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DETERMINATION OF THE OPTIMUM FEED PLATE
LOCATION IN MULTICOMPONENT
DISTILLATION COLUMNS

By

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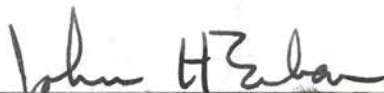
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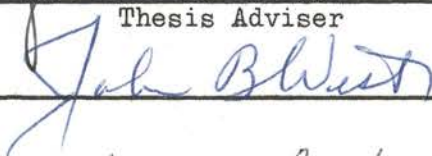
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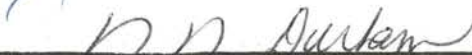
DETERMINATION OF THE OPTIMUM FEED PLATE
LOCATION IN MULTICOMPONENT
DISTILLATION COLUMNS

Thesis Approved:



Thesis Adviser





Dean of the Graduate College

PREFACE

Several methods are available in the literature for locating feed plates in a distillation column operating on a multicomponent mixture. A computer study employing tray-by-tray calculations is used to determine the tray compositions within several distillation columns of varying feed components and product specifications. Using the output data for several different problems, the optimum feed plate is located for each problem by several different criteria and a comparison is made between the different criteria as to reliability. Various reflux ratios are studied for each column configuration and several feed conditions are studied to determine the combined effects of these two parameters on the optimum feed plate location.

I wish to express my deepest gratitude to my thesis adviser, Dr. J. H. Erbar, for his invaluable guidance in keeping me on the objectives of my problem and off the paths of confusion. I would especially like to thank those responsible for the Continental Oil Fellowship which provided funds to make my research possible. I would like to give a special acknowledgement to Mr. Eston W. McGee II for his invaluable drafting talents. I appreciate the concerned attitudes of my parents and parents-in-law. Finally, I would like to thank my wife, Lindsey, for her continued interest in my work and her understanding attitude in my moments of frustration.

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LIST OF SYMBOLS

Major Symbols

English Letters

| | |
|----------------|---|
| a | - mole ratio of liquid to vapor of passing streams in the rectifying section. |
| a' | - mole ratio of liquid to vapor of passing streams in the stripping section. |
| F | - moles of feed/unit time. |
| P | - total pressure of the system. |
| P _o | - vapor pressure of a component at the temperature of the system. |
| r | - key component ratio in the liquid phase. |
| R | - separation ratio for the key components. |
| R' | - mole rate of reflux to the rectifying section. |
| S | - number of stripping trays. |
| V | - moles of vapor/unit time. |
| W | - moles of bottom product/unit time. |
| x | - mole fraction in the liquid phase. |
| X | - moles of liquid/unit time. |
| y | - mole fraction in the vapor phase. |
| Y | - moles of vapor/unit time. |
| z | - feed mole fraction. |

Greek Letters

- α - relative volatility of the key components.
 θ - convergence parameter of the Theta method.

Subscripts

- b - bottom product
d - distillate product
f - feed plate
f+1 - plate above feed plate
f+2 - second plate above feed plate
F - feed liquid
hk - heavy key component
i - a general component in the vapor or liquid;
the intersection of the rectifying and stripping
section operating lines
j - a general plate in the column
lk - light key component
m - the plate just below the feed plate or the
bottom plate in the column in equation [20]
M - minimum reflux
n - a general plate in the rectifying section or the
top plate in the Fenske equations
r - rectifying section
s - stripping section
T - total reflux
w - bottom product
1 - the reboiler

CHAPTER I

INTRODUCTION

In the design of distillation equipment for continuous separation of a complex multicomponent feed mixture, the design engineer is directly confronted with several major problems. Two of these problems of equal complexity are the determination of the number of trays and reflux rate required to separate the feed into the desired products. Another very important problem associated with designing distillation equipment is determining the feed tray location to yield the best separation for a particular column. The location of the optimum feed plate or feed tray in the distillation column is the concern of this study.

The optimum feed plate location can in general be characterized two ways. For a specific separation of key components of a mixture, it is the feed tray in the column which will give the fewest total trays required for that separation. Alternately, for a column with a specific ~~number of trays it is the tray which yields the maximum separation of~~ key components of the mixture, thus the highest product purity. The objective of this work is to (1) evaluate different criteria of the optimum feed plate location; (2) determine which of the literature methods of locating the optimum feed plate are most reliable and (3) develop a set of equations for the optimum feed plate location.

CHAPTER II

LITERATURE REVIEW

The literature on feed plate location in a multicomponent distillation column dates back considerably in time. The problem was considered by Fenske (1) in 1932. His approach studied two ideal extreme cases of operation and the case of actual operation. The two extreme cases of operation, at minimum and total reflux, are derived for a multicomponent column where the mixture of components is treated as a two component mixture of light and heavy key components. Lewis and co-workers were the originators of the concept of key components (2, 3). The terms light key and heavy key components need to be defined to enhance an understanding of the optimization methods employed in this work and in distillation in general. The key components are the two components in the feed mixture where the "split" is to take place. ~~The light key component is the lower boiling or more volatile component of the two key components. It vaporizes more readily at a given temperature. The light key component will tend toward the top of the distillation column as a vapor in a richer or higher concentration than the heavy key component.~~ Conversely, the heavy key component will be more concentrated in the liquid at the bottom of the column. Similarly, components lighter and heavier than the keys will tend toward the distillate and bottom products respectively. The so called "split" is a

measure of how effectively the key components have been separated. If, for example, there is no light key component in the bottom product and no heavy key component in the distillate then a perfect split exists. Generally a small amount of light key exists in the bottom product and some heavy key in the distillate. The greater the proportion of each key component in its respective product stream, relative to the total amount of each key in the feed, the better the separation.

Fenske began his approach with the simplest known relation for vapor-liquid equilibrium, that of Raoult's law, which relates the compositions between vapor and liquid of a single component in a mixture as:

$$y_i^P = x_i^P \frac{P_{oi}}{P} \quad [1]$$

for a general component, i , where P is the total pressure on the system and P_{oi} is the vapor pressure of component i at the system temperature. The fractions of the component i in the liquid and vapor phases are x_i and y_i respectively. Denoting the light and heavy keys with subscripts lk and hk respectively, the equations for the key components are:

$$y_{lk}^P = x_{lk}^P \frac{P_{olk}}{P} \quad [2]$$

$$y_{hk}^P = x_{hk}^P \frac{P_{phk}}{P} \quad [3]$$

If we divide equation [2] by equation [3], the result is the following:

$$\frac{y_{lk}}{y_{hk}} = \alpha \frac{x_{lk}}{x_{hk}} \quad [4]$$

where α is called the relative volatility and is defined as the ratio of vapor pressures of the key components. The general relative volatility need not be between key components, but rather may be a ratio of vapor pressure of any component of the mixture to that of the heavy

key component. In general, the lighter the component, the higher its relative volatility as compared to a heavier component. This is because a light component has a higher vapor pressure than any component heavier than itself at a given temperature (For a well behaved homologous series). Equation [4] gives a relationship for the ratio of the light key to heavy key components between the vapor and liquid phases respectively.

A general material balance around a plate above the feed in a distillation column is:

$$Y_{ni} = X_{n+1i} + X_{di} \quad [5]$$

where n denotes any plate between the feed and distillate, (counted from the bottom up), n+1 denotes the plate just above plate n and d denotes the distillate. Equation [5] applies to all components in the distillation column between the feed and distillate; that is, in the rectification section. At total reflux, no distillate product or bottoms product is removed and therefore X_{di} is zero. (X and Y are mole rates in the liquid and vapor streams respectively, and not the mole fractions x or y).

At total reflux, the material balance equations become:

$$\frac{Y_{nlk}}{Y_{nhk}} = \frac{X_{n+1lk}}{X_{n+1hk}} \quad [6]$$

Substitution of equation [6] into equation [4] containing the relative volatility gives:

$$\frac{X_{n+1lk}}{X_{n+1hk}} = \alpha \frac{X_{nlk}}{X_{nhk}} \quad [7]$$

This is the well known Fenske equation, so named after its author,

which describes the ratio of the key components on one tray in the liquid phase in terms of the ratio of the keys on an adjacent tray in the liquid phase.

The obvious inherent assumption of the Fenske equation is that of total reflux operation. While this mode of operation is unrealistic and totally impractical, equation [7] is a starting point in the design of distillation equipment. When applied between the distillate and the feed plate the following equation results:

$$\frac{x_{nlk}}{x_{nhk}} = \alpha^{n-f} \frac{x_{flk}}{x_{fhk}} \quad [8]$$

where n is the top plate in the column counted from the bottom, f is the feed plate, and $n-f$ is the number of plates in the rectifying section of a distillation column for a given feed and distillate product; that is, a given split at total reflux. Similarly the equation is applicable between the distillate and bottom products to determine the total number of theoretical distillation plates required to effect a particular split.

Thus, it is seen that for total reflux a single optimum feed plate exists for a given key component split. This optimum lies on the plate f , counted from the bottom, such that when the Fenske equation is applied between feed and distillate there are $n-f$ plates above plate f and n total plates in the column. From the definition of mole fractions of a component substitution into equation [8] will obtain the general Fenske equation as:

$$\frac{x_{nlk}}{x_{nhk}} = \alpha^{n-f} \frac{x_{flk}}{x_{fhk}} \quad [9]$$

which expresses ratios of mole fractions of the keys instead of moles.

At this point another inherent assumption of the Fenske equation becomes apparent; that of constant relative volatility of the key components throughout the distillation column. From equation [1] another definition of relative volatility of the key components is:

$$\alpha = \frac{y_{jlk}}{x_{jlk}} \cdot \frac{x_{jhk}}{y_{jhk}} \quad [10]$$

where j is any general plate in the column. It is known that the concentrations of light and heavy key components are changing throughout the liquid and vapor in the column. Noting equation [10] while thinking of the physical aspects of distillation helps to see how the relative volatility of the keys can be approximately constant. Going further up the column, y and x of the light key component increase, while y and x of the heavy key component decrease. If the relative increase in the y/x ratio for the light key component offsets the relative decrease in the x/y ratio for the heavy key component then α is constant.

A similar argument holds for relative volatility of the key components while going down the column. In many instances relative volatility of the key components varies only to a small degree. Using these equations established for the three possible operating conditions of a multicomponent distillation column enables the feed plate to be located, such that the given separation can occur. However, there is no way of determining whether the feed has been located at the optimum tray from the equations as presented by Fenske. Another feed plate could give a better separation of the key components in the mixture, or an equivalent separation with a fewer number of plates in the column. Also, some of Fenske's simplifying assumptions are rather harsh in many operating columns today. One of his assumptions for a multicomponent

column is that of applicability of Raoult's law in determining the equilibrium between the liquid and vapor in the column. Complex systems which exhibit large deviations from Raoult's law equilibrium data abound in the chemical processing industry (4). The famous Fenske equation is the classic beginning of improved methods of designing multicomponent distillation equipment.

An approach by Gilliland (5) to the problem of feed plate location uses the concept of light and heavy key components, and uses the ratio of the light to heavy key components in various parts of the distillation column. His approach uses the concept that the optimum composition for the feed plate should be such that a maximum increase in the ratio of the light to heavy key components is achieved over the entire column, (i.e., per tray), hence the fewest number of trays are required for the separation of the components. Three general cases are solved by Gilliland; (1) the feed may enter as a bubble point liquid, (2) part liquid and part vapor and (3) as a superheated vapor. (The saturated vapor feed is solved analogously to the case of partial vapor feed). Gilliland derives a simple test criterion for determination of the optimum feed plate location for several different feed conditions in multicomponent distillation. All of the criteria are based on the key component ratio defined by equation [23], (Chapter III). For a bubble point feed the criterion for the optimum feed plate is:

$$r_f \leq r_i \leq r_{f+1} \quad [11]$$

where r_i is given from the following equation:

$$r_i = \frac{Fz_{Flk} + \left(\frac{V_n}{V_m} - 1\right) \cdot Wx_{wlk}}{Fz_{Fhk} + \left(\frac{V_n}{V_m} - 1\right) \cdot Wx_{whk}} \quad [12]$$

The quantity r_i is the key component intersection ratio of the operating lines, which should lie between the key component ratio in the liquid phase on the feed plate and the plate just above the feed plate if the feed plate is optimized. Otherwise $r_i \leq r_f$ or $r_i \geq r_{f+1}$ and the feed plate is mislocated. The quantities V_m and V_n in equation [12] are the vapor rates to and from the feed plate respectively. For the case of a partially vaporized feed Gilliland derives the following equation:

$$r_f \leq r_F \leq r_{f+1} \quad [13]$$

The subscript F refers to the feed. Thus the criteria for the optimized feed plate for a partial vapor feed shows that the key component ratio for the liquid portion of the feed should lie between the key component ratio on the feed plate and on the plate just above the feed plate. If the feed plate is mislocated then $r_F \leq r_f$ or $r_F \geq r_{f+1}$. For the case of a saturated vapor feed equation [13] is used with r_F calculated as the liquid that would be in equilibrium with the feed vapor. For the case of a superheated feed, Gilliland derives the equation:

$$r_{f+1} \leq r_i \leq r_{f+2} \quad [14]$$

where r_i is as expressed in equation [12]. All of Gilliland's equations have two inherent assumptions, that of constant molal overflow and that the column operates with theoretical plates. In many cases the constant molal overflow assumption may be too stringent to give a

reliable solution in the determination of the optimum feed plate. In all cases, the ratio of light to heavy key components in the liquid or vapor must be available for the trays directly adjacent to the feed tray or as high as two trays above it for the superheated vapor case. This information is not directly available from the correlations of Gilliland, but must be obtained from a tray-by-tray analysis of the column or other means of obtaining the liquid and vapor profiles directly.

In a method by Scheibel and Montross (6) the light to heavy key ratio on the feed plate is calculated from an empirical equation. It is necessary to know the minimum reflux ratio for the required separation to use the method. However, the case of the minimum reflux ratio can generally be calculated without difficulty. A linear correlation has been determined for calculating the key ratio on the feed plate as:

$$r_f = r_{fT} + \frac{R' M}{R'} (r_{fM} - r_{fT}) \quad [15]$$

where the subscript T refers to total reflux, M refers to minimum reflux and f refers to the feed plate. R' is the operating mole rate of reflux and not to be confused with the separation ratio, R . The key ratio at total reflux on the feed tray has been correlated as a function of the key ratio in the feed and the Murphree tray efficiency of the feed plate. The difficulty with equation [15] lies in the complex equation to calculate r_{fM} , the key ratio on the feed plate at minimum reflux (6). After r_f has been calculated by equation [15], Gilliland's criteria may be used to check for the optimum location of the feed plate. The "exact" optimum feed plate is determined by numerically differentiating the separation ratio with respect to feed tray number

or stripping trays and finding the point corresponding to the fewest total trays required. The most tedious part of the method is in the calculation of the ratio of the key components on the optimum feed tray corresponding to minimum reflux. A very complicated empirical correlation is presented which involves accounting for all components lighter and heavier than the keys, fractional vaporization of the feed, and a pseudo-minimum ratio of the keys in the feed tray liquid from the binary diagram of the key components. In general, this approach is extremely difficult to apply because of the complexity of determining the ratio of the key components on the feed tray at minimum reflux. Very good agreement within the ability to determine the feed tray composition by other methods is claimed.

Floyd and Hipkin (7) have studied in a semiquantitative manner the effects of feed plate location. Tray-by-tray analysis of a depropanizer is performed using the Thiele-Geddes method. The analysis of the optimum feed tray is presented in graphical form for four different depropanizers. The criteria used in the optimization of the feed tray are those of Gilliland (5) and Maxwell (8). Equivalent optimized feed trays are given by either approach for the different columns. A semiquantitative index of the overall degree of fractionation is presented in a plot of the classical Fenske equation to determine the optimum feed plate location. There is another design problem in locating the optimum feed plate in a real distillation column. Floyd and Hipkin discuss three alternatives in transferring the optimum from theoretical trays to actual trays in the column (7). They point out that it is possible in a theoretical sense to determine the optimal feed plate location for theoretical trays, however, insufficient work has been done on plate

efficiencies or column efficiencies to accurately prescribe the optimum feed plate location in terms of actual trays. The standard practice to date is to assume that the ratio of rectifying to stripping trays is the same for actual trays as for theoretical trays in the column. This implies an assumption of constant tray efficiency throughout the column, which is most likely invalid. Tray efficiencies are, however, temperature dependent since they rely on ratios of equilibrium and actual tray compositions. An alternate approach in the determination of the actual optimum feed plate location is suggested. An overall column efficiency is calculated and applied to the total theoretical trays. The rectifying and stripping proportions in terms of theoretical plates are then applied to determine the actual optimum location. This approach assumes a constant tray efficiency throughout the column which is seriously doubtful. The first serious suggestion of Floyd and Hipkin is to abandon the equilibrium stage approach altogether and shift to a turbulent gas-liquid mass transfer approach. Insufficient work has been done to date to make this a worthwhile design method. The thought seems reasonable for future considerations. The second suggestion is to use individual tray efficiencies at each stage in the tray-by-tray calculations to predict automatically the column performance in actual trays rather than theoretical trays. This method would be reasonable except that correlations to predict individual tray efficiencies would need parameters for system properties and tray geometry. This suggestion seems promising. Finally a sectional efficiency approach is suggested, eliminating the constant tray efficiency assumption across the feed plate. If sectional efficiencies can be calculated based on actual to theoretical trays in each section of the column, then the

optimum feed plate can be accurately prescribed by transferring from theoretical to actual trays in individual sections of the distillation column.

In a computer approach to solving for the optimal feed plate location in a distillation column, Waterman, Frazier and Brown, (9), describe a computer solution based on ratios of the key components in the feed and on the feed plate and the plate just above the feed plate. The basis for characterizing the optimum feed plate is the assumption that for multicomponent systems, the ratio of the key components in the feed vapor, (or liquid), may be considered equal to the key component ratio in the vapor, (or liquid), leaving the feed tray. This criterion represents the optimum when the relative volatility of the key components is closer to unity than the relative volatility of the light key and next lighter component or the relative volatility of the heavy key component and the next heavier component. Graphical plots of the key component ratio as a function of plate number in the column are presented to show how the optimum feed plate is located on the basis of key component ratio. When the relative volatility of the key components is not as close to unity as that of the light key and lighter component, or the heavy key and heavier component, the key ratio on the optimum feed plate can be significantly displaced from the key ratio of the feed. A solution still exists, but the key ratio on the optimum feed plate will not be the same as that of the feed. A reliable tray-by-tray method is required for calculating the key component ratios for various feed plate locations.

Optimizing the feed plate location while simultaneously reducing the amount of computer time required to solve a problem is the approach

of Hengstebeck (10). Using rigorous complex tray-by-tray routines requires considerable computer time in that several calculations are required for each feed plate location. Also, several reflux ratios must be studied to characterize a given separation desired. Hengstebeck uses the theory of Fenske (1), which says that the logarithm of the molar concentration ratio of the key components is a linear function of the distillation stage number, provided the relative volatility of the given components is constant. At total reflux this theory is exact. At operating reflux ratios this approximation is good near the terminal ends of the column but becomes asymptotic near the central part of the column, where the flow rates become "pinched in" or nearly constant. According to Hengstebeck, the slope of the line in the stripping section or rectifying section is an indicator of the relative degree of fractionation occurring at some particular section of the column. Therefore, the appropriate optimum feed point would be where the slopes of both sections of the relative fractionation plot become equal. In a graphical sense, Hengstebeck is using the criteria of Gilliland expressed by equation [13]; that is, the key ratio in the liquid on the feed plate should be less than that in the feed, which should be less than on the plate above the feed. Hengstebeck claims the restrictions of the method are those of the Fenske equation, i.e., total reflux and constant relative volatility. However, the method may apply for circumstances which violate these assumptions. For the case of a real operating column, these restrictions may be too severe for the relative fractionation plot. A four step method to give a better approximation for the optimum feed plate is suggested. This approximation is then used in another tray-by-tray run to determine the optimum feed plate.

The method is based on the relative fractionation plot.

An approximate method for finding the concentrations of a multi-component mixture on the feed tray in a distillation column is given by Serov and Zykov (11). In the method employed, no trial and error calculations are required. The more common plate-to-plate type of calculations are avoided. The method is based on approximate relations for the feed plate concentrations which become exact when constant molal overflow exists within the column. This is an assumption of tray-by-tray calculations also, however, it is not a usual assumption. In general for systems involving a wide boiling range feed, this assumption of constant molal overflow is not used. The equations are modified accordingly by non-dimensionalization of the feed plate concentrations and using constants dependent on the rate of reflux and degree of stripping. These constants are dependent on the initial conditions in the column. Once the constants for the concentration relations are determined the feed plate concentrations are easily found. Then one can perform a test using the feed plate optimization criterion of say Gilliland or Maxwell (5,8) to decide if the feed plate has been optimally located. In effect, the method is an application of vapor-liquid equilibrium relations and material balances combined in a single equation to determine the feed plate concentrations. The method is compared with a test case to that of Robinson and Gilliland (12) and compares on 10 out of 15 components of a feed mixture within 10% relative error on the feed plate composition. However, the maximum relative error on one component of the mixture is 41.5%. Therefore the method is as the author prescribes, approximate.

In a paper by Nishino (13) a method which modifies the Thiele-

Geddes tray-by-tray calculation is proposed. A drawback of the traditional type of Thiele-Geddes method is the necessity that the feed plate must be located initially to calculate the terminal column compositions of the components. Thus a trial and error procedure is required to find the feed plate which gives a maximum separation of the key components. In this procedure, a ratio of the light and heavy key component mole fraction in the products from a given section of the column is calculated. The product of these ratios for the rectifying and stripping section is then calculated. This product is useful in the proper location of the feed plate. In the converged Thiele-Geddes calculations, the two feed plate compositions "mesh" so that this product is simply the ratio of the distillate to bottoms mole fractions of the particular component. For the light and heavy key components, the ratio of this quantity should be a maximum for maximum component separation. It is called the separation ratio of the light and heavy key components. Thus the differential of this ratio with respect to the number of stripping trays is equal to zero. This is the key to the determination of the optimum feed plate as applied to the Thiele-Geddes method. Writing the separation ratio for the keys for a column with a total condenser:

$$R = \frac{r_n}{r_1} \quad [16]$$

The quantities r_n and r_1 are given by:

$$r_n = \frac{x_{dlk}}{x_{frlk}} \cdot \frac{x_{fslk}}{x_{dhk}} \quad [17]$$

$$r_1 = \frac{x_{blk}}{x_{frhk}} \cdot \frac{x_{fshk}}{x_{bhk}} \quad [18]$$

where the subscript f refers to the feed plate, s refers to the stripping section and r refers to the rectification section. If R, the separation ratio is plotted as a function of feed plate number, it has a maximum value at the optimum feed plate. Therefore, the derivative of the equation which defines the separation ratio is zero at this maximum. Writing the derivative with respect to number of stripping trays in the stripping section:

$$\frac{dR}{dS} = \frac{d}{dS} \left(\frac{x_{dlk} x_{fslk}}{x_{dhk} x_{frlk}} \cdot \frac{x_{bhk} x_{frhk}}{x_{blk} x_{fshk}} \right) = 0 \quad [19]$$

Now as $\theta \rightarrow 1$. in the Thiele-Geddes method, i.e., as the composition profiles converge to the correct values for all components throughout the column, $x_{fslk} = x_{frlk}$ and $x_{fshk} = x_{frhk}$. (The compositions of the key components exactly mesh at the feed plate). Thus, as the Thiele-Geddes method converges, the differential equation for the separation ratio becomes:

$$\frac{dR}{dS} = \frac{d}{dS} \left(\frac{x_{dlk}}{x_{dhk}} \cdot \frac{x_{bhk}}{x_{blk}} \right) = 0 \quad [20]$$

which is equivalent to equation [25], (Chapter III), after integration. Application of the Thiele-Geddes method to this theory supplies the component compositions in the rectifying and stripping sections of the column into the differential equation of the separation ratio with respect to stripping trays and solving for the optimum feed point. The term "point" is used instead of plate implying the possibility of fractional trays. The optimum feed point is guaranteed by the nature of the differential equation. A maximum must exist in the separation ratio for some particular feed plate in the column, hence the differential equation must have an optimal solution. Given a particular

separation for a given column, with distillate rate specified, a procedure is presented for determination of the optimum feed point. Complex equations involving eigenfunctions are substituted directly into equation [19] and the number of stripping trays is calculated. The method is trial and error and only gives the optimum feed plate when Theta has converged to unity. Using the equations of Nishino on the computer is the only reasonable way to arrive at a reliable solution. The terminal compositions are then obtained. Numerical examples are given, and compared to other solutions by other methods. Close agreement is obtained with the solutions presented in other works.

According to the methods of Maxwell (8), the feed plate location in a column of actual, rather than theoretical plates, is determined by the fraction of "total reflux" plates between the feed and distillate which are required to effect the separation. This fraction of the total theoretical plates in the rectifying section of the column at total reflux is applied to the number of actual plates to determine the number of actual plates above the feed. The Fenske equation (1) is applied between the feed and distillate to obtain the tray requirements in the rectifying section. To account for differences in relative volatility of the key components above and below the feed plate a geometric mean relative volatility is based on the relative volatility of the keys at the dew point of the distillate and the bubble point of the bottoms. For fractional vaporization of the feed, the fraction of rectifying stages is accounted for in the real column by adding the fractional vaporization to the number of total reflux stages in the rectifying section. This is done because 100% vaporization corresponds to a theoretical stage at total reflux. The method insures that the

desired separation can take place in the real column provided sufficient reflux has been accounted for in the reflux calculation, which is separate from the feed plate calculation. However, the method has no assurance that the optimal feed plate has been located in the real column. That is, even though the desired separation can take place, another feed plate could conceivably give an even better separation. An inherent assumption of the method is that of equal tray efficiencies in the real column. This is probably invalid as concluded by Floyd and Hipkin (7).

CHAPTER III

THEORETICAL DISCUSSION

At this point an introduction to the basis for an optimization of the feed plate can be presented. If the relative volatility of the keys can vary throughout the column, then somewhere in the column it will have a maximum value. Equation [7] can be rewritten across the entire column between the distillate and bottom product as:

$$\frac{x_{nlk}}{x_{nhk}} = \alpha^{n-1} \frac{x_{1lk}}{x_{1hk}} \quad [21]$$

where subscript 1 refers to the reboiler, which acts as a theoretical plate, and n refers to the top plate in the column. If a partial condenser is used, this equation would be written as:

$$\frac{x_{dlk}}{x_{dhk}} = \alpha^n \frac{x_{1lk}}{x_{1hk}} \quad [22]$$

Noting the form of equations [21] and [22], if relative volatility takes on a maximum somewhere in the column, then α^{n-1} , (or α^n), will also be at a maximum. If the key component ratio in the liquid on a plate, j, is defined as:

$$r_j = \frac{x_{jlk}}{x_{jhk}} \quad [23]$$

then equations [21] and [22] can be written in a form suitable for determination of the optimal plate to feed on:

$$R = \frac{r_n}{r_1} = \alpha^{n-1} \quad [24]$$

$$R = \frac{r_d}{r_1} = \alpha^n \quad [25]$$

Equation [24] is for a column with a total condenser and equation [25] is for a column with a partial condenser. R is a quantity called the separation ratio for the key components. R is computed by dividing the terminal distillate or top key component ratio by the terminal bottom product key component ratio. (r_1 , r_n , and r_d are all for the liquid phase). In general, the higher the separation ratio the greater the split between the key components. The optimal feed tray is that feed tray which gives the highest separation ratio for the keys. The separation ratio for the key components accounts simultaneously for a material balance and equilibrium relationships over the column.

Fenske has derived equations for actual operating conditions where products are removed from the column (1). For the interval between the feed plate and the condenser, (for a total condenser), the following equation is obtained:

$$\frac{x_{fhk}}{x_{flk}} = (a\alpha)^{n-1} \frac{x_{dhk}}{x_{dlk}} \frac{(\alpha-1)}{(a\alpha-1)} - \frac{(1-a)\alpha}{(a\alpha-1)} \frac{x_{dhk}}{x_{dlk}} \quad [26]$$

where: $a = \frac{X_{n+1k}}{Y_{nlk}} \quad [27]$

The quantity a , is the ratio of moles between passing streams of vapor and liquid in the rectifying section of the column and is assumed to be constant. Rewriting equation [26] in terms of the terminal key component ratio, r_d , the following equation results:

$$r_d = \left[(a\alpha)^{n-1} \frac{(\alpha-1)}{(a\alpha-1)} - \frac{(1-a)\alpha}{(a\alpha-1)} \right] \frac{x_{flk}}{x_{fhk}} \quad [28]$$

where a is as defined in equation [27]. A similar equation for the

key component ratio in the reboiler is the following:

$$r_1 = \left[(a'\alpha)^{n-1} \frac{(\alpha-1)}{(a'\alpha-1)} - \frac{(1-a')\alpha}{(a'\alpha-1)} \right] \frac{x_{f1k}}{x_{fhk}} \quad [29]$$

where $a' = \frac{X_{m+1k}}{Y_{mlk}} \quad [30]$

Using equation [25] which defines the separation ratio for the key components as a ratio of the terminal key component ratios, the separation ratio for the actual operating conditions can be written as:

$$R = \frac{(a\alpha)^{n-1} \frac{(\alpha-1)}{(a\alpha-1)} - \frac{(1-a)\alpha}{(a\alpha-1)}}{(a'\alpha)^{n-1} \frac{(\alpha-1)}{(a'\alpha-1)} - \frac{(1-a')\alpha}{(a'\alpha-1)}} \quad [31]$$

where a and a' are defined in equations [27] and [30]. Equation [31] is based on equilibrium assuming Raoult's law and on material balances above and below the feed plate. For the case of a partial condenser on the column, the exponent on α should be n rather than $n-1$, since the partial condenser acts as another theoretical stage. The maximum value of the separation ratio given by equation [31] is obtained for the optimum feed plate as previously noted in the discussion for the basic Fenske equation. The inherent assumptions of equation [31] are that of constant relative volatility of the keys and a constant component mole ratio between passing streams throughout the column.

CHAPTER IV

RESULTS

The relative fractionation plot of Hengstebeck, (10) is a good indicator of the degree of separation taking place within a multicomponent distillation column. At total reflux, the plot is a straight line between the key component ratio of the distillate and the key component ratio of the bottom product. For real finite reflux rates considerable curvature occurs near the mid-region of the column. The asymptotic effects of the rectifying and stripping sections can be seen from Figure 1. (Figures 1 through 5 are in this chapter, while Figures 6 through 24 are in Appendix A). The dip associated with the rectifying section is caused in general by the sudden increase in key component ratio on the feed plate. For the theoretical total reflux column this dip does not occur because no feed is introduced on any plate. The total reflux relative fractionation plot is always monotonic. That is, no change in the sign of the slope occurs. Due to physical aspects of a real feed, there is nearly always a dip associated in the rectifying section of the column. When no dip occurs the feed has been located such that its key component ratio is less than the plate below and greater than the plate above and a smooth curve results. This is the criterion of Gilliland expressed by equation [13], which only holds for a partially vaporized feed with constant molal overflow. Few columns operate under the constant molal overflow assumption and later

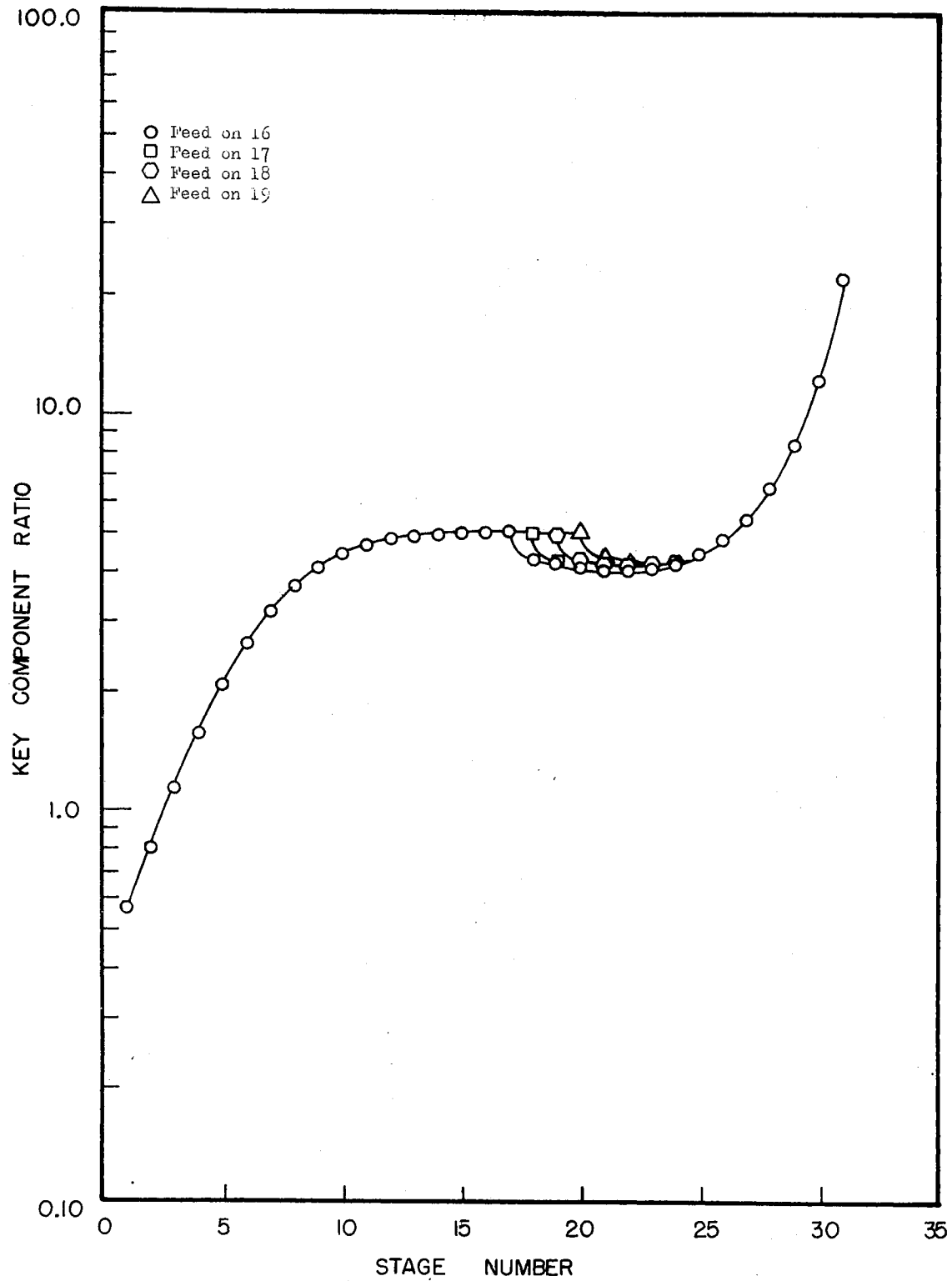


Figure 1. Relative Fractionation Curve- 30 Tray Depropanizer
With Bubble Point Feed, $V/F = 1.12$, $D/F = 0.4854$

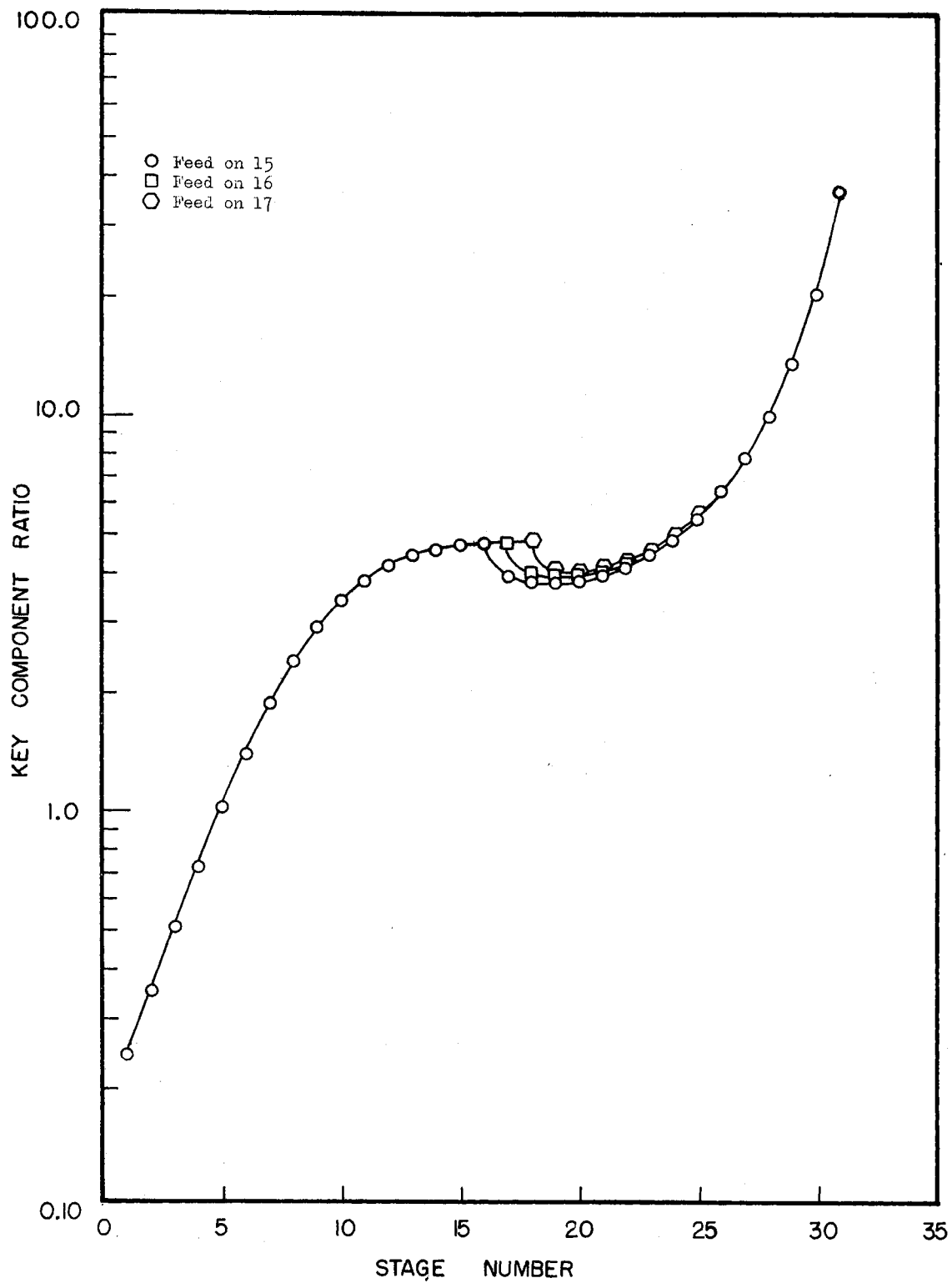


Figure 2. Relative Fractionation Curve- 30 Tray Depropanizer
With Bubble Point Feed, $V/F = 1.26$, $D/F = 0.4854$

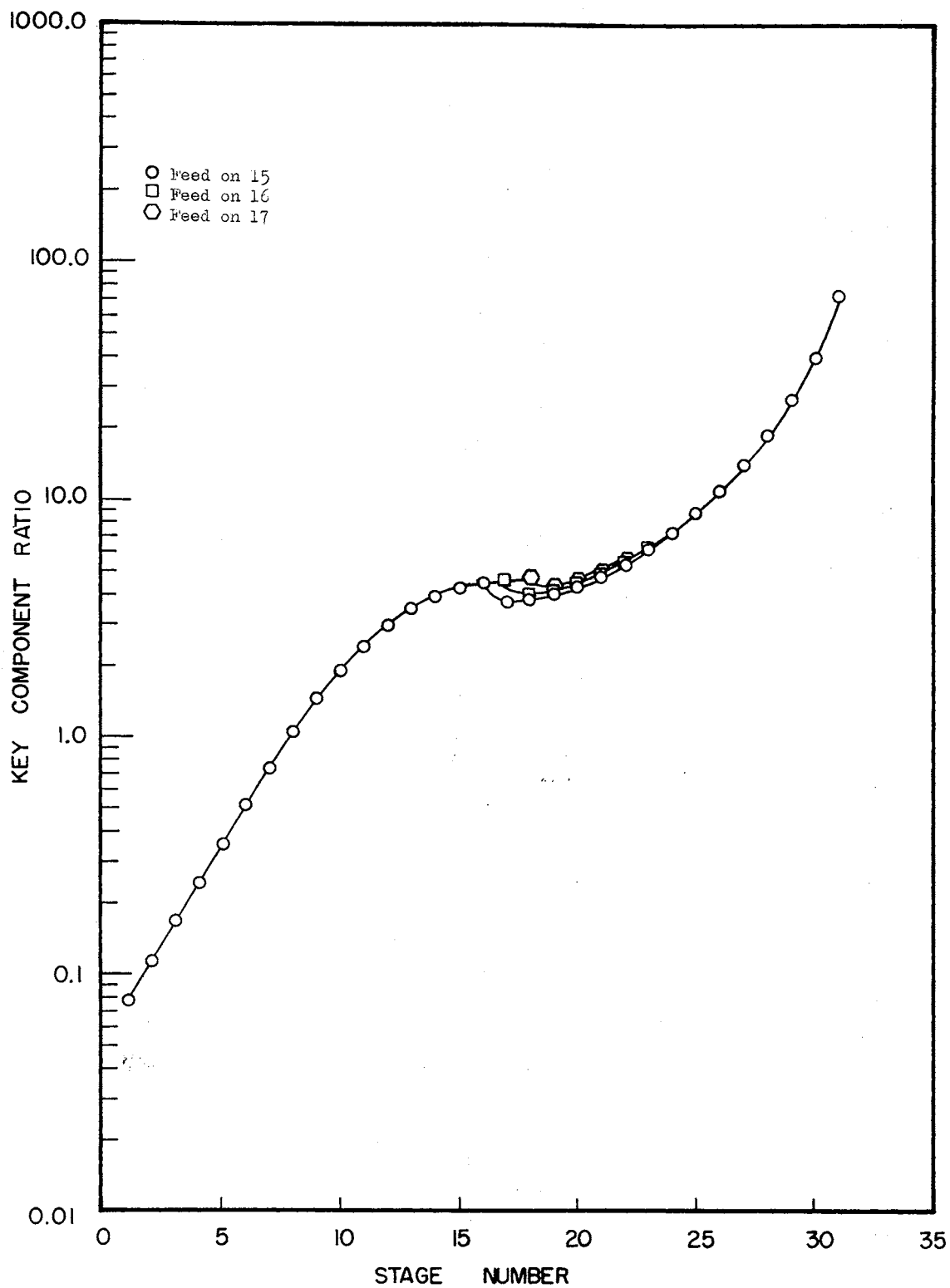


Figure 3. Relative Fractionation Curve- 30 Tray Depropanizer
With Bubble Point Feed, $V/F = 1.40$, $D/F = 0.4854$

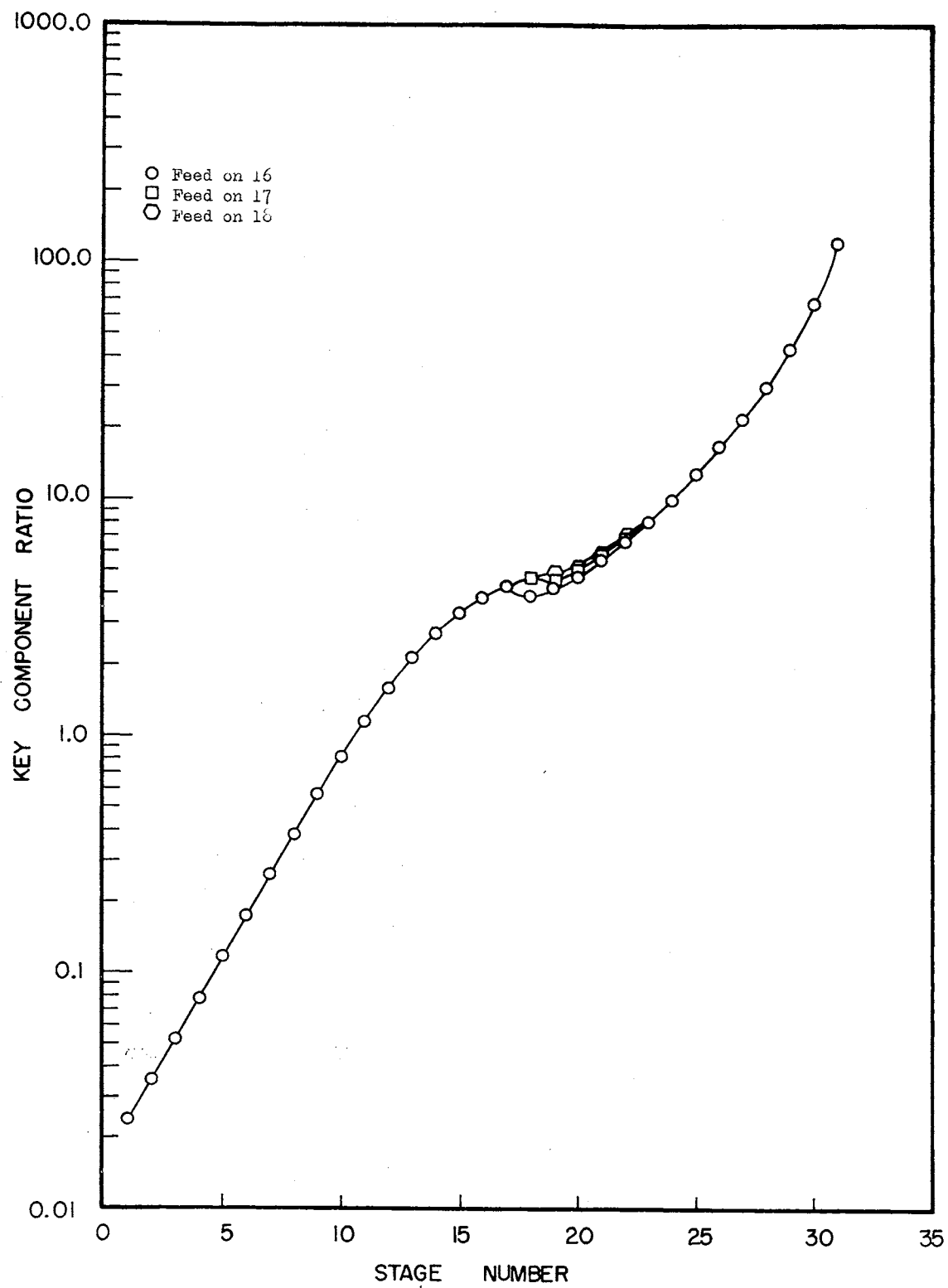


Figure 4. Relative Fractionation Curve- 30 Tray Depropanizer
With Bubble Point Feed, $V/F = 1.54$, $D/F = 0.4854$

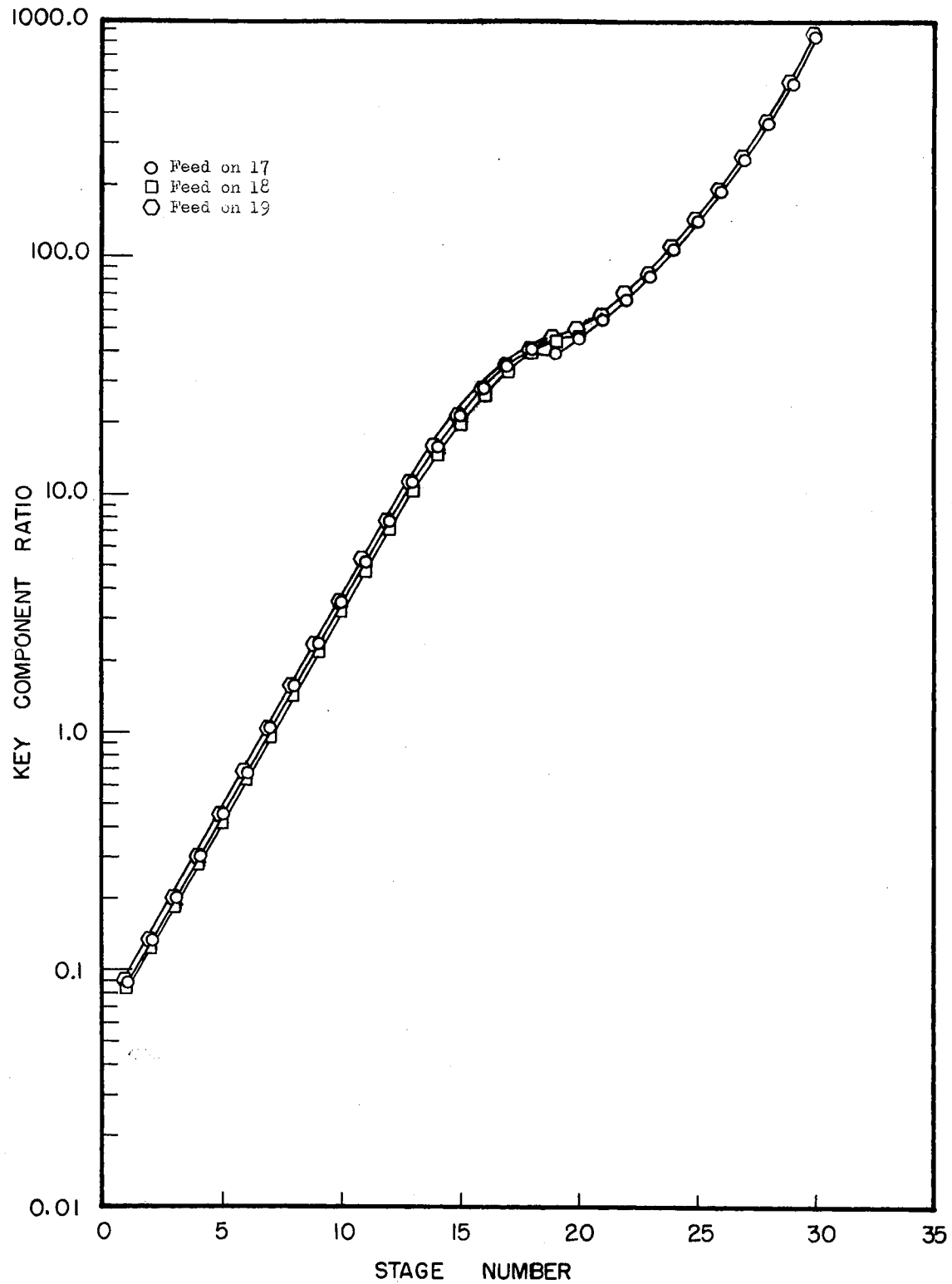


Figure 5. Relative Fractionation Curve- 30 Tray Depropanizer
With Bubble Point Feed, $V/F = 1.68$, $D/F = 0.4854$

Gilliland's criterion will be proved inadequate for the cases tested.

On the relative fractionation plot of Hengstebeck for a depropanizer, (Figure 1), the asymptotic point on the stripping section curve corresponds to the feed plate. As Hengstebeck points out (10) the slope of the curve represents a relative degree of fractionation in the column. Plates near and below the feed plate can be seen to have a very small slope, hence very little fractionation occurs on these plates. They are necessary for the desired separation but they are not helping the separation to a high degree. Removal of these plates however would cause the degree of separation to decrease. The relative amount of dip in the rectifying section of the column does not determine the optimum feed plate except as noted for the case of a column with partial vapor feed operating near constant molal overflow. For real columns operating at finite reflux ratios, the degree of curvature of the relative fractionation plot is highly sensitive to the reflux rate or ratio. As would be expected, the curvature of the entire relative fractionation plot decreases, approaching a straight line as the reflux ratio increases. This can be easily seen by comparing Figure 1 through Figure 5, which correspond to a depropanizer column at low through high reflux ratios. At high reflux ratios the relative fractionation plot becomes very close to linear except very near the feed plate where discontinuities can exist.

The effect of changing feed plates can be seen from Figure 1. Wide deviations exist through the mid-region of the column on trays directly adjacent to the feed plate and near the feed plate. To generalize, as the feed plate shifts further up the column, the relative fractionation curve shifts to the right and the stripping curve

lengthens while the rectifying curve becomes shorter. The degree of depth of the dip in the rectifying section decreases as the feed plate is shifted up higher in the column. This is because the key component ratio is higher near the top of the column than the bottom and the feed causes less discontinuity between trays. The slope of the dip just above the feed plate decreases as the feed plate shifts higher in the column since the relative degree of dip is less. As previously noted, the relative degree of dip in the rectifying section is not an indicator of the optimum feed plate.

In the theory section of Chapter III the separation ratio was noted as an indicator of the total amount of separation taking place in the column. Equations [24] and [25] express the separation ratio in mathematical terms. On the relative fractionation plot, the numerator of these equations corresponds to the terminal point of the rectification curve. Similarly, the denominator corresponds to the terminal point in the stripping section on the relative fractionation plot. So, the separation ratio is expressed as a ratio of the two terminal key component ratios. Thus, the farther the spread of the two terminal points on the relative fractionation plot, the better the separation in the column. In general, the effects of changing feed plate location will not show up in the terminal points on ordinary log paper. The key component ratios at the terminals differ usually only in the third or fourth decimal place for different feed locations. It is generally only possible to read two or three significant figures on log paper so that the terminals usually overlap too closely to distinguish a "best" separation with a relative fractionation plot. However, there are instances where clearly the terminals have wide variation for different

feed locations. This can be noted from Figures 19 and 23 very dramatically. Figure 19 corresponds to a mid-range reflux ratio depropanizer with 25% vaporized feed. Figure 23 corresponds to a demethanizer with mid-range reflux ratio and two feeds. The rectifying section curves separate and clearly the terminal corresponding to plate 11 is the higher in Figure 19. On Figure 23 plate 10 has the lowest terminal in the stripping section. Both terminals for a column must be considered simultaneously as is done by the separation ratio in order to combine effects of maximum recovery of the light key component in the distillate and its minimum recovery in the bottom product. Similarly the converse is true for the heavy key component. All components lighter and heavier than the keys have maximum recovery when the keys have maximum recovery. The separation ratio should be a more "unbiased" estimator of the maximum separation of both key components simultaneously than some other parameter which considers only one key or the other. As noted in the theory of Chapter III, each component of the feed mixture has a different optimum feed plate for maximum recovery in the respective product stream. In general, the lighter component has a lower optimum feed plate location. Thus the light key component has a lower optimum feed location than the heavy key component, and the optimum which maximizes recovery of both key components simultaneously should be somewhere between these two limits. For the cases studied this conclusion is sound.

Another important effect in feed plate optimization is fractional vaporization of the feed. All things being equal, the higher the fractional vaporization of the feed, the lower the separation ratio of the key components. That is, a bubble point feed gives a better key

component separation than a fractionally vaporized feed if both are introduced on the same plate with product rates and reflux ratios equal for both cases. This effect is clearly shown by comparing Figures 1, 6 and 11. These figures all correspond to a depropanizer column with constant product rates and reflux ratio. Figure 1 is for the case of a bubble point feed while Figures 6 and 11 correspond to 25% and 50% vaporization of the feed. Clearly increasing the fractional vaporization of the feed decreases the upper terminal key ratio and increases the lower terminal key ratio, the net effect being a poorer separation. This same effect of fractional vaporization can be seen from the other plots for different reflux ratios. (Figures 1 through 15). The one depropanizer is the only case where fractional vaporization effects were studied specifically.

Three other column configurations were studied for a single feed condition. A six tray stripper with two feeds, (one completely vaporized and one partially vaporized), was briefly studied. Little optimization was required because of the small number of trays. A small demethanizer, (13 trays) with two feeds was studied and a wide boiling depropanizer was studied which contained C2 through NC8. The feeds for these columns are given in Table VIII, Appendix A.

The results are best presented in tabular form. In Table I are the results of the depropanizer column with key components C3 and IC4. (Tables of calculated results are presented in Appendix B). The separation ratio as calculated by equation [24] is presented for various feed plate locations. In addition the light key distillate mole fraction and heavy key bottom product mole fraction are given and their combined product, for various feed plate locations. From the key component mole

fractions, clearly the light key component is optimized at a lower position in the column than the heavy key component. So from the key component mole fractions in the two product streams the optimum feed location can be clearly identified for each key component by looking at its mole fraction in its respective product. The key component separation ratio lies between or on the optimum plate for one key or the other in all cases studied. The product of the light key distillate mole fraction and heavy key bottom product mole fraction should be in close agreement with the separation ratio for the optimum feed tray. The difficulty with this parameter in characterizing the feed plate optimum is that it only considers maximizing the terminal compositions of the keys and neglects small effects of recovery in the inappropriate stream of each key component. However, for most cases recovery of a key component in one stream is proportional to the maximized key component mole fraction in another. A case where this would not be true is unsteady state distillation as in start-up, or oscillation about steady state conditions where accumulation is possible internally. At any rate for the steady state solution the separation ratio should be in agreement with the product of key component compositions in their respective product streams. With reference to Table I, fairly close agreement between the separation ratio and composition product for the keys is obtained over a wide range of reflux ratios and feed conditions.

A difficulty arises in how many figures to trust or accept in this table. The solution to any tray-by-tray analysis is highly dependent on the degree of reliability of the vapor-liquid equilibrium constants, (K-values), and vapor and liquid enthalpy data. For hydrocarbons much data is readily available for vapor and liquid enthalpies and K-values.

The complication comes in the complexity of the equations used in the tray-by-tray analysis to obtain the solution. Enthalpy and K-value data are curve fitted and the coefficients of the curve fitting are then used in the program. In this particular case, the data were curve fitted through three points with a quadratic equation and the coefficients were then used in the tray-by-tray program. It may be reasonable to accept three significant figures in the tables for the parameters estimated from the steady state mole fractions of the problem solution. Certainly four significant figures become somewhat questionable from a simple quadratic equation fit of three points. The accuracy would be perhaps equivalent to trusting a straight line fit of the relative fractionation plot through the terminal points at real reflux rates. In Tables I through V, three significant figures is generally enough to clearly identify the optimum feed plate. In some instances the optimum must be limited to two plates unless the fourth significant figure is considered. Table II gives the results of the wide boiling depropanizer with key components IC4 and NC4. For the lower reflux ratio the optimum could lie on plate 22 or 23 when looking at three significant figures. If four figures are considered, the optimum is clearly plate 23. Table III is a 20 tray configuration of the first depropanizer considered in Table I for a 25% vaporized feed. Good agreement between the separation ratio and component composition product is obtained except at the lowest reflux ratio where there is one tray disagreement. When looking at three significant figures the two lowest reflux ratios limit stating the optimum to one of two trays. If four significant figures are considered, the optimums become clear for these cases. Table IV gives the results of the 13 tray demethanizer.

The total vapor feed is on top in all cases and the fractionally vaporized feed is allowed to vary. The optimum feed plate is clearly identified by the separation ratio and in agreement with the component compositions product. Finally, Table V gives the results of the six tray stripper, which splits between C3 and IC4. Since the vapor is not condensed and no reflux returns to the top tray the separation ratio based on the top tray is zero, as is the compositions product. The tray just beneath the top tray is the optimum. Figures 1 through 24 represent the relative fractionation plots of the four columns considered in this study.

Tables VI and VII are an attempt to correlate the optimum feed plate for the single depropanizer case of Table I by the methods of Gilliland (5). Table VI presents the variables used in equation [13] of Gilliland for a partially vaporized feed. Virtually no agreement is obtained for the criterion of equation [13] over the feed locations tested. The assumption of constant molal overflow is greatly violated with the test case. Table VII reveals the correlation of the optimum feed plate by equation [11] for a bubble point feed. No correlation is obtained again demonstrating that violation of the assumption of constant molal overflow is a severe limitation of Gilliland's criterion.

CHAPTER V

CONCLUSIONS AND RECOMMENDATIONS

In the cases studied, which consist of a wide range of reflux rates, fractional vaporization, components, number of stages and feed tray locations, the separation ratio is an adequate parameter for predicting the optimal feed plate for a desired separation of a multicomponent feed mixture in a distillation column.

The capability to prescribe accurately the optimum feed plate location is dependent on the degree of reliability of the composition profiles throughout a distillation column. A higher order curve fit would be desirable, particularly where more data points as to compositions are available. An instance where more than three data points are available is the case of a complex column with side streams in addition to the usual distillate and bottom product. Here third or fourth order curve fits could be implemented into the tray analysis program to increase reliability of the composition profiles. It is recommended that a study of implementing the separation ratio criterion into the tray analysis program as a call subroutine be made to allow "automatic" determination of the optimum by simply specifying a starting point for the feed point. Third or fourth order convergence techniques could be used to find the maximum in the separation ratio quickly by skipping stages rather than incrementing each, one at a time as the feed plate. The use of a powerful convergence technique, (third or fourth order),

could result in considerable savings of computer time and wasted output in locating the optimal feed plate. The separation ratio criterion needs to be tested for complex columns with multiple product streams.

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APPENDIX A

TABLES OF RESULTS AND

INPUT DATA

TABLE I
SEPARATION RATIO FOR A
TYPICAL DEPROPANIZER

| $\frac{V}{F} = 1.12$ | Feedtray | R | x_{dlk} | x_{bhk} | $x_{dlk} \cdot x_{bhk}$ |
|----------------------|----------|---------|-----------|-----------|-------------------------|
| Bubble | 13 | 36.15 | 0.92168 | 0.18146 | 0.16725 |
| Point | 14 | 37.18 | 0.92238 | 0.18244 | 0.16828 |
| Feed | 15 | 38.10 | 0.92254 | 0.18323 | 0.16903 |
| | 16 | 38.56 | 0.92259* | 0.18382 | 0.16959 |
| | 17+ | 38.91 | 0.92233 | 0.18431 | 0.16999 |
| | 18+ | 38.95* | 0.92193 | 0.18467 | 0.17025 |
| | 19 | 38.79 | 0.92126 | 0.18493* | 0.17037* |
| 25% Vapor Feed | 14 | 12.05 | 0.87827 | 0.14764 | 0.12967 |
| | 15 | 12.12 | 0.87828* | 0.14792 | 0.12992 |
| | 16 | 12.19 | 0.87827 | 0.14822 | 0.13018 |
| | 21+ | 12.44* | 0.87651 | 0.15006 | 0.13153 |
| | 22+ | 12.42 | 0.87548 | 0.15040* | 0.13167* |
| 50% Vapor Feed | 12 | 4.790 | 0.82038 | 0.11414 | 0.09364 |
| | 13 | 4.797 | 0.82039* | 0.11420 | 0.09369 |
| | 14 | 4.803 | 0.82039* | 0.11426 | 0.09374 |
| | 15 | 4.811 | 0.82038 | 0.11434 | 0.09380 |
| | 16 | 4.817 | 0.82037 | 0.11445 | 0.09389 |
| | 17 | 4.831 | 0.82032 | 0.11459 | 0.09400 |
| | 18 | 4.846 | 0.82026 | 0.11477 | 0.09414 |
| | 19 | 4.864 | 0.82015 | 0.11501 | 0.09433 |
| | 20 | 4.889 | 0.81998 | 0.11532 | 0.09456 |
| | 21 | 4.920 | 0.81969 | 0.11572 | 0.09485 |
| | 22 | 4.953 | 0.81923 | 0.11623 | 0.09522 |
| | 23 | 4.986 | 0.81843 | 0.11683 | 0.09562 |
| | 24+ | 5.010* | 0.81703 | 0.11751 | 0.09601 |
| 25 | 5.000 | 0.81456 | 0.11818* | 0.09626* | |

* Optimum of an individual parameter

+ Feedtray optimum, based on three significant figures

lk is C3

hk is IC4

$\frac{D}{F} = 0.4854$

30 total theoretical trays

TABLE I (Continued)

| $\frac{V}{F} = 1.26$ | Feedtray | R | x_{dlk} | x_{bhk} | $x_{dlk} \cdot x_{bhk}$ |
|----------------------|----------|--------|-----------|-----------|-------------------------|
| Bubble | 13 | 127.8 | 0.94957 | 0.20787 | 0.19739 |
| Point | 14 | 139.7 | 0.95084 | 0.20978 | 0.19947 |
| Feed | 15 | 141.3 | 0.95134* | 0.21092 | 0.20069 |
| | 16+ | 152.0* | 0.95111 | 0.23362* | 0.22220* |
| | 17 | 148.7 | 0.95072 | 0.21155 | 0.20112 |
| 25% | 14 | 28.17 | 0.91600 | 0.17020 | 0.15590 |
| Vapor | 15 | 28.68 | 0.91613* | 0.17104 | 0.15669 |
| Feed | 16 | 29.05 | 0.91603 | 0.17176 | 0.15734 |
| | 19+ | 29.34* | 0.91417 | 0.17329 | 0.15842* |
| | 20 | 28.96 | 0.91299 | 0.17350* | 0.15840 |
| | 21 | 28.29 | 0.91134 | 0.17348 | 0.15810 |
| 50% | 13 | 8.164 | 0.86150 | 0.13011 | 0.11209 |
| Vapor | 14 | 8.212 | 0.86151* | 0.13042 | 0.11236 |
| Feed | 15 | 8.264 | 0.86145 | 0.13077 | 0.11265 |
| | 16 | 8.325 | 0.86136 | 0.13117 | 0.11298 |
| | 17 | 8.385 | 0.86118 | 0.13162 | 0.11335 |
| | 18 | 8.454 | 0.86096 | 0.13213 | 0.11376 |
| | 19 | 8.521 | 0.86062 | 0.13270 | 0.11420 |
| | 20 | 8.586 | 0.86010 | 0.13332 | 0.11467 |
| | 21 | 8.641 | 0.85031 | 0.13395 | 0.11510 |
| | 22+ | 8.656* | 0.85810 | 0.13455 | 0.11546 |
| | 23 | 8.615 | 0.85626 | 0.13504* | 0.11563* |

* Optimum of an individual parameter

+ Feedtray optimum, based on three significant figures

lk is C3

hk is IC4

$\frac{D}{F} = 0.4854$

30 total theoretical trays

TABLE I (Continued)

| $\frac{V}{F} = 1.40$ | Feedtray | R | x_{dlk} | x_{bhk} | $x_{dlk} \cdot x_{bhk}$ |
|----------------------|----------|--------|-----------|-----------|-------------------------|
| Bubble | 13 | 620.2 | 0.96510 | 0.22968 | 0.22166 |
| Point | 14 | 787.5 | 0.96652 | 0.23188 | 0.22412 |
| Feed | 15 | 918.5 | 0.96736 | 0.23325 | 0.22564 |
| | 16+ | 969.4* | 0.96762* | 0.23383* | 0.22626* |
| | 17 | 916.9 | 0.96730 | 0.23364 | 0.22600 |
| 25% | 13 | 101.1 | 0.94651 | 0.19985 | 0.18916 |
| Vapor | 14 | 109.4 | 0.94712 | 0.20135 | 0.19070 |
| Feed | 15+ | 114.4* | 0.94726 | 0.20229 | 0.19162 |
| | 16+ | 113.8 | 0.94739* | 0.20280* | 0.19213* |
| | 17 | 112.3 | 0.94653 | 0.20273 | 0.19189 |
| | 18 | 107.2 | 0.94537 | 0.20228 | 0.19123 |
| | 19 | 99.4 | 0.94376 | 0.20141 | 0.19008 |
| | 20 | 90.9 | 0.94159 | 0.20021 | 0.18852 |
| 50% | 13 | 16.82 | 0.90039 | 0.15070 | 0.13569 |
| Vapor | 14 | 17.13 | 0.90049* | 0.15161 | 0.13652 |
| Feed | 15 | 17.43 | 0.90035 | 0.15247 | 0.13728 |
| | 16 | 17.67 | 0.90020 | 0.15333 | 0.13803 |
| | 17 | 17.90 | 0.89977 | 0.15413 | 0.13868 |
| | 18 | 18.05 | 0.89918 | 0.15490 | 0.13928 |
| | 19+ | 18.13* | 0.89837 | 0.15561 | 0.13980 |
| | 20+ | 18.11 | 0.89716 | 0.15617* | 0.14011* |

* Optimum of an individual parameter

+ Feedtray optimum, based on three significant figures

lk is C3

hk is IC4

$\frac{D}{F} = 0.4854$

30 total theoretical trays

TABLE I (Continued)

| $\frac{V}{F} = 1.54$ | Feedtray | R | x_{dlk} | x_{bhk} | $x_{dlk} \cdot x_{bhk}$ |
|----------------------|----------|--------|-----------|-----------|-------------------------|
| Bubble | 15 | 4098. | 0.97307 | 0.24285 | 0.23631 |
| Point | 16 | 4956. | 0.97347 | 0.24357 | 0.23711 |
| Feed | 17+ | 5312.* | 0.97358* | 0.24389* | 0.23745* |
| | 18 | 4822. | 0.97333 | 0.24374 | 0.23724 |
| 25% | 14 | 672.4 | 0.96556 | 0.22705 | 0.21923 |
| Vapor | 15 | 757.9 | 0.96627 | 0.22823 | 0.22053 |
| Feed | 16+ | 772.4* | 0.96635* | 0.22857* | 0.22088* |
| | 17 | 677.8 | 0.96600 | 0.22802 | 0.22027 |
| | 18 | 597.9 | 0.96449 | 0.22681 | 0.21876 |
| | | | | | |
| 50% | 13 | 53.20 | 0.93453 | 0.18180 | 0.16990 |
| Vapor | 14 | 55.43 | 0.93514* | 0.18316 | 0.17128 |
| Feed | 15 | 56.81 | 0.93501 | 0.18395 | 0.17200 |
| | 16 | 57.30 | 0.93473 | 0.18456 | 0.17251 |
| | 17+ | 57.70* | 0.93432 | 0.18484* | 0.17266* |
| | 18 | 55.36 | 0.93292 | 0.18476 | 0.17237 |

* Optimum of an individual parameter

+ Feedtray optimum, based on three significant figures

lk is C3

hk is IC4

$\frac{D}{F} = 0.4854$

30 total theoretical trays

TABLE I (Continued)

| $\frac{V}{D} = 1.68$ | Feedtray | R | x_{dlk} | x_{bhk} | $x_{dlk} \cdot x_{bhk}$ |
|----------------------|----------|---------|-----------|-----------|-------------------------|
| Bubble | 14 | | 0.97431 | 0.24606 | 0.23974 |
| Point | 15 | 10591. | 0.97483 | 0.24689 | 0.24068 |
| Feed | 16 | 9697. | 0.97514 | 0.24744 | 0.24129 |
| | 17 | 17175. | 0.97532 | 0.24782 | 0.24170 |
| | 18+ | 18896.* | 0.97541* | 0.24808 | 0.24198* |
| | 19 | 17308. | 0.97530 | 0.24811* | 0.24198* |
| | | | | | |
| 25% | 14 | 2902. | 0.97214 | 0.23824 | 0.23160 |
| Vapor | 15 | 3708. | 0.97278 | 0.23928 | 0.23177 |
| Feed | 16 | 4284. | 0.97312* | 0.23991 | 0.23346 |
| | 17+ | 4337.* | 0.97309 | 0.24004* | 0.23358* |
| | 18 | 3630. | 0.97265 | 0.23961 | 0.23306 |
| | | | | | |
| 50% | 14 | 324.7 | 0.96047 | 0.21585 | 0.20732 |
| Vapor | 15+ | 344.7* | 0.96075* | 0.21664* | 0.20814* |
| Feed | 16 | 342.6 | 0.96031 | 0.21661 | 0.20801 |
| | 17 | 312.6 | 0.95952 | 0.21590 | 0.20716 |

* Optimum of an individual parameter

+ Feedtray optimum, based on three significant figures

lk is C3

hk is IC4

$\frac{D}{F} = 0.4854$

30 total theoretical trays

TABLE II
SEPARATION RATIO FOR A
TYPICAL WIDE BOILING
DEPROPANIZER

| $\frac{V}{F} = 1.00$ | Feedtray | R | x_{dlk} | x_{bhk} | $x_{dlk} \cdot x_{bhk}$ |
|----------------------|----------|--------|-----------|-----------|-------------------------|
| Bubble | 20 | 3.598 | 0.16256 | 0.22611 | 0.03676 |
| Point | 21 | 3.625 | 0.16259* | 0.22668 | 0.03686 |
| Feed | 22+ | 3.641 | 0.16248 | 0.22725 | 0.03692 |
| | 23+ | 3.644* | 0.16218 | 0.22783 | 0.03695* |
| $\frac{D}{F} = 0.45$ | 24 | 3.629 | 0.16163 | 0.22843 | 0.03692 |
| | 28 | 3.167 | 0.15315 | 0.23134* | 0.03543 |
| $\frac{V}{F} = 3.00$ | | | | | |
| Bubble | 18 | 27.54 | 0.19936 | 0.32038* | 0.06387 |
| Point | 19+ | 27.62* | 0.19955 | 0.32010 | 0.06388* |
| Feed | 20 | 27.44 | 0.19961* | 0.31941 | 0.06376 |
| | 21 | 26.98 | 0.19954 | 0.31841 | 0.06354 |
| | 22 | 26.24 | 0.19933 | 0.31707 | 0.06320 |
| | 23 | 25.22 | 0.19896 | 0.31539 | 0.06275 |
| $\frac{D}{F} = 0.45$ | 25 | 22.37 | 0.19763 | 0.31093 | 0.06145 |

* Optimum of an individual parameter

+ Feedtray optimum, based on three significant figures

lk is IC4

hk is NC4

30 total theoretical trays

TABLE III
SEPARATION RATIO FOR A
TYPICAL DEPROPANIZER

| $\frac{V}{F} = 1.12$ | Feedtray | R | x_{dlk} | x_{bhk} | $x_{dlk} \cdot x_{bhk}$ |
|----------------------|----------|--------|-----------|-----------|-------------------------|
| 25% Vapor | 10 | 11.84 | 0.87525 | 0.14818 | 0.12969 |
| Feed | 11 | 12.08 | 0.87533* | 0.14914 | 0.13055 |
| | 12+ | 12.19* | 0.87480 | 0.14986 | 0.13110 |
| | 13+ | 12.18 | 0.87353 | 0.15033* | 0.13132* |
| $\frac{V}{F} = 1.26$ | | | | | |
| 25% Vapor | 10 | 24.84 | 0.90820 | 0.16933 | 0.15379 |
| Feed | 11+ | 25.66 | 0.90837* | 0.17086 | 0.15520 |
| | 12+ | 25.69* | 0.90737 | 0.17158* | 0.15569* |
| | 13 | 24.95 | 0.90501 | 0.17152 | 0.15523 |
| $\frac{V}{F} = 1.40$ | | | | | |
| 25% Vapor | 10 | 58.35 | 0.93342* | 0.19128 | 0.17854 |
| Feed | 11+ | 61.37* | 0.93342* | 0.19287 | 0.18003* |
| | 12 | 60.41 | 0.93185 | 0.19310* | 0.17994 |
| | 13 | 54.97 | 0.92895 | 0.19195 | 0.17831 |
| $\frac{V}{F} = 1.54$ | | | | | |
| 25% Vapor | 10 | 138.6 | 0.94885 | 0.20901 | 0.19832 |
| Feed | 11+ | 147.3* | 0.94950* | 0.21067* | 0.20003* |
| | 12 | 140.0 | 0.94831 | 0.21050 | 0.19962 |
| | 13 | 122.6 | 0.94508 | 0.20882 | 0.19735 |
| $\frac{V}{F} = 1.68$ | | | | | |
| 25% Vapor | 10 | 291.4 | 0.95807 | 0.22132 | 0.21204 |
| Feed | 11+ | 319.4* | 0.95863* | 0.22284* | 0.21362* |
| | 12 | 304.2 | 0.95788 | 0.22281 | 0.21342 |
| | 13 | 246.9 | 0.95553 | 0.22098 | 0.21115 |

* Optimum of an individual parameter

+ Feedtray optimum, based on three significant figures

lk is C3

hk is IC4

$\frac{D}{F} = 0.4854$ 20 total theoretical trays

TABLE IV
SEPARATION RATIO FOR A
TYPICAL DEMETHANIZER

| $\frac{V}{F} = 0.562$ | Feedtray (second) | R | x_{dlk} | x_{bhk} | $x_{dlk} \cdot x_{bhk}$ |
|--------------------------------|-----------------------|----------|----------------------------|-----------|-------------------------|
| 100% Vapor | 7 | 261.32 | 0.19486 | 0.31209 | 0.06081 |
| Feed-13 | 8 | 460.87 | 0.19486 | 0.31867 | 0.06210 |
| | 9 | 826.39 | 0.19486 | 0.32341 | 0.08636 |
| 50% Second Vapor Feed | 10+ | 1306.3 * | 0.19486* | 0.32727* | 0.08785* |
| | 11 | 57.15 | 0.19486 | 0.22342 | 0.04354 |
| lk is C2 hk is C3 | $\frac{D}{F} = 0.562$ | | 13 total theoretical trays | | |

TABLE V
SEPARATION RATIO FOR A
SIX TRAY STRIPPER

| $\frac{V}{F} = 0.095$ | Feedtray (second) | R** | x_{dlk}^{**} | x_{bhk}^{**} | $x_{dlk} \cdot x_{bhk}^{**}$ |
|---------------------------------|----------------------|--------|----------------|----------------|------------------------------|
| 10% Vapor Feed-3 | 5 | 0. | 0. | 0.02794 | 0. |
| 100% Second Vapor Feed | 6+ | 2.101* | 0.05290* | 0.03002* | 0.00159* |

* Optimum of an individual parameter

+ Feedtray Optimum, based on three significant figures

lk is C3

hk is IC4

** no liquid on tray 6 and no reflux to tray 6

6 total theoretical trays

TABLE VI

GILLILAND CRITERION PARAMETERS FOR A
TYPICAL DEPROPANIZER* WITH
PARTIAL VAPOR FEED

| $\frac{V}{F} = 1.12$ | Feedtray | r_f | r_{f+1} |
|------------------------|----------|--------|-----------|
| 25% Vapor | 14 | 4.0255 | 3.4549 |
| Feed | 15 | 4.0226 | 3.4445 |
| | 16 | 4.0184 | 3.4321 |
| $\frac{D}{F} = 0.4854$ | 21 | 3.9971 | 3.4010 |
| | 22 | 4.0010 | 3.4355 |
| $\frac{V}{F} = 1.26$ | | | |
| 25% Vapor | 14 | 3.8331 | 3.1700 |
| Feed | 15 | 3.8305 | 3.1643 |
| | 16 | 3.8275 | 3.1610 |
| $\frac{D}{F} = 0.4854$ | 19 | 3.8319 | 3.1994 |
| | 20 | 3.8436 | 3.2452 |
| | 21 | 3.8647 | 3.3206 |
| $\frac{V}{F} = 1.40$ | | | |
| 25% Vapor | 13 | 3.4959 | 2.7376 |
| Feed | 14 | 3.5275 | 2.7910 |
| | 15 | 3.5591 | 2.8491 |
| $\frac{D}{F} = 0.4854$ | 16 | 3.5910 | 2.9151 |
| | 17 | 3.6262 | 2.9897 |
| | 18 | 3.6656 | 3.0793 |
| | 19 | 3.7113 | 3.1882 |
| | 20 | 3.7662 | 3.3234 |
| $\frac{V}{F} = 1.54$ | | | |
| 25% Vapor | 14 | 3.2332 | 2.5764 |
| Feed | 15 | 3.3492 | 2.7656 |
| | 16 | 3.4718 | 2.9751 |
| $\frac{D}{F} = 0.4854$ | 17 | 3.6018 | 3.2088 |
| | 18 | 3.6997 | 3.3962 |

$$r_F = 4.3435$$

* 30 tray depropanizer of TABLE I

TABLE VI (Continued)

| $\frac{V}{F} = 1.68$ | Feedtray | r_f | r_{f+1} |
|------------------------|----------|--------|-----------|
| 25% | 14 | 2.8869 | 2.3270 |
| Vapor | 15 | 3.0753 | 2.6188 |
| Feed | 16 | 3.3068 | 2.9907 |
| $\frac{D}{F} = 0.4854$ | 17 | 3.5676 | 3.4288 |
| | 18 | 3.8383 | 3.9086 |
| $r_F = 4.3435$ | | | |
| | 12 | 3.4914 | 2.9859 |
| | 13 | 3.4902 | 2.9826 |
| | 14 | 3.4889 | 2.9782 |
| $\frac{V}{F} = 1.12$ | 15 | 3.4870 | 2.9725 |
| | 16 | 3.4844 | 2.9648 |
| 50% | 17 | 3.4812 | 2.9553 |
| Vapor | 18 | 3.4769 | 2.9440 |
| Feed | 19 | 3.4715 | 2.9304 |
| $\frac{D}{F} = 0.4854$ | 20 | 3.4655 | 2.9161 |
| | 21 | 3.4588 | 2.9035 |
| | 22 | 3.4521 | 2.8980 |
| | 23 | 3.4480 | 2.9103 |
| | 24 | 3.4511 | 2.9623 |
| | 25 | 3.4712 | 3.0954 |
| | 13 | 3.3392 | 2.8087 |
| | 14 | 3.3339 | 2.7960 |
| $\frac{V}{F} = 1.26$ | 15 | 3.3273 | 2.7814 |
| | 16 | 3.3199 | 2.7653 |
| 50% | 17 | 3.3106 | 2.7479 |
| Vapor | 18 | 3.3025 | 2.7309 |
| Feed | 19 | 3.2936 | 2.7169 |
| $\frac{D}{F} = 0.4854$ | 20 | 3.2859 | 2.7094 |
| | 21 | 3.2818 | 2.7151 |
| | 22 | 3.2843 | 2.7441 |
| | 23 | 3.2990 | 2.8131 |
| $r_F = 3.7888$ | | | |

TABLE VI (Continued)

| $\frac{V}{F} = 1.40$ | Feedtray | r_f | r_{f+1} |
|------------------------|----------|--------|-----------|
| 50% | 13 | 3.1402 | 2.5663 |
| Vapor | 14 | 3.1313 | 2.5504 |
| Feed | 15 | 3.1223 | 2.5359 |
| $\frac{D}{F} = 0.4854$ | 16 | 3.1128 | 2.5222 |
| | 17 | 3.1050 | 2.5139 |
| | 18 | 3.0995 | 2.5126 |
| | 19 | 3.0985 | 2.5232 |
| | 20 | 3.1045 | 2.5516 |
| 50% | 13 | 2.8233 | 2.2044 |
| Vapor | 14 | 2.8313 | 2.2185 |
| Feed | 15 | 2.8427 | 2.2398 |
| $\frac{D}{F} = 0.4854$ | 16 | 2.8557 | 2.2668 |
| | 17 | 2.8730 | 2.3041 |
| $\frac{V}{F} = 1.54$ | 18 | 2.8974 | 2.3563 |
| $\frac{V}{F} = 1.68$ | | | |
| 50% | 14 | 2.5617 | 2.0201 |
| Vapor | 15 | 2.6232 | 2.1125 |
| Feed | 16 | 2.6905 | 2.2170 |
| $\frac{D}{F} = 0.4854$ | 17 | 2.7659 | 2.3407 |
| | | | |
| $r_F = 3.7888$ | | | |

TABLE VII
 GILLILAND CRITERION PARAMETERS FOR A
 TYPICAL DEPROPANIZER* WITH
 BUBBLE POINT FEED

| | Feedtray | r_f | r_i | r_{f+1} |
|----------------------|----------|----------|----------|-----------|
| $\frac{V}{F} = 1.12$ | 13 | 4.8519 | 4.9418 | 4.1410 |
| | 14 | 4.8600 | 4.9395 | 4.1547 |
| | 15 | 4.8649 | 4.9375 | 4.1629 |
| | 16 | 4.8668 | 4.9358 | 4.1673 |
| | 17 | 4.8682 | 4.9343 | 4.1741 |
| | 18 | 4.8702 + | 4.8401 + | 4.0880 + |
| | 19 | 4.8751 | 4.9314 | 4.2139 |
| $\frac{V}{F} = 1.26$ | 13 | 4.5227 | 4.8792 | 3.6006 |
| | 14 | 4.5720 | 4.8743 | 3.6996 |
| | 15 | 4.6167 | 4.8709 | 3.7968 |
| | 16 | 4.6573 | 4.8687 | 3.8923 |
| | 17 | 4.6971 | 4.8672 | 3.9954 |
| $\frac{V}{F} = 1.40$ | 13 | 4.0809 | 4.8254 | 3.1192 |
| | 14 | 4.2198 | 4.8199 | 3.3661 |
| | 15 | 4.3718 | 4.6907 | 3.6506 |
| | 16 | 4.5245 | 4.8138 | 4.0934 |
| | 17 | 4.6660 | 4.8262 | 4.2530 |
| $\frac{V}{F} = 1.54$ | 15 | 3.9823 | 4.8000 | 3.3349 |
| | 16 | 4.2581 | 4.7942 | 3.8276 |
| | 17 | 4.5630 | 4.7896 | 4.4103 |
| | 18 | 4.8696 | 4.7870 | 5.0424 |
| $\frac{V}{F} = 1.68$ | 14 | 3.2075 | 4.8047 | 2.4258 |
| | 15 | 3.4471 | 4.7972 | 2.7988 |
| | 16 | 3.7464 | 4.7903 | 3.2864 |
| | 17 | 4.1172 | 4.7845 | 3.9277 |
| | 18 | 4.5732 | 4.7802 | 4.7793 |
| | 19 | 5.0848 | 4.7783 | 5.8295 |

$\frac{D}{F} = 0.4854$

* 30 tray depropanizer of TABLE I

+ agreement with separation ratio of TABLE I

TABLE VIII
FEEDS AND FEED CONDITIONS FOR
VARIOUS DISTILLATION COLUMNS

| Component | Depropanizer* | Wide Boiling Depropanizer+ | Demethanizer** | | Six Tray Strip per ⁺⁺ | |
|------------------------|---|---|--|---|----------------------------------|----------|
| | | | Feed-1 | Feed-2 | Feed-1 | Feed-2 |
| N2 | | | 0.010 | 0.070 | | |
| CO2 | | | 5.750 | 39.710 | 21.320 | 17.070 |
| H2S | | | | | 86.420 | 69.220 |
| C1 | | | 17.460 | 106.590 | 55.310 | 44.300 |
| C2= | | 5.000 | | | | |
| C2 | 0.989 | 5.000 | 22.840 | 205.890 | 120.420 | 96.450 |
| C3= | | 10.000 | | | | |
| C3 | 53.900 | 5.000 | 1.810 | 79.980 | 304.790 | 244.120 |
| IC4= | | 10.000 | | | | |
| IC4 | 10.360 | 10.000 | 0.010 | 10.520 | 85.850 | 68.760 |
| NC4 | 15.483 | 15.000 | | 15.480 | 251.780 | 201.670 |
| IC5 | 7.380 | 10.000 | | 7.380 | 148.650 | 119.060 |
| NC5 | 5.270 | 10.000 | | 5.270 | 221.820 | 177.670 |
| C6 | 10.788 | 10.000 | | 10.830 | 399.280 | 319.800 |
| C7 | 9.440 | 5.000 | | 9.500 | 349.730 | 280.120 |
| C7F | | | 0.500 | 1.090 | | |
| C8 | | 5.000 | | | 360.660 | 288.890 |
| C8F | | | 41.930 | 91.310 | | |
| C9 | | | | | 467.840 | 374.720 |
| C9F | | | 13.970 | 30.400 | | |
| C10 | | | | | 417.720 | 334.570 |
| C10F | | | 11.890 | 25.880 | | |
| C11F | | | 1.040 | 3.140 | | |
| 400F | | | | | 259.270 | 207.670 |
| 450F | | | | | 176.880 | 141.670 |
| 500F | | | | | 222.400 | 178.130 |
| 600F | | | | | 489.160 | 391.790 |
| 600+F | | | | | 1322.290 | 1059.090 |
| Total | 113.610 | 100.000 | 117.210 | 643.040 | 5761.610 | 4614.770 |
| Feed Condi- tion | Bubble Pt., 652.9R 25% Vapor, 670.0R 50% Vapor, 690.0R *Refers to Tables I & III | Bubble Pt., 652.3 R + Refers to Table II | Dew Pt. 50% Vapor 430.0 R 560.0 R ** Refers to Table IV | Dew Pt. 13% Vapor 710.0 R ++ Refers to Table V | Dew Pt. 523.0 R | |

APPENDIX B

RELATIVE FRACTIONATION CURVES

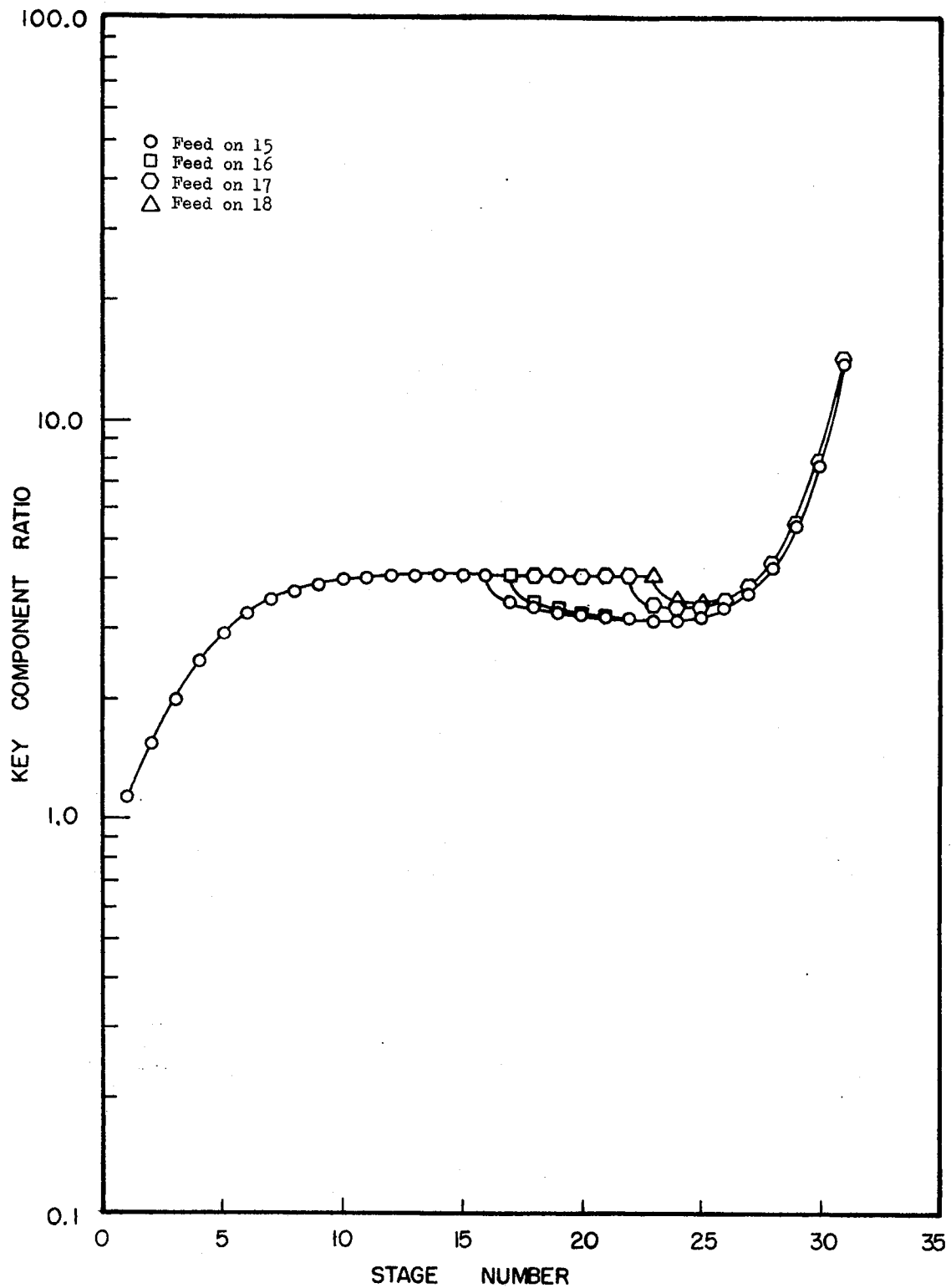


Figure 6. Relative Fractionation Curve- 30 Tray Depropanizer
With 25% Vapor Feed, $V/F = 1.12$, $D/F = 0.4854$

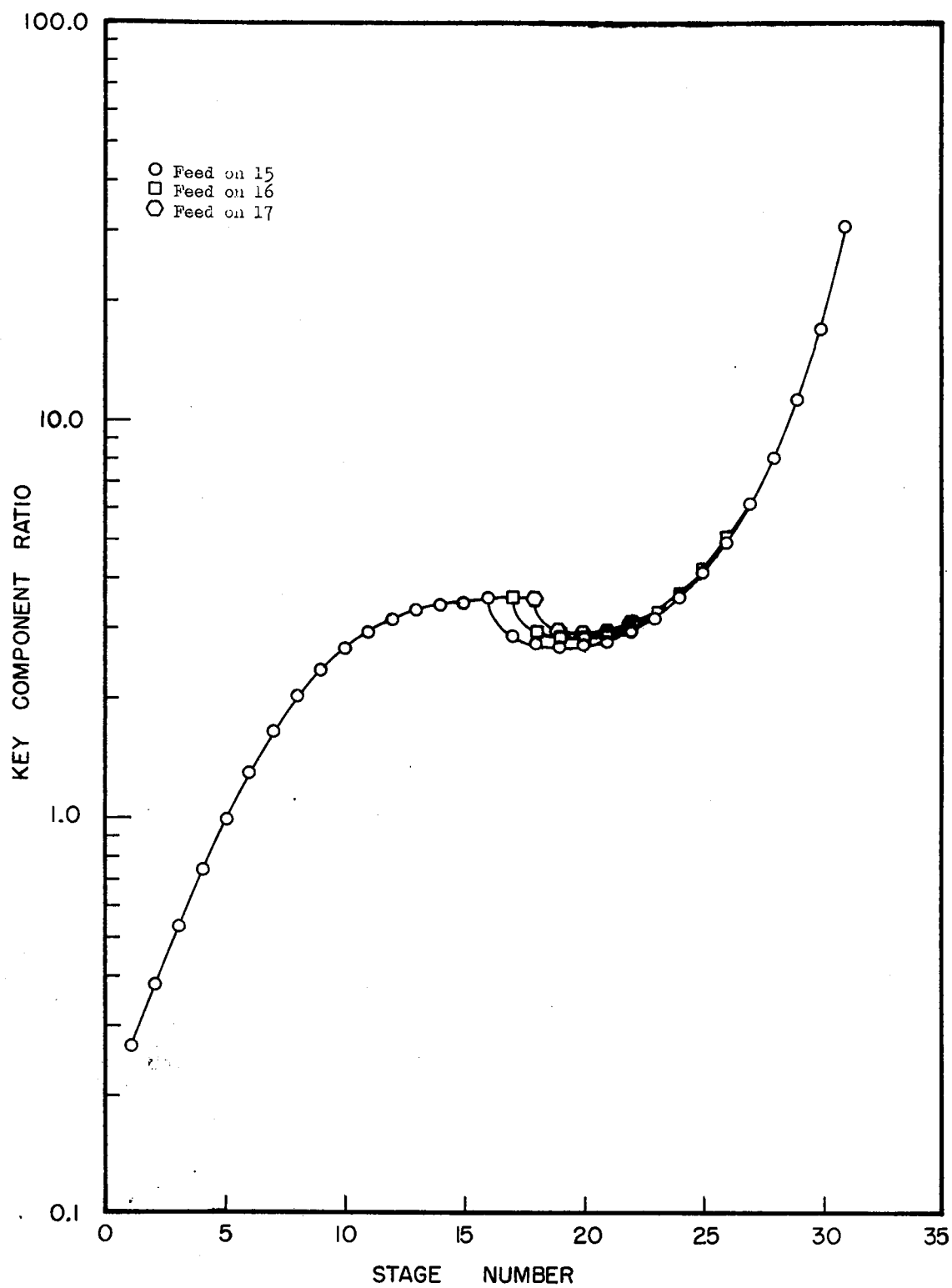


Figure 7. Relative Fractionation Curve- 30 Tray Depropanizer
With 25% Vapor Feed, $V/F = 1.26$, $D/F = 0.4854$

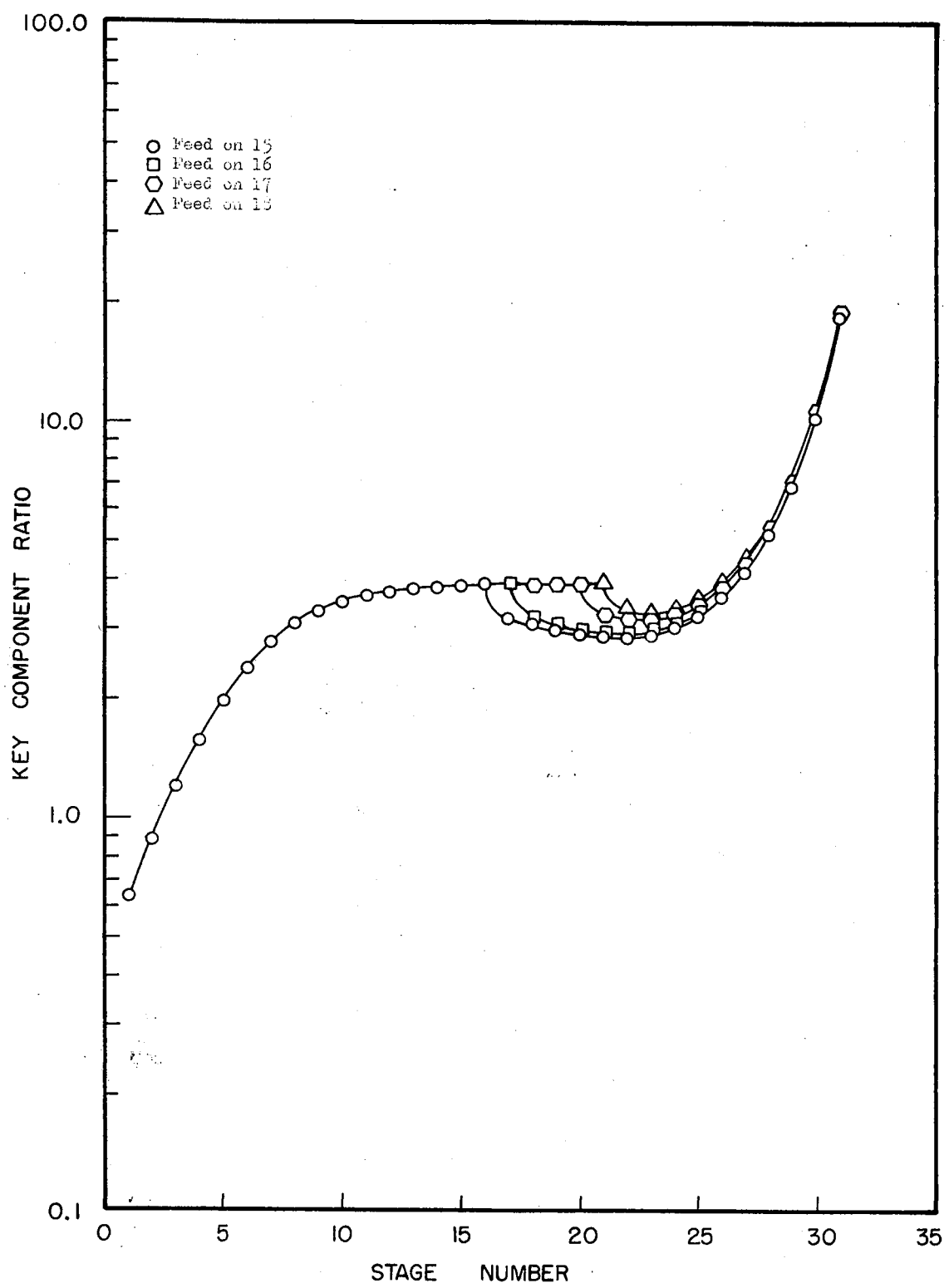


Figure 8. Relative Fractionation Curve- 30 Tray Depropanizer
 With 25% Vapor Feed, $V/F = 1.40$, $D/F = 0.4854$

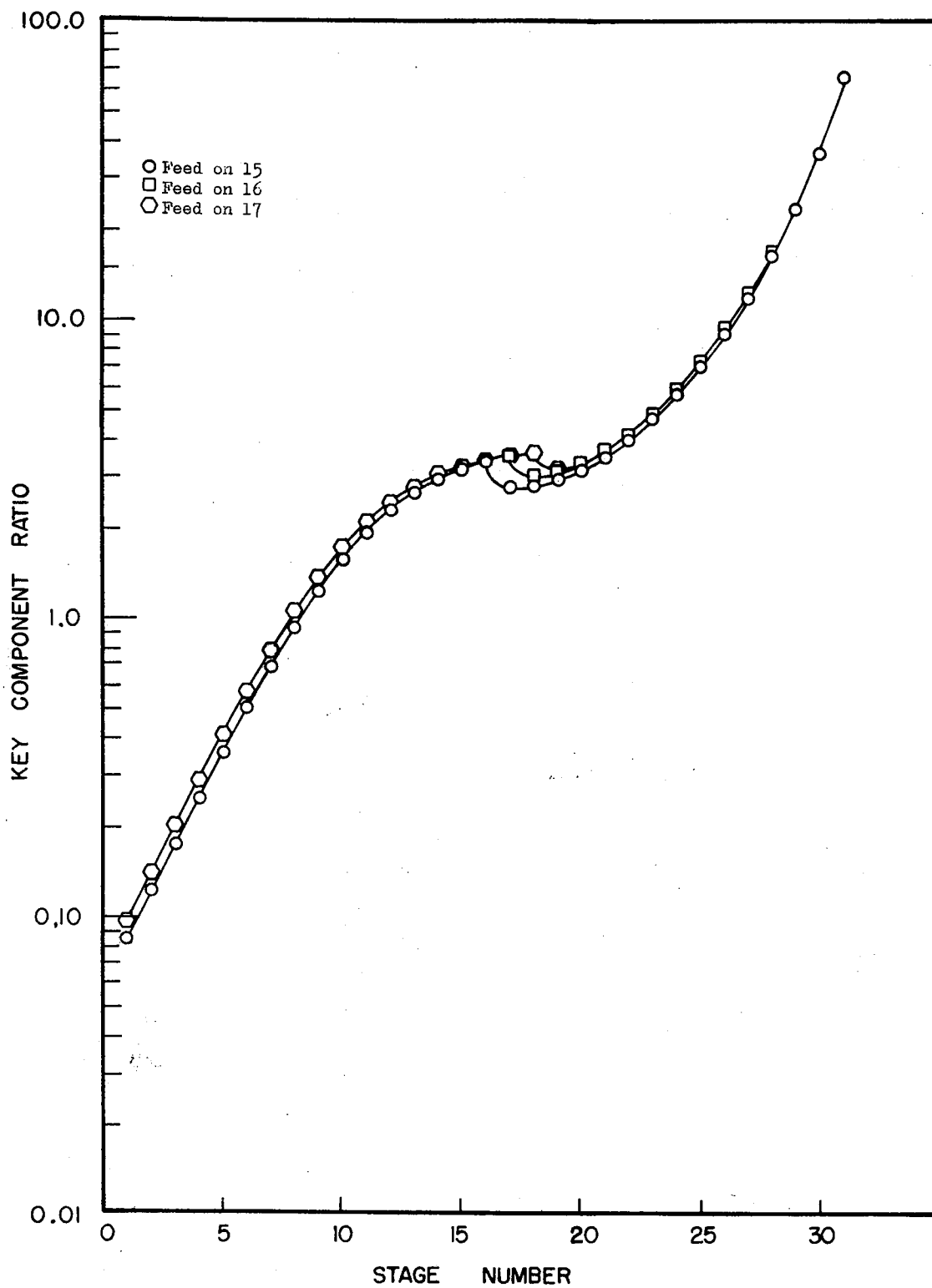


Figure 9. Relative Fractionation Curve- 30 Tray Depropanizer
With 25% Vapor Feed, $V/F = 1.54$, $D/F = 0.4854$

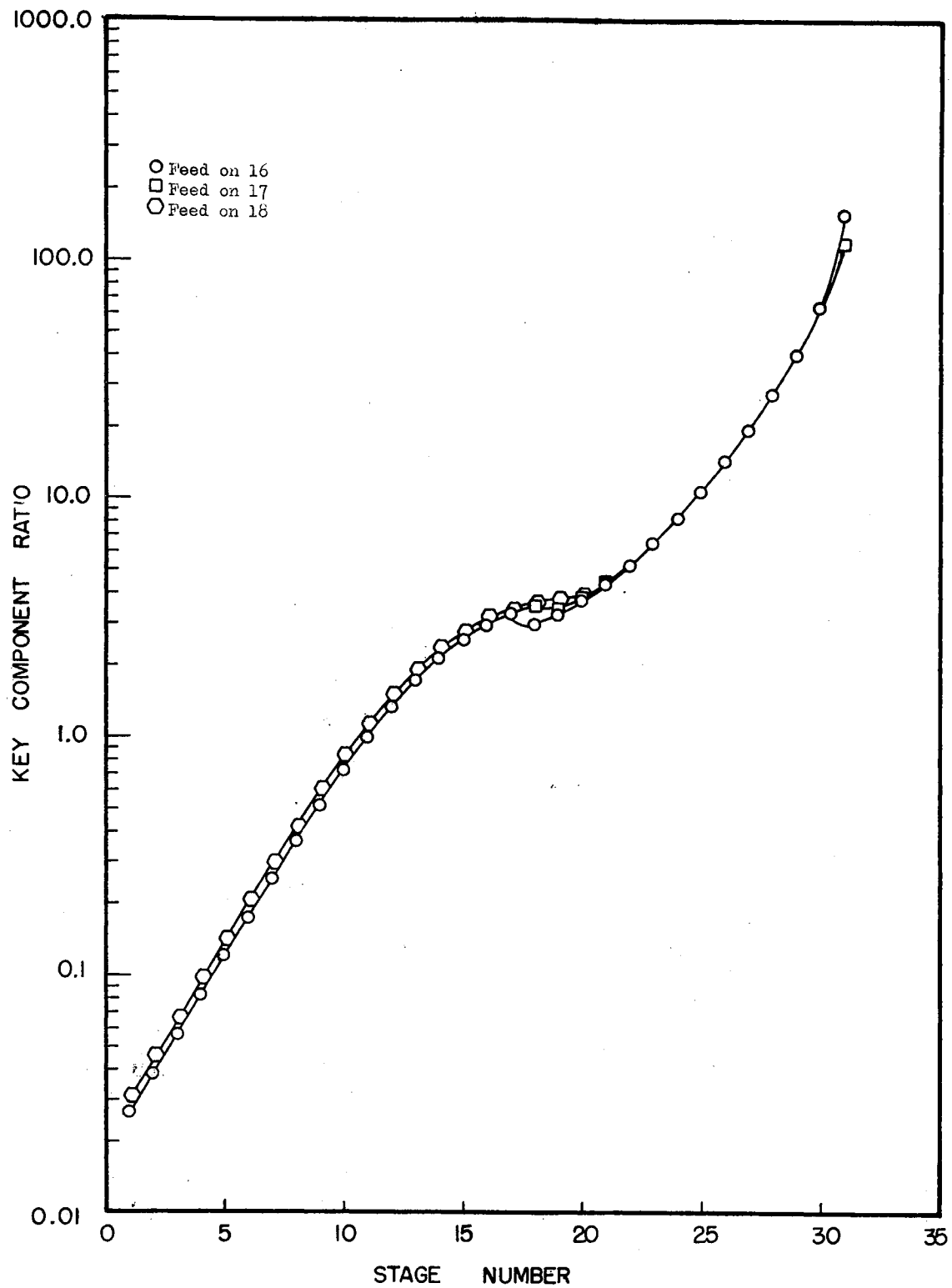


Figure 10. Relative Fractionation Curve- 30 Tray Depropanizer
With 25% Vapor Feed, $V/F = 1.68$, $D/F = 0.4854$

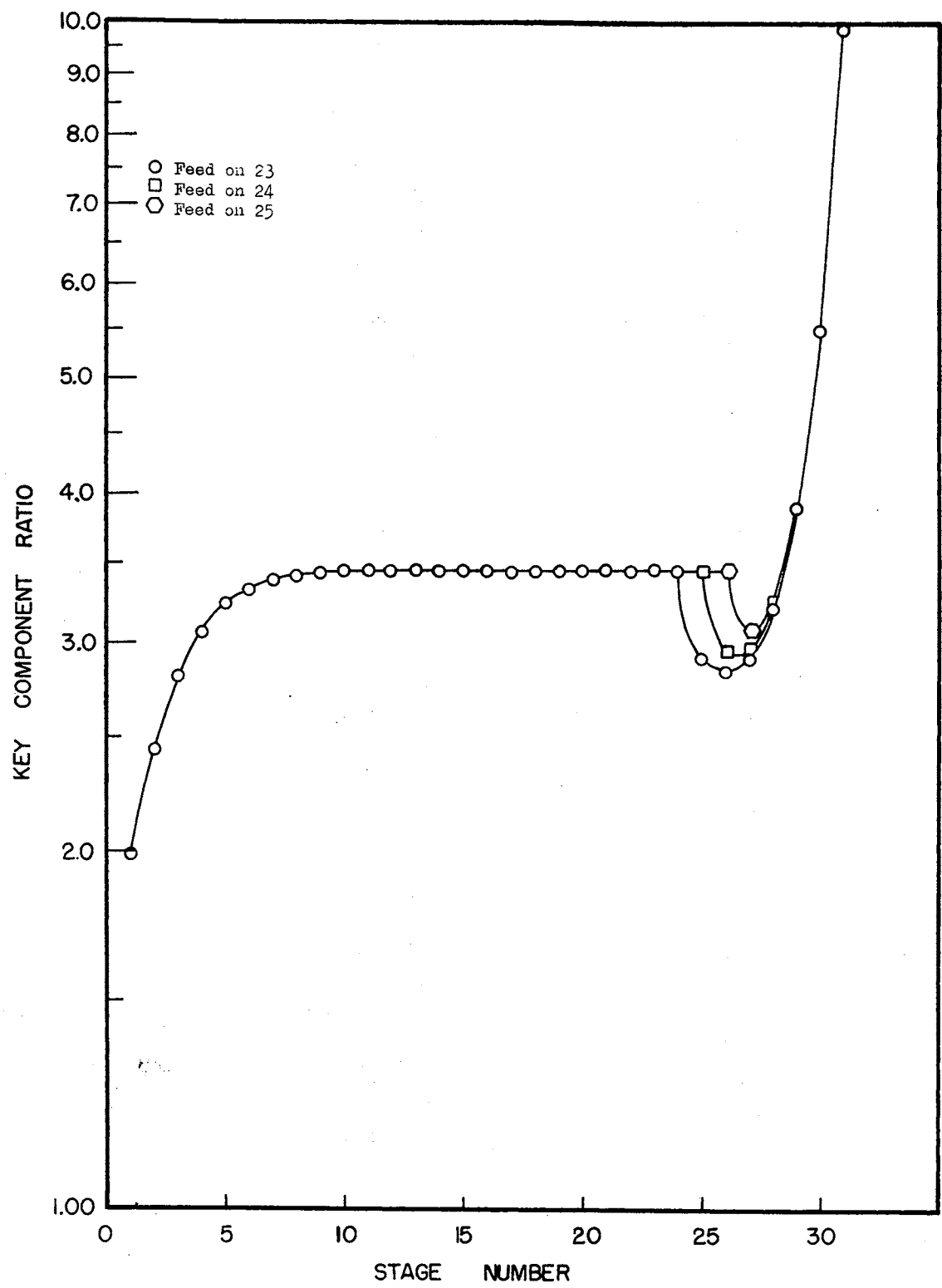


Figure 11. Relative Fractionation Curve- 30 Tray Depropanizer
 With 50% Vapor Feed, $V/F = 1.12$, $D/F = 0.4854$

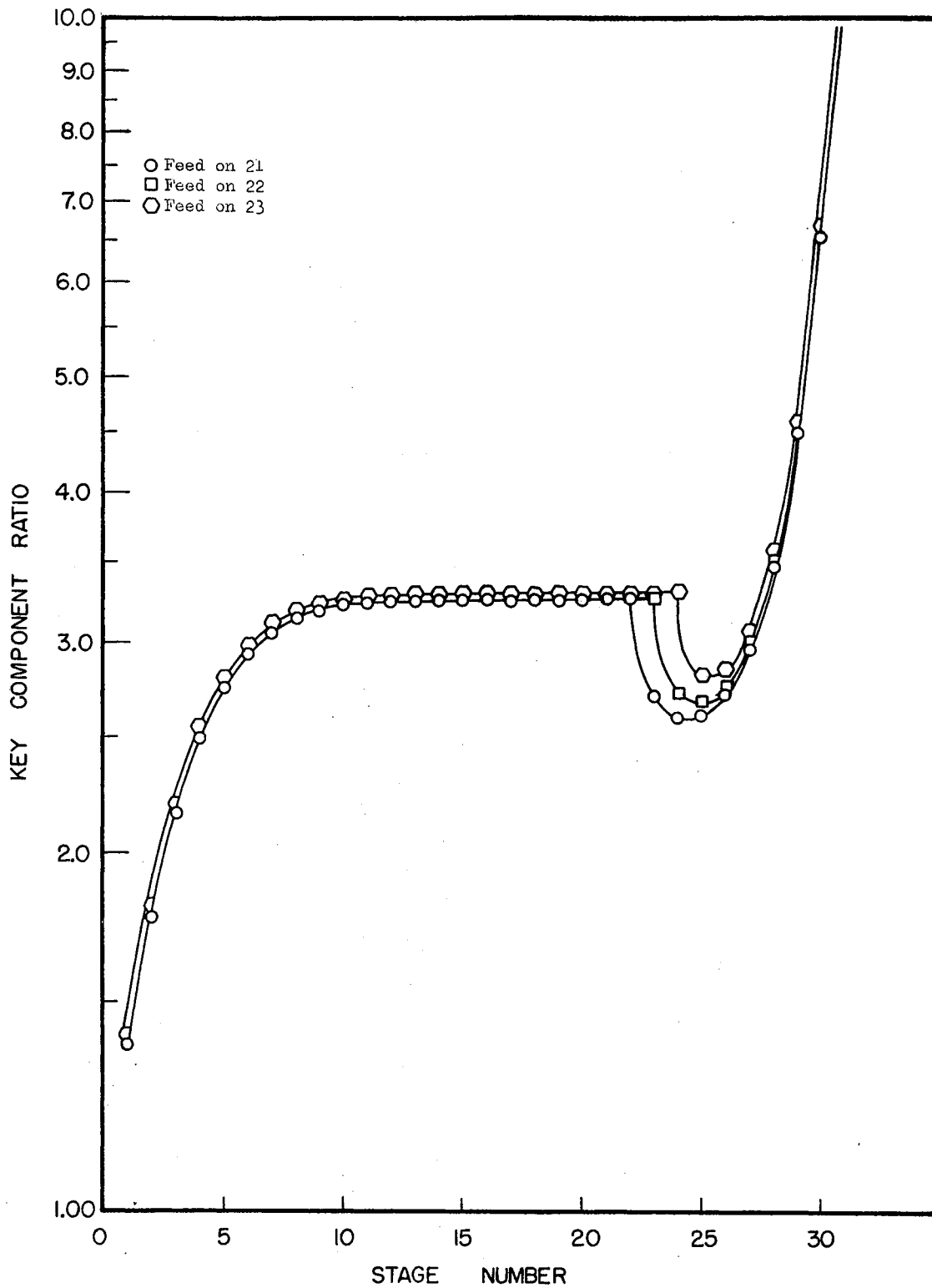


Figure 12. Relative Fractionation Curve- 30 Tray Depropanizer
With 50% Vapor Feed, $V/F = 1.26$, $D/F = 0.4854$

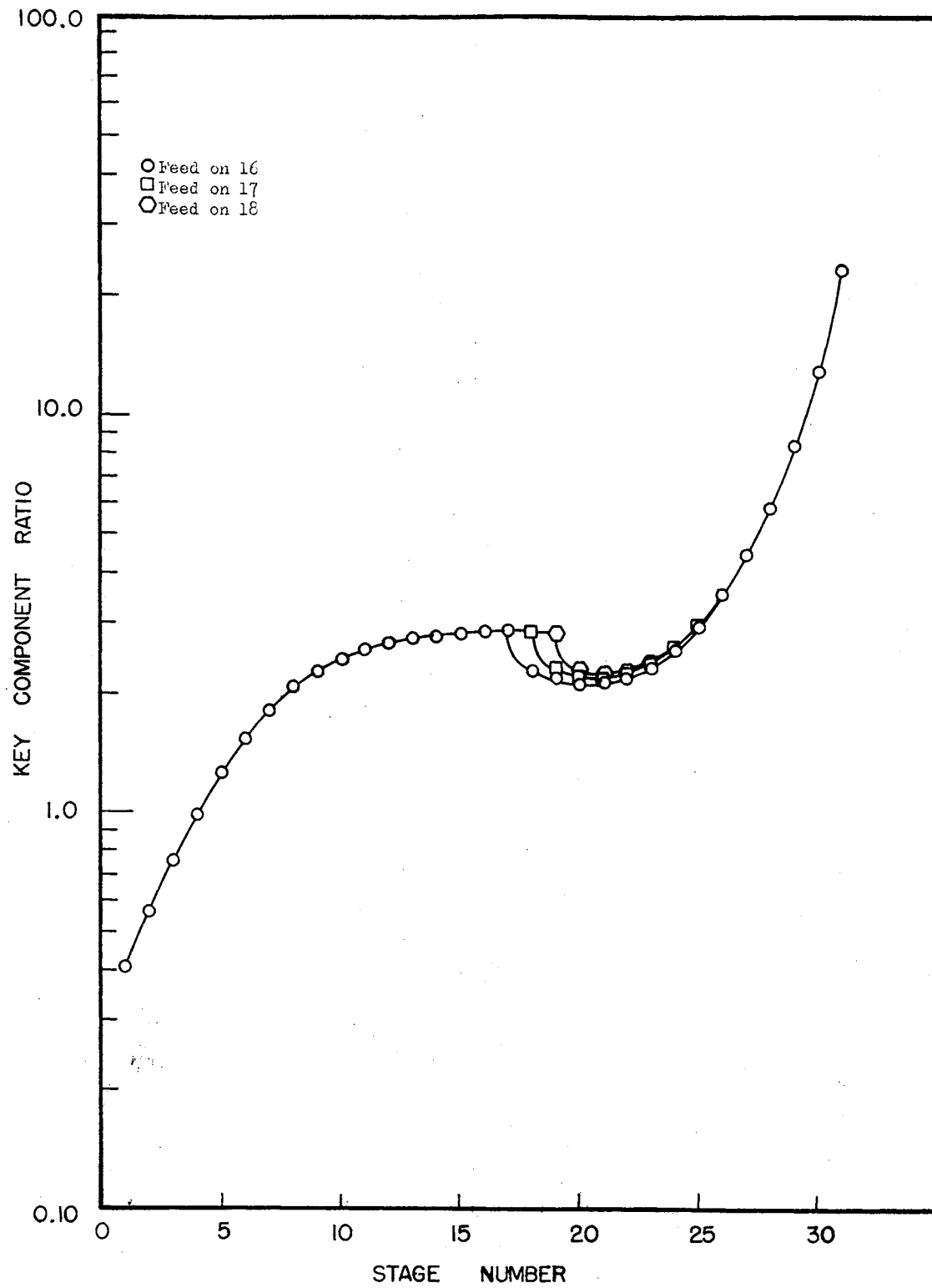


Figure 13. Relative Fractionation Curve- 30 Tray Depropanizer
With 50% Vapor Feed, $V/F = 1.40$, $D/F = 0.4854$

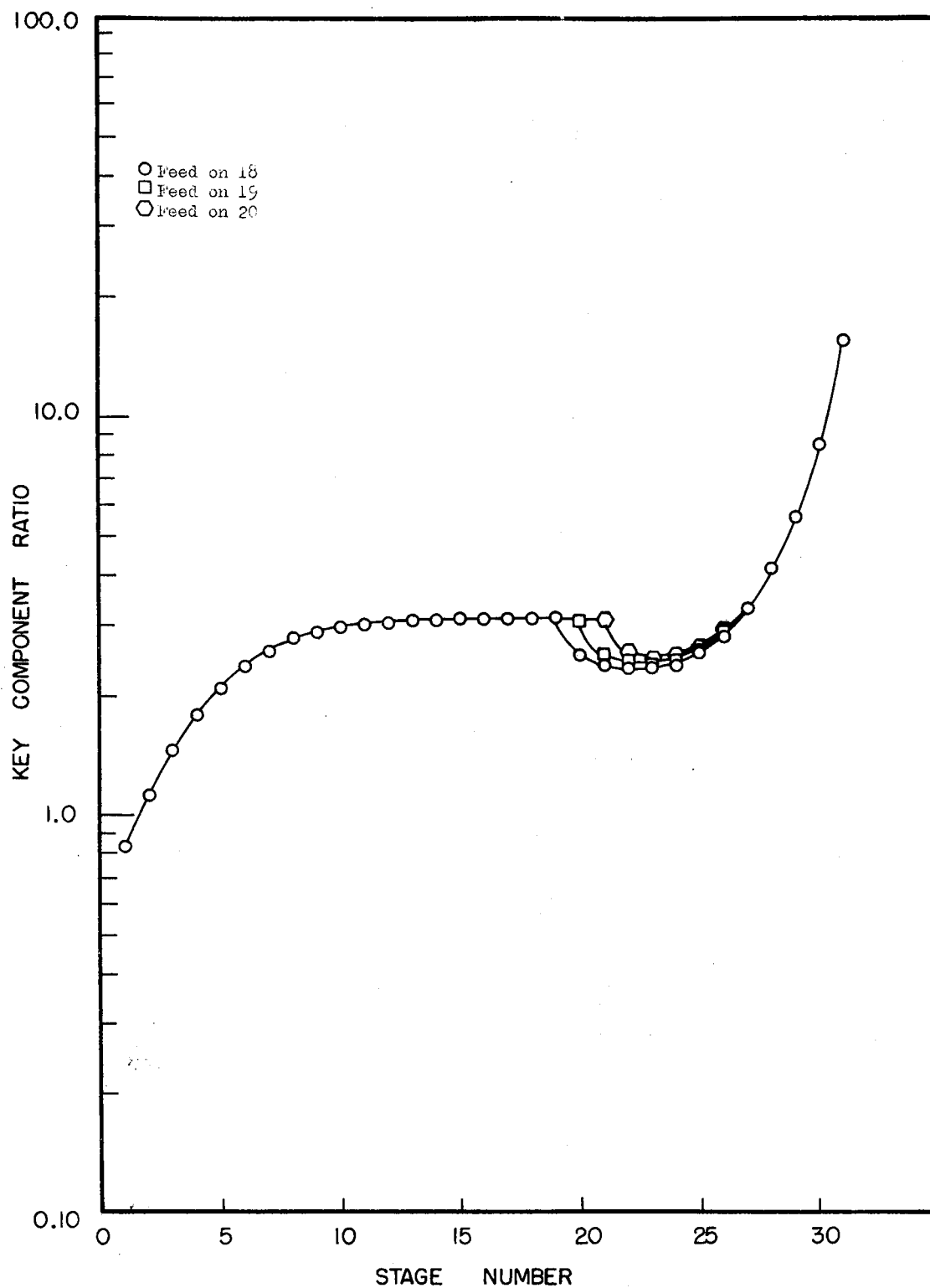


Figure 14. Relative Fractionation Curve- 30 Tray Depropanizer
With 50% Vapor Feed, $V/F = 1.54$, $D/F = 0.4854$

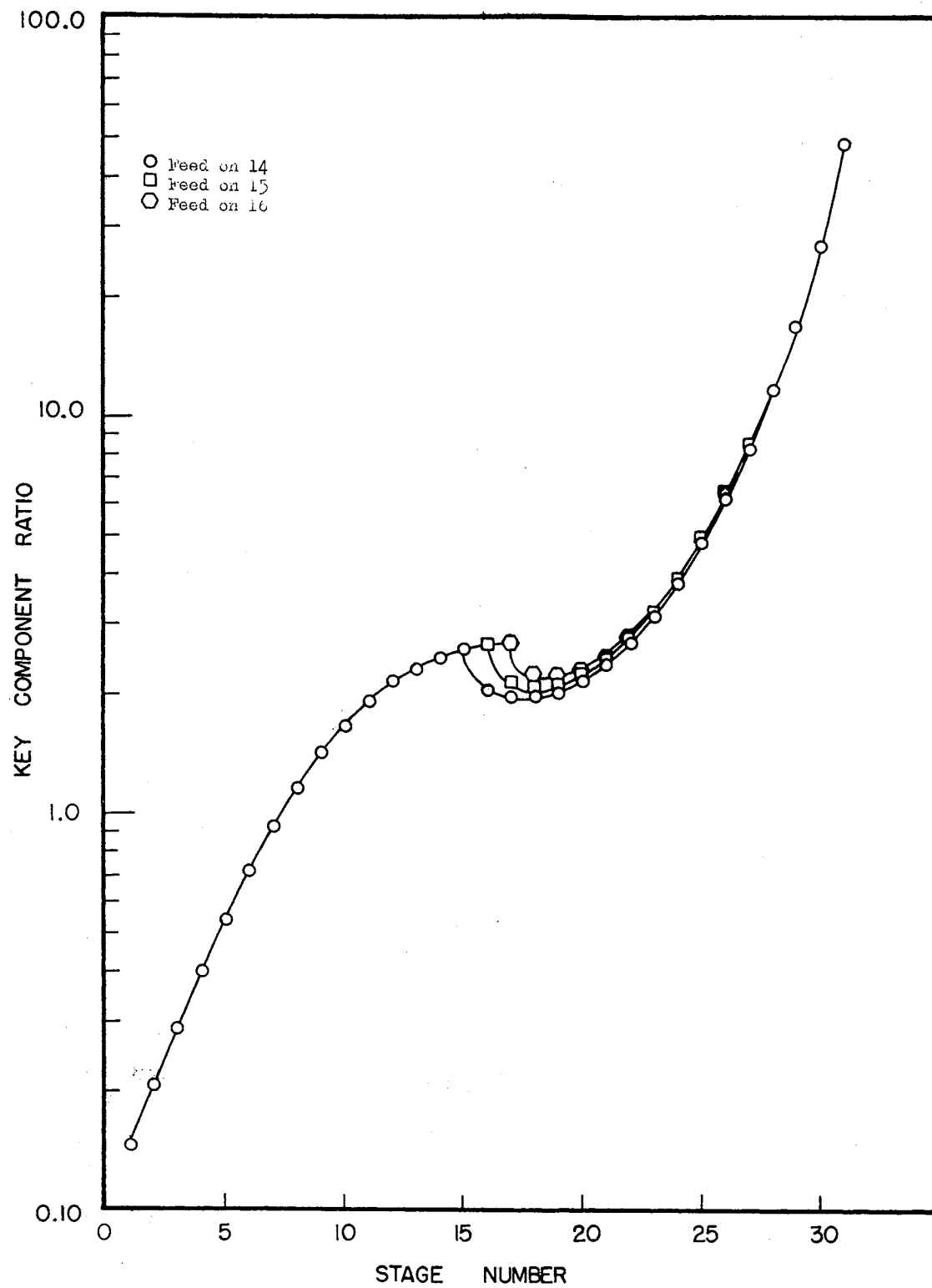


Figure 15. Relative Fractionation Curve- 30 Tray Depropanizer
With 50% Vapor Feed, $V/F = 1.68$, $D/F = 0.4854$

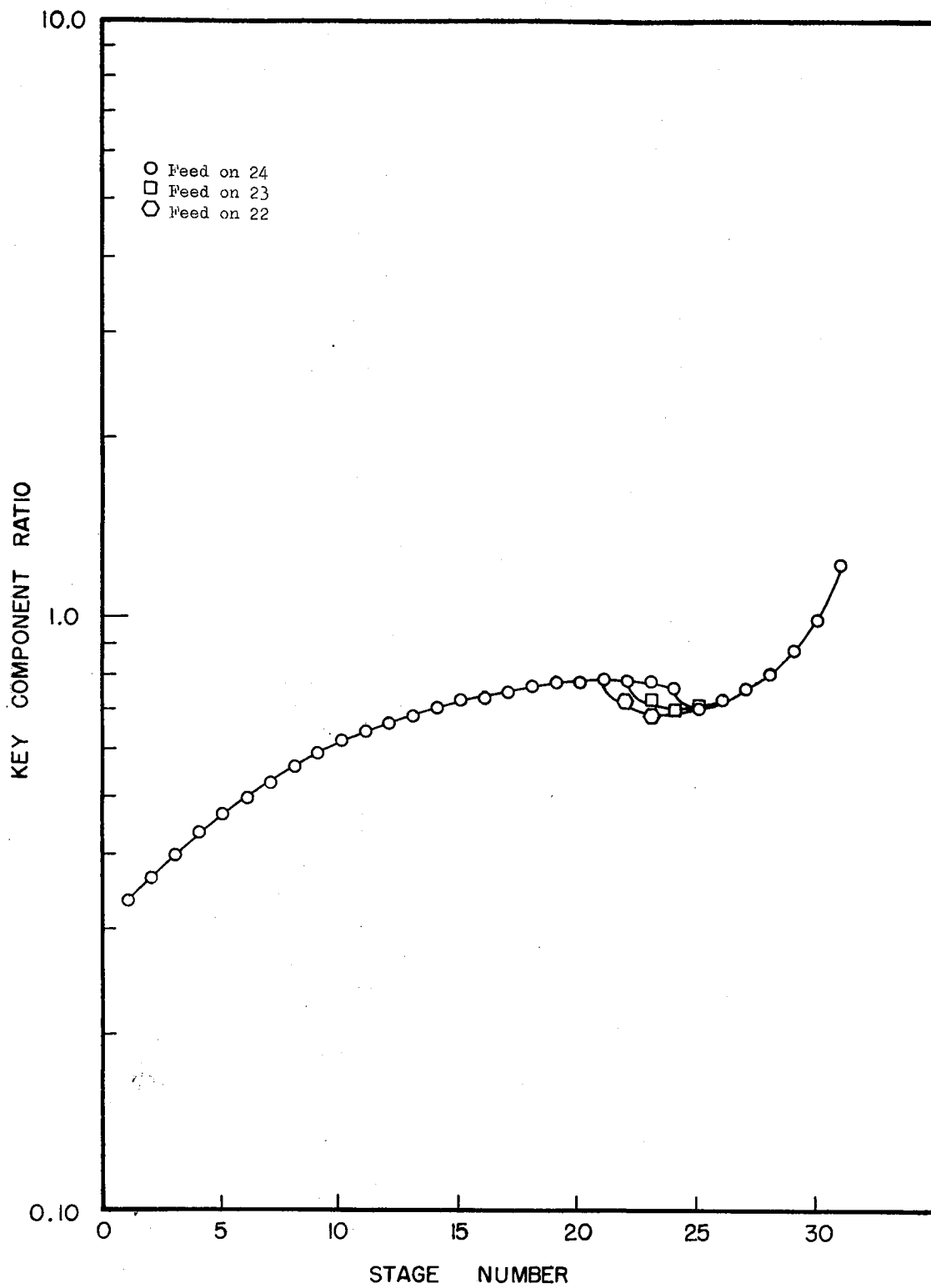


Figure 16. Relative Fractionation Curve- 30 Tray Wide Boiling Depropanizer With Bubble Point Feed, $V/F = 1.00$, $D/F = 0.45$

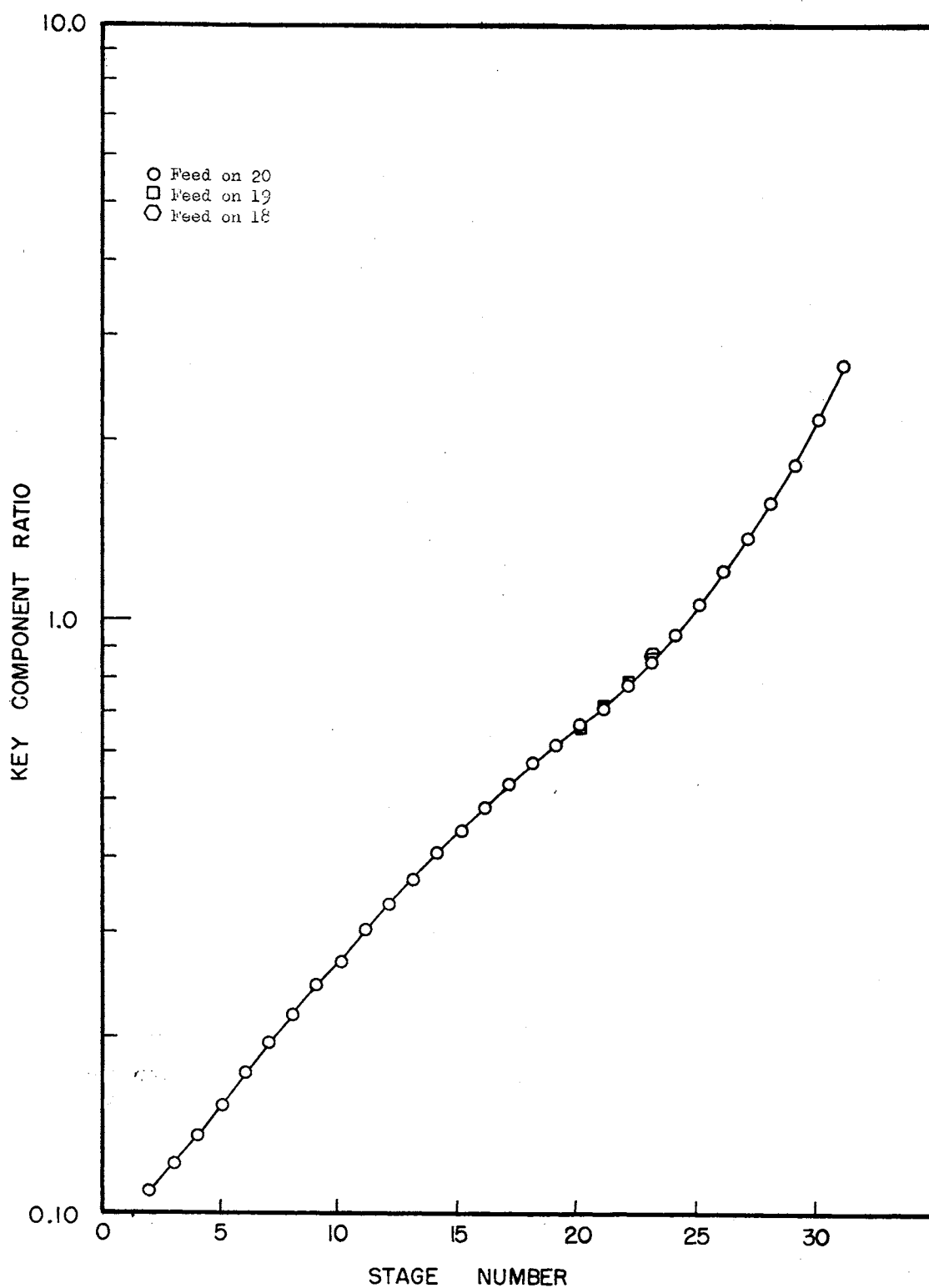


Figure 17. Relative Fractionation Curve- 30 Tray Wide Boiling Depropanizer With Bubble Point
Feed, $V/F = 3.00$, $D/F = 0.45$

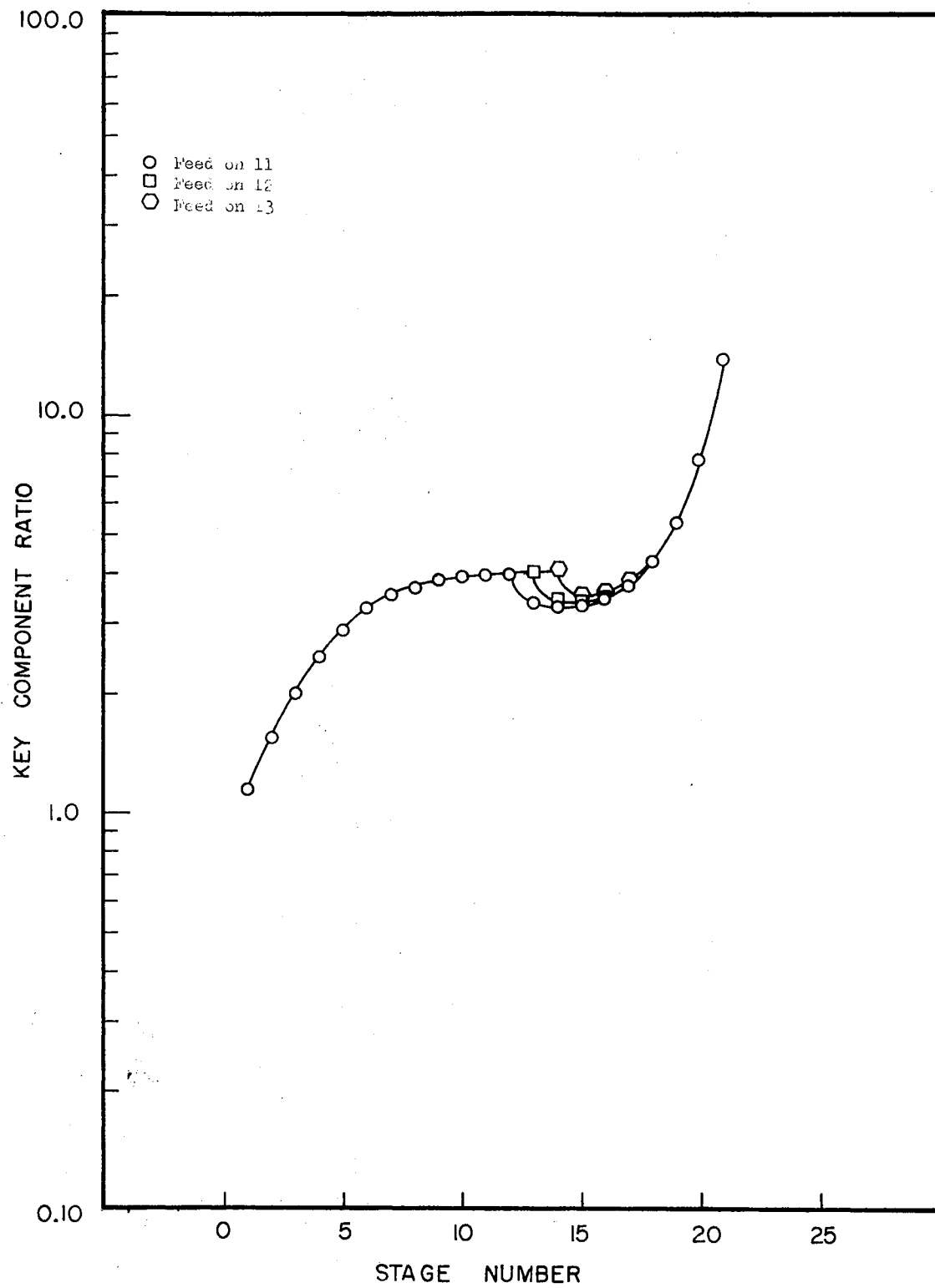


Figure 18. Relative Fractionation Curve- 20 Tray Depropanizer
With 25% Vapor Feed, $V/F = 1.12$, $D/F = 0.4854$

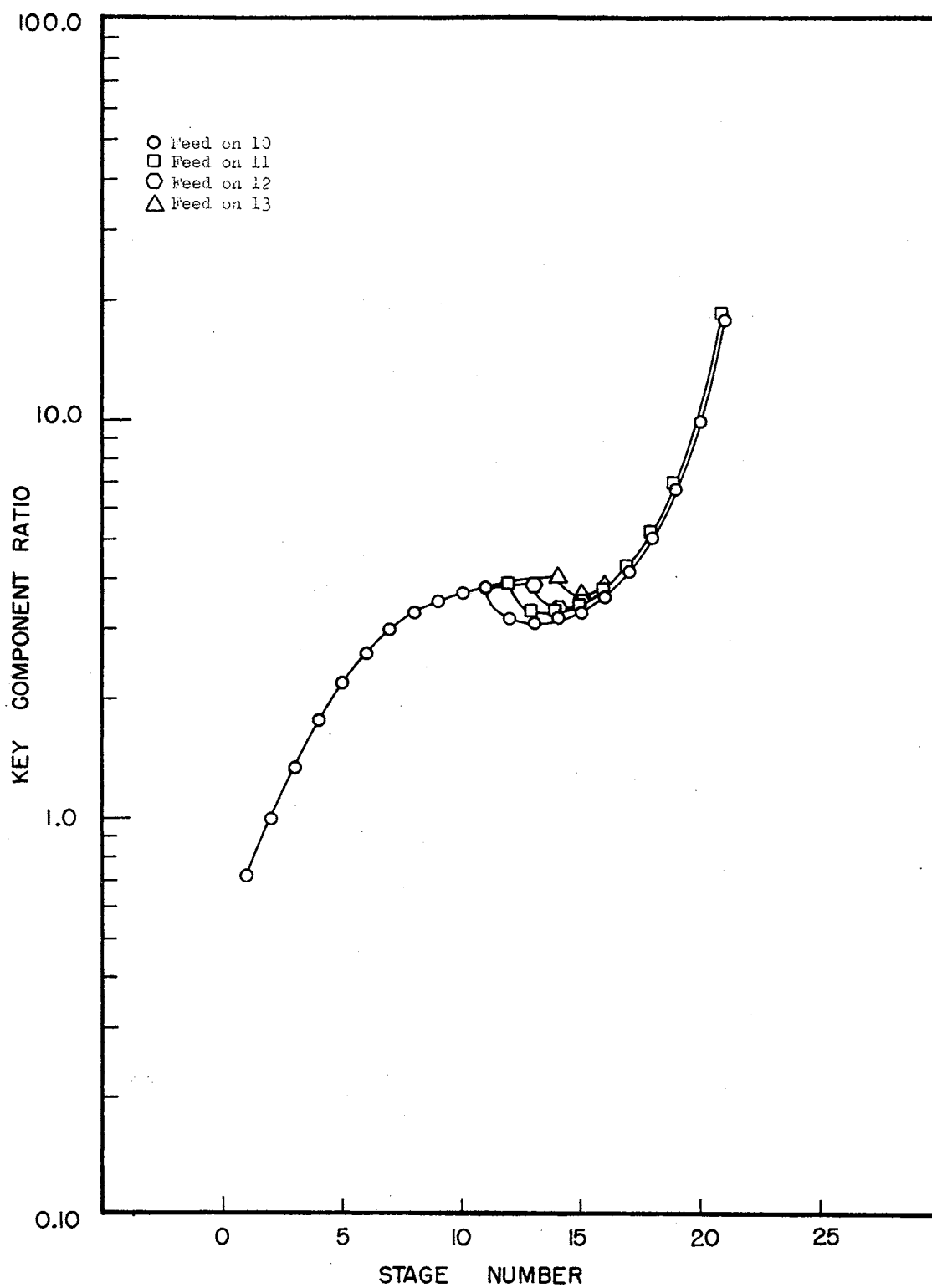


Figure 19. Relative Fractionation Curve- 20 Tray Depropanizer
With 25% Vapor Feed, $V/F = 1.26$, $D/F = 0.4854$

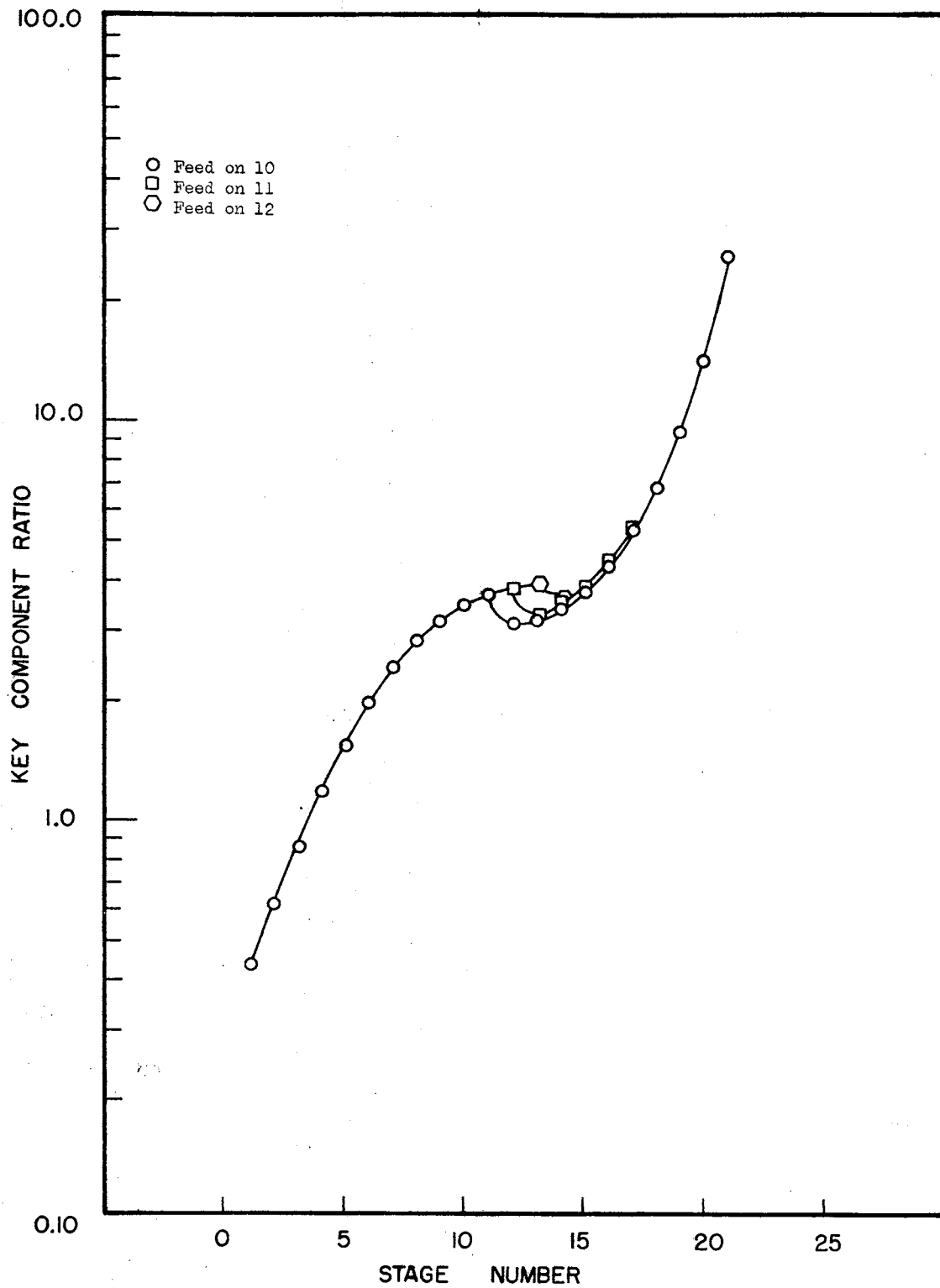


Figure 20. Relative Fractionation Curve- 20 Tray Depropanizer
With 25% Vapor Feed, $V/F = 1.40$, $D/F = 0.4854$

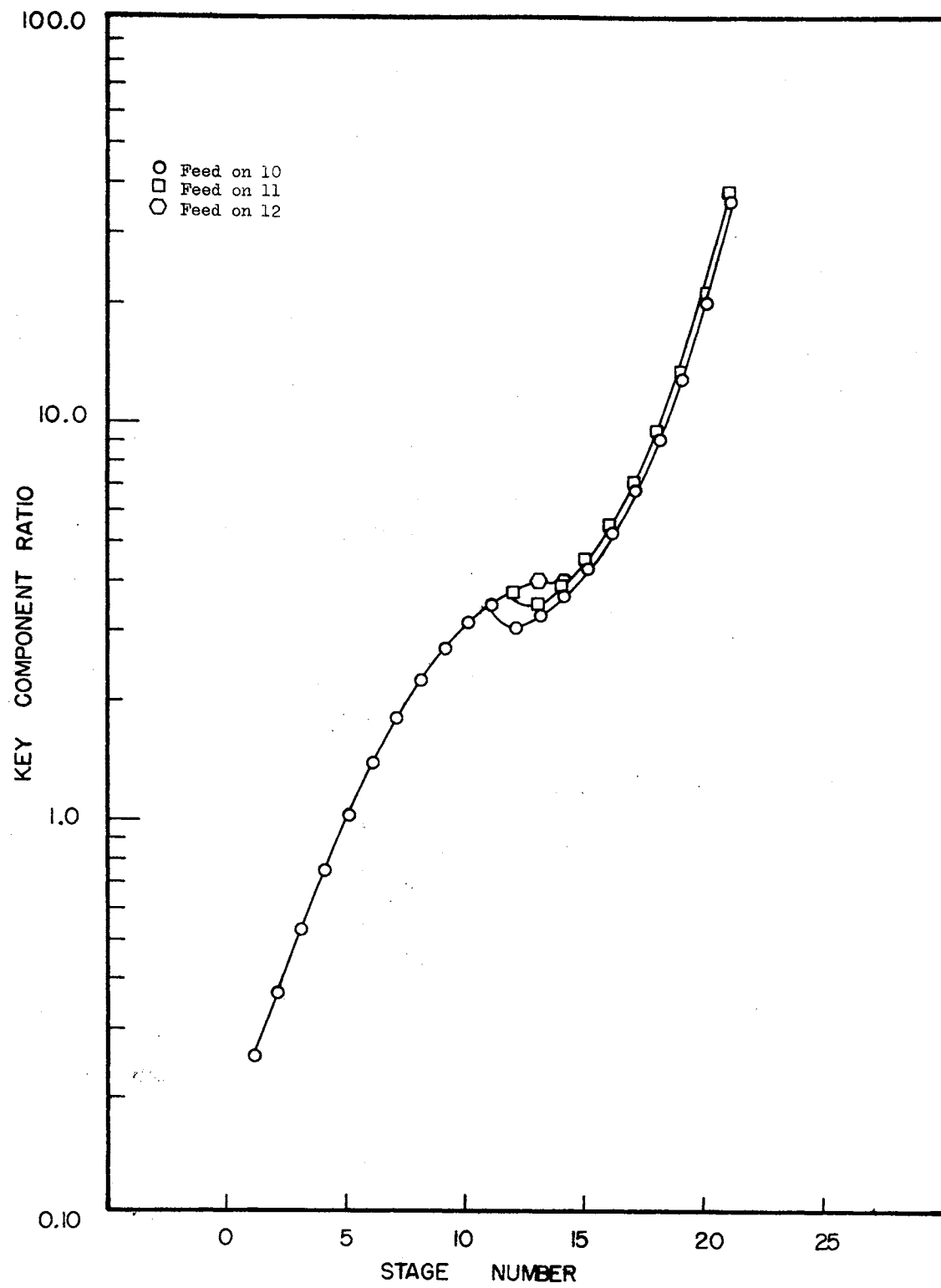


Figure 21. Relative Fractionation Curve- 20 Tray Depropanizer
With 25% Vapor Feed, $V/F = 1.54$, $D/F = 0.4854$

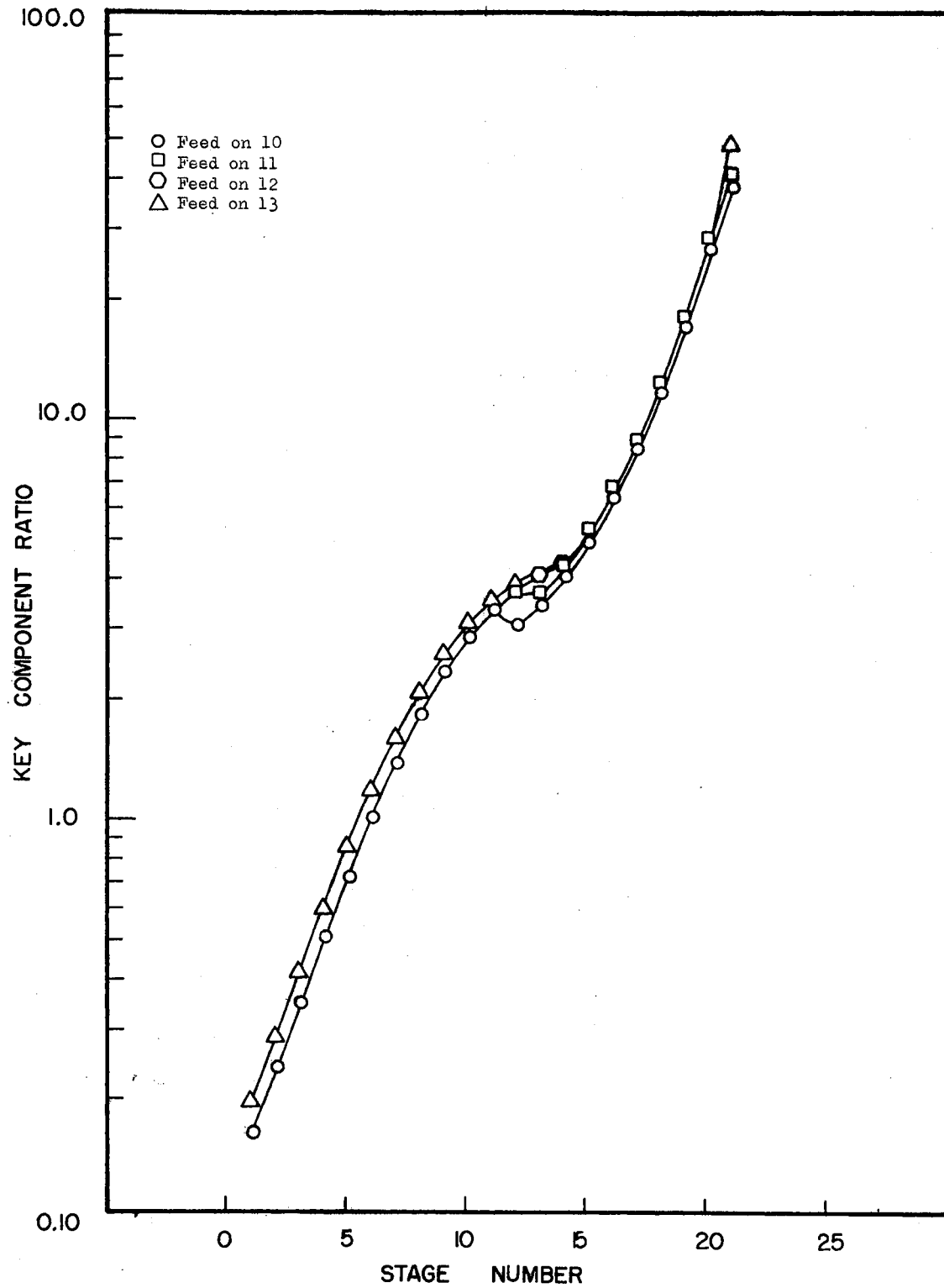


Figure 22. Relative Fractionation Curve- 20 Tray Depropanizer
With 25% Vapor Feed, $V/F = 1.68$, $D/F = 0.4854$

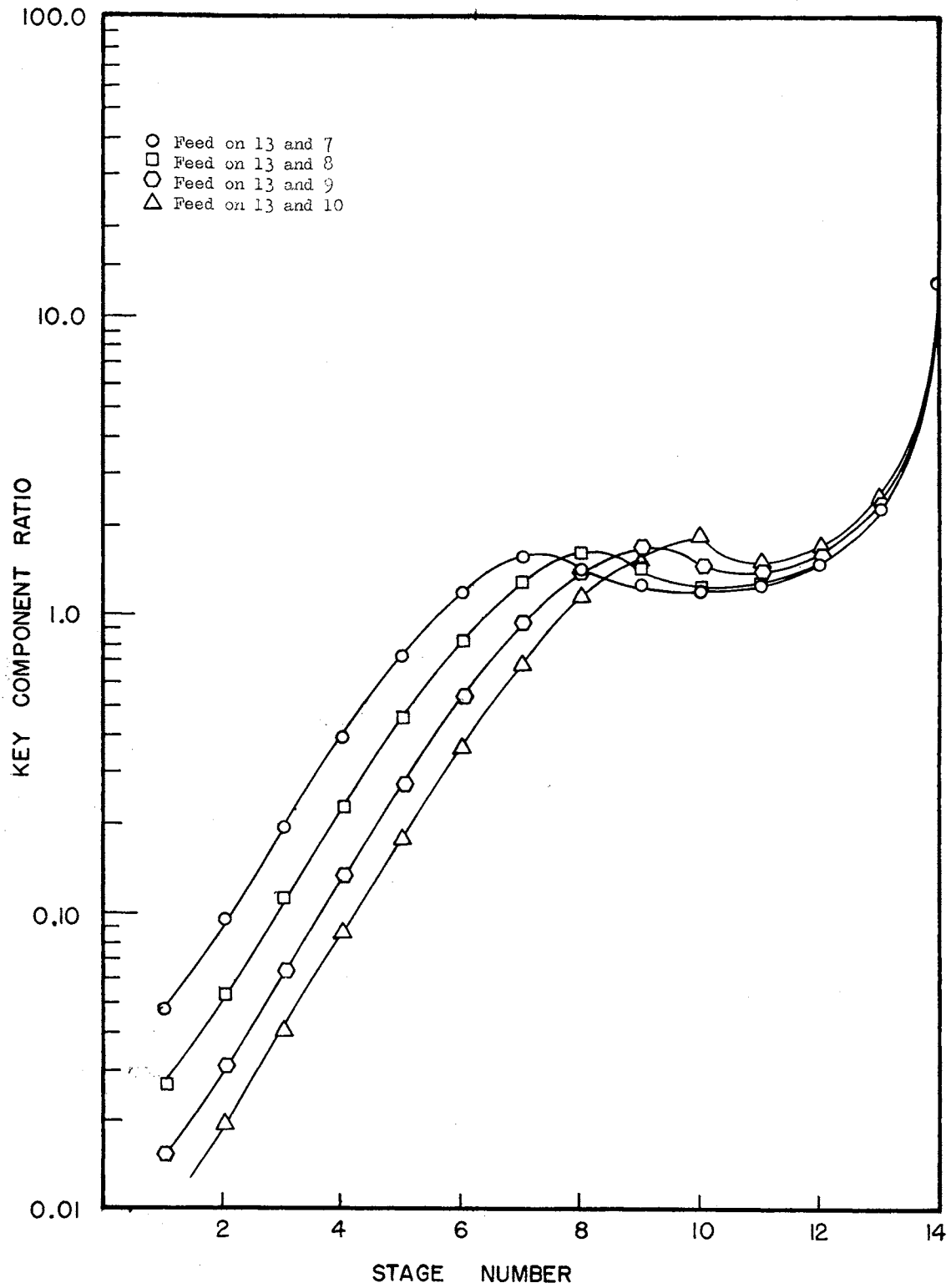


Figure 23. Relative Fractionation Curve- 13 Tray Demethanizer With Two Feeds

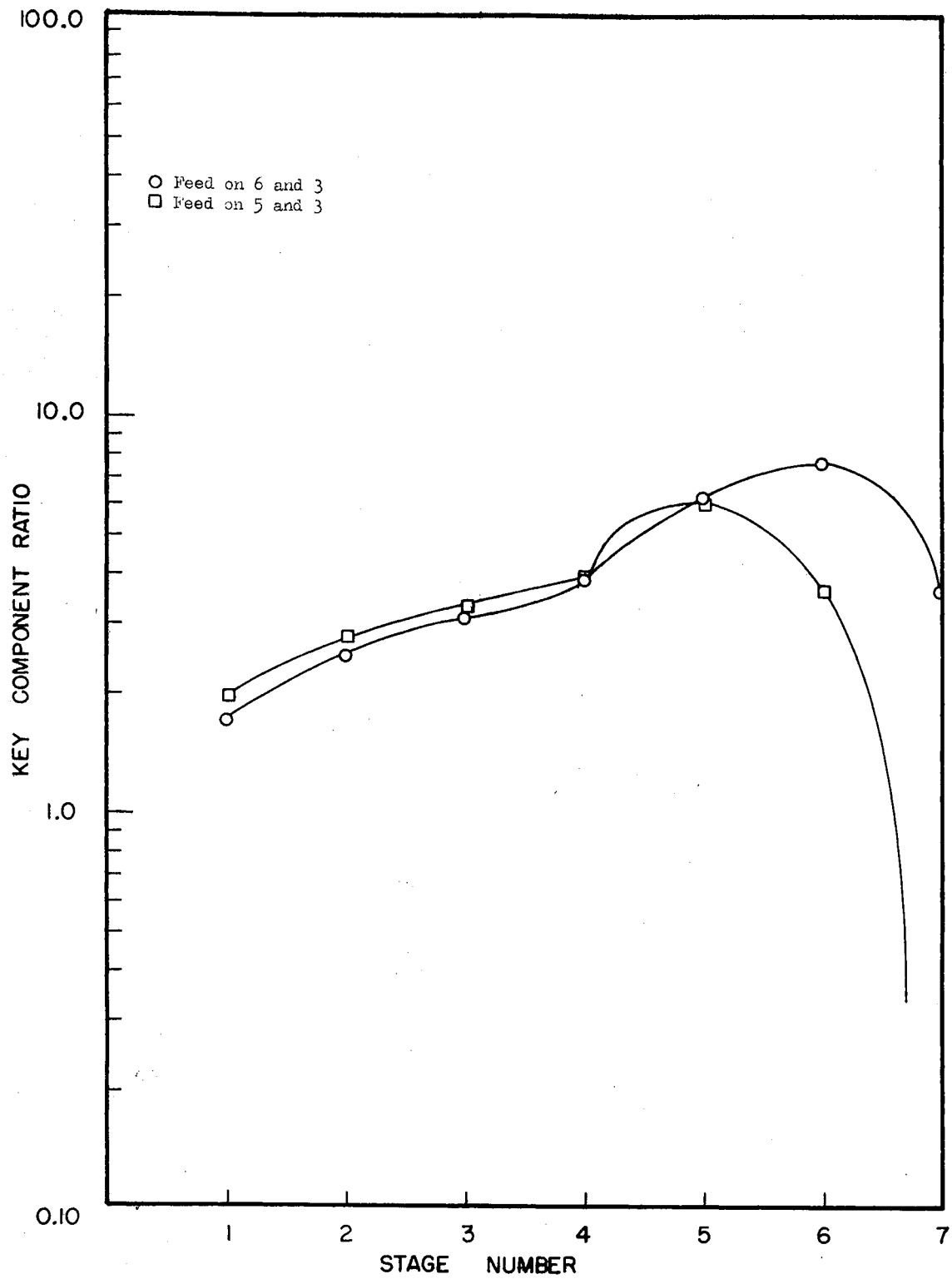


Figure 24. Relative Fractionation Curve- 6 Tray Stripper With Two Feeds

VITA

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