times (X + \delta X) - B = E_2
\]  \hspace{1cm} (5.7)

The process is continued until the \( \delta X \) vector change is smaller than the tolerance.

**Computer Memory Storage Of The Sparse Coefficient Matrix**

In order to minimize the computer memory space, the sparse coefficient matrix \( A \) can be stored in a band shape coefficient matrix. But twice of the memory of the band shape matrix is needed to solve the equation set by using Guassian elimination with total scaled pivoting. Define this bigger storage matrix as \( B \), as shown in FIG 18. Since the matrix \( B \) is in a tilted shape, a new coordinate system must be defined.

The following general relationships define the pressure model coefficient positions in the coefficient matrix \( B \). \( NC \) is the number of process components and \( NS \) is the number of distillation column stages.

(A) **CONDENSER**

(a) Define \( \text{IPX} \), the leftmost row position of the band shape matrix.
FIG 18. Band Shape Coefficient Matrix for the Example Distillation Column.
(1) Reflux drum liquid phase enthalpy balance, (J.34).
   \[ IPX = 4*NC + 15 \]

(2) Condenser process side component material balance II
   (J.32).
   \[ IPX = 4*NC + 15 + II \]

(3) Summation equation 2, (J.33).
   \[ IPX = 5*NC + 15 \]

(4) Reflux drum liquid phase total material balance,
   (J.36) or (J.37).
   \[ IPX = 5*NC + 16 \]

(5) Condenser process side enthalpy balance, (J.35).
   \[ IPX = 5*NC + 17 \]

(6) Summation equation 1, (J.31).
   \[ IPX = 5*NC + 18 \]

(b) Generalized formula for positions of variables.

\[
\begin{align*}
T_1 &= B(IPX, 1) \\
L_1 &= B(IPX-1, 2) \text{ or } SSL \text{ } _1 = B(IPX-1, 2) \\
T_2 &= B(IPX-2, 3) \\
BC &= B(IPX-2-IC, IC+3) \\
X_{IC, 1} &= B(IPX-IC, IC+3)
\end{align*}
\]

The stage two variable coefficient positions can be calculated by using the generalized variable position formulas, which are defined in generalized stage (c), with \( M=0 \) and \( IPYKK=NC+4 \).
(B) STAGE 2

(a) Define IPX, the leftmost row position of the band shape matrix.

(1) Component material balance II, (H.4).

\[ IPX = 5*NC + 18 + II \]

(2) Enthalpy balance, (H.6).

\[ IPX = 6*NC + 18 \]

(3) Summation equation 1, (H.8).

\[ IPX = 6*NC + 19 \]

(4) Summation equation 2, (H.9).

\[ IPX = 6*NC + 20 \]

(5) Tray hydraulic equation, (H.40).

\[ IPX = 6*NC + 21 \]

(b) Generalized formula for positions of variables.

\[
\begin{align*}
T_1 & = B(IPX, 1) \\
L_{1} & = B(IPX-1,2) \text{ or } SSL_{1} = B(IPX-1,2) \\
T_{BC} & = B(IPX-2,3) \\
X_{IC,1} & = B(IPX-2-IC, IC+3)
\end{align*}
\]

The stage two and three variable coefficient positions can be calculated by using generalized variable position formulas, which defined in generalized stage (c), with M = 0 or 1 and IPYKK = NC + 4.
(C) GENERALIZED STAGE J \((3 < J < NS-1)\)

(a) Define \((IPX, IPYKK)\) pair

\[
IPX = 6*NC + 22
\]

\[
IPYKK = (J-2)*(NC+4)
\]

(b) Define \(KK\)

1. Component material balance II, \((H.4)\).

\[
KK = NC + 4 - II
\]

2. Pressure equation, \((H.43)\).

\[
KK = 4
\]

3. Enthalpy balance, \((H.6)\).

\[
KK = 3
\]

4. Summation equation 1, \((H.8)\).

\[
KK = 2
\]

5. Summation equation 2, \((H.9)\).

\[
KK = 1
\]

6. Tray hydraulic equation, \((H.35)\).

\[
KK = 0
\]

(c) Generalized formula for positions of variables.

\[
M = (J-1, J, J+1)
\]

\[
L_{\text{M}} = B((IPX-KK)-(NC+4)*(M-J+1),
\]

\[
IPYKK + (NC+4)*(M-J+1))
\]

\[
V_{\text{M}} = B((IPX-KK)-(NC+4)*(M-J+1)-1,
\]

\[
IPYKK + (NC+4)*(M-J+1)+1)
\]
\[
T = B \left((IPX-KK)-(NC+4)\times(M-J+1)-2,\right. \\
\left.IPYKK+(NC+4)\times(M-J+1)+2\right)
\]

\[
P = B \left((IPX-KK)-(NC+4)\times(M-J+1)-3,\right. \\
\left.IPYKK+(NC+4)\times(M-J+1)+3\right)
\]

\[
X = B \left((IPX-KK)-(NC+4)\times(M-J+1)-(IC+3),\right. \\
\left.IPYKK+(NC+4)\times(M-J+1)+(IC+3)\right)
\]

**REBOILER**

(a) Define \((IPX, IPYKK)\) pair

\[
IPX = 6\times NC + 22 \\
IPYKK = (J-2)\times (NC+4)
\]

(b) Define \(KK\)

1. Total material balance, (I.4).
   \(KK=NC+3\)
2. Component material balance II, (I.6).
   \(KK=(NC-II) +3\)
   \(KK=3\)
4. Summation equation 1, (I.10).
   \(KK=2\)
5. Summation equation 2, (I.9).
   \(KK=1\)
6. Pressure drop equation, (I.40).
   \(KG=0\)

(c) Generalized formula

(same as the generalized stage formulas)
Measure Of Process Interaction

The original approach for characterizing process interactions is the "Relative Gain Array Method" proposed by Bristol (1966). The major advantage of the Relative Gain Array is that it requires only the process gains.

Consider a steady state process model.

\[ \dot{X} = K \times M \]  \hspace{1cm} (5.8)

Where:

- \( \dot{X} \) : vector of steady state changes for \( n \) controlled variables.
- \( M \) : vector of steady state changes for \( n \) manipulated variables.
- \( K \) : \( n \times n \) steady state process gain matrix.

Bristol defined a \( n \times n \) relative gain matrix whose element \( \lambda \) are calculated from,

\[ \lambda_{i,j} = \frac{\partial X_i}{\partial M_{i,j}} \]  \hspace{1cm} (5.9)

Where the \( (\partial X / \partial M)_{i,j} \) is called the open loop gain with all
other manipulated variables held constant except \( \mathbf{M} \).

\( \frac{\partial \mathbf{X}}{\partial \mathbf{M}} \) is called the closed loop gain with all other
measured variables held constant except \( \mathbf{X} \).

It is apparent from equation (5.8) that the open loop
gain \( \mathbf{K}_{i,j} \) between \( \mathbf{X} \) and \( \mathbf{M} \) is an element of the process
\( \mathbf{K} \) gain matrix, \( \mathbf{K} \). Bristol also demonstrated that the close
loop gain is \( \frac{1}{\mathbf{K}} \), where \( \mathbf{K} \) is an element of the matrix
\( \mathbf{K} \) inverse of the transpose of \( \mathbf{K} \).

\[
\tilde{\mathbf{K}} = \{ \tilde{\mathbf{K}}_{i,j} \} = \{ \mathbf{K} \}^{-1 T}
\]

Therefore;

\[
\lambda = \mathbf{K}_{i,j} * \tilde{\mathbf{K}}_{i,j}
\]

The relative gain elements are measures of the influence of
the manipulated variable on the controlled variables, with
the larger (more positive) values indicating the greater
effect. Also, as pointed out by Bristol, each row and
column of the interaction measure sums to one. A value of
interaction one indicates a decoupled situation. For
example,

\[
\begin{array}{ccc}
\mathbf{M} & \mathbf{M} \\
1 & 2 \\
\hline
\mathbf{X} & 0.25 & 0.75 \\
1 & \\
\mathbf{X} & 0.75 & 0.25 \\
2 & \\
\end{array}
\]
This suggested that $X$ should be manipulated through $M$ and
\[ \frac{1}{2}, \frac{2}{1} \]
$X$ should be manipulated through $M$. 

When a relative gain exceeds 1.0, the denominator of equation (5.9) is less than the numerator meaning that the closure of other loops acts to reduce the gain of the pair being evaluated. As pointed out by Bristol (1966), Shinskey (1977) and Gagnepain (1979), a controller pairing which corresponds to negative gain elements should not be selected since it results in an unstable system if integral action is used. The above properties suggested the measure corresponding to the paired variables be positive and as close to one as possible. Numbers negative or much larger than one are to be avoided and large negative numbers are particularly undesirable.

Although the "Relative Gain Array Method" provides useful information, it neglects the process dynamics. In some situations, the dynamic characteristics may be the key factor in determining the correct controller configurations. In this research, the process transfer functions generated at each frequency will be used to calculate the interaction measure. McAvoy (1977) pointed out the following messages for the interaction measure:

"The best way to interpret the interaction measure..... is in the frequency domain. If $s$ is set to $j\omega$ then the interaction measure can be represented as a
magnitude and a phase angle. The magnitude can be interpreted in exactly the same way as the standard relative gain array. It represents the ratio of the effect of one manipulative variable on one controlled variable under two different situations at a particular frequency. The two situations are again an open loop situation and a closed loop situation in which perfect control is achieved at the frequency being considered. Thus, the magnitude of the interaction measure is a measure of the extent of interaction present at a particular frequency. The phase angle is somewhat more difficult to interpret since it is a measure of the time it takes a disturbance to propagate from loop to loop. As such, it is concerned with whether a disturbance arrives at the correct time and is therefore favorable."

The interaction measure used in this research is based on exactly what McAvoy is recommending.
CHAPTER SIX

RESULTS

Introduction

The objective of this research is to set up a general distillation column package in the frequency domain in order to (1) test different model assumptions and their effects, (2) determine the steady state sensitivities, and (3) explore the interaction measure analysis in the frequency domain for the pairings of the distillation column multiple control loops.

In this chapter, interaction measures of the Benzene - Toluene distillation column are calculated based on the generated process transfer functions between the chosen manipulated and controlled variables to predict the pairing difficulties.

The effect of the distillation column model simplification on process transfer functions are compared to determine the effect of the constant pressure and equimolar overflow assumptions. The resulting transfer functions are presented as Bode plots.

A steady state gain analysis of the Benzene - Toluene distillation column was made to demonstrate how the
A computer program can be used to study the steady state sensitivity which can help the design of the control systems.

Two columns are presented in this research, the first distillation column is a Benzene - Toluene system which consists of nineteen trays, a horizontal reflux drum, a horizontal total condenser and a kettle reboiler. The feed introduced into the column at the 14th tray is all liquid which contains 36.29 mole % of Benzene. The second column is a Propane - Butane - Pentane distillation column. There are eleven stages including the horizontal total condenser and the kettle reboiler. The total condenser is stage one and the reboiler is stage eleven. The reflux drum is horizontal. Feed stage is stage five and the feed is all liquid which contains 40 mole % Propane, 25 mole % normal Butane and 35 mole % normal Pentane. The steady state profiles and the mechanical design results for these two columns are listed in the mechanical design manual, Appendix L, as examples one and three.

Steady State Gain Analysis Of The Benzene-Toluene Column

The steady state gain of the process transfer function can be defined as the ratio of the steady state change of the output variable over the change of the input variable.
A positive steady state gain means an increase in the output for an increase in input. A negative steady state gain means a decrease in the output for an increase in input. If the steady state gains of the chosen pairs of controlled variables and manipulated variables are given, the simple "Relative Gain Array" proposed by Bristol (1966), as mentioned in chapter five, can offer a measure of the interaction that exists between the loops.

Tables 2 To 8 constitute the steady state gain analysis of the Benzene - Toluene distillation column. The reflux drum level controller adjusts the reflux and the column base level controller adjusts the residue flow rate. These two level control loops are included as part of the distillation column process. In these Tables, T(1) and L(1) are the steady state gains of the reflux drum temperature and of the reflux flow rate, respectively. TBC is the steady state gain of the process stream temperature leaving the condenser. L, V, T, P and X are respectively the steady state gains of the liquid flow rate, vapor flow rate, temperature, pressure and liquid mole fraction of each tray corresponding to the tray number to the left of each table.

The following physical descriptions of the response of the Benzene - Toluene distillation column subject to different manipulated variables or disturbance variables
are based on the signs of the steady state gains in each corresponding table. The values of the steady state gains were verified by comparison with gains generated by the steady state program.

Table 2: Increase of the reboiler supply steam temperature.

As the reboiler steam side condensation temperature rises, more heat is transferred into the column base. This higher heat transfer rate brings up the column temperature, the column pressure and the vapor flow rate, as shown by the positive signs of the steady state in the table under label T, P and V. Since the distillate and residue flow rates are kept constant, the reflux increases and separation improves.

Table 3: Increase of the coolant flow rate.

When the coolant flow rate increases, more heat is carried out by the coolant, i.e. the condenser duty increases, lowering the column temperature, as shown by the negative signs of the steady state gains in the table under temperature T. Since the column base temperature decreases, the overall temperature difference for the reboiler increases. Therefore, the reboiler heat transfer rate increases in order to meet the column enthalpy balance. Accordingly, more vapor is generated. Since the
### TABLE 2

**B-T SYSTEM. 21 S. L F. P S NO 15. PRESSURE MODEL.**

**DISTILLATION TOWER'S STEADY STATE GAIN FOR INPUT VARIABLE TSO**

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</tr>
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<td>X</td>
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<table>
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<tr>
<th>L</th>
<th>V</th>
<th>T</th>
<th>P</th>
<th>X(BENZENE)</th>
<th>X(TOULENE)</th>
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<td></td>
<td>(L/MOL SEC)/(F)</td>
<td>(L/MOL SEC)/(F)</td>
<td>(F)/(F)</td>
<td>(PSI)/(F)</td>
<td>(MOLE FRACTION)/(F)</td>
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### TABLE 3

**B-T SYSTEM. 21 S. L F. P S NO 15. PRESSURE MODEL.**

DISTILLATION TOWER'S STEADY STATE GAIN FOR INPUT VARIABLE $H_W$

| $T(1)$ = | -20.5897217 |
| $L(1)$ = | 0.0815681 |
| $TBC$ = | -20.5397217 |
| $X$ = | 0.0001274 | -0.0001274 |

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<th>$V$ (LBMOLE/SEC)</th>
<th>$T$ (F)</th>
<th>$P$ (PSI)</th>
<th>$X$ (BENZENE)</th>
<th>$Y$ (TOLENE)</th>
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distillate and residue flow rates are kept constant, the reflux increases and separation improves.

**Table 4: Increase of the distillate product rate.**

The increase of the distillate product rate causes the reflux drum level controller to cut back the reflux. This in turn reduces the liquid flow rate in the whole column. Since the distillate rate increases, more Benzene (light component) leaves the column and the Benzene concentration in the whole column goes down. In order to meet the total material balance, the tower base level controller cuts back the residue flow rate. Therefore, the Toluene (heavy component) composition in the column goes up. More Toluene at the column base with less liquid rate causes the base temperature to go up and the reboiler duty to cut back. Accordingly, less vapor is generated. The column top temperature also goes down to reduce the condenser duty in order to meet the total enthalpy balance.

**Table 5: Increase of the coolant inlet temperature.**

The increase of the coolant inlet temperature reduces the condenser duty and increases the column temperature and pressure. The reboiler duty is then cut back and less vapor is generated. This reduces the reflux and the separation gets worse.
TABLE 4

B-T SYSTEM, 21 S. L.F. P S NO 15. PRESSURE MODEL

DISTILLATION TOWER'S STEADY STATE GAIN FOR INPUT VARIABLE SSL (1)

| T (1) = -384.77954 |
| L (1) = -2.44088  |
| TBC = -384.77954  |
| X = -0.05130      |
| 0.05130           |

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<th>V (lb mole/sec)</th>
<th>T (F)</th>
<th>P (psi)</th>
<th>X (benzene)</th>
<th>X (toluene)</th>
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Table 6: Increase of the feed rate.

The increase of the liquid feed reduces the stripping section temperature. This brings more heat into the column base, but the heat transfer is not enough to do a good separation job if the reboiler steam control loop remains opened. So, the Benzene (light component) mole fraction at the bottom increases, but still more vapor is generated. This brings up the pressure and the reflux, (as shown by the positive signs of the steady state gains in the table under P and L(1)). The column top temperature increases to transfer more heat out from the condenser in order to meet the entire column enthalpy balance.

Table 7: Increase of the feed Benzene concentration

The increase of the feed Benzene concentration causes the Benzene concentration everywhere in the column to go up. Since more light component is in the stripping section, the bubble point temperature goes down, and the reboiler heat duty increases. Therefore, the pressure and vapor flow rate go up. The column top temperature increases to transfer more heat out from the condenser in order to meet the total column enthalpy balance.

Table 8: Increase of the supply steam valve opening.
**TABLE 6**

**B-T SYSTEM. 21 S. L F. F S NO 15. PRESSURE MODEL.**

DISTILLATION TOWER'S STEADY STATE GAIN FOR INPUT VARIABLE FL (15)

\[ T(1) = 127.98434 \]
\[ L(11) = 0.47922 \]
\[ TBC = 127.98434 \]
\[ X = 0.01585 \]

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The increase of the supply steam valve opening causes the temperature, pressure, liquid and vapor flow rates everywhere in the column to go up, (as shown by the positive signs of the steady state gains in the table). Therefore, separation improves.

Inverse Response Of The Ballast Tray Distillation Column

In a binary distillation column, we know that an increase in vapor boilup must drive more low boiling material up the column and therefore decrease the light component in the column bottom.

But as pointed out by Luyben (1973), the tray hydraulics can produce some unexpected results. When the vapor rate through a tray is increased, it tends to (1) back up more liquid in the downcomer and (2) reduce the liquid density on the active tray. The former effect reduces the liquid rates through the column while the liquid holdup in the downcomer is building up. The latter effect tends to increase the liquid rates since the liquid holdup must be depleted. Sieve and Ballast trays are more affected by froth density changes. Therefore, an increase in vapor boilup produces a transient increase in liquid rates down the column. This increase in liquid rate
produce an initial increase in the amount of light component in the bottoms. Eventually, the liquid rates will return to normal when the liquid inventory on the trays has dropped to a new steady state level. Then the effect of the vapor boilup drive the composition of the light component down. This phenomenon is called "INVERSE RESPONSE".

Inverse response can be best characterized by the process in response to a manipulated variable step change, a controlled variable initially moves to the opposite or wrong direction before finally moving in the expected direction. Mathematically, inverse response can be represented by a system that has a transfer function with at least a positive zero, a zero in the right half of the s plane. But the positive zeros do not make the system open loop unstable because the stability of the system depends only on the poles of the transfer function. If the process transfer function contains a negative lead, the system will show inverse response. The inverse response is partly due to tray design or type, and partly due to the froth effect of the tray hydraulics. Also as pointed by Buckley (1975), all the valve tray columns exhibit inverse response over the entire range of normal operation.

The Benzene-Toluene pressure distillation column subject to steam valve opening, CM, is used in this
research to show the inverse response. In this column, the
reflux drum level controller adjusts the distillate flow
rate and the column base level controller adjusts the
residue flow rate. The reflux flow rate is one of the
manipulated variables and is therefore being treated as
constant.

All the tray Benzene mole fraction process transfer
functions in the entire column indicate inverse response,
i.e., a positive zero in the numerator of the process
transfer function. FIG 19 and FIG 20 are the process
transfer functions of Benzene mole fraction of the column
top tray and the column base tray. FIG 20 shows a final
amplitude ratio SLOPE of -80 DB with -540 degrees final
phase lag. The suggested process transfer function is:

\[
X(s) = \frac{K(1-Ts)}{1 - (1+T_1s)(1+T_2s)(1+T_3s)(1+T_4s)(1+T_5s)(1+T_6s)}
\]

The first negative lead - lag pair, \((1-Ts)/(1+Ts)\),
cancels out only each other's amplitude ratio but the
negative lead term, \((1-Ts)\), contributes -90 degrees to the
first order lag, \((1+Ts)\), which makes -180 degrees in
total. Together with the other four first order lags, the
final phase angle lined out at -540 degrees. From the
column bottom tray up to the column top, each additional
FIG 19. Bode Plot of CM-X(1,2), Pressure Model, Benzene-Toulene Column
FIG 20. Bode Plot of CM-X(1,20), Pressure Model, Benzene-Toulene Column.
FIG 21. Bode Plot of CM-X(1,21), Pressure Model, Benzene-Toulene Column
tray will introduce one more lag.

The Benzene mole fraction process transfer functions of reflux drum and reboiler do not show any inverse response mainly because they are not modeled with the froth effect in this research. The process transfer function of the reboiler is shown as FIG 21. Essentially, the transfer function consists of four lags.

**Effect Of The Model Simplification**

In previous chapters, a complete pressure model has already been presented in detail. This pressure model can be simplified to the enthalpy model if one assumes that there is no variation of the pressure profile, (i.e., removes all the pressure relationships from the pressure model). The enthalpy model can be further simplified to equimolar overflow model by removing all the enthalpy balances and the vapor phase summation equations. In this research, we compare the different process transfer functions generated by these three different models. The characteristics for these models are:

(1) Equimolar overflow model.

The equations describing this model are total material balance, component material balances, equilibrium relationships, tray hydraulics and liquid phase
summation equation for each tray. At steady state, vapor flow rates and liquid flow rates in either the rectifying or stripping section are constant. In the dynamic model, if there is a vapor flow rate perturbation, it is assumed to be transferred directly to the top of the column without any delay. Although this model is simple, it is applied strictly only to the separation of structurally similar components. No variations for the pressure profile of the entire column. Since the vapor phase summation equation of each stage is not used in this model, the temperature profile is not calculated.

(2) Enthalpy model.
The equations describing this model are total material balance, component material balances, equilibrium relationships, summation equations, tray hydraulics and enthalpy balance for each tray. Since the enthalpy balance enables the calculation of vapor flow rate and the vapor phase summation equation enables the calculation of bubble point temperature, this model is more rigorous than the equimolar overflow model. The pressure profile is assumed to be constant for the whole column; in other words, all the vapor leaving the column top tray is assumed to be totally condensed at the reflux drum bubble point temperature
in order to maintain the constant column pressure. The condenser and the reboiler are only modeled with total material balance, component material balances, equilibrium relationships, summation equations and enthalpy balance with the heat duty as a whole. The metal heat transfer lags in both the reboiler and condenser are neglected in this model.

(3) Pressure model
Pressure variation through each tray is considered in addition to the enthalpy model. Also, the tray hydraulics, the reboiler and the condenser are modeled in detail.

Benzene-Toluene Column Process Transfer Functions Of Feed Rate Disturbance To Liquid Flow Rate.

In this Benzene - Toulene system, the reflux drum level controller adjusts the reflux flow rate and the column base level controller adjusts the residue flow rate. This set of the results are from FIG 22 to FIG 26. These figures are the process transfer functions Bode plots between the disturbance feed flow rate, FL(10), and the measured liquid flow rate leaving the feed tray, L(10), and the trays below.

In the equimolar overflow dynamic model, since the
FIG 22. Bode Plot of FL(15) - L(15), Equimolar Overflow Model, Benzene-Toulene Column
FIG 23. Bode Plot of FL(15)-L(15), Enthalpy Model, Simple Tray Hydraulics, Benzene-Toluene Column
FIG 24. Bode Plot of FL(15)-L(16), Enthalpy Model, Simple Tray Hydraulics, Benzene-Toluene Column
FIG 25.  Bode Plot of PL(15)-L(15), Pressure Model, Complex Tray Hydraulics, Benzene-Toluene Column
FIG 26. Bode Plot of FL(15) - L(16),  Pressure Model, Complex Tray Hydraulics, Benzene-Toluene Column
vapor flow rate is constant, even though there is a perturbation of the reboiler heat, the vapor flow rate is assumed to be transmitted directly to the column top without any delay. Therefore, if there is a feed liquid flow rate perturbation, the liquid flow rates leaving the trays below the feed tray behave exactly like the liquid flow leaving a series of tanks.

The feed tray liquid flow rate process transfer function, FIG 22, shows the first order lag, i.e., -20 dB slope of the log modulus and -90 degree maximum phase lag. The transfer function of the liquid flow leaving the tray below the feed tray shows the second order lag, i.e., -40 dB slope of the log modulus and -180 degree maximum phase lag and so forth. The tray hydraulics equation used here is the tray hydraulics equation used by Buckley (1975). This tray hydraulic equation indicates the liquid flow rate leaving each tray can be in terms of the vapor flow rate passing through the tray and the liquid flow rate leaving the above tray. For equimolar overflow model, the tray hydraulics is assumed to be only in terms of the liquid flow effect.

FIG 23 and FIG 24 are the process transfer functions Bode plots for the enthalpy model with simplified tray hydraulics. The tray hydraulics equation used here is Buckley's (1975) tray hydraulic equation which is in terms
of both the vapor and the liquid effect.

In FIG 23 and FIG 24, although the final log modulus slope is -20, -40 DB respectively, at low frequency, there is a lead-lag pair with almost the same break point frequency. Similar Bode plots have been observed for the feed disturbance of the Propane - Butane - Pentane system. This can be traced to the enthalpy balance since it is not present in the equimolar overflow model.

FIG 25 and FIG 26 are the process transfer functions Bode plots for the pressure model. At intermediate frequency, the log modulus slope is -20 DB in FIG 25 and -40 DB in FIG 26. At higher frequency, the amplitude ratio final slope of both figures are -20 DB but the phase angles line out at -270 and -450 degree individually. In other words, the pressure model introduces an additional negative lead - lag pair per tray. The proposed models for these two transfer functions are:

\[
\begin{align*}
L(s) &= \frac{K*(1 + Ts)*(1 - Ts)}{10} \\
F_L(s) &= \frac{(1 + Ts)*(1 + Ts)*T^2}{10} \\
L(s) &= \frac{K*(1 + Ts)*(1 - Ts)*(1 - Ts)}{10} \\
F_L(s) &= \frac{(1 + Ts)*(1 + Ts)*T^2}{10}
\end{align*}
\]

(6.2)

For equation (6.2), the first lead - lag pair,
\(\frac{1 + Ts}{1 + Ts}\), cancels out each other's amplitude ratio and phase angle. The second negative lead-lag pair, \(\frac{1 - Ts}{1 + Ts}\), cancels out each other's amplitude ratio but the negative lead term, \(1 - Ts\), contributes \(-90^\circ\) degrees to the first order lag, \(1 + Ts\), which makes \(-180^\circ\) degrees in total. Together with the first order lag, \(1 + Ts\), the final phase angle lined out at \(-270^\circ\) degrees.

For equation (6.3), there is one more negative lead-lag pair; therefore, the phase angle levels out at \(-450^\circ\) degrees.

Physically, in this particular system, since the liquid feed temperature is lower than the feed tray bubble point temperature, this reduces the vapor flow rate, temperature and pressure on the feed tray. Less vapor being generated causes the froth on all the trays to shrink. In other words, the liquid level height over the weir is reduced, correspondingly, less liquid flow leaves the tray. However, more liquid feed coming into the distillation column will end up this phenomenon. By observing the pressure model transfer function Bode plots FIG 25 and FIG 26 again, one can find that the inverse response happens at rather high frequency. This high frequency inverse response can usually be neglected.

If the simple tray hydraulics equation in the enthalpy model is switched to the one derived in chapter four with the pressure term being neglected, the liquid flow process transfer function will also indicate inverse response.
Benzene-Toluene Tower Pressure Process Transfer Functions By

Manipulated Variable Distillate Product Flow Rate

The reflux drum level controller of this Benzene -
Toluene tower is assumed to adjust the reflux flow rate and
the column base level controller to adjust the residue flow
rate.

FIG 27 is a column top tray pressure, P(2), process
transfer function Bode plot for manipulated distillate
product rate, SSL(1), of the Benzene - Toulene system.
This column top tray pressure process transfer function
shows a final amplitude ratio slope of -40 DB with -540
degree phase lag. The suggested process transfer function
is:

\[
P(s) = \frac{K*(1 - T*s)*(1 - T*s)}{1 + T*s} \]

\[
SSL(s) = \frac{(1 + T*s)*(1 + T*s)*(1 + T*s)*(1 + T*s)*(1 + T*s)}{1} \]

(6.4)

From top tray down to the column base, for each additional
tray, the pressure process transfer function will have one
more negative lead - lag pair which makes no changes of the
final transfer function amplitude ratio slope but adds -180
degree for the phase angle.

Physically, the increase of the distillate flow rate
causes the reflux drum level controller to cut back the
FIG 27. Bode Plot of SSL(1)-P(2), Pressure Model, Benzene-Toluene Column
reflux into the distillation column. Since the reflux drum temperature is lower than the column top tray temperature, the less reflux gets into the column top tray means less cool reflux to mix with the liquid holdup on the column top tray. In other words, the pressure and temperature of the column top tray will rise. But since more light component, Benzene, has been pulled out through the distillate, as time goes by, the Toluene concentration in the whole column rises and this brings up the bubble point temperature everywhere. When this effect reaches the column base, the reboiler heat duty will be eventually reduced, less vapor flow rate will be generated and the column pressure will finally be reduced.

The inverse response shown by this case is not due to the tray froth effect, but instead, from the heat and material balance.

Benzene-Toluene Tower Pressure and Benzene Mole Fraction Process Transfer Functions By Manipulated variable Reflux Flow Rate

If the Benzene - Toluene distillation column reflux drum level controller adjusts the distillate flow rate, then the reflux rate becomes a manipulated variable.

FIG 28, FIG 29 and FIG 30 are the Benzene mole
FIG 28. Bode Plot of L(1)-X(1,1), Pressure Model, Benzene-Toluene Column
FIG 29. Bode Plot of $L(1)-X(1,2)$, Pressure Model, Benzene-Toluene Column
FIG 30. Bode Plot of L(1)-X(1,21), Pressure Model, Benzene-Toluene Column.
fraction transfer function Bode plots of the distillate, column top tray and residue. FIG 28, the distillate Benzene mole fraction Bode plot indicates two lags and FIG 29, the column top tray Benzene mole fraction Bode plot indicates only one lag. Generally speaking, from the column top tray to the column base, there is at least one negative lead - Lag pair per tray being introduced into the process transfer function.

FIG 31 and FIG 32 are the column top tray and the second tray pressure transfer function Bode plot. From the second tray below to the column base, each tray introduces one more negative lead - lag pair to the process transfer function.

Physically, since the reflux drum temperature is lower than the column top tray temperature. The immediate increase in the flow of the cooler reflux causes the reduction of the column top tray temperature. Consequently, less vapor will be generated from the column top tray and the pressure goes down. But more reflux going into the column causes the reflux drum level controller to cut back the distillate flow rate. In other words, less light component, Benzene, will be pulled out from the distillation column. This makes the Benzene concentration to slowly rise everywhere in the column and correspondingly decrease the bubble point temperature on each tray. When
FIG 31. Bode Plot of L(1)-P(2), Pressure Model, Benzene-Toluene Column
FIG 32. Bode Plot of L(1)-P(3), Pressure Model, Benzene-Toluene Column
this effect reaches the column base, more heat will be brought into the column, eventually increasing the column pressure.

Propane-Butane-Pentane Tower Pressure Process Transfer Functions by Manipulated Variable cooling water flow rate.

The reflux drum level controller of this Propane-Butane-Pentane tower is assumed to adjust the distillate flow rate and the column base level controller to adjust the residue flow rate.

FIG 33 is a column top tray pressure, $P(2)$, process transfer function Bode plot for the manipulated variable, cooling water flow rate, $FW$, of the Propane-Butane-Pentane system. This column top tray pressure process transfer function indicates a final amplitude ratio of $-60$ DB with $-270$ degree final phase lag. From top tray down to the column base, each additional tray introduces one more negative lead-lag pair into the process transfer function which makes no changes of the transfer function amplitude ratio slope but adds $-180$ degree for the phase angle.

Physically speaking, the increase of the coolant flow rate reduces the reflux drum temperature. The same amount of the reflux with lower temperature gets into the column top tray reducing the column top tray pressure and
**BODE PLOT - FW-P(2)**

**Omega (Rad/Sec)**

**Phase Angle in Deg**

**Amplitude Ratio in DB**

**PBX SYSTEM. 11 T. L. F. FT NO. 5. PRESSURE MODEL.**

**Steady Gain** = -0.13979E+02  
**T.C. Lead** = 0.27646E+03

**FIG 33. Bode Plot of FW-P(2), Pressure Model, Propane-Butane-Pentane Column**
temperature. Since the vapor rate leaving the column top tray is reduced, more light component gets into the trays below which decreases the bubble point temperature. More vapors are generated for a short period by the energy depletion on the trays and the vapor pressure rises. Although the column temperature tends to decrease and more heat is transferred at the reboiler, eventually the pressure will drop because the vapor generated by this effect is not high enough to raise the column pressure.

Frequency Response Interaction Measure

In designing a multiloop control system, a key design decision is to determine the proper pairings of controlled and manipulated variables. If an incorrect pairing is used, the resulting control system may perform poorly or even be inoperable. Normally, the Relative Gain Array method proposed by Bristol is used to determine the multiloop control pairings based on the steady state gains. In this research, the Benzene - Toulene distillation column is used as an example to show the dynamic effects of the interaction measures based on the process transfer functions between the chosen manipulated and controlled variables as a function of frequency.

The reflux drum level controller of this Benzene -
Toluene column adjusts the reflux and the column base level controller adjusts the residue flow rate. These two level control systems are included as part of the distillation process. There are three open loops in this particularly chosen distillation column, the steam control valve, the distillate flow rate control valve and the condenser cooling water flow rate control valve. This means one can pick up to three controlled variables to pair these with. Since the distillation column is used to separate components of different volatility, the desired distillate or residue component composition are certainly the controlled variables. Besides these, the column top pressure has been chosen to be the third controlled variable because the column pressure determines the energy consumption.

FIG 27, FIG 34 and FIG 35 are the process transfer function Bode plots of manipulated distillate flow rate, SSL(1), with either the column top tray pressure P(2), or distillate Benzene concentration X(1,1), or residue Toluene concentration X(2,21). FIG 36 to FIG 38 are the process transfer function Bode plots of manipulated variable cooling water flow rate, FW, with P(2), X(1,1), X(2,21). FIG 39 to FIG 41 are the process transfer function Bode plots of manipulated variable steam valve opening, CM, with P(2), X(1,1), X(2,21). Based on the discussion of the
FIG 34. Bode Plot of SSL(1)-X(1,1), Pressure Model, Benzene-Toluene Column
FIG 35. Bode Plot of SSL(1)-X(2,21), Pressure Model, Benzene-Toluene Column
FIG 36. Bode Plot of FW-P(2), Pressure Model, Benzene-Toluene Column.
FIG 37. Bode Plot of FW-X(1,1), Pressure Model, Benzene-Toluene Column
FIG 38. Bode Plot of FW-X(2,21), Pressure Model, Benzene-Toluene Column.
FIG 39. Bode Plot of CM-P(2), Pressure Model, Benzene-Toluene Column
BODE PLOT - CM-X(1,1)

OMEGA (RAD/SEC)

AMPLITUDE RATIO IN DB

-20.00 -10.00 0.00 10.00

×10^2

-10.00 0.00 10.00

FIG. 40. Bode Plot of CM-X(1,1), Pressure Model, Benzene-Toluene Column
FIG 41. Bode Plot of CM-X(2,21), Pressure Model, Benzene-Toluene Column
frequency response interaction measure in chapter five, the interaction measures are calculated by Equation (5.10) and the results are shown from FIG 42 to FIG 50.

The following conclusions show the dynamic effects of the interaction measures. At low frequency, that is the long term or steady state effects, the interaction measures are summarized in TABLE 9. The interaction measures in TABLE 9 are calculated by Equation (5.10) which based on the process gains of the corresponding manipulated variable – measured variable pairs. From TABLE 9, the residue Toluene concentration is suggested to be adjusted by the distillate flow rate. The distillate Benzene concentration is suggested to be adjusted by the cooling water flow rate and the column top pressure is adjusted by the steam valve opening.

In order to show the dynamic imbalance of the pairings in multivariable system which are not indicated by the static relative gain method. The interaction measures at a frequency of 10 RAD/SEC are chosen and summarized in the TABLE 10. At this high frequency, the steam valve is suggested to adjust the residue Toluene concentration, the cooling water flow rate to adjust the distillate Benzene concentration and the distillate flow rate to adjust the column top pressure. This indicates that the short term effects are different from the steady state effects and
FIG 42. Interaction Measure of P2-SSL1
FIG 43. Interaction Measure of XD-SSL1
FIG 44.  Interaction Measure of IB-SSL1
FIG 45. Interaction Measure of P2-FW
FIG 46. Interaction Measure of XD-FW
FIG 47. Interaction Measure of XB-FW
FIG 48. Interaction Measure of P2-CM
FIG 49. Interaction Measure of XD-CM
FIG 50. Interaction Measure of XB-CM
### TABLE 9.

<table>
<thead>
<tr>
<th></th>
<th>SSL(1)</th>
<th>FW</th>
<th>CM</th>
</tr>
</thead>
<tbody>
<tr>
<td>P(2)</td>
<td>-0.50402E-5</td>
<td>0.055215</td>
<td>0.94479</td>
</tr>
<tr>
<td>X(1,1)</td>
<td>0.003275</td>
<td>0.9416</td>
<td>0.055121</td>
</tr>
<tr>
<td>X(2,1)</td>
<td>0.99673</td>
<td>0.0031813</td>
<td>0.8846E-4</td>
</tr>
</tbody>
</table>

### TABLE 10.
The Interaction Measures Of The Benzene-Toluene System At Frequency Of 10 RAD/SEC. The Reflux Drum Level Controller Adjusts The Reflux Flow Rate.

<table>
<thead>
<tr>
<th></th>
<th>SSL(1)</th>
<th>FW</th>
<th>CM</th>
</tr>
</thead>
<tbody>
<tr>
<td>P(2)</td>
<td>1.0</td>
<td>0.0</td>
<td>0.0</td>
</tr>
<tr>
<td>X(1,1)</td>
<td>0.0</td>
<td>1.0</td>
<td>0.0</td>
</tr>
<tr>
<td>X(2,1)</td>
<td>0.0</td>
<td>0.0</td>
<td>1.0</td>
</tr>
</tbody>
</table>
that dynamic decoupling is probably necessary. This program provides an invaluable tool to design such a decoupler, as the Bode plots can be used to determine the necessary dynamic compensation.

If the reflux drum level controller of this Benzene - Toluene distillation column adjusts the distillate flow rate and the column base level controller adjusts the residue flow rate, there are still three open loops in this distillation column, the steam control valve, the reflux flow rate control valve and the condenser cooling water flow rate control valve. If the distillate and residue composition and the column top pressure are chosen as the controlled variables, the steady state interaction measures are summarized in TABLE 11. Also for this particular case, the interaction measures at a frequency of 100 RAD/SEC are chosen and summarized in TABLE 12.

Summary Of The Results

The effects of the distillation column model simplification on process transfer functions are compared among the pressure model, the enthalpy model, and the equimolar overflow model. In the enthalpy and pressure models, since the detail modeled Ballast tray hydraulics in the frequency domain is used, results indicate inverse
### TABLE 11.
The Interaction Measures Of The Benzene-Toluene System At Steady State. The Reflux Drum Level Controller Adjusts The Distillate Flow Rate.

<table>
<thead>
<tr>
<th></th>
<th>L(1)</th>
<th>PW</th>
<th>CM</th>
</tr>
</thead>
<tbody>
<tr>
<td>P(2)</td>
<td>0.524</td>
<td>0.090796</td>
<td>0.38516</td>
</tr>
<tr>
<td>X(1,1)</td>
<td>24.027</td>
<td>-11.731</td>
<td>-11.295</td>
</tr>
<tr>
<td>X(2,21)</td>
<td>-23.551</td>
<td>12.641</td>
<td>11.91</td>
</tr>
</tbody>
</table>

### TABLE 12.
The Interaction Measures Of The Benzene-Toluene System At Frequency Of 100 RAD/SEC. The Reflux Drum Level Controller Adjusts The Distillate Flow Rate.

<table>
<thead>
<tr>
<th></th>
<th>L(1)</th>
<th>PW</th>
<th>CM</th>
</tr>
</thead>
<tbody>
<tr>
<td>P(2)</td>
<td>1.0</td>
<td>0.0</td>
<td>0.0</td>
</tr>
<tr>
<td>X(1,1)</td>
<td>0.0</td>
<td>1.0</td>
<td>0.0</td>
</tr>
<tr>
<td>X(2,21)</td>
<td>0.0</td>
<td>0.0</td>
<td>1.0</td>
</tr>
</tbody>
</table>
response. This inverse response phenomena of the Ballast tray column has also noted by Luyben (1969) and Buckley (1975).

In designing the multiloop control systems, normally, the steady state interaction measures calculated by Bristol's relative gain array method is widely used. In the second part of the results, the steady state gain analysis of the Benzene - Toluene column are studied.

In the last part of the results, the dynamic effects of the interaction measures are calculated as a function of frequency to provide a better insight into the multiloop control pairings through the Benzene - Toluene pressure model.
SUMMARY AND CONCLUSIONS

The objective of this research was to develop a flexible frequency response package for the distillation columns. There are three applications of it:
(1) Test the model assumptions and their effects.
(2) Determine the steady state sensitivities.
(3) Calculate the interaction measure as a function of frequency.

A complete set of computer programs are set up in this research which includes a steady state distillation program, a mechanical distillation column design program, a generalized distillation column pressure model in the frequency domain, a process transfer function Bode plot program and an interaction measure calculation program. The steady state distillation program is needed to facilitate the set up of the steady state profile around which the model equations are linearized. The mechanical design program provides the mechanical parameters of the tray column, the reboiler, the condenser and the reflux drum. The distillation column pressure model is used to generate the process transfer functions between the chosen manipulated and controlled variables. Based on the generated process transfer functions, the interaction measure program calculates the interaction measures as a
function of frequency. The results are then presented as Bode plots. Although the interaction measure example in chapter six is a three manipulated variables with three controlled variables system, the interaction measure calculations may be generated to n dimensions if there is a distillation column with n manipulated variables and n controlled variables. The dynamic interaction measure results indicate a discrepancy between steady state and dynamic pairings. The program developed in this research can help develop dynamic decouplers to handle this situation.

In the past, the majority of the authors set up distillation column equimolar overflow models or enthalpy models for a particular system, but all of them are not flexible enough to handle different types of columns. In this research, both the steady state distillation column model and the dynamic pressure model are derived with the assumption that each tray has liquid feed and vapor feed coming in, liquid product and vapor product going out and the interstage heater or cooler.

Also most papers in the past dealing with the analysis of distillation columns have assumed the pressure to remain constant and have been consequently unable to study the pressure as a controlled variable. The dynamic pressure distillation column model used in this research was based
on the total and component material balances, enthalpy balances, summation equations, equilibrium relationships as well as the pressure relationships and the detailed modeling of the tray hydraulics, condenser, reflux drum and reboiler. The pressure model can be simplified to the enthalpy model by removing all the pressure relationships. The enthalpy model can be simplified to the equimolar overflow model by removing all the enthalpy balances and the vapor phase summation equations. [The variable pressure model permits the study of the effect of the pressure control loop on the column interaction analysis. This improvement over models currently reported in the literature is an important contribution of this work.]

Recommendations For Future Work

This study was considered as the first step in the research program, laying the groundwork for the future study. The recommendations for the future work are:

(1) The interaction measures of different possible measured variables in the control pairings.

(2) The study of the feasibility of the decoupling of the distillation column.

(3) The interaction measures of the more complex distillation column, e.g., distillation column with
multiple feeds, multiple sidestreams, inter stage heaters / coolers.

(4) The determination of whether the interactions for a particular system dynamically are favorable or not.

(5) The study of the different control pairings between the high purity control and low purity control.
REFERENCES


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APPENDIX A

FLASH FEED INTO THE DISTILLATION COLUMN

As shown in FIG 2, pressure at the upstream and downstream of the restriction valve is assumed constant. The equations describing this system are steady state total and component material balances, enthalpy balance, summation equations and equilibrium relationships.

(A) feed at the upstream of the restriction valve

1) total material balance.
\[ F = F_{UL} + F_{UV} \] (A.1)

2) component material balance (NC-1 equations).
\[ F \cdot Z = (F_{UL} \cdot F_{UX}) + (F_{UV} \cdot F_{UY}) \] (A.2)

3) equilibrium relationships (for i=1,NC).
\[ F_{UY} = K_{U} \cdot F_{UX} \] (A.3)

4) enthalpy balance.
\[ F \cdot H_{F} = (F_{UL} \cdot H_{UL}) + (F_{UV} \cdot H_{UV}) \] (A.4)

5) thermal properties.
\[ H_{FL} = f(T_{U}, F_{UX}) \] (A.5)
\[ H_{FV} = f(T_{U}, F_{UY}, P_{U}) \] (A.6)
(6) summation equations.

\[ \sum_{i} F_{i} = 1.0 \]  \hspace{1cm} (A.7)
\[ \sum_{i} F_{i} = 1.0 \]  \hspace{1cm} (A.8)

(B) feed at the downstream of the restriction valve

(1) total material balance.

\[ F = F_{L} + F_{V} \]  \hspace{1cm} (A.9)

(2) component material balance (NC-1 equations).

\[ F_{i} * Z_{i} = (F_{L} * Z_{L}) + (F_{V} * Z_{V}) \]  \hspace{1cm} (A.10)

(3) equilibrium relationships (for i=1,NC).

\[ Z_{V_{i}} = K_{D} * Z_{L_{i}} \]  \hspace{1cm} (A.11)

(4) enthalpy balance.

\[ F_{i} * H_{i} = (F_{L} * H_{L_{i}}) + (F_{V} * H_{V_{i}}) \]  \hspace{1cm} (A.12)

(5) thermal properties.

\[ H_{L_{i}} = f(T_{F_{i}}, Z_{L_{i}}) \]  \hspace{1cm} (A.13)

\[ H_{V_{i}} = f(T_{F_{i}}, Z_{V_{i}}, P_{i}) \]  \hspace{1cm} (A.14)

(6) summation equations.

\[ \sum_{i} Z_{L_{i}} = 1.0 \]  \hspace{1cm} (A.15)
\[ \sum_{i} Z_{V_{i}} = 1.0 \]  \hspace{1cm} (A.16)

First, the upstream working equations can be derived.
Substituting $F_{UV}/F$ for $FU$ and solving for $FUX$;

$$Z_{i,n} = \frac{FUX}{(1.0-FU) + (FU \times KU)_{i,n}} \quad (A.17)$$

$$\Sigma_{i,n} FUX = \Sigma_{i,n} \frac{(1.0-FU) + (FU \times KU)_{i,n}}{i,n} = 1.0 \quad (A.18)$$

Similarly,

$$\Sigma_{i,n} FUY = \Sigma_{i,n} \frac{FU + (1.0-FU)/KU}{i,n} = 1.0 \quad (A.19)$$

Once the $FU$ is solved from either equation (A.18) or (A.19), the amount of liquid and vapor and the corresponding composition, enthalpy for the upstream feed can then be calculated. But by taking a closer examination, if either (A.18) or (A.19) is solved by trial and error approach for $FU$, there exists two possible solutions in either equation. For equation (A.18), the $FU$ might converge to zero or a value in-between zero and one. For equation (A.19), the $FU$ might converge to one or a value in-between zero and one. Instead, if the following equation is solved, then the solution will be unique.

$$\Sigma_{i,n} FUX - \Sigma_{i,n} FUY = 0.0 \quad (A.20)$$
Similar equation can be derived for the downstream of the restriction valve.

\[ \sum_{ZL} - \sum_{ZV} = 0.0 \]  \hspace{1cm} (A.21)

\[ \sum_{i,n}^{Z_{i,n}} - \sum_{i,n}^{Z_{i,n}} = \sum_{i,n}^{\frac{Z_{i,n}}{(1.0-FD)+(PD*KD)}} - \sum_{i,n}^{\frac{Z_{i,n}}{PD+(1.0-PD)/KD}} \]

The flash feed calculation procedure is now summarized as follows:

1. Based on the upstream feed temperature, feed pressure, total feed rate, feed composition, solve \( F_0 \) from equation (A.20) by using the Newton-Raphson method.

2. Based on \( F_0 \), calculate the upstream total feed enthalpy.

3. Make the downstream bubble and dew point calculation according to the downstream pressure and the total feed composition.

4. Calculate the two reference enthalpy by assuming either the feed is all liquid at downstream bubble point temperature or all vapor at downstream dew point temperature.

5. If the upstream feed enthalpy is less than the reference downstream bubble point temperature enthalpy, the downstream feed is then all liquid.
Calculate the downstream liquid temperature.

(6) If the upstream feed enthalpy is greater than the reference downstream dew point temperature enthalpy, then, the downstream feed is all vapor. Calculate the downstream vapor temperature.

(7) If the upstream feed enthalpy is in-between the reference bubble point and dew point enthalpy, this means the downstream feed is a flash feed which consists of both liquid and vapor. Find out the fraction vaporized and temperature by:

(a) Use bi-section method, choose a temperature between the reference bubble point and dew point temperature.

(b) Solve for fraction vaporized from eq (A.21).

(c) Calculate the downstream flash feed enthalpy.

(d) Check the calculated flash feed enthalpy with the total feed enthalpy, if not matched, go back to (a). Otherwise, exit.
NOMENCLATURE FOR APPENDIX A

\( F_n \) : total feed rate, LBMOLE/SEC, to stage \( n \).

\( F_{Dn} \) : fraction vaporized of downstream feed of stage \( n = \frac{F_{Vn}}{F_n} \).

\( F_{Ln} \) : downstream liquid feed rate, LBMOLE/SEC, to stage \( n \).

\( F_{Un} \) : fraction vaporized of upstream feed of stage \( n = \frac{F_{UVn}}{F_n} \).

\( F_{Vn} \) : downstream vapor flow rate, LBMOLE/SEC, to stage \( n \).

\( F_{ULn} \) : upstream liquid feed rate, LBMOLE/SEC, to stage \( n \).

\( F_{UVn} \) : upstream vapor flow rate, LBMOLE/SEC, to stage \( n \).

\( F_{UX_{i,n}} \) : mole fraction of component \( i \) of \( F_{ULn} \).

\( F_{UY_{i,n}} \) : mole fraction of component \( i \) of \( F_{UVn} \).

\( H_{Pn} \) : enthalpy, BTU/LBMOLE, of \( F_n \).

\( H_{F_{PLn}} \) : enthalpy, BTU/LBMOLE, of \( FL_n \).

\( H_{F_{PVn}} \) : enthalpy, BTU/LBMOLE, of \( F_{Vn} \).
HUL\_n \quad : \text{enthalpy, BTU/LBMOLE, of FUL} \_n

HU\_n \quad : \text{enthalpy, BTU/LBMOLE, of FUV} \_n

KD\_i,n \quad : \text{equilibrium K of component i of downstream feed of stage n.}

KU\_i,n \quad : \text{equilibrium K of component i of upstream feed of stage n.}

P\_n \quad : \text{pressure, PSI, at the downstream of the restriction valve of stage n.}

PU\_n \quad : \text{pressure, PSI, at the upstream of the restriction valve of stage n.}

TP\_n \quad : \text{feed temperature, F, at the downstream of the restriction valve of stage n.}

TU\_n \quad : \text{feed temperature, F, at the upstream of the restriction valve of stage n.}

ZL\_i,n \quad : \text{mole fraction of component i of FL} \_n

ZV\_i,n \quad : \text{mole fraction of component i of FV} \_n
APPENDIX B

DESIGN PROCEDURES OF VERTICAL THERMOSYPHON REBOILER

The structure of the vertical thermosyphon reboiler designed by this research is limited to one shell and one tube pass. The design of the vertical thermosyphon reboiler can be divided into two cases. The first case selects the optimum recirculation ratio based on the fixed tube length. The second case selects the optimum tube length based on the fixed recirculation ratio.

The design procedures are summarized as follows:

1. Evaluate the heat load $Q$.
2. Evaluate the LMTD by assuming reboiler is under isothermal boiling.
3. Calculate the required steam mass flow rate.
4. From the input design flux, calculate the heat transfer area required. As suggested by Kern (1950), the maximum flux for the organic materials is limited to $12000.0 \text{ BTU/HR, FT}^2$. Experience has shown that a flux of 6000 to 8000 is a good starting value for the organics.

$$\frac{Q}{\text{FLUX}} = \text{--------}$$  \hspace{1cm} (B.1)

5. For case one, the tube length is fixed. For case two,
the tube length is assumed.

(6) From the tube length, tube outside diameter, tube BWG, tube layout pitch, find the tube count and shell size from TABLE 13. If the heat transfer area is too big, reboilers in parallel will be assumed.

(7) Evaluate an operating overall heat transfer coefficient.

\[
\frac{Q}{U} = \frac{D}{A*\text{LMTD}} \tag{B.2}
\]

(8) For case one, a recirculation ratio is assumed. For case two, the recirculation ratio is specified in the input data.

(9) Determine the material balance around the unit.

Suppose the recirculation ratio is M : 1, then:

- total weight of recirculated liquid
  
  \[
  = M \times \text{(desired vapor rate, V)}
  \]

- vapor = desired vapor rate, V

- total = \((M+1)\times\text{(desired vapor rate, V)}\)

(10) Pressure balance across reboiler

(a) Static pressure of reboiler leg.

\[
\frac{L*\rho}{144.0} = \frac{2.3*L}{144.0*(V-V)} \times \log\left(\frac{V}{V}ight), \text{ PSI} \tag{B.3}
\]

where:

- \(L\) : length of reboiler tube, FT.
TABLE 13

TUBE SHEET LAYOUTS (TUBE COUNTS).
Triangular Pitch

<table>
<thead>
<tr>
<th>Shell</th>
<th>1-P</th>
<th>2-P</th>
<th>4-P</th>
<th>6-P</th>
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</table>

1 in. OD tubes on 1\(\frac{1}{4}\)-in. triangular pitch

<table>
<thead>
<tr>
<th>Shell</th>
<th>1-P</th>
<th>2-P</th>
<th>4-P</th>
<th>6-P</th>
<th>8-P</th>
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</tbody>
</table>

1\(\frac{1}{4}\) in. OD tubes on 1\(\frac{1}{4}\)-in. triangular pitch

<table>
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<tr>
<th>Shell</th>
<th>1-P</th>
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V₀ : specific volume of fluid at the outlet of reboiler, FT³/LBM.

Vᵢ : specific volume of fluid at the inlet of reboiler, FT³/LBM.

(b) Friction resistance to flow inside tubes

\[
\text{pressure drop} = \frac{2 f G L n}{t} = \frac{-\text{---}}{5.22E10 D s \phi}, \text{ PSI}
\]

(B.4)

where:

G : flow into tubes, LB/HR, FT², cross sectional area of tubes.

Dᵢ : tube inside diameter, FT.

ϕᵢ : 1.0.

f : friction factor for flow in tubes.

If Reynolds number Re < 2000,
use Hagen-Poiseuille equation.

\[
f = \text{----} \quad \text{(B.5)}
\]

If Reynolds number Re > 2000,
use McAdams and Seltzer relationship.

\[
f = 0.0035 + \text{-----} \quad \text{(B.6)}
\]

s : mean specific gravity in tube as average of inlet liquid and outlet vapor-liquid mixture.
(11) Total resistance to flow

\[ \text{Total resistance to flow} = \text{static pressure of reboiler leg} \]
\[ + \text{pressure drop through tubes} \]
\[ + \text{frictional resistance of inlet piping} \]
\[ + \text{frictional resistance of outlet piping} \]
\[ + \text{expansion loss} \]

(12) Driving force

\[
\text{driving force} = \frac{\Delta Z \cdot \rho \cdot L}{144.0}, \text{PSI} \tag{B.7}
\]

where:
\[ \Delta Z \] : height of liquid level in column above reboiler bottom tube sheet, FT.
\[ \rho \] : density of liquid, LB/FT**3.

(13) If the driving force, step (12) does not equal or slightly exceed the total resistance in step (11), the unit should be rebalanced. For case 1, lower recirculation ratio will give less pressure drop. For case 2, shorter tubes will give less pressure drop.

(14) After a pressure drop balance has been obtained to +0.1 to +0.2 psi, compute the heat transfer coefficient as follows:

(a) Shell side: As suggested by Kern (1950), the steam side heat transfer coefficient can be taken
to be 1500 BTU/FT**2, HR, F for the design purpose.

(b) Tube side: determine heat transfer coefficient from Sieder and Tate correlation. As suggested by Ludwig (1965), if the h calculated exceeds 300 for organics, use value of 300, and correct to outside coefficient h_i o.

Gas and liquid heat transfer inside tubes has been studied by Sieder and Tate (1936) and is shown in FIG 51. The equations representing portion of the graph are:

(b.1) For viscous streamline flow of organic liquids, water solutions (not water) and gases with Reynolds number less than 2100 in horizontal or vertical tubes (deviation ± 12 %):

\[ h_{i} \frac{K}{D} \frac{u^{2/3}}{2/3} \frac{u^{1/3}}{1/3} < u_{0.14} \]

\[ C \times G \quad D \times G \quad \frac{u \times C}{L} \quad \frac{u}{W} \]

(b.8)

(b.2) For turbulent flow of viscous fluids as organics liquids, water solutions (not water) and gases with Reynolds number greater than 10,000 in horizontal or vertical tubes (deviation ± 15% - -10 %):
FIG 51. Tube Side Heat Transfer, Sieder And Tate Correlation.
$\frac{hD}{k} \cdot \frac{D}{G} \cdot 0.8 \cdot C \cdot \mu^{1/3} \cdot \lambda \cdot 0.14 = 0.027 \cdot \frac{\mu}{\mu} \cdot \frac{k}{\lambda} \cdot \frac{a}{\lambda} \cdot \frac{a}{\lambda}$

(\text{B.9})

Where:

- $C$: heat capacity of fluid at constant pressure, BTU/LB,°F.
- $G$: fluid mass velocity, LB/HR*FT**2 of tube cross sectional area.
- $L$: heated straight tube length, FT.
- $\mu$: viscosity of fluid, LB/HR,FT.
- $D$: inside diameter of the tube, FT.
- $k$: thermal conductivity of fluid at average bulk temperature of fluid, BTU/HR,FT**2/(°F/FT).
- $h$: film heat transfer coefficient inside tube, BTU/HR,FT**2,°F.
- $\lambda$: viscosity of fluid at the tube wall temperature, LB/HR,FT.
(15) Calculate the clean overall heat transfer coefficient by:

\[
U = \frac{1}{C D D * \ln\left(\frac{D}{D_o}\right)} + \frac{1}{D * h} + \frac{1}{2k} + \frac{1}{h} \]  

(B.10)

where:

- \( D \): outside diameter of the tube, FT.
- \( k \): thermal conductivity of the reboiler tube metal, BTU/HR,FT,F.

(16) Calculate the resistance of fouling, \( R \), required for the operation of reboiler.

\[
R = \frac{U - U_o}{C D} \]  

(B.11)

(17) If this fouling seems too low for the service, then the unit must be redesigned.
APPENDIX C

DESIGN PROCEDURES OF THE FORCED CIRCULATION REBOILER

A forced circulation reboiler is shown in FIG 8. In this system, a circulating pump is used to assure the desired recirculation ratio and prevent the formation of excessive deposit.

The structure of the forced circulation reboiler designed by this research has one shell and one tube pass. Tube length, tube outside diameter, tube BWG and tube layout pitch must be provided by the user.

Design procedures are summarized as follows:

1. Evaluate the heat load $Q$.
2. Evaluate the LMTD by assuming that the reboiler is under isothermal boiling.
3. Calculate the required steam mass flow rate.
4. From the input design flux, calculate the heat transfer area required. As suggested by Kern (1950), the maximum flux for the organic materials is limited to $12000.0 \text{ BTU/HR, FT}^2$. Experience has shown that a flux of 6000 to 8000 is a good starting value for the organics.

\[
Q = \frac{A}{FLUX} \quad \text{(C.1)}
\]
(5) From the tube length, tube outside diameter, tube BWG, tube layout pitch, find the tube count and the shell size from TABLE 13. If the heat transfer area is too big, reboilers in parallel will be assumed.

(6) Evaluate an operating overall heat transfer coefficient.

\[
Q \quad U = \frac{Q}{D \cdot A \cdot \text{LMTD}} \quad (C.2)
\]

(7) From the designed recirculation ratio, evaluate the total enthalpy at the column base, point A in FIG 8.

(8) Since flash occurring when fluid passes through the control valve E in FIG 8, it is a isenthalpic process between the tube bundle outlet of the reboiler and the column base. Since the column base, point A, condition is known from the steady state design, therefore; the vapor composition at the reboiler tube bundle outlet can be back calculated.

(9) Calculate the fluid mean density at the outlet of the reboiler tube bundle.

(10) Static pressure of the reboiler leg

(equation B.3)

(11) Calculate the inlet and outlet piping frictional resistance to flow.

(a) The inlet piping friction loss is calculated based on the equivalent length of 100 feet of piping,
three fully opened gate valves, one swing check valve and two 90 degree long sweep elbows, as shown in FIG 8. The fluid velocity in the piping is assumed to be 6 FT/SEC.

(b) The outlet piping friction loss is calculated based on the equivalent length of 100 feet of piping, one globe valve, one control valve, two 90 degree long swing elbows, two gate valves and two tees, as shown in FIG 8. The fluid velocity in the piping is assumed to be 6 FT/SEC.

(c) Friction resistance to flow inside tubes
(equation 3.4)

(12) Static pressure driving force
(equation 3.7)

(13) Calculate the pump horsepower from the mechanical energy balance.

Mechanical energy balance:

$$\Delta P + \frac{g}{c} \left( \rho * dz + \frac{1}{c} \right) \rho * V * dV \mid \frac{g}{L} + \frac{g}{L} \frac{L}{L} + \frac{g}{c} \frac{L}{c} \rho * dF = 0$$

(C.3)

The kinetic energy in this case can be neglected.

Refer to FIG 8, the mechanical energy equation can be re-written to be:
\[(P - P) + \int_{BC} \rho * dF + \int_{CD} \rho * dF + \frac{g}{g} \int_{C TB} \rho * dZ\]

\[+ \frac{g}{g} \int_{C} \rho * (H - H) - \rho * \frac{g}{g} \int_{C} W = 0 \quad \text{(C.4)}\]

where:

- \(P - P\) : pressure difference between point D and point A, PSI. The pressure drop through the control valve E and the friction loss by piping.

- \(\int_{BC} \rho * dF\) : friction loss by piping in section BC.

- \(\int_{CD} \rho * dF\) : friction loss by reboiler tubes.

- \(\frac{g}{g} \int_{C} \rho * dZ\) : static pressure of the reboiler leg.

- \(\frac{g}{g} \int_{C} (H - H)\): static pressure of the reboiler leg.

- \(\rho \frac{g}{g} \int_{C} W\) : pump shaft work.

(14) Compute the heat transfer coefficient

(equation B.8 or B.9).

(15) Calculate the clean overall heat transfer coefficient.

(equation B.10).

(16) Calculate the resistance of fouling, \(R\), required for the operation of reboiler.
(equation B.11).

(17) If this fouling seems too low for the service, then the unit must be redesigned.
APPENDIX D

DESIGN PROCEDURES OF THE KETTLE REBOILER

A kettle reboiler is shown in FIG 7. The kettle unit used in reboiling service usually has an internal weir to maintain a fixed liquid level and tube coverage. The residue drawoff is from the weir section. In this research, the U-tube kettle reboiler is designed by the following procedures: (1) Evaluate the heat load \( Q \). (2) Evaluate the LMTD by assuming the reboiler is under isothermal boiling.

(3) Calculate the steam mass flow rate.

(4) From the input design flux, calculate the approximate heat transfer area required.

\[
\frac{Q}{A} = \frac{\text{FLUX}}{\text{FLUX}} \quad (D.1)
\]

(5) From the input design tube length, tube outside diameter, BWG, pitch layout and tube passes, estimate a unit size (number of tubes, size of shell, etc). If the heat transfer area is too big, reboilers in parallel will be assumed. As pointed out by E. E. Ludwig (1965), one problem in the designing of U-tube heat exchanger area is to determine the effective
length of the tubes. The effective tube length of the tube bundle is the mean of the tube lengths between the outside tubes and the inside tubes as shown in FIG 52. For example, when U-tube bundle are fabricated from nominal tube length 12 ft tubes, the maximum tube length in the bundle is 12 ft, the inside tube is the shortest and less than 12 ft in length. A Chart, FIG 54, of tube effective length vs. number of tubes is prepared by E. E. Ludwig (1965) to avoid tedious calculation. The values read from this chart are not more than 1% lower than the detailed calculated values, except where the curves are extrapolated to lower tube count. The basis for the chart preparation and the correlations used in this research are summarized in TABLE 14. Therefore;

\[
\text{total effective bundle outside surface area} = (\text{effective length, FT}) \times (\text{number of tubes}) \times (\text{tube outside surface area per foot of length})
\]

(6) As suggested by Kern, the steam side (tube side) heat transfer coefficient can be taken to be 1500 BTU/HR,F,FT**2 for the design purpose.

(7) From the Levy's correlation,

(a) Evaluate the tube wall temperature.
(b) Determine the shell side coefficient.

Levy (1958) presented a correlation showing good
FIG 52. U-Tube Bundle.

FIG 53. Determination Of Coefficient $B_L$ From Boiling Heat Transfer Data By Levy.
FIG 54. Effective Tube Length Correlations For Calculating Outside Surface Area Of U-Tube Bundle.
TABLE (14).

Effective Length Of U-tubes

(1) E.L. = 31.50241245 - (0.00811674*NTUBE) + (5.35E-6*NTUBE)

(2) E.L. = 31.48453127 - (0.00771423*NTUBE) + (1.137E-5*NTUBE)

(3) E.L. = 23.59712783 - (0.01045344*NTUBE) + (1.824E-5*NTUBE)

(4) E.L. = 23.48453127 - (0.00771423*NTUBE) + (1.137E-5*NTUBE)

(5) E.L. = 15.59712783 - (0.01045344*NTUBE) + (1.824E-5*NTUBE)

(6) E.L. = 15.48401339 - (0.00773832*NTUBE) + (1.178E-5*NTUBE)

(7) E.L. = 11.52138647 - (0.00842008*NTUBE) + (6.1E-6*NTUBE)

(8) E.L. = 11.48401339 - (0.00773832*NTUBE) + (1.178E-5*NTUBE)

(9) E.L. = 7.58981565 - (0.0099268*NTUBE) + (1.1E-5*NTUBE)

(10) E.L. = 7.48004521 - (0.00742572*NTUBE) + (8.38E-6*NTUBE)

Where:
**E.L. = effective tube length, FT.**

**NTUBE = number of U-tubes.**

**Basis for the U-tube.**

<table>
<thead>
<tr>
<th>Tube O.D.</th>
<th>Minimum Radius Bend</th>
<th>Pitch</th>
</tr>
</thead>
<tbody>
<tr>
<td>3/4 &quot;</td>
<td>1.5 Tube Diameter</td>
<td>1 &quot;</td>
</tr>
<tr>
<td>1 &quot;</td>
<td>2.5 Tube Diameter</td>
<td>1.25&quot;</td>
</tr>
</tbody>
</table>
agreement for pool boiling and nucleate boiling heat transfer flux \((Q/A)\) below the critical delta \(T\) for subcooled and vapor containing liquids. This covers the pressure range of sub- and above- atmospheric and is obtained for inside and outside tube boiling.

\[
\frac{Q}{A} = \frac{2}{k} \frac{C}{L} \frac{\rho}{L} \frac{\Delta T}{\delta} (\frac{\delta}{T \cdot (\rho - \rho_v)} \frac{B}{S \cdot L \cdot V \cdot L})
\]

where:

- \(X\): vapor quality of fluid, zero for pool boiling.
- \(k\): thermal conductivity of saturated liquid, \(\text{BTU/HR,F,FT}\).
- \(C\): specific heat of liquid, \(\text{BTU/LB,F}\).
- \(\rho\): liquid density, \(\text{LB/FT}^3\).
- \(\rho_v\): vapor density, \(\text{LB/FT}^3\).
- \(\delta\): surface tension of liquid, \(\text{BTU/FT}^2\).
- \(\Delta T\): temperature difference, \(T - T_s\), \(\text{R}\).
- \(T_s\): saturation temperature of liquid, \(\text{R}\).
- \(h\): latent heat of vaporization, \(\text{BTU/LB}\).
- \(B\): refer to FIG 53. It can be calculated as follows.

The curve correlated is as follows:
\[ X = \log\left(\frac{\rho * h}{V f g}\right) \]

\[ Y = -4.38541408 + (14.5047494 * x) - (13.7760137 * x^2) \]
\[ + (6.11743947 * x^3) - (1.2116551 * x^4) + (0.08614632 * x^5) \]

\[ \frac{1.E6}{Y} \frac{1}{B} \frac{1}{L} = 10 \]  \hspace{1cm} (D.3)

8) Calculate the required area, based on film coefficient of step (6) and (7) together with fouling and tube wall resistance.

\[ Q = \frac{A}{U \Delta T} \]  \hspace{1cm} (D.4)

9) If the approximated unit does not have sufficient area, select a large size unit, and repeat the above procedure until the approximated unit is satisfactory.
APPENDIX E

DESIGN PROCEDURES OF

THE TOTAL CONDENSER OR THE PARTIAL CONDENSER

The design procedures for either the total condenser or partial condenser are summarized as follows:

(1) Establish the physical properties of process fluid at the average of process inlet and outlet temperature, shell side.

(2) Set an allowable temperature rise for coolant, (input data of the design program). Coolant is assumed to be in liquid phase at both the inlet and outlet of the tube bundle.

(3) Establish the physical properties of coolant at the average of coolant inlet and outlet temperature, tube side.

(4) (a) Determine the temperature range of condensation.

(b) Compute the condensing curve.

(c) Compute the weighted temperature.

(d) Calculate the heat load of condensation.

(5) Calculate the coolant rate, W.

\[
W = \frac{Q}{C \cdot \Delta T} \quad \text{LBS/HR (E.1)}
\]
where:

\[ Q \] : BTU/HR, heat load of condensation.
\[ \Delta T \] : temperature rise in water, F.
\[ C \] : heat capacity of coolant, BTU/LB,F.
\[ P \]

(6) Estimate number of tubes per pass to maintain minimum coolant velocity.

(a) Set minimum coolant velocity in tubes at 3.5–6.0 FT/SEC.

(b) Calculate the necessary total cross sectional area for the coolant flow.

(c) From the input tube outside diameter, tube BWG, calculate the flow area per tube.

(d) Estimate the number of tubes per pass.

(7) Assume a unit

(a) Estimate overall heat transfer coefficient \( U \). For the first iteration, \( U \) is the input data. After the first iteration, \( U \) is adjusted by step (k).

(b) Estimate heat transfer area, \( A \)

\[
A = \frac{Q}{U \times \text{weighted } T} \quad \text{(E.2)}
\]

(c) Total tube footage required, FTL

\[
FTL = \frac{A}{\text{sq. ft tube outside surface}} \quad \text{(E.3)}
\]

\[
\text{ft of tube length}
\]
(d) Number of tube passes

\[
\frac{FTL}{\text{(tubes per pass)} \times \text{(tube length)}} = \quad \text{(E.4)}
\]

If this value is not reasonable, re-assume the tube length, and/or the size of the tubes. If necessary, condensers in parallel will be assumed.

(e) From TABLE 13, pick an exchanger shell diameter which closely contains the required number of tubes at the required number of tube passes.

(f) From the actual tube count selected, establish the actual number of tubes/pass.

(g) Re-adjust the flow area per pass.

(h) Calculate coolant velocity in tubes again.

(i) Calculate the film coefficient tube side, \( h_i \).

\[ (\text{equation B.8 or B.9}). \]

(j) Calculate the film coefficient shell side, \( h_o \).

(j.1) Vertical tube bundle

\[
\frac{4 \cdot G'}{u} \quad \text{For} \quad \frac{\alpha}{u} < 2000, \text{Ludwig (1965).}
\]

\[
\begin{align*}
K \cdot \rho \cdot g & = k \cdot \rho \cdot g^* \pi^* D \cdot F^{1/3} \cdot f^{1/3} \\
\frac{h}{cm} & = 0.945 \left( \frac{\alpha \cdot G'}{u} \right) = 0.945 \left( \frac{\alpha \cdot W}{f} \right)
\end{align*}
\]
\[
G' = \frac{W}{\pi N * D}, \text{ LB/HR, linear foot} \quad (E.5)
\]

\[
4 \cdot G' = 0
\]

For \( \frac{C}{10} \) > 2000, Ludwig (1965).

\[
\begin{align*}
\frac{3}{2} & \quad \frac{K \cdot \rho \cdot g}{f \cdot f^{1/3} \cdot 4.0 \cdot W} \cdot 0.4 \\
\end{align*}
\]

\[
\begin{align*}
\text{cm} & \quad \frac{2}{f^{1/3}} \quad \frac{u \cdot \pi \cdot D}{f} & 0
\end{align*}
\]

usually applicable to long tubes and high flow rates, average film coefficient.

(j.2) Horizontal tube bundle, Ludwig (1965).

\[
4 \cdot G'' = 0
\]

For \( \frac{C}{10} \) < 2100

\[
\begin{align*}
\frac{3}{2} & \quad \frac{K \cdot \rho \cdot g}{f \cdot F^{1/3} \cdot 2/3} \\
\end{align*}
\]

\[
\begin{align*}
\text{cm} & \quad \frac{u \cdot g^*}{f} \quad \frac{u \cdot W}{f} \\
\end{align*}
\]

\[
\begin{align*}
G = \frac{W}{L \cdot N \cdot 2/3}, \text{ LB/HR, linear foot} \quad (E.7)
\end{align*}
\]

where:

\[
\begin{align*}
h & : \text{average value of condensing film coefficient, cm} \\
\text{cm} & \quad \text{BTU/hr, FT}^{**2}, \text{F.}
\end{align*}
\]
\[ U = \frac{1}{\frac{D \ln(\frac{D}{D_1})}{h_{io} + \frac{1}{r_{io}} + \frac{1}{r_{o}} + \frac{1}{h_{io}} + \frac{1}{2k_{io}}}} \]  

(E.8)

where:

- \( h \): \( h \) *(tube I.D.)/(tube O.D.)
- \( k \): thermal conductivity of the condenser tube metal, BTU/HR, FT, F.
- \( r_o \): shell side fouling factor.
- \( r_{io} \): tube side fouling factor.

- \( K_f \): thermal conductivity at film temperature, BTU/HR, FT, F.
- \( \rho_f \): density, LB/FT**3, at film temperature T_f.
- \( g \): acceleration of gravity, FT/HR**2.
- \( \mu_f \): viscosity at film, LB/FT, HR.
- \( W \): process flow rate, LB/HR.
- \( D_0 \): outside diameter of tubes, FT.
- \( N_t \): total number of tubes in bundle used for condensation.
- \( L \): tube length, FT.
(1) Area required.

\[
A = \frac{Q}{U \times (\text{weighted } T)} \quad \text{(E.9)}
\]

(2) Compared, and if the available area is equal to or larger than the required area, the select unit will perform satisfactorily. If the required area is greater than the available area, select a new unit with more tubes, longer tubes, larger diameter tubes or some combination and return to (7.a).
APPENDIX F

DESIGN OF THE REFLUX DRUM

For partial condenser's reflux drum, in the vapor, the droplet separation is attained by limiting vapor velocity which can be calculated by Newton's law.

\[
v = K \sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \approx K \sqrt{\frac{\rho_L}{\rho_V}} \quad (F.1)
\]

where:

- \( \rho_L \): liquid density, LB/FT**3, at design condition.
- \( \rho_V \): vapor density, LB/FT**3, at design condition.
- \( v \): vapor velocity, FT/SEC.
- \( K \): velocity constant,
  - (1) 0.13 FT/SEC for horizontal vessel.
  - (2) 0.26 FT/SEC for horizontal vessel having a wire-mesh demister.

The cross section of the reflux drum is shown as FIG 9.

If no second liquid phase is to be settled, L1 is situated above vessel bottom at a distance D-hL equal to 10% of the drum diameter. If possible, this distance should be greater than 5 inches. This circular segment of height 10% of the
diameter has an area equal to 5% of the total vessel cross sectional area. If we define $x_L$ as the percent of the drum cross sectional area occupied by the liquid at its Lh and calling $x_V$ the minimum free, vapor passage area above the Lh, expressed as percent of the drum's cross sectional area; then,

$$x_L + x_V = 95.0$$  \hspace{1cm} (F.2)

If the header volume is neglected; then,

vapor velocity $V$

$$\frac{M}{V} \left( \frac{3600 \cdot \rho}{V} \right)$$

$$= \frac{2}{(x_L /100) \cdot (\pi \cdot D /4)}$$

$$= \frac{\text{FT/SEC}}{V}$$ \hspace{1cm} (F.3)

holding time of liquid $t$

$$\frac{2}{(x_L /100) \cdot L \cdot (\pi \cdot D /4)}$$

$$= \frac{\text{MIN}}{L}$$ \hspace{1cm} (F.4)

where:

$D$ : reflux drum outside diameter, FT.

$L$ : reflux drum length, FT.

$M_L$ : liquid flow rate, LB/HR.

$M_V$ : vapor flow rate, LB/HR.

$t_s$ : $t$. 


Combine equation (F.3) and (F.4) and rearrange,

\[
\frac{x}{v} = \frac{L*M*p}{V*L} \quad \frac{v}{L} = \frac{V*t*M*p*60}{L^2} \quad (F.5)
\]

Substitute equation (F.1) into the above expression:

\[
\frac{x}{v} = \frac{L*M*a}{V} \quad \frac{v}{L} = \frac{K*t*M*60}{L^2} \quad (F.6)
\]

where \( a = \frac{p}{p_L} \)

Combine equation (F.2) with equation (F.6),

\[
\frac{x}{v} = \frac{60*t*K*M}{v} \quad L = \left( \frac{\text{---------}}{95-x} \right) * \left( \frac{\text{---------}}{a*M} \right) \quad (F.7)
\]

Solve eq. (F.4) for L, and combine with eq. (F.7).

\[
D = 0.188*\left( \frac{M*a}{v} \right)^{1/2} \quad 1 \quad (---), \quad PT \quad (F.8)
\]

Combine equation (F.7) and (F.8) by eliminating \( x \).

\[
\frac{2}{95.0*K*L*p} * \frac{D - 0.188*(a*M*L+60*t*K*M)}{v} = 0 \quad (F.9)
\]

The dimensions of the vessel are set by economic criteria, a good approximation might be the minimum vessel weight, with a correction factor for the heads, a typical
formula for vessel with 2:1 elliptical head is,

\[
P \frac{4}{2S*E - P} \left( \frac{1}{D} \right)^4 + 0.083D^3 - 0.0905V = 0 \quad (F.10)
\]

where:
- \(V\): vessel volume, GAL.
- \(P\): design pressure, PSIG.
- \(S\): allowable stress of the shell material, PSI.
- \(E\): joint efficiency, less than 1.

For a (2:1) ellipsoidal heads cylindrical vessel:

\[
V = 7.48 \left( \frac{\pi D}{4} \right)^2 \left( L + \frac{D}{3} \right), \text{ GAL (F-11)}
\]

Combining equation (F.10) with (F.11) by eliminating \(V\).

\[
L = \frac{P*D}{0.053*(2S*E-P)} + 1.227*D, \text{ FT (F.12)}
\]

Equation (F.9) and equation (F.12) are two nonlinear equations with two unknowns \(L\) and \(D\). Newton’s method is used in this research to solve the \(L\) and \(D\). Then the \(L\) and \(D\) are rounded off to the next commercial size.

Since there is no vapor stream for the reflux drum of total condenser; therefore, only the liquid surge time, i.e., the liquid holdup time between \(L_h\) and \(L_l\), should be concerned. The reflux drum volume between \(hH\) (20% \(D\)) and \(D-hL\) (10% \(D\)) in FIG 9 is ,
\[ V_h = 0.2014 \pi L D \]  

where \( V_h \) is the liquid holdup for the corresponding liquid surge time suggested by Sigales in TABLE 1. Equation (F.12) and equation (F.13) are still two nonlinear equations with two unknowns \( L \) and \( D \). Newton's method is used to solve for \( L \) and \( D \). Then, the \( L \) and \( D \) are rounded off to the next commercial size.
APPENDIX G

DESIGN OF THE TRAY COLUMN

The design of individual trays and tray column used in this research is based on the work published by Economopoulos (1978).

Tray Hydraulics

The steady state tray hydraulics are described as follows:

(1) dry tray pressure drop, $h_d$

Hughmark and O'Connell (1957) used the orifice equation to estimate the dry tray pressure drop.

$$h = \frac{0.186 \text{ CFS}}{\frac{V}{d^2}} \left[ \frac{\rho A}{2h} \right]$$

Where $C$ is the orifice coefficient, defined by:

$$C = \frac{880.6 - 67.7 h^2 + 7.32 h^3}{1000 t}$$

$$d \quad d \quad d \quad d \quad h \quad h \quad h \quad t \quad t \quad t$$

$$0 \quad 1000$$

(G.2)
(2) surface tension head, $h_{\sigma}$.

The surface tension head is defined as the necessary pressure to form a vapor bubble through the tray hole and is given by:

$$ h_{\sigma} = 0.04*0/\left( \rho * d \right) \quad (G.3) $$

(3) liquid crest over weir, $h_{ow}$.

The liquid crest over the weir can be estimated by the Francis weir formula:

$$ h_{ow} = 0.092*F*(L /GPM)^{2/3} \quad (G.4) $$

the correction factor, $F$, for multipass trays can be assumed to be 1.0.

(4) clear liquid height, $h_{l}$.

Fair (1963) correlated the effective liquid head to the operating liquid seal at the tray outlet weir ($h_{ow} + h_{w}$) by means of an aeration factor.

$$ h_{l} = \beta (h_{w} + h_{ow}) \quad (G.5) $$

the $\beta$ factor is estimated as a function of a vapor kinetic energy parameter, $F_s$

$$ F_s = (V / A)^{0.5} \quad (G.6) $$
the β is correlated as follows:

$$\beta = 0.977 - 0.619*F + 0.341*(F)^2 - 0.0636*(F)^3$$  \hspace{1cm} (G.7)

(5) head loss under the downcomer, $h_{ud}$

The liquid head loss for flow under the downcomer apron is calculated by:

$$h = \frac{L_{GPM}^2}{0.558 \times \frac{448.8*A_{ud}}{A_{ud}}}$$ \hspace{1cm} (G.8)

Normal construction practice yields approximately:

$$A = 0.42 \times A_{ud}$$  \hspace{1cm} (G.9)

Design Of A Single Tray

To obtain the necessary tray size, the following factors should be considered: number of flow passes; weir length; flow path length and width of flow path; downcomer area; active area; tray spacing; tray flood factor; and total tray area. The design is based on the steady state profile generated from the steady state program outlined in chapter two. Also the desired tray spacing and flood factor should be specified as the input data.

(1) Number of Flow Passes, $N_p$
The number of tray passes affects the weir length of the tray, hence, the liquid loading. For a given tray, the number of flow passes is increased, as necessary, to bring the weir loading within the proper design limits.

\[ \frac{L}{W} < 96 \quad \text{GPM} \quad L \quad (G.10) \]

Depending on the size of the tray, there is a limit to the number of flow passes. This limit is necessary in order to be able to construct internal manways and obtain a satisfactory tray efficiency. The maximum number of flow passes, \( N \), recommended by Economopoulos is given by:

\[ N = \frac{0.377 \times \sqrt{A}}{P_{\text{max}}} \quad (G.11) \]

where \( N \) is rounded to the next higher integer, but \( P_{\text{max}} \) never less than 1.

(2) Weir Length, \( W \), Flowpath Length, \( F \), and Width of Flowpath, \( W \).

Interior downcomers is assumed to have rectangular shapes and are placed in positions to divide the tray into equal flowpath lengths. The following relationships are recommended by Economopoulos:
\[
W = W_L + D \times (N - 1) \quad \text{(G.12)}
\]

\[
\begin{align*}
&\frac{A_d}{d} = 0.054 \quad \text{d (sd)} \\
&\frac{D - 2H}{t \sd} = -2 \left( \frac{N - 1}{D - P} \right) (1 - \frac{A}{t}) \\
&F = \frac{\text{-----}}{\text{----------}} \quad \text{(G.13)}
\end{align*}
\]

The side downcomers chord height, \( H \), is calculated from the following nonlinear equation:

\[
A \\
\frac{d (sd)}{t} = \cos \left( \frac{1 - 2 (1 - \frac{H}{sd})}{D} \right) - 2 \left( \frac{1 - 2 (1 - \frac{H}{sd})}{t} \right) \quad \text{(G.14)}
\]

The weir length, \( W \), of the sidedowncomer is calculated as follows:

\[
W = 2 \sqrt{\frac{H (D - H)}{L (sd) \sd t}} \quad \text{(G.15)}
\]

The average width of flow can be calculated as a function of the tray active area and the flowpath length.

\[
W = \frac{A}{FP} \quad \text{(G.16)}
\]

(3) Downcomer area, \( \frac{A}{d} \).


The design correlation recommended by the Glitsch design manual 4900 for sizing the ballast tray downcomer is
summarized as follows:

(1) \( A' \)

\[ A = \text{the larger of } \frac{d}{(a) \ 2A'} \quad \text{or} \quad \frac{d}{(b) \ 0.11A} \]

where:

\[ A' = \frac{L}{V \cdot F} \]  \hspace{1cm} \text{(G.17)}

\[ V = \text{the smaller of } \frac{(2) \ 41 \sqrt[3]{\frac{\rho - \rho S}{L \cdot V \cdot F}}}{d} \quad \text{(G.18)} \]

\[ (3) \ 7.5 \sqrt[3]{\frac{t' \cdot (\rho - \rho S)}{s \cdot L \cdot V \cdot F}} \]  \hspace{1cm} \text{(G.19)}


\[ A = \frac{L}{V \cdot F} \]  \hspace{1cm} \text{(G.20)}

where:

(1) \( 8.578t' \cdot S \) \hspace{1cm} \text{(G.21)}

\[ V = \text{the smaller of } \frac{0.533t' \cdot (\rho - \rho S)}{s \cdot L \cdot V \cdot F} \]

\[ t' = \text{the smaller of } \frac{t}{s} \quad \text{and} \quad 30. \]

\[ S : \text{system derating factor, that derates the tray's } \frac{F}{\text{loading handling capacity.}} \]
Fair (1963) related this factor to liquid surface tension:

\[ S = \frac{0.2}{F} \]  
\[ (G.22) \]

In Glitsch and Koch manuals, \( S \) is related to the foaming tendency of the system.

**TABLE 15. Derating Factors.**

<table>
<thead>
<tr>
<th>Derating Factor, ( S )</th>
<th>System Condition</th>
<th>example</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.0</td>
<td>Nonfoaming</td>
<td>Regular</td>
</tr>
<tr>
<td>0.85</td>
<td>Moderate foaming</td>
<td>Oil absorbers, amine and glycol regenerator.</td>
</tr>
<tr>
<td>0.73</td>
<td>Heavy foaming</td>
<td>Amine and glycol contactor.</td>
</tr>
<tr>
<td>0.60</td>
<td>Severe foaming</td>
<td>Methyl ethyl ketone units.</td>
</tr>
<tr>
<td>0.30</td>
<td>Form stable</td>
<td>Caustic regenerator.</td>
</tr>
</tbody>
</table>

(4) Active Area.


\[ V + L \ (F/1083) \]
\[ \text{load GPM PL} \]
\[ A = \frac{C *S *F}{AFO \ F \ f} \]  
\[ (G.23) \]

The Graph for the capacity factor, \( C \), has been correlated by Economopoulos as follows:
\( C = \text{the smaller of the} \quad \text{AF0} \)

\[
\begin{align*}
(1) & \quad \frac{0.65}{t} \quad \frac{0.167}{\rho} \quad \frac{1}{12.0} \\
(2) & \quad 0.3174 + 0.04122(t - 12) - 10 \quad \frac{\rho}{V} \quad (245 + 661t) \\
(3) & \quad 0.595 - 0.0596 \quad \frac{\rho}{V}
\end{align*}
\]

The vapor load factor, \( V \), is defined as

\[
V = V \frac{\rho}{\sqrt{\frac{\rho}{V} - \frac{\rho}{V}}}
\]

\[(G.25)\]

(4. B) Flexitray of the Koch Company.

\[
A = V \quad \frac{\rho}{(C \quad S \quad F)}
\]

\[
\text{a load} \quad \text{AF0} \quad \text{F} \quad f
\]

where:

\( C = \text{the smaller of} \quad \text{AF0} \)

\[
(1) \quad \frac{0.1667}{t} \quad \frac{1}{82.3}
\]

\[
\frac{c}{L} \quad \frac{0.533}{GPM}
\]

\[
(2) \quad c \quad (t) \quad \frac{1}{s} \quad (0.533 \quad \text{------------------------}) \quad \frac{c}{2}
\]

\[
479*W \quad \frac{F*c}{L} \quad F \quad 1 \quad s
\]

\[(G.27)\]

\( c = 0.153 \quad \text{for} \quad \rho < 1.5 \quad \frac{1}{V} \)
$0.172$ for $\frac{\rho}{V} > 1.5$

$c = 0.587$ for $\frac{\rho}{V} < 1.5$

$0.546$ for $\frac{\rho}{V} > 1.5$

(4.C) Fair's method.

Fair estimates the tray active area by using a flow parameter, $F$, instead of the vapor load factor, $\frac{V_{lv}}{V_{load}}$.

The flow parameter accounts for the liquid/vapor kinetic energy effects and is defined by:

$$F = \frac{L_{GPM}}{448.8*V_{CFS}} \sqrt{\frac{\rho}{V_{LV}}}$$

The active area is estimated from equation (G.26) and $S_F$ from equation (G.22). The capacity factor, $C_F$, is correlated by Economopoulos from Fair's flood capacity graph.

$$C_F = \text{the smaller of}$$

$$\begin{align*}
(1) & \quad 0.118 * \exp(0.0479*t) \\
(2) & \quad 0.425 * \exp(0.0479*t) * (0.1092 - 0.058*\ln(F_{lv}))
\end{align*}$$

The active area calculated by the Fair's method is independent of the number of the tray passes.
(5) total column area

When straight downcomers are used, or when using sloped or stepped downcomers with recessed inlet areas or drawoff sumps, the column area is given by:

\[ A = A + 2A \frac{t}{d} \] (G.30)

A can be considered as the average of the downcomer areas of odd numbered and even numbered of trays.

**Tray Hole Area**

The tray hole area affects the tray operating flexibility through the dry tray pressure drop. Reduction in the hole area reduces the weep point and, up to a certain point, increases the tray turndown ratio. The turndown ratio is defined as the ratio of maximum allowable flooding to minimum allowable operating throughput. Beyond this point, it can cause premature flooding due to downcomer backup and/or undesirably high pressure drop.

Operation with vapor loads below those at the weep point is not predictable. Hence, the weep point is considered the lower design limit of a given tray. The flood point is the upper limit.

The hole area corresponds to a given overall pressure drop can be solved from equation (G.1) together with C from
equation (G.2).

\[
A = \frac{\frac{V}{CFS}}{h} \sqrt{\frac{\sqrt{\left(\frac{\frac{CFS}{2}}{2L} \cdot \frac{\left(h - h - h\right)}{\left(h - h - h\right)}\right) + 5.38C}}{\frac{\left(h - h - h\right)}{\left(h - h - h\right)}}}
\]

\[ (G.31) \]

**Maximum Tray Hole Area**

Since the weep point is the lower design limit of a given tray, the minimum pressure drop required to maintain the operating conditions at the weep point is given by Economopoulos as:

\[
h_{(\text{min})} = h + 0.35(h + h + h)
\]

\[ (G.32) \]

The maximum permissible hole area that will keep the tray above the weep point at the minimum design vapor rates is given by:

\[
A_{(\text{max})} = \frac{\frac{F \cdot V}{CFS}}{h} \sqrt{\frac{\frac{F \cdot V}{w \cdot CFS}}{\left(\frac{\left(h - h - h\right) \cdot \left(h - h - h\right)}{\left(h - h - h\right)}\right) + 5.38C}}
\]

\[ (G.33) \]

For tray hydraulic stability, the tray hole area is not allowed to exceed 15% of the tray active area. Thus:
\[ A = \text{the smaller of } A \text{ and } 0.15*A \]
\[ h(\text{max}) \quad h(\text{max}) \quad a \]

**Minimum Tray Hole Area**

To avoid premature tray flooding, the maximum design pressure drop has to be calculated at the maximum expected liquid rate (i.e. \( \frac{L}{F} \)). The equation is derived by Economopoulos:

\[
h = \frac{h_{\text{ow}}}{\rho} - \frac{\rho L V}{2/3 w} \left( \frac{h}{t(\text{md})} - \frac{2}{\rho L} \right) \quad (\text{F}) \quad (\text{F}) \]  

\[
= (0.5*t -0.5*h - \frac{\rho L V}{2/3 w}) \left( \frac{h}{t(\text{md})} - \frac{2}{\rho L} \right) \quad (\text{F}) \quad (\text{F}) \]  

The minimum hole area, which for the maximum vapor flow will cause a tray pressure drop equal to \( h \), is:

\[
A = \frac{V}{F} \frac{CFS}{f} \left( \frac{V}{CFS} \frac{2}{\rho L} \left( \frac{h}{t(\text{md})} - \frac{2}{\rho L} \right) + 5.38*(C) \right) \left( \frac{h}{a} - \frac{h}{h} \right) \]  

\[
= \frac{V}{F} \frac{CFS}{f} \left( \frac{V}{CFS} \frac{2}{\rho L} \left( \frac{h}{t(\text{md})} - \frac{2}{\rho L} \right) + 5.38*(C) \right) \left( \frac{h}{a} - \frac{h}{h} \right) \]  

\[
= \frac{V}{F} \frac{CFS}{f} \left( \frac{V}{CFS} \frac{2}{\rho L} \left( \frac{h}{t(\text{md})} - \frac{2}{\rho L} \right) + 5.38*(C) \right) \left( \frac{h}{a} - \frac{h}{h} \right) \]  

For towers operating with low pressure or equipped with many stages, a maximum permissible pressure drop across each tray is specified. This create the additional vapor head loss
limitation:

\[
h = \left(\frac{1728}{\rho} \right) \frac{\Delta P}{L_{\text{max}}} \quad (G.36)
\]

The corresponding minimum tray hole area can be calculated from equation (G.31) as:

\[
\frac{V}{CFS} = \frac{\rho}{h(\text{mc})} \sqrt{\frac{V}{CFS 2} + 5.38(C) \left(\frac{2}{L_{\text{max}}} \frac{h_{\text{mc}}}{t(\text{mc})} - h - h_{\text{md}} \right)}
\]

\[a \quad \text{(G.37)}\]

For operating stability, the hole area is not allowed to be less than 5% of the active area. Then:

\[A = \text{the larger of } A_{\text{h(min)}}, A_{\text{h(md)}}, 0.05 \cdot A_{\text{a}}
\]

**Minimum Tray Spacing**

Assuming that \(A > A_{\text{min}}\) and also \(h(\text{max}) > h(\text{min})\):

\[A_{\text{h(min)}} < A < A_{\text{h(max)}}\]

In this region, the tray weep factor is equal to or better than the design value, and the tray pressure drop does not exceed the maximum specified. In the event that \(A < A_{\text{min}}\), set \(A = A_{\text{h(max)}}\).

so that the tray does not weep prematurely.
The minimum tray spacing, \( t \), is calculated as

\[
s_{\text{md}}(\text{m/d})
\]

follows to avoid premature tray flooding by downcomer backup, when the tray hole area is \( A \).

\[
h = h + h + h_{\text{md}} + \frac{\rho}{1 - \sqrt{\frac{V}{F}}}
\]

\[
A \frac{V}{F} \frac{1}{2} \frac{\rho}{1 - \sqrt{\frac{V}{F}}}
\]

\[
C \frac{h}{L} \frac{A}{a}
\]

Introducing \( h \) from equation (G.38) into equation (G.34) and solving for \( t \), the minimum tray spacing is calculated as follows:

\[
t = 2 \left( \frac{h}{\rho - \rho_{\text{md}}} \right) + 2 \frac{h_{\text{ud}}}{\rho} + \frac{h_{\text{ow}}}{w} + \frac{h_{\text{ow}}}{w^{2/3}}
\]

\[
s_{\text{md}}(\text{m/d}) \rho \frac{L}{V} \frac{t_{\text{md}}}{2 \sqrt{\frac{F}{f}}} + \frac{h_{\text{ud}}}{\rho} \frac{h_{\text{ow}}}{w} \frac{h_{\text{ow}}}{w^{2/3}}
\]

(G.39)

Design Of Individual Trays

The flow chart for the design of individual trays by Economopoulos is shown in FIG 55 and FIG 56.

The design of individual trays starts by assuming a one pass tray, and estimating the tray active area by Fair's correlation (1963), using input only the tray overflow
FIG 55.

Single Tray Design Routine Proposed By Economopoulos.
Approximate estimation of $A_i$ from Eq. (36) and (37)

Call AD
Calculate $A_d$

Call TG
Calculate $W_L$, $F_{Lc}$, $W_{pp}$

Call AA
Calculate $A_a$

Yes

$A_L = A_d + 2A_a$

Yes

$N_p < N_{p_{max}}$ ?

Yes

$N_p = N_p + 1$

Stop

No

$N_p < N_{p_{max}}$ ?

Yes

$\frac{L_{min}}{W_{pp}} < 240$ ?

Yes

$N_p = N_p + 1$

Stop

No

$\frac{L_{min}}{W_{pp}} < 95$ ?

No

$\frac{L_{min}}{W_{pp}} < 95$ ?

Yes

Return

Fig. 56 Subroutine TSG for Tray Size and Geometry Proposed by Economopoulos
rates, phase densities, and liquid surface tension.

Following this, the tray active area and the downcomer area is estimated by either the Ballast tray method or the Flexitray method. Then, calculate the associated weir and flowpath lengths. Due to the interaction between the number of flow passes, the weir and flowpath lengths, the downcomer area and the active area, the calculation procedure is iterative and proceeds until convergence.

Liquid loading of the weir is calculated in the final step to assess whether it falls within the recommended design limits. If not, the number of tray passes will have to be increased by one, provided this is permitted by the size of the tray which is calculated from equation (G.10). With a new \( N \), the calculations for the tray active and downcomer areas and weir and flowpath lengths are repeated.

Having fixed the tray size and geometry, the maximum and minimum hole areas are estimated, as the flowchart shows. If \( A_{\text{p}} < A_{\text{h(min)}} \), the tray is designed with the minimum permissible hole area for maximum operating flexibility. Otherwise, the tray hole area is made equal to \( A_{\text{h(max)}} \), so that the desirable operating flexibility is maintained while the tray spacing if necessary.
Design Of Tray Columns

The tray column design flowchart is shown in FIG 57. The procedures are summarized by Economopoulos as follows:

1. The size and geometry of each individual tray is calculated by the individual tray design procedures. The design requires the system properties and tray spacing \( t \), as specified by the user. The tower cross sectional area is set equal to the larger tray size and is rounded off to correspond to a standard diameter.

2. Tray zones with identical number of flow passes, and active and downcomer areas are established. The zone, and active and downcomer areas are made equal to the largest active and downcomer areas of the trays in the zone. The extent of the zone is determined by the condition that the resulting area of the zone tray does not exceed the tower cross sectional area. Following this, the calculated zone areas are normalized, and the associated weir length, flowpath length, and width of flowpath of the zone tray are determined.

3. The increase in the size of smaller trays in (1) which made them equal to the size of the larger trays in the column, has proportionally enlarged their capacity. Advantage is taken of this extra capacity by reducing
FIG 57. Tray Column Design Flowchart Proposed by Economopoulos
the tray spacing to the minimum required. Hence all trays in the column will operate with a uniform flood factor. This reduces the overall height of the tower vessel.

(4) Tray zones with identical design parameters, including the hole area, are established. The maximum allowable hole area of the tray zone, \( A_{h(\text{max})} \), is equal to the smaller of the maximum hole areas of the individual trays in the zone. Similarly, the minimum permitted hole area of the tray zone, \( A_{h(\text{min})} \), is equal to the larger of the minimum hole areas of the individual trays in the zone. The extent of the zone, which contains trays with identical design parameters, is established by the condition

\[
A_{h(\text{max})} > A_{h(\text{min})}
\]

(5) Based on the calculated hole area of the tray zone, the minimum tray spacing required to prevent premature downcomer backup, \( t_{s(md)} \), is calculated for each tray. The tray spacing, \( t_s \), established in (3) is adjusted if necessary.

The calculations continue from step (2) until all stages of the tray column have been classified in a tray zone.
Nomenclature For the Tray Column Design

A : Active tray area, FT**2.
a
A : downcomer area, FT**2.
d
A : hole area of the tray, FT**2.
h
A : total tray area, FT**2.
t
A : free area for liquid flow under the downcomer apron, ud
    FT**2.

C : vapor discharge coefficient for dry tray. dimensionless.
0
C : tray capacity factor, as defined by equation (G.24), AP0
    (G.27) or equation (G.29).

D : tray diameter, FT.
t
F : tray flood factor.
f
F : length of liquid flow path on the tray, FT.
PL
F : tray weep factor.
w
h : clear or effective liquid height in the tray, INCH.
l
H : height of liquid crest over weir, INCH.
ow
\( h \) : head loss due to liquid flow under downcomer apron, in.
\( ud \) : INCH.
\( h \) : weir height, INCH.
\( w \)
\( L \) : liquid overflow rate, GPM.
\( GPM \)
\( N \) : number of tray flow passes.
\( p \)
\( t \) : tray deck thickness, INCH.
\( d \)
\( t \) : tray spacing, INCH.
\( s \)
\( V \) : vapor flow rate, FT**2/SEC.
\( CPS \)
\( W \) : average width of liquid flowpath on the tray, FT.
\( PP \)
\( W \) : weir length, FT.
\( L \)
In Laplace transformation notation, the generalized equilibrium tray equations become:

(1) total material balance:

\[ M(S) * s = L_n(S) + V_n(S) + P_{n-1}L_n(S) + FV_n(S) - L_n(S) \]

\[ - SS(S) - V_n(S) - SSV_n(S) \]

\[(H.1)\]

(2) component material balances, (for \(i=1, NC-1\)).

After the equation (4.19) is linearized and Laplace transformed, substitute \(M(S) * s\) by equation (H.1)
And \(Y_{i,n}(s)\) and \(Y_{i,n+1}(s)\) by,

\[ Y_{i,n}(s) = K * X_{i,n}(s) + X_{i,n} * C1 * T_{i,n}(s) \]

\[ + X_{i,n} * C8 * P_{i,n}(s) \]

\[(H.2)\]

\[ Y_{i,n+1}(s) = K * X_{i,n+1}(s) + X_{i,n+1} * C1 * T_{i,n+1}(s) \]

\[ + X_{i,n+1} * C8 * P_{i,n+1}(s) \]

\[(H.3)\]

Then, the Laplace transformed component material balance becomes:
\[
L(s)(\bar{X} - \bar{X}) + X(s)(\bar{L}) + V(s)(\bar{X} - \bar{Y})
\]
\[
- T(s)\left(\bar{X} * C1 * (V + SSV)\right) - P(s)\left((\bar{V} + SSV) * \bar{X} * C8\right)
\]
\[
- X(s)(L + SSL + M * s + K * \bar{V} + K * SSV)
\]
\[
+ V(s)(\bar{Y} - \bar{X}) + T(s)(\bar{X} * C1 * \bar{V})
\]
\[
+ P(s)(\bar{V} * \bar{X} * C8) + X(s)(\bar{K} * \bar{V})
\]
\[
= - FL(s)(ZL - \bar{X}) - FV(s)(\bar{ZV} - \bar{X})
\]
\[
- SSV(s)(\bar{X} - \bar{Y}) - ZL(s)(FL) - ZV(s)(FV)
\]
\[
(3) \text{enthalpy balance}
\]
After the Equation (4.20) is linearized and Laplace transformed, substitute \(M(s)*s\) by equation (H.1).

Also,
\[
HL(s) = C6 * T(s) + \Sigma C7 * X(s)
\]
\[
HL(s) = C6 * T(s) + \Sigma C7 * X(s)
\]
\[ H_V (s) = (\sum_{i=2}^{n} c_{i} \cdot c_{i,n} + c_{4} \cdot c_{i,n} \cdot c_{n}) \cdot T(s) + \sum_{i=3}^{n} x_{i} (s) + \sum_{i=2}^{n} c_{i} \cdot c_{8} \cdot p(s) \]

\[ H_V (s) = (\sum_{i=2}^{n+1} c_{i} \cdot c_{1,n+1} + c_{4} \cdot c_{i,n+1} \cdot c_{n+1}) \cdot T(s) + \sum_{i=3}^{n+1} x_{i} (s) + \sum_{i=2}^{n+1} c_{i} \cdot c_{8} \cdot p(s) \]

\[ H_{FL} (s) = (c_{6} \cdot f_{i,i,n} \cdot f_{i,n} \cdot f_{i,n}) \cdot T_{F}(s) + \sum_{i=7}^{n} z_{i} (s) \]

\[ H_{FV} (s) = (c_{2} \cdot f_{i,i,n} \cdot f_{i,n} \cdot f_{i,n}) \cdot T_{F}(s) + 2c_{3} \cdot z_{i} (s) \]

Then, the Laplace domain enthalpy balance becomes:

\[ L(n-1) (\sum_{i=1}^{n} L_{i} - L_{i,n-1}) + T(n-1) (L \cdot c_{6}) \]

\[ + \sum_{i=n-1}^{n} L_{i} \cdot c_{7} \cdot x_{i} (s) + V(n) (\sum_{i=n}^{n} H_{L} - H_{V}) \]

\[ + T(s) (-\sum_{i=n}^{n} V + SSV) (\sum_{i=n}^{n} c_{2} \cdot c_{1} + c_{4}) - (L + S_{SL}) \cdot c_{6} \]

\[ - s*(\sum_{i=n}^{n} M \cdot c_{6} + \sum_{i=n}^{n} c_{M} c_{C}) \]

\[ + P(s) (-\sum_{i=n}^{n} V + SSV) (\sum_{i=n}^{n} c_{2} \cdot c_{8}) \]


\[ + \sum \left[ -(\bar{V} + SSV_C^3)_{i,n} - (\bar{L} + SSL_C^7)_{i,n} - s*(\bar{m} + \bar{c}^7)_{i,n} \right] X(s)_{i,n} \]

\[ + V_{n+1}(s)*(\bar{H}_V - \bar{H}_L)_{i,n} \]

\[ + T_{n+1}(s)\left[ \bar{V}_{i,n+1} \left( \sum C^2_{i,n+1} + C^4_{i,n+1} \right) \right] \]

\[ + P_{n+1}(s)\left[ \bar{V}_{i,n+1} \left( \sum C^2_{i,n+1} + C^8_{i,n+1} \right) \right] \]

\[ = F_L(s)*(\bar{H}_L - HFL)_{i,n} + F_V(s)*(\bar{H}_V - HFL)_{i,n} + SSV(s)*(\bar{H}_V - HFL)_{i,n} \]

\[ - T_F(s)*\left( F_L * C^6_{i,n} + F_V * C^2_{i,n} \right) - F_L \left( \sum C^7_{i,n} * Z_L(s) \right) \]

\[ - F_V \left( \sum C^3_{i,n} * Z_V(s) \right) - Q(s)_{i,n} \]

(4) equilibrium relationships.

\[ Y(s) = K * X(s)_{i,n} + X * C^1_{i,n} * T(s)_{i,n} + X * C^8_{i,n} * P(s)_{i,n} \]

(5) summation equations.

\[ \Sigma X(s)_{i,n} = 0_{i,n} \]

\[ \Sigma \left( K * X(s)_{i,n} + X * C^1_{i,n} * T(s)_{i,n} + X * C^8_{i,n} * P(s)_{i,n} \right) = 0_{i,n} \]

(6) feed split equations.

\[ F(s)_{n} = F_L(s)_{n} + F_V(s)_{n} \]
\[
\bar{Z} * F (s) + \bar{F} * Z (s) = \bar{ZL} * FL (s) + \bar{FL} * ZL (s)
\]
\[
i, n \ n \ n \ i, n \ n \ n \ i, n
\]
\[\]
\[+ \bar{ZV} * FV (s) + \bar{FV} * ZV (s)
\]
\[i, n \ n \ n \ i, n \]

(7) thermal properties.

\[H_L (s) = C_6 * T (s) + \sum_{i} C_7 * X (s)
\]
\[n \ n \ n \ i \ i, n \ n \ n \ i, n \]

\[H_V (s) = (\sum_{i} C_2 * C_1 + C_4) * T (s) + \sum_{i} C_3 * X (s)
\]
\[n \ n \ n \ i \ i, n \ n \ n \ i \ i, n \ i, n \]
\[+ \sum_{i} C_2 * C_8 * P (s)
\]
\[i \ i, n \ i, n \ n \]

(8) tray hydraulics.

(a) downcomer liquid level.

\[\frac{W (s) - W_{DC} (s)}{n - 1} = \frac{H_{DC} (s)}{n \ \rho_{DC} * A_{DC} * s}
\]

(b) downcomer pressure drop and flow.

Equation (4.31) may be Laplace transformed to be:

\[\Delta P_{DC} (s) = \frac{q}{n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n}
\[\frac{\rho_{DC} * H_{DC} (s) - \rho_{TR} * H_{TR}' (s) - H_{TR}' * \rho_{TR} (s)}{n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n}
\[c - (P' (s) - P' (s))
\]

\[n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n \ n\]
In Laplace domain, equation (4.34) becomes,

\[
\frac{\partial WDC}{\partial \Delta PDC} \cdot \frac{n}{n} WDC (s) = (\frac{- \Delta PDC (s)}{\partial \Delta PDC} \cdot \frac{n}{n}) (H.16)
\]

and

\[
\frac{\partial WDC}{\partial \Delta PDC} \cdot \frac{2}{n} \frac{\partial ADM}{\partial \rho DC} \cdot \frac{g}{c} \cdot \frac{n}{n} = 0.5 \cdot \frac{\Delta PDC (s)}{\partial \Delta PDC} \cdot \frac{0.10072}{n} \cdot \frac{WDC (s)}{g} (H.17)
\]

\[
1.75139 \cdot \frac{\partial ADC}{\partial \rho DC} \cdot \frac{g}{c} \cdot \frac{n}{n} = \frac{2}{n} \cdot \frac{WDC (s)}{g} (H.17)
\]

(c) aerated liquid holdup and gradient on tray.

\[
\frac{VTR (s)}{A TR} = \frac{n}{n} \cdot \frac{HOW (s) + HTR' (s)}{2} (H.18)
\]

(d) inlet liquid height over weir.

\[
\frac{HOW (s)}{n} = \frac{HTR' (s)}{n} (H.19)
\]

(e) volume of liquid on the active tray.

\[
\frac{\partial VTR}{\partial WTR} \cdot \frac{n}{n} VTR (s) = \frac{\partial VTR}{\partial \rho TR} \cdot \frac{n}{n} VTR (s) + \frac{\partial VTR}{\partial \rho TR} \cdot \frac{n}{n} \rho TR (s) (H.20)
\]
where:

\[
\begin{align*}
\frac{\partial V_{TR}}{\partial \rho_{TR}} \frac{1}{n} &= \frac{\partial V_{TR}}{\partial \rho_{TR}} \frac{1}{n} - \frac{\partial W_{TR}}{\partial \rho_{TR}} \frac{2}{n} \frac{V_{TR}}{n} \\
\end{align*}
\]

(f) active tray total material balance.

\[
W_{TR}(s) = \left( W_{V}(s) - W_{V}(s) + W_{DC}(s) - W_{L}(s) + W_{FL}(s) \right. \\
+ W_{PV}(s) - W_{SSL}(s) - W_{SSV}(s) \left. \right) / s
\]

(g) aerated liquid density as a function of vapor velocity.

\[
\begin{align*}
\rho_{TR}(s) &= \frac{\partial \rho_{TR}}{\partial F} * \frac{\partial F}{n} * W_{V}(s) \\
\end{align*}
\]

where:

\[
\begin{align*}
\frac{\partial F}{\partial \rho_{V}} \frac{0.5}{n} &= \frac{\partial F}{\partial \rho_{V}} \frac{0.5}{n} \\
\frac{\partial W_{V}}{\partial A_{TR}} \frac{A_{TR}}{n} &= \frac{\partial W_{V}}{\partial A_{TR}} \frac{A_{TR}}{n} \\
\frac{\partial \rho_{TR}}{\partial F} \frac{-1.238 + 1.364 \bar{F} - 0.3816 \bar{F}^2}{n} &= \frac{\partial \rho_{TR}}{\partial F} \frac{-1.238 + 1.364 \bar{F} - 0.3816 \bar{F}^2}{n} \rho_{L}
\end{align*}
\]
(h) liquid overflow from tray.

\[ \frac{\partial W}{\partial \rho_{TR}} n + \frac{\partial W}{\partial \rho_{H}} n = W(s) = -\rho_{TR}(s) + \rho_{H}(s) (H.25) \]

where:

\[ \frac{\partial W}{\partial \rho_{TR}} n = KK \frac{3/2}{\rho_{TR}} n \]

\[ \frac{\partial W}{\partial \rho_{H}} n = KK \frac{0.5}{\rho_{H}} n \]

(j) total tray pressure drop.

\[ P'(s) - P'(s) = \Delta P_{TR}'(s) \]

\[ \frac{\partial \Delta P_{TR}'}{\partial W_{V}} n - \frac{\partial \Delta P_{TR}'}{\partial W_{V}} n-1 = \frac{\partial \Delta P_{TR}'}{\partial W_{V}} n + \frac{\partial \Delta P_{TR}'}{\partial W_{V}} n-1 \]

where:

(1) for tray valve unit part opened.

\[ \frac{\partial \Delta P_{TR}'}{\partial W_{V}} n - \frac{\partial \Delta P_{TR}'}{\partial W_{V}} n-1 = 0.0333 \frac{g}{\rho_{V}} n c \frac{\rho_{V} * AH}{n(n-1)} \]

(1) for tray valve unit fully opened.
At a first step in deriving an overall tray equation, a signal flow diagram can be constructed from equation (H.14), (H.15), (H.16), (H.18), (H.19), (H.20), (H.22), (H.23), (H.25) and (H.27). The signal flow diagram for a generalized tray (not including the column top tray) is shown in FIG 58. By successive reductions, this signal flow diagram can be reduced until the following equation is derived:

\[
W(s) = \frac{1 + 2*\xi*TR*s + 2}{n}\]

\[
\frac{\partial \Delta PTR}{\partial W} = 0.0175 * \frac{g}{n} \frac{n}{n} * \frac{g}{n} \frac{2}{2} \frac{c}{c} \frac{\rho V*AH}{n} \frac{n-1}{n-1} \tag{H.29}
\]

\[
\frac{\partial \Delta PTR}{\partial W} = 0.248525 * \frac{g}{n-1} \frac{n-1}{n-1} * \frac{\rho L}{W} \frac{n-1}{n-1} \tag{H.30}
\]
\[ * (TVV \cdot ADC \cdot s + 1) \]
\[ + WV (s) * \]
\[ \begin{cases} 
\text{where:} \\
\begin{align*}
1.5 & \cdot \frac{\text{TVV}}{\text{ADC}} \\
\rho_{\text{TR}} & \cdot \text{ATR} \\
\text{TVV} & = \frac{c}{c} \\
\text{TR} & = \frac{\text{ADT} \cdot \text{PDC}}{\text{WDC}} \\
\text{TVV} & = \frac{c}{c} \\
\end{align*}
\end{cases} \]
FIG 58. Signal Diagram For Generalized Tray Material Balance Dynamics.
In Laplace transformation notation, the mass flow rate of vapor or liquid can be related to the molar flow rate of vapor or liquid through the mixture's molecular weight as follows:

\[
W(s) = \sum_{i=1}^{n} \sum_{i,n=1}^{n-1} L(s)(\Sigma X_{i,n} \cdot MW_{i}) + \sum_{i=1}^{n} \sum_{i,n=1}^{n-1} L(s)\left(\sum_{i=1}^{n} MW_{i} \cdot X_{i,n}(s)\right)
\]

\[
W(s) = \sum_{i=1}^{n} \sum_{i=1}^{n} L(s)(\Sigma X_{i,n} \cdot MW_{i}) + \sum_{i=1}^{n} \sum_{i=1}^{n} L(s)\left(\sum_{i=1}^{n} MW_{i} \cdot X_{i,n}(s)\right)
\]

\[
WV(s) = \sum_{i=1}^{n} \sum_{i=1}^{n} V(s)(\Sigma Y_{i,n} \cdot MW_{i}) + \sum_{i=1}^{n} \sum_{i=1}^{n} V(s)\left(\sum_{i=1}^{n} MW_{i} \cdot Y_{i,n}(s)\right)
\]

\[
WV(s) = \sum_{i=1}^{n+1} \sum_{i=1}^{n+1} V(s)(\Sigma Y_{i,n+1} \cdot MW_{i}) + \sum_{i=1}^{n+1} \sum_{i=1}^{n+1} V(s)\left(\sum_{i=1}^{n+1} MW_{i} \cdot Y_{i,n+1}(s)\right)
\]

\[
WPL(s) = \sum_{i=1}^{n} \sum_{i=1}^{n} PL(s)(\Sigma Y_{i,n} \cdot MW_{i}) + \sum_{i=1}^{n} \sum_{i=1}^{n} PL(s)\left(\sum_{i=1}^{n} MW_{i} \cdot Y_{i,n}(s)\right)
\]

\[
WPL(s) = \sum_{i=1}^{n} \sum_{i=1}^{n} PL(s)(\Sigma Y_{i,n} \cdot MW_{i}) + \sum_{i=1}^{n} \sum_{i=1}^{n} PL(s)\left(\sum_{i=1}^{n} MW_{i} \cdot Y_{i,n}(s)\right)
\]

\[
WFV(s) = \sum_{i=1}^{n} \sum_{i=1}^{n} FV(s)(\Sigma Y_{i,n} \cdot MW_{i}) + \sum_{i=1}^{n} \sum_{i=1}^{n} FV(s)\left(\sum_{i=1}^{n} MW_{i} \cdot Y_{i,n}(s)\right)
\]

\[
WFV(s) = \sum_{i=1}^{n} \sum_{i=1}^{n} FV(s)(\Sigma Y_{i,n} \cdot MW_{i}) + \sum_{i=1}^{n} \sum_{i=1}^{n} FV(s)\left(\sum_{i=1}^{n} MW_{i} \cdot Y_{i,n}(s)\right)
\]
\[ WSSV(s) = SSV(s) \left( \sum_{i,n} Y(s) * MW \right) + SSV \left( \sum_{i,n} MW * Y(s) \right) \]

\[ WSSL(s) = SSL(s) \left( \sum_{i,n} X(s) * MW \right) + SSL \left( \sum_{i,n} MW * X(s) \right) \]

(H.33)

Substitute the equation set (H.33) into equation (H.32).

and \( Y_i(s) \) and \( Y_{i+1}(s) \) by,

\[ Y_i(s) = K_i \cdot X_i(s) + C_i \cdot T_i(s) + C_i \cdot P_i(s) \]

\[ Y_{i+1}(s) = K_{i+1} \cdot X_{i+1}(s) + C_{i+1} \cdot T_{i+1}(s) + C_{i+1} \cdot P_{i+1}(s) \]

Then, the equation (H.32) becomes to be,

\[ L(s) \left( \sum_{i,n} A * MW \right) \left( \sum_{i,n} Y(s) \right) \left( 1 - \frac{1}{\sum_{i,n} X(s)} \right) \]
\[ + P(s) \{(\Sigma(-AA*BB*SSV*MW + AA*CC*MW*V)*X*C8) \} \\
+ \Sigma((-AA*BB*SSV*MW + AA*CC*MW*V)*\bar{X}*)K \]
\[ = P(s) \{(\Sigma(-AA*BB*SSV*MW + AA*CC*MW*V)*X*C8) \} \\
+ \Sigma((-AA*BB*SSV*MW + AA*CC*MW*V)*\bar{X}*)K \]

\[ + V(s) \{(\Sigma(-AA*BB*SSV*MW + AA*CC*MW*V)*X*C8) \} \\
+ T(s) \{(\Sigma(-AA*BB*SSV*MW + AA*CC*MW*V)*X*C8) \} \\
+ P(s) \{(\Sigma(-AA*BB*SSV*MW + AA*CC*MW*V)*X*C8) \} \\
+ \Sigma((-AA*BB*SSV*MW + AA*CC*MW*V)*\bar{X}*)K \]

where:

\[ AA = \frac{1}{n^2} \]
\[
\begin{align*}
BB &= \left( -\frac{\partial P}{\partial g} \right)_{\text{ADC}} * s + 1 \\
&= \left( -\frac{\partial W D C}{\partial g} \right)_{\text{ADC}} * s + 1 \\
&= \left( -\frac{\partial \Delta P D C}{\partial g} \right)_{\text{ADC}} * s + 1 \\
&= \left( -\frac{\partial F}{\partial g} \right)_{\text{ADC}} * s + 1 \\
&= \left( -\frac{\partial \rho T R}{\partial g} \right)_{\text{ADC}} * s + 1 \\
CC &= \left( -\frac{\partial V R}{\partial W V} \right)_{\text{ADC}} * s + 1 \\
&= \left( -\frac{\partial F}{\partial W V} \right)_{\text{ADC}} * s + 1 \\
&= \left( -\frac{\partial \rho T R}{\partial W V} \right)_{\text{ADC}} * s + 1 \\
&= \left( -\frac{\partial F}{\partial W V} \right)_{\text{ADC}} * s + 1 \\
\end{align*}
\]

Since the column top tray does not have the downcomer, the tray hydraulics is different from the other generalized trays. The Laplace domain equations which describe the top tray are:

(c) aerated liquid holdup and gradient on tray. (H.18).

(d) inlet liquid height over weir. (H.19).
FIG 59. Signal Diagram For Distillation Tower Top Tray Material Balance Dynamics.
(e) volume of liquid on the active tray. (H.20).
(f) active tray total material balance. (H.22).
(g) aerated liquid density as a function of vapor density. (H.23).
(h) liquid overflow from tray. (H.25).

A signal diagram, FIG 59, can be constructed from these equations. By successive reduction, this signal diagram can be reduced until the following equation is derived.

\[
W(s) = \frac{1}{2} \left( \begin{array}{c}
\frac{1}{\beta H} \\
\rho T R
\end{array} \right) \left( \begin{array}{c}
\frac{1}{\beta H} \\
\rho T R *s
\end{array} \right)
\]

\[
W(s) = \frac{1}{2} \left( \begin{array}{c}
\frac{1}{\beta H} \\
\rho T R *s
\end{array} \right) + \frac{1}{2} \left( \begin{array}{c}
\frac{1}{\beta H} \\
\rho T R *s
\end{array} \right)
\]

\[
W(s) * \left( \begin{array}{c}
\frac{1}{\beta H} \\
\rho T R *s
\end{array} \right) + \frac{1}{2} \left( \begin{array}{c}
\frac{1}{\beta H} \\
\rho T R *s
\end{array} \right) + \frac{1}{2} \left( \begin{array}{c}
\frac{1}{\beta H} \\
\rho T R *s
\end{array} \right)
\]
\[
\left\{\begin{array}{c}
\bar{W} \\
1 + \frac{1}{\frac{\rho_{TR} * s}{ATR}} \left(1.5 \frac{1}{HOW}\right) \left(1.5 \frac{1}{HOW}\right) \\
1 + \frac{1}{\frac{\rho_{TR} * s}{ATR}} \left(1.5 \frac{1}{HOW}\right)
\end{array}\right. \\
\right\} \\
(\text{H.37})
\]

Let:

\[
\frac{1}{\frac{\rho_{TR} * s}{ATR}} \left(1.5 \frac{1}{HOW}\right)
\]

\[
AA2 = \frac{1}{\rho_{TR} * s} + \frac{1}{\frac{\rho_{TR} * s}{ATR}} \left(1.5 \frac{1}{HOW}\right)
\]

\[
BB2 = \left\{\begin{array}{c}
\frac{\partial F}{\partial \rho_{TR}} \\
\frac{\partial F}{\partial \rho_{TR}} \\
\frac{\partial F}{\partial \rho_{TR}} \\
\frac{\partial F}{\partial \rho_{TR}}
\end{array}\right. \\
\frac{\partial F}{\partial \rho_{TR}} \frac{\partial F}{\partial \rho_{TR}} \\
\frac{\partial F}{\partial \rho_{TR}} \frac{\partial F}{\partial \rho_{TR}} \\
\frac{\partial F}{\partial \rho_{TR}} \frac{\partial F}{\partial \rho_{TR}}
\]

\[
\frac{\bar{W}}{\rho_{TR} * s} + \frac{1}{\frac{\rho_{TR} * s}{ATR}} \left(1.5 \frac{1}{HOW}\right)
\]

\[
(\text{H.38})
\]

If the mass flow rate of vapor or liquid is related to the molar flow rate of vapor or liquid through the mixture's molecular weight. Then, equation (\text{H.37}) becomes,
\[
L(s)\left(\Sigma AA2*MW \ *X\right) + \Sigma AA2*MW \ *L \ *X\ (s) - L(s)\left(\Sigma MW \ *X\right)
\]

\[
+ V(s)\left(\Sigma BB2*MW \ *Y\right)
\]

\[
+ T(s)\left(\Sigma (BB2*MW \ *V - AA2*MW \ *SSV) \ *X \ *C1\right)
\]

\[
+ P(s)\left(\Sigma (BB2*MW \ *V - AA2*MW \ *SSV) \ *X \ *C0\right)
\]

\[
= FL(s)\left(- \Sigma AA2*MW \ *ZL\right) - \Sigma AA2*MW \ *FL \ *ZL(s)
\]

\[
+ FV(s)*(- \Sigma AA2*MW \ *ZV) - \Sigma AA2*MW \ *FV \ *ZV(s)
\]

\[
+ SSL(s)\left(\Sigma AA2*MW \ *X\right) + SSV(s)\left(\Sigma AA2*MW \ *Y\right)
\]

\[
\Delta PTR = P(s) - P(s) \quad n-1 \quad n \quad n-1
\]

(9) pressure drop through the tray.
\[
\frac{\partial \Delta \text{PTR}}{\partial n} \quad \text{but:}
\]

\[
W(s) = L(s) * (\sum \mathbf{W} * \mathbf{X}(s)) + \bar{L} * (\sum \mathbf{W} * \mathbf{X}(s))
\]

\[
WV(s) = V(s) * (\sum \mathbf{W} * \mathbf{Y}(s)) + \bar{V} * (\sum \mathbf{W} * \mathbf{Y}(s))
\]

\[
Y(s) = \bar{K} * X(s) * X * \mathbf{C1} * T(s) + \bar{X} * \mathbf{C8} * P(s)
\]

Combine equation (H.41) with (H.40) and re-arrange.

\[
\frac{\partial \Delta \text{PTR}}{\partial n} \quad \text{but:}
\]

\[
W(s) = L(s) * (\sum \mathbf{W} * \mathbf{X}(s)) + \bar{L} * (\sum \mathbf{W} * \mathbf{X}(s))
\]

\[
WV(s) = V(s) * (\sum \mathbf{W} * \mathbf{Y}(s)) + \bar{V} * (\sum \mathbf{W} * \mathbf{Y}(s))
\]

\[
Y(s) = \bar{K} * X(s) * X * \mathbf{C1} * T(s) + \bar{X} * \mathbf{C8} * P(s)
\]
\[ \frac{\partial \Delta PTR}{\partial n} = \frac{P(s)}{WV} \left( \sum_{i=1}^{n} \left( \frac{X_i}{C_0 i} \right) - 1 \right) \]

\[ \frac{\partial \Delta PTR}{\partial n} = \frac{1}{WV} \left( \sum_{i=1}^{n} \frac{X_i}{C_0 i} \right) = 0 \]  
(\text{H.43})

where:

1. For tray valve unit part opened.

\[ \frac{\partial \Delta PTR}{\partial n} = \frac{g}{WV} \left( \frac{V}{C_0} \rho V \right)^{2/3} \]  
\[ \text{(H.44)} \]

2. For tray valve unit fully opened.

\[ \frac{\partial \Delta PTR}{\partial n} = \frac{g}{WV} \left( \frac{1}{C_0} \rho V \right)^{2/3} \]  
\[ \text{(H.45)} \]
APPENDIX I

REBOILER LAPLACE DOMAIN RELATIONSHIPS

(A) reboiler process side.

(1) total material balance.

After the time domain total material balance (4.51) is

Laplace transformed,

\[
S \ast M (s) = L_j (s) - V (s) - L_j (s) \tag{I.1}
\]

then, combine with the controller equation by eliminating

\[
L_j (s), \quad j
\]

\[
L_j (s) = KR \ast H_j (s) \tag{I.2}
\]

and column base accumulation by eliminating \( H_j (s) \).

\[
M_j (s) = A_j \ast \rho \ast H_j (s) \tag{I.3}
\]

The process side total material balance becomes:

\[
S \ast A \ast \rho \ast \frac{1}{j} - L_j (s) + X_i (s) = 0 \tag{I.4}
\]

(2) component material balances.

After the time domain component material balance (4.53) is

linearlized and Laplace transformed, substitute \( S \ast M_j (s) \)

by (I.1) and \( Y_i (s) \) by.

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\[ Y_{ij}(s) = \overline{K}_{ij} \cdot X_{ij}(s) + \overline{X}_{ij} \cdot C1 \cdot T(s) + \overline{X}_{ij} \cdot C8 \cdot P(s) \]

Then, the Laplace domain component balance becomes,

\[ L_{ij}(s) (X_{ij} - X_{i,j-1}) - X_{ij}(s) (L_{ij} - V_{ij} + V_{i,j-1}) + T(s) (V_{i,j} \cdot X_{ij} + P(s) (V_{i,j} \cdot X_{ij} + C8) \]

\[ + X_{ij}(s) (\overline{H} s + \overline{L} + \overline{V} \cdot \overline{K}) = 0 \quad (I.6) \]

(3) Enthalpy balance.

After the time domain process enthalpy balance (4.54) is linearized and Laplace transformed, substitute \( s \cdot H(s) \) by \( (I.1) \) and also \( HL(s) \), \( HL_{ij}(s) \) and \( HV(s) \) by:

\[ HL_{ij}(s) = C6 \cdot T(s) + \sum C7 \cdot X_{ij}(s) \]

\[ HL_{ij}(s) = C6 \cdot T_{ij-1}(s) + \sum C7 \cdot X_{ij-1}(s) \]

\[ HV(s) = (\sum C2 \cdot C1 + C4) \cdot T(s) + \sum C3 \cdot X_{ij}(s) \]

\[ + \sum C2 \cdot C8 \cdot P(s) \]

Then, the Laplace domain enthalpy balance becomes:

\[ L_{ij}(s) (HL_{ij} - HL_{i,j-1}) - T_{ij}(s) (L_{ij} - C6) \]
- \( \bar{L} \left[ \overline{2C7} \right. \left. *X \right] \left( s \right) \left. \right) + V \left( s \right) \left( H \overline{V} - \overline{H}L \right) \)
+ \( T \left( s \right) \left( U * A + \left( \overline{H} * S + \overline{L} \right) *C6 + \overline{V} \left( \overline{2C2} *C1 + \overline{C4} \right) \right) \)
+ \( P \left( s \right) \left( \overline{V} * \left( \Sigma C2 *C8 \right) \right) - T \left( s \right) \left( U \right) \left( A \right) \)
+ \( \Sigma \left( \overline{V} * \overline{C3} + \left( \overline{H} * S + \overline{L} \right) *C7 \right) *X \left( s \right) = 0 \) (I.8)

(4) summation equations.

\( \Sigma (X \left( s \right) - T \left( s \right) * (X \overline{C1}) \) \left. \right) + P \left( s \right) * (X \overline{C8}) = 0 \) (I.9)

\( \Sigma X \left( s \right) = 0 \) (I.10)

(5) reboiler steam side.

In Laplace domain, equation (4.63), (4.64) and (4.65) becomes,

\( V *s* \rho \left( s \right) = \left( \begin{array}{cc} F \left( s \right) \left( S \right) & -F \left( s \right) \left( C \right) \end{array} \right) \) (I.11)

\[ \begin{array}{ccc} \bar{P} \left( s \right) & \bar{P} \left( s \right) & \bar{P} \left( s \right) \\ \bar{S} & \bar{S} & \bar{S} \\ \bar{S} & \bar{S} & \bar{S} \end{array} \]

\[ \begin{array}{ccc} \bar{F} \left( s \right) = \left( \begin{array}{ccc} \bar{S} & \bar{S} & \bar{S} \\ \bar{S} & \bar{S} & \bar{S} \\ \bar{S} & \bar{S} & \bar{S} \end{array} \right) \left( \begin{array}{ccc} \bar{S} & \bar{S} & \bar{S} \\ \bar{S} & \bar{S} & \bar{S} \\ \bar{S} & \bar{S} & \bar{S} \end{array} \right) \left( \begin{array}{ccc} \bar{S} & \bar{S} & \bar{S} \\ \bar{S} & \bar{S} & \bar{S} \\ \bar{S} & \bar{S} & \bar{S} \end{array} \right) \]
Where:

\[
\frac{\partial F}{\partial S} = -0.0088 \ast CV \ast \overline{CM} \ast (\overline{P} - \overline{P}) \ast \rho
\]

\[
\frac{\partial F}{\partial \rho} = 0.0176 \ast CV \ast \overline{CM} \ast (\overline{P} - \overline{P}) \ast \rho
\]

\[
\frac{\partial F}{\partial \lambda} = 0.0088 \ast CV \ast \overline{CM} \ast (\overline{P} - \overline{P}) \ast \rho
\]

\[
\frac{\partial F}{\partial U} = -U \ast A
\]
\[ \frac{\partial P}{\partial \lambda} = U \cdot A \cdot \frac{(\bar{T} - \bar{T})}{S} \cdot \left( \frac{1}{MR} \right) \cdot \frac{1}{\lambda^2} \]

But, in Laplace domain,

\[ \frac{dP}{s} = \frac{P(s)}{S} \cdot \frac{dP}{dT} \cdot \frac{1}{S} \]

\[ P(s) = \frac{1}{s} \cdot T(s) \]

\[ \frac{dP}{Su} = \frac{P(s)}{SU} \cdot \frac{dP}{dT} \cdot \frac{1}{SU} \]

\[ P(s) = \frac{1}{s} \cdot T(s) \]

\[ \frac{d\rho}{SU} = \frac{\rho(s)}{SU} \cdot \frac{d\rho}{dT} \cdot \frac{1}{SU} \]

\[ \rho(s) = \frac{1}{s} \cdot T(s) \]

\[ \frac{d\rho}{SC} = \frac{\rho(s)}{SC} \cdot \frac{d\rho}{dT} \cdot \frac{1}{S} \]

\[ \rho(s) = \frac{1}{s} \cdot T(s) \]

where:

\[ \frac{dP}{S} = 0.028981108 + (1.55062675E-10) \frac{\bar{T}^4}{S} \frac{dT}{S} \]

\[ - (1.15515528E-13) \frac{\bar{T}^5}{S} \frac{d\rho}{SU} \]

\[ \frac{dP}{SU} = 0.028981108 + (1.55062675E-10) \frac{\bar{T}^4}{SU} \frac{dT}{SU} \]

\[ - (1.15515528E-13) \frac{\bar{T}^5}{SU} \frac{d\rho}{SU} \]
\[
\begin{align*}
\frac{d\rho}{dT}^{\text{SC}} &= (0.68170932\times10^{-10})T^3 - (1.64926055\times10^{-14})T^4 \\
\quad &- (1.95524436\times10^{-16})T^5 \\
\frac{d\rho}{dT}^{\text{SU}} &= (0.68170932\times10^{-10})T^3 - (1.64926055\times10^{-14})T^4 \\
\quad &+ (1.95524436\times10^{-16})T^5
\end{align*}
\] (I.26, I.27)

Combine equation (I.20), (I.21), (I.22) with equation (I.12).

\[
F'(s) = T(s) \cdot \left( \frac{\partial F}{\partial \rho} \frac{d\rho}{dT}^{\text{SC}} + \frac{\partial F}{\partial \rho} \frac{d\rho}{dT}^{\text{SU}} \right) + CM(s) \cdot \left( \frac{\partial F}{\partial \rho} \frac{d\rho}{dT}^{\text{SU}} \right)
\] (I.28)

Also:

\[
\lambda'(s) = \left( \frac{\partial \lambda}{\partial \rho} \right) \frac{d\rho}{dT}^{\text{SC}} + T(s) \frac{\partial \lambda}{\partial \rho} \frac{d\rho}{dT}^{\text{SU}}
\] (I.29)

where:
\[ \frac{d\lambda}{dT} = -0.59359378 + (0.2465785E-4)T - (0.5864035E-8)T^3 \]

(I.30)

Combine equation (I.29) with equation (I.13), then,

\[ P(S) = T(S) \left( \frac{\partial F}{C} + \frac{\partial F}{C} \frac{\partial \lambda}{V} \right) + \left( \frac{\partial F}{S} \right) \frac{dT}{T} \]

(I.31)

Substitute equation (I.28) and (I.31) into (I.11) and solve for \( T(S) \).

\[ T(S) = \frac{1}{CTS} \left( \frac{\partial F}{S} \frac{dP}{SU} + \frac{\partial F}{S} \frac{d\rho}{SU} \right) \]

\[ - \left( \frac{1}{CTS} \frac{\partial F}{\partial T} \right) T(S) + \left( \frac{1}{CTS} \frac{\partial F}{\partial CM} \right) CM(S) \]

where:

\[ CTS = \]

\[ \frac{d\rho}{SC} \frac{dF}{S} \frac{dP}{S} \frac{d\rho}{C} \frac{dF}{C} \frac{d\lambda}{V} \]

(I.33)
(5) enthalpy balance for the reboiler metal.

In Laplace domain, equation (4.62) is,

\[ s^*M^*C^*T^*(s) = U^*A^*T^*(s) - U^*A^*T^*(s) - U^*A^*T^*(s) + U^*A^*T^*(s) \]  

\[ B^*B^*M^*R^*B^*B^*j \]  

(1.34)

If substitute \( T(s) \) in the above equation by (1.32) and solve for \( T(s) \); then,

\[ T(s) = \]

\[ \frac{U^*A}{s} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \]

\[ \frac{1}{s} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \]

\[ \frac{CMR}{CTS} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \]

\[ \frac{1}{s} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \]

\[ \frac{CMR}{CTS} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \]

\[ \frac{1}{s} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \]

\[ \frac{CMR}{CTS} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \frac{\delta F}{\delta T} \]

where:


\[ \frac{M^*R^*S^*B^*B^*C^*S^*T^*J^*}{M^*R^*S^*B^*B^*C^*S^*T^*J^*} \]

(1.35)

Eliminate \( T(s) \) from (1.8) by using (I.36).

\[ L_{MR} \]

\[ \frac{(\overline{HL} - \overline{HL}) - T_{MR}(\overline{L} \overline{C^6})}{j-1 j-1 j-1 j-1 j-1} \]

\[ \frac{-L_{MR}(\overline{\Sigma C^7} + \overline{\Sigma C^7})}{j-1 i i, j-1 i, j-1 j j} \]

\[ + V_{MR}(\overline{HV} - \overline{HL}) \]

\[ j-1 j j j j \]
\[ + T(s) \left[ V \left( \Sigma C_2 \right) + C_1 + C_4 \right] + U \cdot A + L \cdot C_6 \]

\[ + \left( M \cdot C_6 \cdot s - \text{-------------} \right) \]

\[ + P(s) \left( V \left( \Sigma C_2 \right) + C_8 \right) \]

\[ + \Sigma \left[ V \left( C_3 \right) + L \cdot C_7 \right] + s \cdot (M \cdot C_7 \cdot s) \cdot X(s) \]

\[ = CM(s) \left[ \left( \text{-------------} \right) \left( \text{-------------} \right) \left( \text{-------------} \right) \right] \]

\[ \Delta P_{TR} (s) = P(s) - P(s) \]

\[ \left( \text{-------------} \right) \left( \text{-------------} \right) \left( \text{-------------} \right) \left( \text{-------------} \right) \]

\[ \left( \text{-------------} \right) \left( \text{-------------} \right) \left( \text{-------------} \right) \left( \text{-------------} \right) \]
but:

\[
W(S) = L(S) * (\Sigma_{i,j}X(s)) + L(S) * (\Sigma_{i,j}X(s))
\]

\[
WV(S) = V(S) * (\Sigma_{i,j}Y(s)) + V(S) * (\Sigma_{i,j}Y(s))
\]

\[
Y(s) = K * X(s) + X(s) * C^T * T(s) + X(s) * C_8 * P(s)
\]

Combine equation (I.39) with equation (I.38) and re-arrange.

\[
\frac{\partial \Delta P_{TR}}{\partial t} = L(S) * \left[ \left( \frac{-\Delta P_{TR}}{\partial w} \right) * (\Sigma X) * M(W) \right] + P(S)
\]

\[
+ \frac{\partial \Delta P_{TR}}{\partial t} = V(S) * \left[ \left( \frac{-\Delta P_{TR}}{\partial w} \right) * (\Sigma Y) * M(W) \right]
\]

\[
+ T(S) * \left[ \left( \frac{-\Delta P_{TR}}{\partial w} \right) * \Sigma \bar{V} * (\Sigma M(W) * X) \right]
\]
\[
\begin{align*}
\Delta \text{PTR} & = \frac{1}{j-1} + P(s) \left( \begin{array}{c}
\frac{V}{j} \\
\omega
\end{array} \right) \left( \sum_{\mu} \sum_{\nu} \sum_{\rho} E^{\mu \nu \rho} \right) - 1 \\
& \quad + \left( \begin{array}{c}
\frac{V}{j} \\
\omega
\end{array} \right) \left( \sum_{\mu} \sum_{\nu} \sum_{\rho} E^{\mu \nu \rho} \right) = 0 \quad (I.40)
\end{align*}
\]

where:

(1) for tray valve unit part opened.

\[
\begin{align*}
\Delta \text{PTR} & = \frac{1}{j-1} g_j \left( \begin{array}{c}
\frac{V}{j} \\
\omega
\end{array} \right) \left( \sum_{\mu} \sum_{\nu} \sum_{\rho} E^{\mu \nu \rho} \right)  \\
& \quad = 0.00023148 \left( \begin{array}{c}
\frac{V}{j} \\
\omega
\end{array} \right) \left( \sum_{\mu} \sum_{\nu} \sum_{\rho} E^{\mu \nu \rho} \right)  \quad (I.41)
\end{align*}
\]

(2) for tray valve unit fully opened.

\[
\begin{align*}
\Delta \text{PTR} & = \frac{1}{j-1} g_j \left( \begin{array}{c}
\frac{V}{j} \\
\omega
\end{array} \right) \left( \sum_{\mu} \sum_{\nu} \sum_{\rho} E^{\mu \nu \rho} \right)  \\
& \quad = 0.001215 \left( \begin{array}{c}
\frac{V}{j} \\
\omega
\end{array} \right) \left( \sum_{\mu} \sum_{\nu} \sum_{\rho} E^{\mu \nu \rho} \right)  \quad (I.42)
\end{align*}
\]
APPENDIX J

CONDENSER LAPLACE DOMAIN RELATIONSHIPS

(1) total material balance.

\[ V (s) = L (s) + V (s) \]  \hspace{1cm} (J.1)

(2) component material balances.

\[ \bar{V} * Y_i (s) + \bar{Y} * V (s) = \bar{L} * X_i (s) + \bar{X} * L (s) \]

\[ + \bar{V} * Y_i (s) + \bar{Y} * V (s) \]  \hspace{1cm} (J.2)

(3) enthalpy balance.

\[ \bar{H}V * V (s) + \bar{V} * H V (s) = \bar{L} * H L (s) + \bar{H}L * L (s) + \bar{V} * H V (s) \]

\[ + \bar{H}V * V (s) + U * A * (T (s) - T (s)) \]  \hspace{1cm} (J.3)

(4) equilibrium relationships.

\[ Y (s) = \bar{K} * X_i (s) + \bar{X} * C1 * T (s) + \bar{X} * C8 * P (s) \]

\[ C_i \quad C_i \quad C_i \quad C_i \quad C_i \quad BC \quad C_i \quad C_i \quad i \]  \hspace{1cm} (J.4)

(5) summation equations.

\[ \Sigma X_i (s) = 0 \]  \hspace{1cm} (J.5)

\[ \Sigma \bar{K} * X_i (s) + T (s) * (\Sigma \bar{X} * C1) + \]

\[ i \quad C_i \quad C_i \quad BC \quad i \quad C_i \quad C_i \]
(6) pressure relationships.

\[ P(s) = P(s) - \Delta P(s) \]

\[ \Delta P(s) = \frac{2 \cdot CP \cdot \left( \sum_{i=1}^{n} W_i \cdot Y_{i,2} \right)}{s \cdot V_{i,2}} \]

(7) process side temperature approximation.

\[ T(s) = T(s) \]

(8) process thermal properties.

\[ H_L(s) = \sum_{i=1}^{n} C_6 \cdot T_{i,2} + \sum_{i=1}^{n} C_7 \cdot X_{i,1} \]

\[ H_V(s) = \sum_{i=1}^{n} (C_2 \cdot C_1 + C_4) \cdot T_{i,2} + \sum_{i=1}^{n} C_3 \cdot X_{i,1} \]

\[ + \sum_{i=1}^{n} (C_2 \cdot C_8) \cdot P(s) \]

(8) enthalpy balance for the condenser metal.

If the time domain metal enthalpy balance is linearized,

Laplace transformed and solve for \( T(s) \); then,

\[ T(s) = C_{TC} \cdot T(s) + C_{TW} \cdot T_W(s) \]

where:

\[ C_{TC} = \frac{U \cdot A}{C \cdot C} \]

\[ C_{TW} = \frac{M \cdot C \cdot S + U \cdot A + U \cdot A}{MC \cdot MC \cdot C \cdot C \cdot W \cdot W} \]
(C) Condenser coolant side enthalpy balance.

If the time domain metal enthalpy balance is linearized, Laplace transformed and solve for $T(s)$; then,

$$T(s) = CTW_0 \ast TW(s) + CTFW \ast F(s) + CTMC \ast T(s)$$

where:

$$CTW_0 = \frac{\rho \ast F \ast C}{W \ast W \ast W}$$

$$CTMC = \frac{U \ast A}{W \ast W}$$

$$CTFW = \frac{\rho \ast C \ast (TW - \overline{T})}{W \ast W \ast W \ast W \ast W \ast W}$$

If substitute the $T(s)$ in (J.15) by (J.12), then,

$$T(s) = CTW_1 \ast TW(s) + CTW_4 \ast F(s) + CTW_2 \ast T(s)$$

where:

$$(J.19)$$
(D) reflux drum equations.

(a) vapor phase.

(1) total material balance.

If both equation (4.89) and equation (4.90) are linearized and Laplace transformed, these two equations can combine with each other by eliminating the $M(s)$.

$$V_D V^{p_D} D 1 \left[ \frac{s*}{s*} * P(s) - \frac{s*}{s*} * T(s) \right] = V(s) + S SV(s)$$

$$= 0 \quad (J. 21)$$

(2) component material balances.

$$Y_{i,1}^C = Y_{i,1}^C \quad (J. 22)$$

(3) enthalpy balance.

$$T(s) = T(s) \quad (J. 23)$$

(4) thermal property.

$$HV(s) = HV(s) \quad (J. 24)$$
(b) liquid phase.

(1) total material balance.
\[ s * M (s) = L (s) - L (s) - SSL (s) \]  \hspace{1cm} (J.25)

(2) component material balances.
\[ (\bar{\bar{M}} * s + \bar{L} + \overline{SSL}) * X (s) - L * X (s) = 0 \]  \hspace{1cm} (J.26)

(3) enthalpy balance.
\[ (\bar{\bar{M}} * s + \bar{L} + \overline{SSL}) * HL (s) - \bar{L} * HL (s) = 0 \]  \hspace{1cm} (J.27)

(4) level controller equations.
(case 1) \[ SSL (s) = KCL * H (s) \]  \hspace{1cm} (J.28)
(case 2) \[ L (s) = KCL * H (s) \]  \hspace{1cm} (J.29)

(5) thermal property.
\[ HL (s) = C6 * T (s) + \sum C7 * X \]  \hspace{1cm} (J.30)

The above Laplace domain condenser equations set can be further simplified by eliminating \[ X (s), HL (s), HV (s), \]
\[ T (s), P (s), \Delta P, L (s), V (s), M (s), HL (s), HV (s), \]
\[ T (s), Y (s) \] and \[ Y (s). \]

Then the resulting equations are listed below:

(1)
\[ \sum X (s) = 0 \]  \hspace{1cm} (J.31)
(2) For $i = 1$ to $NC-1$.

\[
\begin{align*}
T \ (s) & \equiv (-\overline{V - L}) \overline{X} \overline{C} \overline{i} \ - \ s \cdot \left( -\frac{-\overline{L}}{s} + \frac{-\overline{L}}{s} \right) \\
& \text{BC} \end{align*}
\]

\[
\begin{array}{l}
\left\{ \begin{array}{l}
\overline{M} \cdot s \\
- X \ (s) \left[ \left( \frac{-\overline{L}}{s} + 1 \right) \cdot (\overline{L} + \frac{\overline{L}}{s}) \right] \\
i,1 \ & \ \left[ \begin{array}{l}
\overline{C} \\
\overline{U} \\
\overline{C} \\
\end{array} \right] \\
\end{array} \right. \\
\end{array}
\]

\[
\begin{align*}
& + V \ (s) \left[ \overline{Y} - \overline{Y} + (-\frac{\overline{C}}{s}) \left( \overline{X} \overline{C} \overline{i} \right) \left( \overline{V} - \overline{L} \right) \right] \\
& \text{BC} \\
& + \frac{V \cdot s}{D} \frac{2 \cdot CP}{2 \cdot CP} \\
& \text{BC}
\end{align*}
\]

\[
\begin{align*}
& + T \ (s) \left[ \left( \overline{V} \overline{X} \overline{C} \overline{i} \right) \cdot \left( \overline{V} \overline{L} \right) \right] \\
& \text{BC} \\
& + \frac{V \cdot s}{D} \frac{2 \cdot CP}{2 \cdot CP} \\
& \text{BC}
\end{align*}
\]

\[
\begin{align*}
& - \overline{C} \overline{A} \cdot \left( \frac{-\overline{L}}{s} + \frac{-\overline{L}}{s} \right) \cdot R^* \overline{X} \overline{C} \overline{i} \\
& \text{BC} \\
& + P \ (s) \left[ \left( \overline{V} \overline{X} \overline{C} \overline{i} \right) - \left( \overline{V} \overline{L} \right) \cdot \left( \overline{X} \overline{C} \overline{i} \right) \right] \\
& \text{BC}
\end{align*}
\]

\[
\begin{align*}
& - \left( \frac{-\overline{L}}{s} + \frac{-\overline{L}}{s} \right) \cdot \left( \overline{X} \overline{C} \overline{i} \right) \\
& \text{BC}
\end{align*}
\]
\[ V \ast s \]
\[ m=m,2, \; 2 \leq c \; c,i \; c,i \; i \; \ast R \ast T \; BC \]

\[ *BBB \]

\[ V \ast s \]
\[ + \; x \; (s) \ast \left\{ \frac{V \ast K}{i,2} + \left( \frac{V - L}{2} \ast (\frac{X \ast C8}{2} - CCA \ast \ldots) \right) \right\} \; c,i \; c,i \; i \; \ast R \ast T \; BC \]

\[ *BBB \ast m \ast k \]
\[ i \; i,2 \]

\[ = S S V \; (s) \ast (-CCA) \]  
\[ 1 \; i \]  

(3)
\[ \frac{M \ast s}{1} \]
\[ T \; (s) \ast (\Sigma X \ast \Sigma C1) + \Sigma X \ast \ldots + 1 \ast X \; (s) \]  
\[ BC \; i \; c,i \; c,i \; i \; c,i \; \frac{L}{c} \; i,1 \]

\[ + \; V \; (s) \ast \left\{ \frac{2 \ast CP}{i} \right\} \]
\[ + \; v \; (s) \ast \left\{ \frac{V \ast (X \ast C8)}{2} \right\} \]
\[ + \; T \; (s) \ast \left\{ \frac{2 \ast C1}{i} \right\} \]
\[ + \; P \; (s) \ast \left\{ \frac{2 \ast C8}{i} \right\} \]

\[ + \; \Sigma (X \ast \ast \ast) \ast \frac{\Sigma (X \ast C8)}{2} \ast (\Sigma (X \ast C8)) \]

\[ - \; (\Sigma (X \ast \ast \ast)) \ast \Sigma (X \ast \ast \ast) \]
\[ 1 \; i \; i \; i,2 \; i,2 \]  

(J.33)
(4)

\[ T(s) \widetilde{M} + T(s)\left(\overline{C}_6 \widetilde{M} \overline{L} \overline{C}_7 \right) + \overline{X}(s) = 0 \]

(5)

\[ T(s)\left( - \overline{L} \overline{C}_6 - (\overline{V} - \overline{L}) \{ \overline{C}_2 \overline{C}_1 + \overline{C}_4 \} \right) - U * A \]

\[ + U * A * T C + U * A * C T W * C T W^2 + C C B * ( - - - - - - - - ) \]

\[ + V(s) \overline{C}_2 \overline{C}_8 \{ \overline{C}_2 \overline{C}_8 \} * ( - - - - - - - - ) \]

\[ + \overline{C} B \overline{C} B \{ \overline{C} B \overline{C} B \} \]
(case 1) the reflux drum level controller controls the distillate flow rate.
(case 2) the reflux drum level controller controls the reflux flow rate.

\[
\begin{align*}
    T(s) &+ \frac{V_s}{D} \\
    \text{BC} &- \frac{(-\cdots\cdots) \cdot \overline{BBB} \cdot (\Sigma MW \cdot \theta_X \cdot \theta_C)}{2} \\
    \text{BC} &+ \frac{P(s) \cdot \left(\frac{1}{\theta_B - \overline{BBB} \cdot (\Sigma MW \cdot \theta_X \cdot \theta_C)}\right)}{2}
\end{align*}
\]
\[ V \ast s \quad D \quad A \ast \rho \ast s \quad \frac{1 \ast 1}{1 \ast 1} \]

\[ - \left( \sum_{i} R \ast T_{i} \ast i, i, 2 \right) \ast \left( \sum_{i, 2} X \ast (s) \right) + \left( L \ast (s) \right) \ast (1 + \frac{1}{KCL}) \]

\[ = - SSL (s) - SSV (s) \quad \frac{1}{1} \quad \text{(J. 37)} \]

Where:

\[ \overline{BBB} = \frac{2 \ast CP}{\sum_{l, l, l, 2} \overline{MW} \ast \overline{Y}} \quad \text{(J. 38)} \]

\[ \overline{AAA} = \sum_{m, C_{m}, C_{m}} \overline{C_{2}} \ast \overline{C_{8}} \quad \text{(J. 39)} \]

\[ \overline{CCA} = \overline{X} - \overline{Y} \quad \text{(J. 40)} \]

\[ \overline{CCB} = \overline{HV} - \overline{HL} \quad \text{(J. 41)} \]
APPENDIX K

STEADY STATE DISTILLATION COLUMN DESIGN MANUAL

This steady state distillation column program was first set up by Kurt Kominek on August 30, 1979. Then it is re-modified by Chien Wang at Jan. 22, 1980.

This program performs rating calculations for a complex multi-component distillation column, multiple feed, products, inter-stage heaters or coolers. The mixture to be separated can have any degree of liquid phase non-ideality. The vapor phase is assumed to be ideal. Since enthalpy data are required, it is not necessary to assume an equimolal overflow model. Three types of condensers are possible, total, partial or none. The column must have a partial reboiler. The number of equilibrium stages, the column pressure profile, the feed rate, and the feed composition must be given. In addition the rate of each product but one and reflux ratio must be specified.

The basic solution method is based on the Wang-Henke method (Hydrocarbon Processing, VOL. 45, NO 8, August 1966). This method requires an initial column profile of temperature, liquid and vapor rates, and liquid and vapor compositions. The convergence of this iterative procedure
depends on how good the initial profiles are. Convergence is aided by the Holland's Theta method and a partial substitution damping factor. If the system is a new process, equimolar overflow model should be run first in order to get an ideal of the column profile of the enthalpy model.

The input data must be prepared as follows:

(A) DATA RECORD A:
   (a) One record.
   (b) FORMAT: I2,19A4,A2
   (c) This is a header record for the physical property data set. The first two columns contain the number of components (right justified). The next 78 columns will be printed at the top of the first page. This record should identify the components as well as indicate any special attributes of the components.

(B) DATA RECORD SET B:
   (a) physical property data record set, three records for each component.
   (b) fields for the first data record:
       FORMAT: 2A4,2X,7E10.0
       (1) name of the component.
       (2) molecular weight.
(3)-(8) Antoine coefficients, ANTA, ANTB, ANTC,
for calculating vapor pressure with vapor
pressure in MMHG and temperature T in F.

\[ \ln(\text{vapor pressure}) = \text{ANTA} - \text{ANTB}/(T+\text{ANTC}) \]

(c) fields for the second data record:

**FORMAT**: 8E10.0

(1) component critical temperature, K.
(2) component normal boiling temperature, K.
(3) component heat of vaporization at the normal
boiling temperature, CAL/GMOLE.
(4)-(8) coefficients C1, C2, C3, C4, C5 for
calculating vapor enthalpy HV, with temperature
T in F and HV in BTU/LBMOLE.

\[ HV = C1 + (C2\times T) + (C3\times T^2) + (C4\times T^3) + (C5\times T^4) \]

(d) fields for the third data record:

**FORMAT**: 8E10.0

(1)-(6) coefficients B1, B2, B3, B4, B5, B6 for
calculating liquid enthalpy HL, with temperature
T in F and HL in BTU/LBMOLE.

\[ HL = B1 + (B2\times T) + (B3\times T^2) + (B4\times T^3) + (B5\times T^4) + (B6\times T^5) \]

(C) DATA RECORD SET C

(a) initial mole fraction estimation for the top and
bottom liquid compositions, two data records per
set.
(b) fields for the first data record

FORMAT: 8E10.0

(1)-(8) initial estimate for the distillate compositions.

(c) fields for the second data record

FORMAT: 8E10.0

(1)-(8) initial estimate for the residue compositions.

(D) DATA RECORD D:

(a) one record

(b) FORMAT: 20A4

(c) This is a header record. All 80 characters will be printed at the top of each new page.

(E) DATA RECORD E:

(a) control data record.

(b) fields for record E:

FORMAT: 8I5

(1) number of stages.

(2) number of components.

(3) maximum number of iterations.

(4) flag for equimolar overflow.

0 means equimolar overflow model is assumed.

1 means enthalpy model is assumed.

(5) flag for inter-stage heaters or coolers.

0 means none.

1 means one or more heaters or coolers.
(6) flag for print option of final solution.
    0 means print only stages with feeds, products,
    heaters or coolers.
    1 means print information for every stage.

(7) flag for plot option
    0 means no plots.
    1 means plot vapor and liquid composition (mole
     fraction) vs. stages.

(8) flag for Holland* Theta method.
    0 means use Theta method to aid the convergence.
    1 means do not use the Theta method.

(F) DATA RECORD F:

(a) data to initialize the column.

(b) fields for record F:

    FORMAT : 8E10.0

    (1) reflux ratio. (reflux/(dist liq+dist vap)
    (2) stage 1 pressure, PSI.
    (3) reboiler pressure, PSI.
    (4) stage 1 estimated temperature, F.
    (5) reboiler estimated temperature, F.
    (6) pressure drop from condenser to stage 1, PSI.

(G) DATA RECORD G:

(a) information for the column convergence.

(b) fields for record G:

    FORMAT : 8E10.0
(1) damping factor for temperature, (0.1-2.0).
(2) damping factor for liquid/vapor flow rate, (0.1-2.0).
(3) relative error tolerance for temperature.
(4) relative error tolerance for vapor flow.
(5) relative error tolerance for liquid mole fraction.
(6) minimum temperature, F. If left blank, this is the stage one temperature minus 100.
(7) maximum temperature, F. If left blank, this is the stage one temperature plus 100.

(H) DATA RECORD H:

(a) Murphee efficiency.

(b) fields for record H:

FORMAT : I1,9X,2(I3,7X),E10.0

(1) flag to indicate another stage efficiency data record.
0 means there is another record.
1 means there are no more data records.

(2) stage number for the beginning of the plate efficiency.

(3) stage number for the end of the plate efficiency.

(4) Murphee efficiency (fraction) for the above section stages.

(I) DATA RECORD SET I:

(a) feed stream data records, two records for each feed
stream.

(b) fields for the first data record:

**FORMAT**: I1,9X,I3,7X,2E10.0

(1) flag to indicate another feed stream data record.
   0 means there is another stream.
   1 means there are no more streams.

(2) stage number where feed enters.

(3) For equimolar overflow model, this is the feed liquid fraction. For enthalpy model, this is the upstream total pressure, PSI.

(4) upstream feed temperature, \( F \), if enthalpy model is used.

(b) fields for the second data record:

**FORMAT**: 8E10.0

(1)-(8) feed rate of each component, LBMOLE/TIME.

(J) DATA RECORD J:

(a) product stream data record.

(b) fields for record J:

**FORMAT**: I1,I9,2E10.0

(1) flag to indicate another product stream data record.
   0 means there is another record.
   1 means there are no more data records.

(2) stage number of product stream.

(3) liquid flow rates, (LBmole/TIME).
(4) vapor flow rates, (LBmole/TIME).

(K) DATA RECORD K:

(a) inter-stage heater or cooler data record.

(b) fields for record K:

FORMAT : I1, I9, E10.0

(1) flag to indicate another inter-stage heater or cooler data record.
   0 means there is another record.
   1 means there are no more data records.

(2) stage number of inter-stage heater or cooler.

(3) duty for the inter-stage heater or cooler, BTU/TIME.

NOTE 1:

The liquid phase activity coefficients are calculated through a user supplied subroutine GAMMA. The following subroutine is used if the user does not supply one. In this case, an ideal liquid phase is assumed.

SUBROUTINE GAMMA (T, J, NC, GAM, X, IL, IH)

C------NOMENCLATURE
C
C GAM(I) = LIQUID ACTIVITY COEFFICIENTS OF COMPONENT I
C NC = NUMBER OF PROCESS COMPONENTS IN THE
C       DISTILLATION COLUMN
C J = STAGE NUMBER.
C X(I,J) = LIQUID MOLE FRACTION OF COMPONENT I OF STAGE J.
C IL = DUMMY VARIABLE.
C IH = DUMMY VARIABLE.
NOTE 2:

The data base of the entire computer program structure is shown in FIG (1). Data base 1 is the necessary information to run this steady state program. Two sets of steady state profile, data base 2 and 3, are generated by this steady state program through program output unit FT02F001 and FT03F001. The data base 2 is used for the mechanical design program and the data base 3 is used for the constants generation program of the process frequency response analysis.
EXAMPLE 1:

This is the enthalpy model for de-isobutanizer. There are thirty three stages including the condenser and the reboiler. Total condenser is stage one and the reboiler is stage 33. Feed stage is 17 and the feed is all liquid.

<table>
<thead>
<tr>
<th>Stage</th>
<th>DEISOBUTANIZER</th>
</tr>
</thead>
<tbody>
<tr>
<td>2</td>
<td>1-BUTANE</td>
</tr>
<tr>
<td></td>
<td>58.124</td>
</tr>
<tr>
<td></td>
<td>15.5381</td>
</tr>
<tr>
<td></td>
<td>2032.73</td>
</tr>
<tr>
<td></td>
<td>-33.15</td>
</tr>
<tr>
<td>33</td>
<td>N-BUTANE</td>
</tr>
<tr>
<td></td>
<td>58.124</td>
</tr>
<tr>
<td></td>
<td>15.6782</td>
</tr>
<tr>
<td></td>
<td>2154.9</td>
</tr>
<tr>
<td></td>
<td>-34.42</td>
</tr>
</tbody>
</table>

DEISOBUTANIZER ENTHALPY MODEL, P(PSI) T(F) FLOW RATE(MOLE/TIME)

<table>
<thead>
<tr>
<th>Stage</th>
<th>P</th>
<th>T</th>
<th>Flow Rate</th>
</tr>
</thead>
<tbody>
<tr>
<td>6.0</td>
<td>180.56</td>
<td>183.5</td>
<td>160.0</td>
</tr>
<tr>
<td>1.0</td>
<td>1.0</td>
<td>0.01</td>
<td>0.01</td>
</tr>
<tr>
<td>1</td>
<td>2</td>
<td>33</td>
<td>0.8</td>
</tr>
<tr>
<td>1</td>
<td>17</td>
<td>190.0</td>
<td>112.0</td>
</tr>
<tr>
<td>300.0</td>
<td>300.0</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1</td>
<td>1300.0</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
EXAMPLE 2:

This is the enthalpy model for benzene - toluene - p-xylene system. There are seven stages including the condenser and the reboiler. Total condenser is stage one and reboiler is stage 7. Feed stage is stage 4 and the feed is a flash feed which consists of both liquid and vapor.

P.P DATA FOR BTX SYSTEM

<table>
<thead>
<tr>
<th>Component</th>
<th>P (PSIA)</th>
<th>T (°F)</th>
<th>X</th>
<th>Y</th>
</tr>
</thead>
<tbody>
<tr>
<td>Benzene</td>
<td>78.105</td>
<td>15.9008</td>
<td>2788.51</td>
<td>-52.36</td>
</tr>
<tr>
<td></td>
<td>4788.8</td>
<td>8171735.8157189</td>
<td>8.0223E-11</td>
<td></td>
</tr>
<tr>
<td>Toluene</td>
<td>92.134</td>
<td>16.0137</td>
<td>3096.52</td>
<td>-53.67</td>
</tr>
<tr>
<td></td>
<td>5309.5501740.7176138</td>
<td>0.276423E-4</td>
<td></td>
<td></td>
</tr>
<tr>
<td>P-Xylene</td>
<td>106.16</td>
<td>16.0963</td>
<td>3346.65</td>
<td>-57.84</td>
</tr>
<tr>
<td></td>
<td>11515.282467.43940480.04931496</td>
<td>0.02952951-.70052E-5.698302E-9</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>0.7924</td>
<td>0.1778</td>
<td>0.62972</td>
<td></td>
</tr>
<tr>
<td></td>
<td>0.2073</td>
<td>0.4223</td>
<td>0.3704</td>
<td></td>
</tr>
</tbody>
</table>

BTX SYSTEM TEMP (°F) : FLOW (MOLES) : PRESSURE (PSIA)

<table>
<thead>
<tr>
<th>Stage</th>
<th>Temp</th>
<th>Flow</th>
<th>Pressure</th>
</tr>
</thead>
<tbody>
<tr>
<td>7</td>
<td>50</td>
<td>15.7</td>
<td>226.0</td>
</tr>
<tr>
<td>0.2</td>
<td>50.0</td>
<td>156.218.7</td>
<td>220.0</td>
</tr>
<tr>
<td>0.5</td>
<td>2</td>
<td>6.01</td>
<td>0.8</td>
</tr>
<tr>
<td>1</td>
<td>4</td>
<td>201.0</td>
<td></td>
</tr>
<tr>
<td>501.0</td>
<td>1501.0</td>
<td>201.0</td>
<td></td>
</tr>
</tbody>
</table>
**EXAMPLE 3:**

This is the enthalpy model for benzene - toluene system. There are 21 stages including the condenser and the reboiler. Total condenser is stage one and reboiler is stage 21. Feed stage is stage 15 and the feed is all liquid.

| P.P DATA FOR BENZENE-TOLUENE SYSTEM (SEPERSION HANDBOOK 1-36) |
|-------------------|----------------------------------|---------------------------|
| BENZENE 78.105    | 15.9008 2788.51 -52.36          |
|                   | 562.10 353.30 7352.0 19572.227 |
|                   | 1643156E2 221683E-1 -60703E-5 730024E-9 |
| TOLUENE 92.134    | 16.0137 3096.52 -53.67          |
|                   | 591.7 383.8 7930.22308.21       |
|                   | 21.34517 0.0252609 -58701E-5 502829E-9 |

| BENZENE-TOLUENE SYSTEM TEMP (F) : FLOW (MOLES/HR) : PRESSURE (PSIA) |
|-------------------|-----------------------------|--------------------------|
| 21                | 2 150 1                     |
| 5.9449            | 23.0 25.0 204.0 285.0 0.5  |
| 0.3               | 0.3 0.001 0.001 0.001       |
| 1                 | 2 20 1.0                    |
| 1                 | 15 26.0 230.0               |
| 221.21            | 388.34                      |
| 1                 | 1 213.63                    |
EXAMPLE 4:

This is the enthalpy model for propane - butane - pentane system. There are eleven stages including the condenser and the reboiler. Total condenser is stage one and reboiler is stage eleven. Feed stage is stage five and the feed is a liquid feed.

<table>
<thead>
<tr>
<th>03 PROPANE-BUTANE-PENTANE SYSTEM</th>
<th>(A)</th>
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<tbody>
<tr>
<td>PROPANE 44.09 15.726 1872.46 -25.16</td>
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<tr>
<td>369.8 231.1 4487.012127.86 1533851E2.153578E-1 -32925E-5 .32913E-9</td>
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<td>4654.1085928.6057208 -0.0356083-6.7468E-5</td>
<td>(B.3)</td>
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<td>N-BUTANE 58.12 15.6782 2154.9 -34.42</td>
<td>(B.1)</td>
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<td>425.2 272.7 5352.016073.388 2077065E2.182591E-1 -26701E-5 28892E-10</td>
<td>(B.2)</td>
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<tr>
<td>6060.5720722.22105717.58299E-2 -1.6415E-7</td>
<td>(B.3)</td>
</tr>
<tr>
<td>N-PENTANE 72.15 15.8333 2477.07 -39.94</td>
<td>(B.1)</td>
</tr>
<tr>
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<td>(B.2)</td>
</tr>
<tr>
<td>6743.7925539.07053850.02837482</td>
<td>(B.3)</td>
</tr>
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<td>(C.1)</td>
</tr>
<tr>
<td>0.0051 0.3949 0.6</td>
<td>(C.2)</td>
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PBP SYSTEM

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<td>3.0 180. 181.0 87.0 250.0 0.1</td>
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<tr>
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<tr>
<td>1 2 10 1.0</td>
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<tr>
<td>1 5 180.17 119.0</td>
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<tr>
<td>240.0 150.0 210.0</td>
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<tr>
<td>1 1 250.0</td>
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APPENDIX L

MECHANICAL DESIGN MANUAL

This distillation column mechanical design program and manual were designed by Chien Wang in April 1981. This program allows the following distillation mechanical design options:

(a) reboiler
   (1) kettle reboiler, vertical thermosyphon reboiler or forced circulation reboiler.

(B) condenser
   (1) total condenser or partial condenser.
   (2) vertical condenser or horizontal condenser.

(c) horizontal reflux drum.
   (1) drum for total condenser or for partial condenser.

(d) tray column.
   (1) single tray design or tray column design.
   (2) ballast tray designed by the Glitsch company or Flexitray designed by the Koch company.

The input data must be prepared as follows:

(A) DATA RECORD A:
   (a) one record.
   (b) fields for record A.
(1) number of process components in the distillation column.

(2) number of stages in the distillation column. The first stage is the condenser and the last stage is the reboiler.

(B) DATA RECORD B:

(a) physical property data record set, two records for each process component.

REFERENCE: "Properties Of Liquid and Gases" by Reid, Prausnitz and Sherwood.

(b) fields for the first data record.

FORMAT: 2A1*,2X,7E10.0

(1) the name of the process component \( i \), maximum allowed eight characters.

(2) the Parachor of process component \( i \). The Parachor is a temperature independent parameter defined by Sugden in calculating the liquid surface tension.

(3) the liquid density reference temperature of component \( i \), \( K \).

(c) fields for the second data record.

FORMAT: 8E10.0

(1) the critical pressure of component \( i \), ATM.

(2) Pitzer's Acentric Factor of component \( i \).

(3) specific volume of component \( i \), \( \text{CM}^{**3}/\text{GMOLE} \).
(4) critical compressibility of component i.
(5) the saturated liquid density at the liquid density reference temperature of component i. gramm/cm**3.
(6) parameter that depends on the molecular structure of component i to calculate thermal conductivity. The liquid thermal conductivity is calculated by the method of Robbins and Kingrea.
(7) liquid viscosity constant, B, of component i. The liquid viscosity is calculated by the method of Van Velzen, Cardozo and Langenkamp.
(8) liquid viscosity constant, T0, of component i.

(C) DATA RECORD C :
(a) one record.
(b) fields for data record C.

FORMAT : 8E10.0

(1) maximum allowable material stress for distillation column construction, PSI.
(2) joint efficiency of the distillation tower.
(3) distillation column construction material density, lbm/in**3.
(4) dirt factor inside the reboiler tube.
(5) dirt factor outside the reboiler tube.
(6) thermal conductivity of the reboiler tube metal,
(7) thermal conductivity of the condenser tube metal, BTU/HR, FT, F.

(D) DATA RECORD D:
(a) reboiler design data record, one record.
(b) fields for data record D.

FORMAT: 4E10.0, I2, 8X, 2E10.0

(1) re-circulation ratio for either thermosyphon reboiler design case one or forced circulation reboiler. If the kettle reboiler is designed, leave this field blank.

(2) desired reboiler design heat flux, BTU/HR, FT**2.

(3) reboiler tube length, FT. For kettle reboiler, tube length is the U-tube nominal tube length.

(4) reboiler tube outside diameter, IN.

(5) reboiler tube BWG.

(6) reboiler tube sheet layout pitch, IN. (only triangular tube layout allowed).

(7) reboiler supplied steam pressure, if left blank, the steam pressure is the pressure corresponds to the reboiler process temperature plus 60 F.

(8) pressure drop through the forced circulation reboiler control valve. Leave this field blank if forced circulation reboiler design is not required.
(E) DATA RECORD E :

(a) flag data record, one record.

(b) fields for data record E.

FORMAT : 7(I1,9X)

(1) flag for the design of reflux drum.
   1 means design horizontal reflux drum without wire mesh.
   2 means design horizontal reflux drum with wire mesh.

(2) flag for the design of reflux drum.
   1 means product to storage, 2 min surge time.
   2 means product to heat exchanger, 5 min surge time.
   3 means product to heater, 10 min surge time.

(3) flag for the design of reflux drum.
   1 means design reflux drum for total condenser.
   2 means design reflux drum for partial condenser.

(4) flag for the design of reboiler.
   1 means design thermosyphon reboiler with re-circulation ratio fixed, calculate the reboiler tube length.
   2 means design thermosyphon reboiler with tube length fixed, calculate the re-circulation ratio.
   3 means design the kettle reboiler.
If the force circulation reboiler is designed, leave this field blank.

(5) flag for the unit of flow rate.
   1 means the flow rate unit in steady state profile is LBMOL/MIN.
   2 means the flow rate unit in steady state profile is LBMOL/HR.

(6) flag for the design of reboiler.
   1 means design the thermosyphon reboiler.
   2 means design the force circulation reboiler.
   If the kettle reboiler is designed, leave this field blank.

(F) DATA RECORD F :

(a) distillation column tray design data record.

(b) fields for the data record F.

   FORMAT : I1,9X,2F5.3,6E10.0

   (1) flag for the tray design.
      1 means Ballast tray designed by the Glitsch company.
      2 means Flexitray designed by the Koch company.

   (2) tray flood factor, fractional.

   (3) tray weep factor, fractional.

   (4) hole or perforation diameter, IN.

   (5) tray deck thickness, IN.

   (6) maximum allowable pressure drop per tray, PSI.
(7) weir height of tray, IN.
(8) system derating factor. If leaves blank, program will use the Fair's method.

(G) DATA RECORD G:
(a) tray spacing data record, one record. This data record specifies the maximum allowable tray spacing of each tray zone. But, if the calculated minimum tray spacing is larger than this one, the calculated tray spacing will be used.
(b) fields for data record G:
  FORMAT: I1,9X,2(I3,7X),E10.0
  (1) flag for tray spacing data record.
        0 means there is another data record.
        1 means there is no more data record.
  (2) the first tray of the tray spacing zone.
  (3) the last tray of the tray spacing zone.
  (4) desired tray spacing, IN, of the tray zone.

(H) DATA RECORD H:
(a) condenser design data record, two records.
(b) fields for the first data record.
  FORMAT: I3,7X,3(I1,9X),I2,8X,2E10.0,I1
  (1) number of grid points involved in the calculation of multicomponent vapor condensation weighted temperature.
  (2) flag for the design of condenser.
1 means design the vertical condenser.
2 means design the horizontal condenser.

(3) flag for the design of condenser.
0 means the condenser tube passes will be adjusted by the program.
1 means tube passes are specified by the user.

(4) number of tube passes desired for the condenser.
if (3) is zero, leave this field blank.

(5) condenser tube BWG.
(6) condenser tube outside diameter, IN.
(7) condenser tube sheet layout pitch, IN. (only triangular tube layout allowed).

(8) number of coolant components.

(c) fields for the second data record.
FORMAT : 8E10.0
(1) allowable temperature rise for the coolant, F.
(2) approximated overall heat transfer coefficient for the condenser, BTU/HR,FT**2,F.
(3) condenser tube length, FT.
(4) condenser tube side fouling, 1/ BTU/HR, FT**2,F.
(5) condenser shell side fouling, 1/ BTU/HR, FT**2,F.
(6) coolant inlet temperature, F.
(7) desired coolant velocity in tube, FT/sec.

(I) DATA RECORD I :
(a) physical properties data record for the condenser
coolant. Three records per coolant component.

(b) fields for the first data record:

FORMAT: 2A4,2X,7E10.0

(1) name of the coolant component i.
(2) normal boiling temperature of coolant component i, K.
(3) critical temperature of coolant component i, K.
(4) Pitzer's Acentric Factor of coolant component i.
(5) the liquid density reference temperature of coolant component i, K.

(c) fields for the second data record:

FORMAT: 6E10.0

(1) to (6) coefficients B1, B2, B3, B4, B5, B6 in calculating liquid enthalpy HL of coolant component i, with temperature T in F and HL in BTU/LBMOLE.

\[ HL = B1 + (B2 \times T) + (B3 \times T^2) + (B4 \times T^3) + (B5 \times T^4) + (B6 \times T^5) \]

(d) fields for the third data record:

FORMAT: 8E10.0

(1) parameter, depends on the molecular structure, to calculate the thermal conductivity of coolant component i.
(2) heat of vaporization of coolant component i at
the normal boiling point, CAL/GMOLE.

(3) saturated liquid density at the liquid density
reference temperature, GRAM/CM**3.

(4) molecular weight of coolant component i.

(5) critical pressure of coolant component i, ATM.

(6) liquid viscosity, B, of coolant component i.

(7) liquid viscosity, C0, of coolant component i.

(8) mole fraction of coolant component i.

(J) DATA RECORD J :

(a) one record. This data record is used for the dynamic
program. If only mechanical design required, put a
blank record instead.

(b) fields for data record J.

FORMAT :

(1) heat capacity, BTU/LBM,F, of distillation tower
construction metal.

(2) reboiler upstream steam supplied pressure, PSI.

(3) flag for the reflux drum liquid level controller.
   1 means the liquid level controller controls the
distillate flow rate.
   2 means the liquid level controller controls the
   reflux flow rate.

(4) flag for the tray valve opening.
   0 means tray valve part opened.
   1 means tray valve fully opened.
NOTE:

The liquid phase activity coefficients are calculated through a user supplied subroutine GAMMA. The following subroutine is used if the user does not supply one. In this case, an ideal liquid phase is assumed.

SUBROUTINE GAMMA (T, J, NC, GAM, X, IL, IH)

C
C--------NOMENCLATURE
C
C  GAM(I) = LIQUID ACTIVITY COEFFICIENTS OF COMPONENT I
C  NC    = NUMBER OF PROCESS COMPONENTS IN THE
C          DISTILLATION COLUMN
C  J      = STAGE NUMBER.
C  X(I,J) = LIQUID MOLE FRACTION OF COMPONENT I OF STAGE
C          J.
C  IL     = DUMMY VARIABLE.
C  IH     = DUMMY VARIABLE.
C
DIMENSION GAM(16), X(IL, IH)
DO 10 I=1, NC
10  GAM(I)=1.0
RETURN
END
EXAMPLE 1:

This is the mechanical design for the Benzene - Toluene system. The reflux drum for the total condenser is a horizontal cylindrical vessel without the wire mesh. The liquid product goes to the heat exchanger and there is only one liquid phase in the drum. The tray column consists of Ballast trays sized by the Glitch design manual 4900. Kettle reboiler is designed with 12 FT nominal tube length. The carbon steel tube outside diameter is 0.75 IN, tube layout pitch 1 IN and tube BWG 16. Total condenser is horizontal. The carbon steel tube length is 10 FT, tube layout pitch 0.9375 IN and tube BWG 16. The coolant is cooling water with inlet temperature 80 F.

<table>
<thead>
<tr>
<th></th>
<th>(A)</th>
<th>(B.1)</th>
<th>(B.2)</th>
<th>(C)</th>
<th>(D)</th>
<th>(E)</th>
<th>(F)</th>
<th>(G)</th>
<th>(H.1)</th>
<th>(H.2)</th>
<th>(I.1)</th>
<th>(I.2)</th>
<th>(I.3)</th>
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<td>293.0</td>
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STEADY STATE PROFILE OF THE DISTILLATION COLUMN

NUMBER OF STAGES = 21, THE FIRST STAGE IS CONDENSER, THE LAST STAGE IS REBOILER.

TOTAL CONDENSER

<table>
<thead>
<tr>
<th>TEMP (F)</th>
<th>PRESSURE (PSI)</th>
<th>REFUX FLOW RATE (LB/HR)</th>
<th>VAPOR PRODUCT RATE (LB/HR)</th>
<th>LIQUID PRODUCT RATE (LB/HR)</th>
<th>LIQUID DENSITY (LB/FT**3)</th>
<th>VAPOR DENSITY (LB/FT**3)</th>
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<tbody>
<tr>
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MOL FRACTION

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<thead>
<tr>
<th>BENZENE</th>
<th>TOLUENE</th>
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<tbody>
<tr>
<td>0.9999E+00</td>
<td>0.1000E+01</td>
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STAGE 2

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<th>TEMP (F)</th>
<th>PRESSURE (PSI)</th>
<th>LIQUID FLOW RATE (GPM)</th>
<th>VAPOR FLOW RATE (CFS)</th>
<th>LIQUID DENSITY (LB/FT**3)</th>
<th>VAPOR DENSITY (LB/FT**3)</th>
<th>SURFACE TENSION (DYNE/CM)</th>
</tr>
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<tbody>
<tr>
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<td>0.2466E+03</td>
<td>0.1181E+03</td>
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<td>0.272E+00</td>
<td>0.5303E+01</td>
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MOL FRACTION

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<th>TOLUENE</th>
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<tbody>
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STAGE 3

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<th>PRESSURE (PSI)</th>
<th>LIQUID FLOW RATE (GPM)</th>
<th>VAPOR FLOW RATE (CFS)</th>
<th>LIQUID DENSITY (LB/FT**3)</th>
<th>VAPOR DENSITY (LB/FT**3)</th>
<th>SURFACE TENSION (DYNE/CM)</th>
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</thead>
<tbody>
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<td>0.1181E+03</td>
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MOL FRACTION

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STAGE 4

<table>
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<tr>
<th>TEMP (F)</th>
<th>PRESSURE (PSI)</th>
<th>LIQUID FLOW RATE (GPM)</th>
<th>VAPOR FLOW RATE (CFS)</th>
<th>LIQUID DENSITY (LB/FT**3)</th>
<th>VAPOR DENSITY (LB/FT**3)</th>
<th>SURFACE TENSION (DYNE/CM)</th>
</tr>
</thead>
<tbody>
<tr>
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<td>0.2366E+02</td>
<td>0.2468E+03</td>
<td>0.1177E+03</td>
<td>0.503E+02</td>
<td>0.274E+00</td>
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MOL FRACTION

<table>
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<tr>
<th>BENZENE</th>
<th>TOLUENE</th>
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<tbody>
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<td>0.9996E+00</td>
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MOL FRACTION

- **STAGE 15**
  - Benzene: 0.3416E+00
  - Toluene: 0.6584E+00

- **STAGE 16**
  - Benzene: 0.2513E+00
  - Toluene: 0.7487E+00

- **STAGE 17**
  - Benzene: 0.1709E+00
  - Toluene: 0.8291E+00

- **STAGE 18**
  - Benzene: 0.1085E+00
  - Toluene: 0.9952E+00

- **STAGE 19**
  - Benzene: 0.6690E-01
  - Toluene: 0.3951E+00
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**MOL FRACTION**

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**REBOILER**

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**MOL FRACTION**

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REFLUX DRUM DESIGN

(A) DESIGN CONDITION FOR REFLUX DRUM OF TOTAL CONDENSER

(1) LIQUID PRODUCT TO HEAT EXCHANGER
(2) REFLUX DRUM IS A HORIZONTAL CYLINDRICAL SHAPE VESSEL WITHOUT A WIRE MESH

D ORIGINAL = 0.58989E+01

(B) REFLUX DRUM DESIGN SPECIFICATION

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<th>DRUM DIAMETER (FT)</th>
<th>DRUM WALL THICKNESS (IN)</th>
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## Tray Column Sizing by Glitsch Manual Ballast Tray Manual 4000

### Column Description

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<th>Diameter of Deck Tray (in)</th>
<th>Tray Pressure Drop (psi)</th>
<th>Mat Shear Stress (psi)</th>
<th>Height Tower (ft)</th>
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KETTLE REBOILER DESIGN

(A) (1) USER SUPPLIES THE TUBE LENGTH
(2) ASSUMING THE PRESSURE DROP IS NEGLIGIBLE

(B) MECHANICAL DESCRIPTION

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<th>TUBE PASS</th>
<th>NUMBER OF TUBES</th>
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<th>TUBE OUTSIDE DIAMETER (IN)</th>
<th>Bwg</th>
<th>PITCH (IN)</th>
<th>LENGTH (FT)</th>
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(C) STEAM PROPERTIES

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<th>STEAM PRES (PSI)</th>
<th>STEAM TEMP (F)</th>
<th>STEAM LATENT HEAT (BTU/LB)</th>
<th>STEAM MASS FLOW RATE (LB/HR)</th>
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<tr>
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(D) PROCESS DESCRIPTION OF THE KETTLE REBOILER

<table>
<thead>
<tr>
<th>REBOILER HEAT DUTY (BTU/HR)</th>
<th>VAPOR MASS FLOW RATE (LB/HR)</th>
<th>PROCESS TEMPERATURE (F)</th>
<th>SAFETY FACTOR</th>
<th>CLEAN HEAT TRANSFER COEFF (BTU/HR,F,FT2)</th>
<th>DIRTY HEAT TRANSFER COEFF (BTU/HR,F,FT2)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.8895E+07</td>
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<td>0.2649E+03</td>
<td>0.1154E+00</td>
<td>0.2928E+03</td>
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</tbody>
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TOTAL CONDENSER DESIGN

(A) HORIZONTAL CONDENSER

(B) MECHANICAL DESCRIPTION

<table>
<thead>
<tr>
<th>NUMBER OF CONDENSER</th>
<th>SHELL PASS</th>
<th>TUBE PASS</th>
<th>NUMBER OF TUBES</th>
<th>SHELL DIAMETER (IN)</th>
<th>TUBE OUTSIDE DIAMETER (IN)</th>
<th>TUBE BWG</th>
<th>TUBE PITCH (IN)</th>
<th>TUBE LENGTH (FT)</th>
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<tr>
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<td>0.9375E+00</td>
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</table>

(C) PROCESS DESCRIPTION

<table>
<thead>
<tr>
<th>REQUIRED COOLANT, LB/HR TEMP (F)</th>
<th>COOLANT IN TEMP (F)</th>
<th>COOLANT OUT TEMP (F)</th>
<th>TUBE PRESSURE DROP (PSI)</th>
<th>VELOCITY IN TUBE (FT/SEC)</th>
<th>PROCESS IN TEMP (F)</th>
<th>PROCESS OUT TEMP (F)</th>
<th>CLEAN H.T.C. (BTU/HR, FT2)</th>
<th>DIRTY H.T.C. (BTU/HR, FT2)</th>
<th>TEMPERATURE (F)</th>
<th>CONDENSER HEAT DUTY (BTU/HR)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.4349E+06</td>
<td>0.8000E+02</td>
<td>0.1200E+03</td>
<td>0.4906E+01</td>
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<td>0.9678E+02</td>
<td>0.1041E+03</td>
<td>0.1738E+08</td>
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</tbody>
</table>
EXAMPLE 2:

This is the mechanical design for the Benzene - Toluene system. The reflux drum for the total condenser is a horizontal cylindrical vessel with wire mesh. The liquid product goes to storage and there is only one liquid phase in the drum. The tray column consists of Flexitrays sized by the Koch Flexitray design manual. Thermosyphon reboiler is designed with fixed 10 FT tube length and the re-circulation ratio is calculated. The carbon steel tube outside diameter is 0.75 IN, tube layout pitch 1 IN and tube BWG 16. Total condenser is vertical. The carbon steel tube length is 10 FT, tube layout pitch 0.9375 IN and tube BWG 16. The coolant is cooling water with inlet temperature 80 F.

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<th>(B.1)</th>
<th>(B.2)</th>
<th>(B.3)</th>
<th>(C)</th>
<th>(D)</th>
<th>(E)</th>
<th>(F)</th>
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<th>(H.1)</th>
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REFLUX DRUM DESIGN

(A) DESIGN CONDITION FOR REFLUX DRUM OF TOTAL CONDENSER

(1) LIQUID PRODUCT TO STORAGE
(2) REFLUX DRUM IS A HORIZONTAL CYLINDRICAL SHAPE VESSEL WITH A WIRE MESH

D ORIGINAL = 0.58989E+01

(B) REFLUX DRUM DESIGN SPECIFICATION

<table>
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<tr>
<th>DRUM LENGTH (FT)</th>
<th>DRUM DIAMETER (FT)</th>
<th>DRUM WALL THICKNESS (IN)</th>
<th>LIQUID LEAVE DRUM (LB/HR)</th>
<th>VAPOR LEAVE DRUM (LB/HR)</th>
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<tr>
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<td>WEIR LENGTH (FT)</td>
<td>DOWNCOMER AREA (FT**2)</td>
<td>ACTIVE TRAY AREA (FT**2)</td>
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THERMOSPHERIC REBOILER DESIGN

(A) (1) - TUBE LENGTH IS FIXED. CALCULATE THE RECIRCULATION RATIO
(2) - PIPING LOSS IS CALCULATED ACCORDING TO THE EQUIVALENT LENGTH SUGGESTED BY THE LUDWIG

(B) MECHANICAL DESCRIPTION

<table>
<thead>
<tr>
<th>NUMBER OF REBOILER</th>
<th>SHELL PASS</th>
<th>TUBE PASS</th>
<th>NUMBER OF TUBES</th>
<th>SHELL DIAMETER (IN)</th>
<th>TUBE OUTSIDE DIAMETER (IN)</th>
<th>BWG</th>
<th>PITCH (IN)</th>
<th>TUBE LENGTH (FT)</th>
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(C) STEAM PROPERTIES

<table>
<thead>
<tr>
<th>STEAM FRES (PSI)</th>
<th>STEAM TEMP (F)</th>
<th>STEAM LATENT HEAT (BTU/LBM)</th>
<th>STEAM MASS FLOW RATE (LB/HR)</th>
</tr>
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<tbody>
<tr>
<td>0.9598E+02</td>
<td>0.3249E+03</td>
<td>0.6911E+03</td>
<td>0.1996E+05</td>
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(D) PROCESS DESCRIPTION OF REBOILER.

<table>
<thead>
<tr>
<th>REBOILER HEAT DUTY (BTU/HR)</th>
<th>VAPOR FLOW RATE (LB/HR)</th>
<th>PROCESS TEMPERATURE (F)</th>
<th>RECIRCULATE RATIO</th>
<th>RESISTANCE TOTAL (PSI)</th>
<th>CLEAN H.T.C. (BTU/HR, F, FT²)</th>
<th>DIRTY H.T.C. (BTU/HR, F, FT²)</th>
<th>TOTAL DRIVING FORCE (PSI)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.1779E+08</td>
<td>0.1249E+06</td>
<td>0.2649E+03</td>
<td>0.2417E+02</td>
<td>0.3330E+01</td>
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TOTAL DRIVING FORCE (PSI)

0.3334E+01
TOTAL CONDENSE DESIGN

(A) VERTICAL CONDENSER

(B) MECHANICAL DESCRIPTION

<table>
<thead>
<tr>
<th>NUMBER OF CONDENSER</th>
<th>SHELL PASS</th>
<th>TUBE PASS</th>
<th>NUMBER OF TUBES</th>
<th>SHELL DIAMETER (IN)</th>
<th>TUBE OUTSIDE DIAMETER (IN)</th>
<th>BWG</th>
<th>PITCH (IN)</th>
<th>LENGTH (FT)</th>
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<td>0.2325E+02</td>
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<td>0.9375E+00</td>
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(C) PROCESS DESCRIPTION

<table>
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<tr>
<th>REQUIRED COOLANT IN</th>
<th>COOLANT OUT</th>
<th>TUBE PRESSURE DROP (PSI)</th>
<th>VELOCITY IN TUBE (FT/SEC)</th>
<th>PROCESS IN TEMP (F)</th>
<th>PROCESS OUT TEMP (F)</th>
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<tr>
<td>0.4349E+06</td>
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<td>0.1200E+03</td>
<td>0.2854E+01</td>
<td>0.2048E+03</td>
<td>0.2034E+03</td>
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<table>
<thead>
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<th>CLEAN H.T.C. (BTU/HR, F, FT²)</th>
<th>DIRTY H.T.C. (BTU/HR, F, FT²)</th>
<th>TEMPERATURE LMTD (F)</th>
<th>CONDENSER HEAT DUTY (BTU/HR)</th>
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EXAMPLE 3:

This is the mechanical design for the propane - butane - pentane system. The reflux drum for the total condenser is a horizontal cylindrical vessel without the wire mesh. The liquid product goes to storage and there is only one liquid phase in the drum. The tray column consists of Ballast trays sized by the Glitsch's design manual 4900. Kettle reboiler is designed with 8 FT nominal tube length. The carbon steel tube outside diameter is 0.75 IN, tube layout pitch 1 IN and tube BWG 16. Total condenser is horizontal. The carbon steel tube length is 16 FT, tube layout pitch 0.9375 IN and tube BWG 16. The coolant is cooling water with inlet temperature 72 F.

<p>| | | | | | | | | |</p>
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<td>304.0</td>
<td>0.262</td>
<td>0.626</td>
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<td>N-PENTANE</td>
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</tbody>
</table>

(A)   (B.1)   (B.2)   (B.1)   (B.2)   (C)   (D)   (E)   (F)   (G)   (H.1)   (H.2)   (I.1)   (I.2)   (I.3)   (J)
STADY STATE PROFILE OF THE DISTILLATION COLUMN

NUMBER OF STAGES = 11, THE FIRST STAGE IS CONDENSER. THE LAST STAGE IS REBOILER.

TOTAL CONDENSATES

<table>
<thead>
<tr>
<th>TEMP (F)</th>
<th>PRESSURE (PSI)</th>
<th>REFLUX FLOW</th>
<th>VAPOR PRODUCT</th>
<th>LIQUID PRODUCT</th>
<th>LIQUID DENSITY</th>
<th>VAPOR DENSITY</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.9941E+02</td>
<td>0.1800E+03</td>
<td>0.3351E+05</td>
<td>0.0</td>
<td>0.1117E+05</td>
<td>0.2942E+02</td>
<td>0.0</td>
</tr>
</tbody>
</table>

MOL FRACTION

<table>
<thead>
<tr>
<th>X</th>
<th>Y</th>
</tr>
</thead>
<tbody>
<tr>
<td>PROPANE</td>
<td>0.9581E+00</td>
</tr>
<tr>
<td>N-BUTANE</td>
<td>0.4163E-01</td>
</tr>
<tr>
<td>N-PENTAN</td>
<td>0.2663E-03</td>
</tr>
</tbody>
</table>

STAGE 2

<table>
<thead>
<tr>
<th>TEMP (F)</th>
<th>PRESSURE (PSI)</th>
<th>LIQUID FLOW</th>
<th>VAPOR FLOW</th>
<th>LIQUID DENSITY</th>
<th>VAPOR DENSITY</th>
<th>SURFACE TENSION</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.1052E+03</td>
<td>0.1801E+03</td>
<td>0.1361E+03</td>
<td>0.7714E+01</td>
<td>0.2970E+02</td>
<td>0.1609E+01</td>
<td>0.5477E+01</td>
</tr>
</tbody>
</table>

MOL FRACTION

<table>
<thead>
<tr>
<th>X</th>
<th>Y</th>
</tr>
</thead>
<tbody>
<tr>
<td>PROPANE</td>
<td>0.8633E+00</td>
</tr>
<tr>
<td>N-BUTANE</td>
<td>0.1399E+00</td>
</tr>
<tr>
<td>N-PENTAN</td>
<td>0.2766E-02</td>
</tr>
</tbody>
</table>

STAGE 3

<table>
<thead>
<tr>
<th>TEMP (F)</th>
<th>PRESSURE (PSI)</th>
<th>LIQUID FLOW</th>
<th>VAPOR FLOW</th>
<th>LIQUID DENSITY</th>
<th>VAPOR DENSITY</th>
<th>SURFACE TENSION</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.1176E+03</td>
<td>0.1802E+03</td>
<td>0.1289E+03</td>
<td>0.7552E+01</td>
<td>0.3015E+02</td>
<td>0.1609E+01</td>
<td>0.5631E+01</td>
</tr>
</tbody>
</table>

MOL FRACTION

<table>
<thead>
<tr>
<th>X</th>
<th>Y</th>
</tr>
</thead>
<tbody>
<tr>
<td>PROPANE</td>
<td>0.6887E+00</td>
</tr>
<tr>
<td>N-BUTANE</td>
<td>0.2936E+00</td>
</tr>
<tr>
<td>N-PENTAN</td>
<td>0.1711E-01</td>
</tr>
</tbody>
</table>

STAGE 4

<table>
<thead>
<tr>
<th>TEMP (F)</th>
<th>PRESSURE (PSI)</th>
<th>LIQUID FLOW</th>
<th>VAPOR FLOW</th>
<th>LIQUID DENSITY</th>
<th>VAPOR DENSITY</th>
<th>SURFACE TENSION</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.1375E+03</td>
<td>0.1803E+03</td>
<td>0.1248E+03</td>
<td>0.7311E+01</td>
<td>0.3061E+02</td>
<td>0.1609E+01</td>
<td>0.5723E+01</td>
</tr>
</tbody>
</table>

MOL FRACTION

<table>
<thead>
<tr>
<th>X</th>
<th>Y</th>
</tr>
</thead>
<tbody>
<tr>
<td>PROPANE</td>
<td>0.4722E+00</td>
</tr>
<tr>
<td>N-BUTANE</td>
<td>0.4511E+00</td>
</tr>
<tr>
<td>N-PENTAN</td>
<td>0.7675E-01</td>
</tr>
<tr>
<td>Stage</td>
<td>Temp (°F)</td>
</tr>
<tr>
<td>-------</td>
<td>-----------</td>
</tr>
<tr>
<td>5</td>
<td>0-160</td>
</tr>
<tr>
<td>6</td>
<td>0-180</td>
</tr>
<tr>
<td>7</td>
<td>0-180</td>
</tr>
<tr>
<td>8</td>
<td>0-205</td>
</tr>
<tr>
<td>STAGE 9</td>
<td>TEMP (F)</td>
</tr>
<tr>
<td>---------</td>
<td>----------</td>
</tr>
<tr>
<td></td>
<td>0.2134E+03</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>MOL FRACTION</th>
</tr>
</thead>
<tbody>
<tr>
<td>PROPANE</td>
</tr>
<tr>
<td>N-BUTANE</td>
</tr>
<tr>
<td>N-PENTANE</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>STAGE 10</th>
<th>TEMP (F)</th>
<th>PRESSURE (PSI)</th>
<th>LIQUID FLOW RATE (GPH)</th>
<th>VAPOR FLOW RATE (CFS)</th>
<th>LIQUID DENSITY (LB/FT³)</th>
<th>VAPOR DENSITY (LB/FT³)</th>
<th>SURFACE TENSION (DYN/E CM)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.2231E+03</td>
<td>0.1809E+03</td>
<td>0.3729E+03</td>
<td>0.9636E+01</td>
<td>0.3023E+02</td>
<td>0.1861E+01</td>
<td>0.4720E+01</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>MOL FRACTION</th>
</tr>
</thead>
<tbody>
<tr>
<td>PROPANE</td>
</tr>
<tr>
<td>N-BUTANE</td>
</tr>
<tr>
<td>N-PENTANE</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>REBOILER</th>
<th>TEMP (F)</th>
<th>PRESSURE (PSI)</th>
<th>LIQUID FLOW RATE (LB/HR)</th>
<th>VAPOR FLOW RATE (LB/HR)</th>
<th>LIQUID DENSITY (LB/FT³)</th>
<th>VAPOR DENSITY (LB/FT³)</th>
<th>SURFACE TENSION (DYN/E CM)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.2369E+03</td>
<td>0.1810E+03</td>
<td>0.2328E+05</td>
<td>0.6713E+05</td>
<td>0.3020E+02</td>
<td>0.1898E+01</td>
<td>0.4720E+01</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>MOL FRACTION</th>
</tr>
</thead>
<tbody>
<tr>
<td>PROPANE</td>
</tr>
<tr>
<td>N-BUTANE</td>
</tr>
<tr>
<td>N-PENTANE</td>
</tr>
</tbody>
</table>
REFLUX DRUM DESIGN

(A) DESIGN CONDITION FOR REFLUX DRUM OF TOTAL CONDENSER

(1) LIQUID PRODUCT TO STORAGE
(2) REFLUX DRUM IS A HORIZONTAL CYLINDRICAL SHAPE VESSEL WITHOUT A WIRE MESH

D ORIGINAL = 0.43942E+01

(B) REFLUX DRUM DESIGN SPECIFICATION

<table>
<thead>
<tr>
<th>DRUM LENGTH (FT)</th>
<th>DRUM DIAMETER (FT)</th>
<th>DRUM WALL THICKNESS (IN)</th>
<th>LIQUID LEAVE DRUM (LB/HR)</th>
<th>VAPOR LEAVE DRUM (LB/HR)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.7956E+01</td>
<td>0.4500E+01</td>
<td>0.1540E+00</td>
<td>0.4468E+05</td>
<td>0.0</td>
</tr>
<tr>
<td>TRAY NO</td>
<td>PASS</td>
<td>TRAY SPACE (IN)</td>
<td>WEIR LENGTH (FT)</td>
<td>DOWNSPOUT AREA (FT²)</td>
</tr>
<tr>
<td>---------</td>
<td>------</td>
<td>----------------</td>
<td>------------------</td>
<td>----------------------</td>
</tr>
<tr>
<td>2</td>
<td>1</td>
<td>0.1800E+02</td>
<td>0.3878E+01</td>
<td>0.3724E+01</td>
</tr>
<tr>
<td>3</td>
<td>1</td>
<td>0.1800E+02</td>
<td>0.3878E+01</td>
<td>0.3724E+01</td>
</tr>
<tr>
<td>4</td>
<td>1</td>
<td>0.1800E+02</td>
<td>0.3878E+01</td>
<td>0.3724E+01</td>
</tr>
<tr>
<td>5</td>
<td>1</td>
<td>0.1800E+02</td>
<td>0.3878E+01</td>
<td>0.3724E+01</td>
</tr>
<tr>
<td>6</td>
<td>1</td>
<td>0.1801E+02</td>
<td>0.3878E+01</td>
<td>0.3724E+01</td>
</tr>
<tr>
<td>7</td>
<td>1</td>
<td>0.1801E+02</td>
<td>0.3878E+01</td>
<td>0.3724E+01</td>
</tr>
<tr>
<td>8</td>
<td>1</td>
<td>0.1801E+02</td>
<td>0.3878E+01</td>
<td>0.3724E+01</td>
</tr>
<tr>
<td>9</td>
<td>1</td>
<td>0.1801E+02</td>
<td>0.3878E+01</td>
<td>0.3724E+01</td>
</tr>
<tr>
<td>10</td>
<td>1</td>
<td>0.1801E+02</td>
<td>0.3878E+01</td>
<td>0.3724E+01</td>
</tr>
</tbody>
</table>

**TRAY COLUMN SIZING BY GLITSCH MANUAL BALLAST TRAY MANUAL 4900**

**COLUMN DESCRIPTION**

<table>
<thead>
<tr>
<th>FLOOD FACTOR</th>
<th>TRAY PRESSURE</th>
<th>MAT SHEAR</th>
<th>TRAY HEIGHT</th>
<th>DIAMETER OF</th>
<th>DECK</th>
<th>TRAY PRESSURE</th>
<th>WEIR (IN)</th>
<th>TOWER (FT)</th>
<th>DIAMETER OF</th>
<th>HEIG</th>
<th>JOINT</th>
<th>CONSTR MAT</th>
<th>EFFICIENCY</th>
<th>DEN (LB/FT³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.800</td>
<td>0.600</td>
<td>0.3750E+00</td>
<td>0.0780000</td>
<td>0.1000E+00</td>
<td>0.1500E+05</td>
<td>2.000000</td>
<td>0.5500E+01</td>
<td>0.2438E+02</td>
<td>0.050</td>
<td>0.2840E+00</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

*Note: Values are in scientific notation.*
KETTLE REBOILER DESIGN

(A) (1) USER SUPPLIES THE TUBE LENGTH
(2) ASSUMING THE PRESSURE DROP IS NEGLIGIBLE

(B) MECHANICAL DESCRIPTION

<table>
<thead>
<tr>
<th>NUMBER OF</th>
<th>SHELL PASS</th>
<th>TUBE PASS</th>
<th>NUMBER OF TUBES</th>
<th>SHELL DIAMETER (IN)</th>
<th>TUBE OUTSIDE DIAMETER (IN)</th>
<th>TUBE BWG</th>
<th>TUBE PITCH (IN)</th>
<th>TUBE LENGTH (FT)</th>
</tr>
</thead>
<tbody>
<tr>
<td>2</td>
<td>1</td>
<td>2</td>
<td>317</td>
<td>0.300E+02</td>
<td>0.750E+00</td>
<td>16</td>
<td>0.100E+01</td>
<td>0.800E+01</td>
</tr>
</tbody>
</table>

(C) STEAM PROPERTIES

<table>
<thead>
<tr>
<th>STEAM PRESS (PSI)</th>
<th>STEAM TEMP (F)</th>
<th>STEAM LATENT HEAT (BTU/LBM)</th>
<th>STEAM MASS FLOW RATE (LB/HR)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.6399E+02</td>
<td>0.2969E+03</td>
<td>0.9544E+03</td>
<td>0.4613E+04</td>
</tr>
</tbody>
</table>

(D) PROCESS DESCRIPTION OF THE KETTLE REBOILER

<table>
<thead>
<tr>
<th>REBOILER HEAT DUTY (BTU/HR)</th>
<th>VAPOR MASS FLOW RATE (LBM/HR)</th>
<th>PROCESS TEMPERATURE (F)</th>
<th>SAFETY FACTOR</th>
<th>CLEAN HEAT TRANSFER COEFF (BTU/HR,F,FT2)</th>
<th>DIRTY HEAT TRANSFER COEFF (BTU/HR,F,FT2)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.4403E+07</td>
<td>0.6713E+05</td>
<td>0.2369E+03</td>
<td>0.5388E+00</td>
<td>0.7753E+03</td>
<td>0.3040E+03</td>
</tr>
</tbody>
</table>
TOTAL CONDENSER DESIGN

(A) HORIZONTAL CONDENSER

(B) MECHANICAL DESCRIPTION

<table>
<thead>
<tr>
<th>NUMBER OF SHELL PASS</th>
<th>TUBE PASS</th>
<th>NUMBER OF TUBES</th>
<th>SHELL DIAMETER (IN)</th>
<th>TUBE OUTSIDE DIAMETER (IN)</th>
<th>TUBE BWG</th>
<th>TUBE LAYOUT PITCH (IN)</th>
<th>TUBE LENGTH (Ft)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>1</td>
<td>8</td>
<td>1212</td>
<td>0.3900E+02</td>
<td>0.7500E+00</td>
<td>16</td>
<td>0.9375E+00</td>
</tr>
</tbody>
</table>

(C) PROCESS DESCRIPTION

<table>
<thead>
<tr>
<th>REQUIRED COOLANT, LB/HR</th>
<th>COOLANT IN TEMP (F)</th>
<th>COOLANT OUT TEMP (F)</th>
<th>TUBE PRESSURE DROP (PSI)</th>
<th>VELOCITY IN TUBE (FT/SEC)</th>
<th>PROCESS IN TEMP (F)</th>
<th>PROCESS OUT TEMP (F)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.3672E+06</td>
<td>0.7200E+02</td>
<td>0.9200E+02</td>
<td>0.2035E+02</td>
<td>0.4900E+01</td>
<td>0.1052E+03</td>
<td>0.9941E+02</td>
</tr>
</tbody>
</table>

CLEAN H.T.C. (BTU/HR,F,FT2) | DIRTY H.T.C. (BTU/HR,F,FT2) | TEMPERATURE LMTO (F) | CONDENSER HEAT DUTY (BTU/HR) |
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>0.1110E+03</td>
<td>0.0191E+02</td>
<td>0.1833E+02</td>
<td>0.6939E+07</td>
</tr>
</tbody>
</table>
EXAMPLE 4:
This is the mechanical design for the propane - butane - pentane system.
The reflux drum for the total condenser is a horizontal cylindrical vessel
without the wire mesh. The liquid product goes to storage and there is only one
liquid phase in the drum. The tray column consists of Ballast trays sized by
the Glitsch's design manual 4900. Force circulation reboiler is designed with
10 FT tube length and re-circulation ratio 10. The carbon steel tube outside
diameter is 0.75 IN, tube layout pitch 1 IN and tube BWG 16. Total condenser is
horizontal. The carbon steel tube length is 16 FT, tube layout pitch 0.9375 IN
and tube BWG 16. The coolant is cooling water with inlet temperature 72 F.

| 3 11 | PROPA N E | 151.0 | 231.0 |
| 41.9 | N - BUTANE | 191.0 | 293.0 |
| 37.5 | N - PENTANE | 231.0 | 293.0 |
| 33.3 | 0.251 | 304.0 | 0.262 | 0.626 | 0. | 313.66 | 182.48 |
| 15000.0 | 0.85 | 0.284 | 0.001 | 0.001 | 30.0 | 30.0 |
| 10.0 | 8000.0 | 8.0 | 0.75 | 16 | 1.0 | 0. | 10.0 |
| 1 | 1 | 1 | 2 | 2 |
| 1 | 0.8 | 0.6 | 0.375 | 0.078 | 0.1 | 2.0 |
| 1 | 1 | 10 | 18.0 |
| 11 | 2 | 0 | 6 | 16 | 0.75 | 0.9375 | 1 |
| 20.0 | 120.0 | 16.0 | 0.002 | 0.001 | 72.0 | 5.0 |
| WATER | 373.2 | 647.3 | 0.344 | 293.0 |
| 3603.3954718.0007638-.81065E-4 |
| 0. | 9717.0 | 0.998 | 18.015 | 217.6 | 658.25 | 283.16 | 1.0 |
| 0.107 | 150.0 | 2 |
FORCE CIRCULATION REBOILER DESIGN

(A) (1) TUBE LENGTH AND RE-CIRCULATION RATIO ARE PROVIDED BY THE USER
(2) THE INLET PIPING FRICTION LOSS IS CALCULATED BASED ON 100 FT PIPING, 3 FULLY OPENED GATE VALVE, 1 SWING CHECK VALVE AND 2 90 DEGREE LONG SWEEP ELBOW
(3) THE OUTLET PIPING FRICTION LOSS IS CALCULATED BASED ON 100 FT PIPING, 1 GLOBE VALVE, 1 CONTROL VALVE, TWO 90 DEGREE LONG SWING ELBOW, 2 GATE VALVE AND 2 TEES

(B) MECHANICAL DESCRIPTION

<table>
<thead>
<tr>
<th>NUMBER OF SHELL</th>
<th>SHELL TUBE</th>
<th>SHELL NUMBER</th>
<th>SHELL OUTSIDE TUBE</th>
<th>SHELL TUBE LAYOUT</th>
<th>TUBE LENGTH(FT)</th>
</tr>
</thead>
<tbody>
<tr>
<td>REBOILER PASS</td>
<td>PASS TUBES</td>
<td>DIAMETER( IN)</td>
<td>DIAMETER( IN)</td>
<td>PITCH( IN)</td>
<td></td>
</tr>
<tr>
<td>1</td>
<td>1</td>
<td>745</td>
<td>0.3100E+02</td>
<td>0.7500E+00</td>
<td>16</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0.1000E+01</td>
</tr>
</tbody>
</table>

(C) STEAM PROPERTIES

<table>
<thead>
<tr>
<th>STEAM PRESS (PSI)</th>
<th>STEAM TEMP (F)</th>
<th>STEAM LATENT HEAT (BTU/LBH)</th>
<th>STEAM MASS FLOW RATE (LB/HR)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.6714E+02</td>
<td>0.3013E+03</td>
<td>0.9100E+03</td>
<td>0.9676E+04</td>
</tr>
</tbody>
</table>

(D) PROCESS DESCRIPTION OF REBOILER

<table>
<thead>
<tr>
<th>REBOILER VAPOR MASS</th>
<th>PROCESS HEAT DUTY</th>
<th>RECIRCULATE RATIO</th>
<th>RESISTANCE TOTAL(PSI)</th>
<th>CLEAN H.T.C. (BTU/HR,F,FT2)</th>
<th>DIRTY H.T.C. (BTU/HR,F,FT2)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.8655E+07</td>
<td>0.6714E+05</td>
<td>0.2401E+03</td>
<td>0.1000E+02</td>
<td>0.2305E+01</td>
<td>0.2042E+03</td>
</tr>
<tr>
<td>PUMP HORSEPOWER</td>
<td>FLOW RATE</td>
<td>PASS PUMP (GPM)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>0.1890E+02</td>
<td>0.3048E+04</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
EXAMPLE 5:
This is the mechanical design for the propane - butane - pentane system. The reflux drum for the total condenser is a horizontal cylindrical vessel without the wire mesh. The liquid product goes to storage and there is only one liquid phase in the drum. The tray column consists of Flexitrays sized by the Koch Flexitray design manual. Thermosyphon reboiler is designed with fixed re-circulation ratio 8 and the tube length is calculated. The carbon steel tube outside diameter is 0.75 IN, tube layout pitch 1 IN and tube BWG 16. Total condenser is vertical. The carbon steel tube length is 16 FT, tube layout pitch 0.9375 IN and tube BWG 16. The coolant is cooling water with inlet temperature 72 F.

| 11 | 151.0 | 231.0 | (A) |
| 41.9 | 0.152 | 203.0 | 0.281 | 0.582 | 0. | 222.67 | 133.41 | (B.1) |
| N-BUTANE | 191.0 | 293.0 | (B.1) |
| 37.5 | 0.193 | 255.0 | 0.274 | 0.579 | 0. | 265.84 | 160.2 | (B.2) |
| N-PENTANE | 231.0 | 293.0 | (B.1) |
| 33.3 | 0.251 | 304.0 | 0.262 | 0.626 | 0. | 313.66 | 182.48 | (B.2) |
| 15000.0 | 0.85 | 0.284 | 0.001 | 0.001 | 30.0 | 30.0 | (C) |
| 8.0 | 9000.0 | 20.0 | 0.75 | 16 | 1.0 | 0. | 10.0 | (D) |
| 1 | 1 | 1 | 2 | 1 | (E) |
| 2 | 0.8 | 0.6 | 0.375 | 0.078 | 0.1 | 2.0 | (F) |
| 1 | 10 | 20.0 | (G) |
| 11 | 1 | 0 | 6 | 16 | 0.75 | 0.9375 | 1 | (H.1) |
| 20.0 | 600.0 | 16.0 | 0.002 | 0.001 | 72.0 | 5.0 | (H.2) |
| WATER | 373.2 | 647.3 | 0.344 | 293.0 | (I.1) |
| 3603.3954718.0007638-.81065E-4 | (I.2) |
| 0. | 9717.0 | 0.998 | 18.015 | 217.6 | 658.25 | 283.16 | 1.0 | (I.3) |
| 0.107 | 150.0 | 2 | (J) |
REFLUX DRUM DESIGN

(A) DESIGN CONDITION FOR REFLUX DRUM OF TOTAL CONDENSER

(1) LIQUID PRODUCT TO STORAGE

(2) REFLUX DRUM IS A HORIZONTAL CYLINDRICAL SHAPE VESSEL WITHOUT A WIRE MESH

\[ D \text{ ORIGINAL} = 0.43942 \times 10^1 \]

(B) REFLUX DRUM DESIGN SPECIFICATION

<table>
<thead>
<tr>
<th>DRUM LENGTH (FT)</th>
<th>DRUM DIAMETER (FT)</th>
<th>DRUM WALL THICKNESS (IN)</th>
<th>LIQUID LEAVE DRUM (LB/HR)</th>
<th>VAPOR LEAVE DRUM (LB/HR)</th>
</tr>
</thead>
<tbody>
<tr>
<td>(0.7956 \times 10^1)</td>
<td>(0.4500 \times 10^1)</td>
<td>(0.1540 \times 10^0)</td>
<td>(0.4468 \times 10^5)</td>
<td>0.0</td>
</tr>
</tbody>
</table>
## COLUMN DESCRIPTION

<table>
<thead>
<tr>
<th>TRAY NO</th>
<th>TRAY PASS</th>
<th>WEIR</th>
<th>DOWNCOMER</th>
<th>ACTIVE TRAY</th>
<th>HOLE AREA</th>
<th>LIQUID CREST</th>
<th>LIQUID FLOW</th>
<th>CONSTRUCTION MASS (LB)</th>
</tr>
</thead>
<tbody>
<tr>
<td>2</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
<td>0.1971E+02</td>
<td>0.8518E+00</td>
<td>0.9367E+00</td>
<td>0.4295E+01</td>
</tr>
<tr>
<td>3</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
<td>0.1971E+02</td>
<td>0.8518E+00</td>
<td>0.9367E+00</td>
<td>0.4295E+01</td>
</tr>
<tr>
<td>4</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
<td>0.1971E+02</td>
<td>0.8518E+00</td>
<td>0.9367E+00</td>
<td>0.4295E+01</td>
</tr>
<tr>
<td>5</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
<td>0.1971E+02</td>
<td>0.8518E+00</td>
<td>0.9367E+00</td>
<td>0.4295E+01</td>
</tr>
<tr>
<td>6</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
<td>0.1971E+02</td>
<td>0.8518E+00</td>
<td>0.9367E+00</td>
<td>0.4295E+01</td>
</tr>
<tr>
<td>7</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
<td>0.1971E+02</td>
<td>0.8518E+00</td>
<td>0.9367E+00</td>
<td>0.4295E+01</td>
</tr>
<tr>
<td>8</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
<td>0.1971E+02</td>
<td>0.8518E+00</td>
<td>0.9367E+00</td>
<td>0.4295E+01</td>
</tr>
<tr>
<td>9</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
<td>0.1971E+02</td>
<td>0.8518E+00</td>
<td>0.9367E+00</td>
<td>0.4295E+01</td>
</tr>
<tr>
<td>10</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
<td>0.1971E+02</td>
<td>0.8518E+00</td>
<td>0.9367E+00</td>
<td>0.4295E+01</td>
</tr>
</tbody>
</table>

## COLUMN SIZING BY KCCH FLEXITRAY MANUAL

<table>
<thead>
<tr>
<th>FLOOD FACTOR</th>
<th>WEEP FACTOR</th>
<th>DIAMETER OF HOLE (IN)</th>
<th>DECK THICK (IN)</th>
<th>TRAY PRESSURE DROP (PSI)</th>
<th>MAX SHEAR STRESS (PSI)</th>
<th>MAX WEIR (IN)</th>
<th>TOWER HEIGHT (FT)</th>
<th>JOINT EFFICIENCY DRN (LB/IN³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.600</td>
<td>0.600</td>
<td>0.3750E+00</td>
<td>0.7500E+00</td>
<td>0.1000E+00</td>
<td>0.1500E+00</td>
<td>2.0000E+00</td>
<td>0.6000E+01</td>
<td>0.2414E+02</td>
</tr>
</tbody>
</table>

### TRAY COLUMN WEIR, SPACE, LENGTH, AREA

<table>
<thead>
<tr>
<th>TRAY NO</th>
<th>TRAY PASS</th>
<th>WEIR</th>
<th>LENGTH</th>
<th>AREA (FT²)</th>
</tr>
</thead>
<tbody>
<tr>
<td>2</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
</tr>
<tr>
<td>3</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
</tr>
<tr>
<td>4</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
</tr>
<tr>
<td>5</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
</tr>
<tr>
<td>6</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
</tr>
<tr>
<td>7</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
</tr>
<tr>
<td>8</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
</tr>
<tr>
<td>9</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
</tr>
<tr>
<td>10</td>
<td>1</td>
<td>0.2000E+02</td>
<td>0.4190E+01</td>
<td>0.4272E+01</td>
</tr>
</tbody>
</table>
THERMOSYPHON REBOILER DESIGN

(A) (1) RECIRCULATION RATIO IS FIXED, CALCULATE THE TUBE LENGTH
(2) PIPING LOSS IS CALCULATED ACCORDING TO THE EQUIVALENT LENGTH SUGGESTED BY THE LUDWIG

(B) MECHANICAL DESCRIPTION

<table>
<thead>
<tr>
<th>NUMBER OF REBOILER</th>
<th>SHELL PASS</th>
<th>TUBE PASS</th>
<th>NUMBER OF TUBES</th>
<th>SHELL DIAMETER (IN)</th>
<th>TUBE OUTSIDE DIAMETER (IN)</th>
<th>TYPE</th>
<th>PITCH (IN)</th>
<th>TUBE LENGTH (FT)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>1</td>
<td>1</td>
<td>470</td>
<td>0.2500E+02</td>
<td>0.7500E+00</td>
<td>16</td>
<td>0.1000E+01</td>
<td>0.1200E+02</td>
</tr>
</tbody>
</table>

(C) STEAM PROPERTIES

<table>
<thead>
<tr>
<th>STEAM PRESS (PSI)</th>
<th>STEAM TEMP (F)</th>
<th>STEAM LATENT HEAT (BTU/LBM)</th>
<th>STEAM MASS FLOW RATE (LB/HR)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.6399E+02</td>
<td>0.2969E+03</td>
<td>0.9124E+03</td>
<td>0.9651E+04</td>
</tr>
</tbody>
</table>

(D) PROCESS DESCRIPTION OF REBOILER

<table>
<thead>
<tr>
<th>REBOILER HEAT CUTO (BTU/HR)</th>
<th>VAPOR MASS FLOW RATE (LBM/HR)</th>
<th>PROCESS TEMPERATURE (F)</th>
<th>RECIRCULATE RATIO</th>
<th>RESISTANCE TOTAL (PSI)</th>
<th>CLEAN H.T.C. (BTU/HR,F,FT2)</th>
<th>DIRTY H.T.C. (BTU/HR,F,FT2)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.8605E+07</td>
<td>0.6713E+05</td>
<td>0.2369E+03</td>
<td>0.8000E+01</td>
<td>0.2414E+01</td>
<td>0.2042E+03</td>
<td>0.1325E+03</td>
</tr>
</tbody>
</table>

TOTAL DRIVING FORCE (PSI)

0.2517E+01
### TOTAL CONDENSER DESIGN

#### (A) VERTICAL CONDENSER

#### (B) MECHANICAL DESCRIPTION

<table>
<thead>
<tr>
<th>NUMBER OF SHELL PASS</th>
<th>SHELL TUBES</th>
<th>NUMBER OF PASS TUBES</th>
<th>SHELL DIAMETER (IN)</th>
<th>TUBE OUTSIDE DIAMETER (IN)</th>
<th>TUBE PITCH (IN)</th>
<th>TUBE LENGTH (FT)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>1</td>
<td>4</td>
<td>640</td>
<td>0.290E+02</td>
<td>0.750E+00</td>
<td>16</td>
</tr>
</tbody>
</table>

#### (C) PROCESS DESCRIPTION

<table>
<thead>
<tr>
<th>REQUIRED COOLANT, LB/HR</th>
<th>COOLANT IN TEMP (°F)</th>
<th>COOLANT OUT TEMP (°F)</th>
<th>TUBE PRESSURE DROP (PSI)</th>
<th>VELLOCITY IN TUBE (FT/SEC)</th>
<th>PROCESS IN TEMP (°F)</th>
<th>PROCESS OUT TEMP (°F)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.3472E+06</td>
<td>0.7200E+02</td>
<td>0.9200E+02</td>
<td>0.8105E+01</td>
<td>0.4625E+01</td>
<td>0.1052E+01</td>
<td>0.9941E+02</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>CLEAN H.T.C. (BTU/HR, °F, FT²)</th>
<th>DIRTY H.T.C. (BTU/HR, °F, FT²)</th>
<th>TEMPERATURE LMTD (°F)</th>
<th>CONDENSER HPAT DUTY (BTU/HR)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.6163E+03</td>
<td>0.2074E+03</td>
<td>0.1833E+02</td>
<td>0.6939E+07</td>
</tr>
</tbody>
</table>
CHIEN WANG was born on August 2, 1954 in Taichung, Taiwan, Republic of China and received his early education in Taichung. He graduated from Taichung 1st Senior High School in 1972.

His undergraduate work was at National Taiwan University, graduating with a Bachelor of Science degree in Chemical Engineering in June 1976. In December 1978, he received the degree of master of Science in Chemical Engineering at Louisiana State University. Also, he received the degree of Master of Science in System Science at LSU in May 1980.

He is a member of Phi-Kappa-Phi Honor Society and Phi-Lamda-Phi Honor Society.

He is married to the former Miss CHIH-CHUN LIU in July, 1977.

At present, he is a candidate for the degree of Doctor of Philosophy in Chemical Engineering.
Candidate: Chien Wang
Major Field: Chemical Engineering
Title of Thesis: Frequency Response Analysis of Distillation Column Interaction

Approved:

[Signatures]

Dean of the Graduate School

EXAMINING COMMITTEE:

[Signatures]

Date of Examination:

July 10, 1981