MULTICOMPONENT DISTILLATION WITHOUT EXTERNAL REFLUX

By

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Bachelor of Science

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1954

Submitted to the faculty of the Graduate School of
the Oklahoma State University in partial
fulfillment of the requirements
for the degree of
MASTER OF SCIENCE
August, 1958

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PREFACE

In this thesis I have attempted to add to the understanding of the variables which affect the design and operation of a nonexternally refluxed distillation column. Variations of pressure, number of stages, and fraction of the feed removed as overhead product were studied. The purpose was to determine the effect of pressure and number of stages upon the recovery from a distillation column operating without external reflux.

I sincerely appreciate the guidance and constructive criticism offered by Dr. Robert N. Maddox.

I wish to thank the Dow Chemical Company and the Continental Oil Company for their fellowship grants which made this work possible.

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CHAPTER I

THE PROBLEM

Statement of the Problem

Use of the non-refluxed distillation column has grown tremendously in recent years. Development of the high pressure distillate fields of the Gulf Coast has led to many new developments in natural gas processing. One of the new developments is the "low temperature" type liquid recovery system. In a low temperature plant, liquid from the primary separator is available at temperatures as low as $0^{\circ}F$. When these cold liquids are used as feed to a distillation column, sufficient internal reflux is generated to provide a reasonably sharp separation and satisfactory raw gasoline product from the bottom of the tower.

The non-refluxed tower may be described as the lower portion or stripping section of the normally envisioned distillation column. A schematic diagram of a typical non-refluxed tower is shown in Figure I.

In the design of distillation columns for use in the natural gas industry certain variables have become fairly well accepted as being functions of the desired separation. Some of these variables are pressure, overhead and bottoms

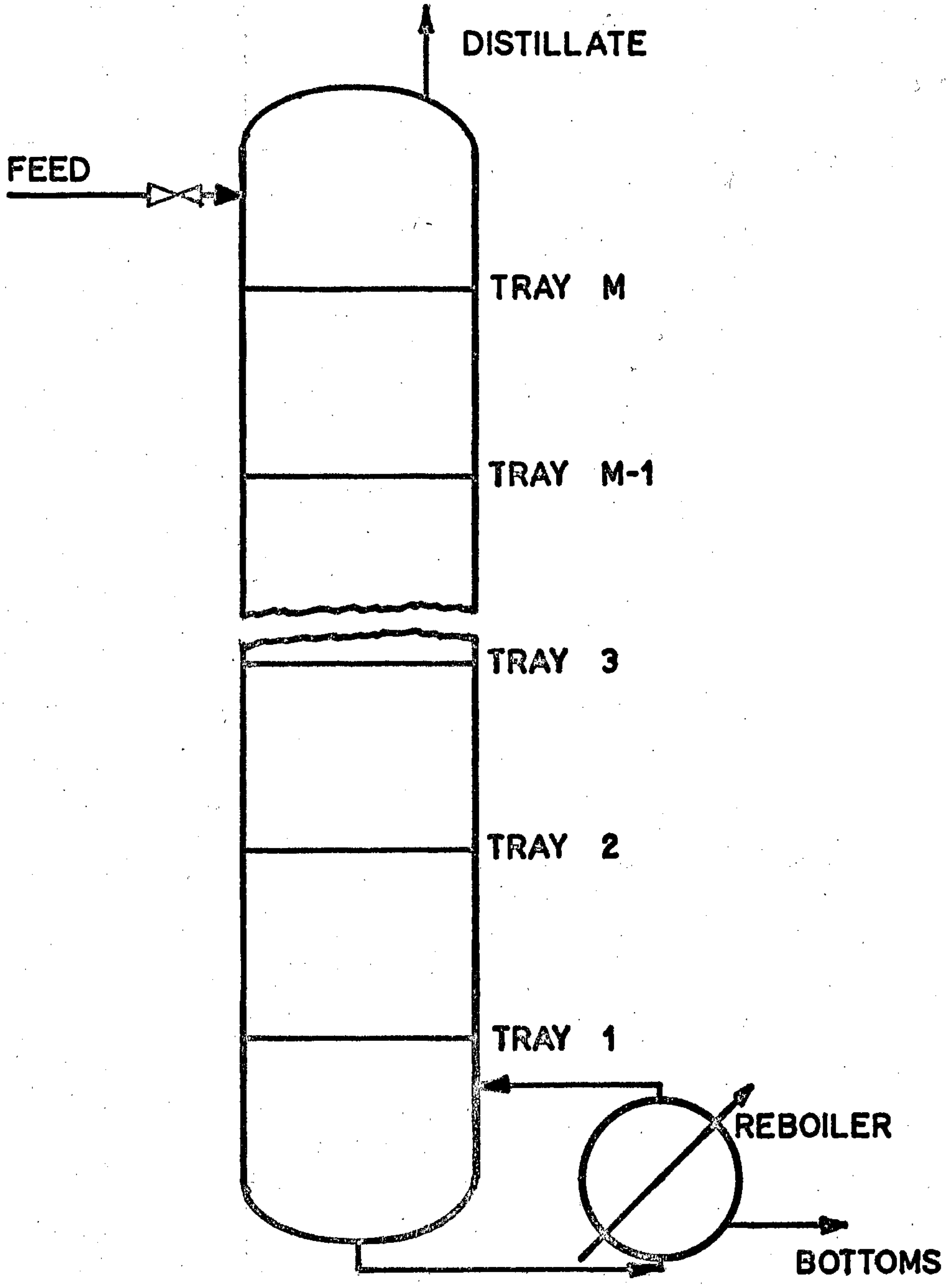


FIGURE 1

SCHEMATIC DIAGRAM OF A NON-REFLUXED DISTILLATION COLUMN temperatures. Although relative volatility data is a function of both temperature and pressure, the effect of pressure is very small. It can then be said that the relative volatility data can be considered as a function of the separation. Purity specifications for the products have been established for the industry. Composition of the feed going to a column making given separation are fairly uniform. These coupled with the desire to achieve the maximum economical recovery, indicate that the minimum number of stages and minimum reflux ratio can be considered as a function of the separation the tower is designed to make.

If these generalizations are applied to the Fenske (4) relationship

$$\alpha^{S_m} = \left(\frac{X_d}{X_w}\right)_{LK} \left(\frac{X_w}{X_d}\right)_{HK} \qquad (1)$$

it is readily apparent that α^{S_m} must equal a constant which is a function of the separation. If either $\left(\frac{X_d}{X_w}\right)_{LK}$ or $\left(\frac{X_w}{X_d}\right)_{HK}$ is considered constant for a given separation, the other must also be constant for the same separation. It can then be said that for each separation there are pseudo "equilibrium" constants which will describe the separation.

It can be shown that the flash equation as applied to the two key components will resolve to the Fenske relationship with a minimum number of stages equal to unity. The separation of the non-refluxed tower may be described by application of the appropriate pseudo equilibrium constants and the flash calculation.

Rigorous plate to plate calculations are carried out by means of a trial and error procedure. It is necessary to assume a composition for either the overhead or bottoms product. The composition of the second stream may then be determined by a material balance. These assumptions are used as the basis of plate by plate calculations from the condenser and reboiler to the feed plate. If the feed plate composition from the two sets of calculations is not the same, adjustments are made in the assumed product distribution and the procedure outlined above repeated. It is readily apparent that this is a time consuming procedure when the calculations are performed manually.

The relaxation technique should apply to the non-refluxed tower as well as to the more conventional tower operating with external reflux. It is difficult to predict the product composition for a non-refluxed tower because all components in the feed are present in the overhead product.

The purpose of this investigation was to study the effect of pressure and number of stages upon recovery in a non-refluxed tower. A secondary aim was the development of a technique to reduce the calculation time necessary to predict a satisfactory performance for the non-refluxed tower.

Limitations of the Study

The study was confined to an analysis of the effects of pressure and number of stages upon the feed composition of Table I.

TABLE I

FEED COMPOSITION

Component	Moles/Hr.	Mol. Percent
C_1	18.63	11.70
C_2	14.92	9.36
C_3	34.70	21.78
iC_4	10.30	6.46
nC_4	30.87	19.38
iC_5	13.04	8.07
nC_5	9.28	5.86
$C_{6+}(C_7)$	27.71	17.39
Total	159.45	100.00

It was further limited to three distillate rates expressed as fraction of feed going overhead.

The pressures used were seventy-five, two hundred, and three hundred psia and the number of stages five, seven, and ten theoretical stages. The distillate rates were 0.21, 0.30 and 0.50 fractions of feed as distillate which correspond to approximate de-ethanization, depropanization, and de-isobutanization respectively.

CHAPTER II

REVIEW OF LITERATURE

The field of multicomponent distillation is replete with design techniques. These techniques may be divided into two broad general classes: (1) the rigorous tray to tray techniques and (2) the "short-cut" techniques. While all techniques differ somewhat in their approach, certain basic assumptions must be made in each procedure. These basic assumptions are: (1) steady state operation, ie, the feed input rate and the product removal rates are constant, and (2) no heat of mixing of any of the components (9).

One of the most widely known rigorous methods is the Lewis and Matheson (6) tray to tray calculation. This method is based on Sorel's method together with the basic assumptions of multicomponent distillation. With the assumption of a reflux ratio and by the application of Hausbrand's operating lines, equilibrium, material, and energy balances, complete tray compositions and traffics may be established for each tray in the column. The number of trays required to effect a given separation at the assumed reflux ratio is found by determining the tray on which the liquid composition best approximates the feed plate composition for both the stripping and rectifying section calculations, with the required number of trays being the summation of the number of trays in each section. The achievement of proper feed plate match is usually accomplished by arbitrarily introducing the non-distributed components at various points in the column. The points at which they are introduced are based upon judgement, relative column position and the physical properties of the components.

Determination of the minimum number of stages or minimum reflux ratio to effect a given separation involves a procedure similar to the calculation sequence for a finite number of trays and reflux ratios. The minimum number of trays will occur at total reflux, so it is assumed that all of the overhead product is returned to the column as reflux (9). The calculations are analogous to those mentioned. As the minimum reflux ratio occurs at an infinite number of plates, it becomes somewhat more difficult to determine this value by the tray to tray procedure. The derivation of short-cut techniques relies upon the assumption that in multicomponent distillation a zone of constant composition occurs in both the rectifying and stripping sections. The appearance of this zone at the feed plate indicates the minimum reflux ratio has been attained. This assumption is based on a proven occurence in ideal binary mixtures. The calculation sequence is the same as mentioned before but involves a trial and error method of assuming a reflux ratio and checking to determine if the constant composition zone develops in the proper place.

The tray to tray calculation provides a complete picture of the internal operation of the column and gives the engineer data necessary for column design. The major disadvantage is the extended time required for manual solution, usually on the order of thirty minutes per component per tray.

As a result of the time requirements necessary for the solution of the tray to tray calculation, so called "short-cut" techniques have been developed. In addition to the assumptions mentioned, the short-cut procedures further assume constant molal overflow which involves the tacit assumption of equal molal latent heats of vaporization, and, when applied to multicomponent distillation, a constant relative volatility of the components. There are graphical methods (11) for the solution of multicomponent mixtures, but these as a general rule are quite complex and time consuming, and are not considered in this study.

One of the better known "short-cut" techniques for predicting product compositions is that proposed by Fenske (4). From a consideration of conditions in a tower operating at total reflux, he derived Equation 1. The product distribution of any feed stream can be determined by the proper substitution of relative volatilities and feed components.

The determination of the minimum reflux ratio can be accomplished by any one of several methods. Gilliland (5) has proposed an equation for the determination of minimum reflux based upon the appearance of a zone wherein the ratio of the concentration of the key components remains constant. Maxwell (7) has modified this equation, taking into consideration the influence of feed condition and the influence of feed components other than the light and heavy keys.

The calculation of the finite operating stages and reflux ratio from the minimum values is based upon a correlation set forth by Gilliland (8). This correlation is based upon the assumption that there is a definite relationship between operating and minimum reflux ratio and operating and minimum trays. This correlation is usually presented in the form of a graph.

The Fenske relationship coupled with the Maxwell minimum reflux ratio equation and Gilliland correlation present an excellent design technique. This method gives a very good combination when considering the time spent in solution of the problem, reliability of results, and applicability of the assumptions made in the problem solution.

The advent of the high speed digital computer has to a large degree eliminated the time factor in the solution of the tray to tray procedure. Bonner (2,3) has developed a plate to plate calculation routine for the I.B.M. 650 Digital Computer. This routine follows basically the Lewis and Matheson approach to multicomponent distillation. Certain small modifications were made in the calculation sequence to facilitate machine calculation. The routine based upon a given number of stages, feed plate location, reflux ratio, and percentage overhead product will calculate the overhead and bottoms compositions, and a heat balance around the entire column. Supplementary calculated data includes temperature and vapor rate profiles and, if desired, individual tray compositions. It has been estimated that the time required for a computer solution is approximately four and one half hours, including data set-up time, for an eight component feed, forty tray column. The approximate time required for manual solution of the same problem would be about 185 hours.

The choice of vapor-liquid equilibrium constants to be used in multicomponent distillation is in itself a major problem. Vapor liquid equilibrium of the individual components may be affected by variation in composition, polarity of the components, and the physical conditions to which the system may be subjected (9).

Many investigators have published equilibrium charts, however, failure to recognize the numerous factors affecting vapor-liquid equilibrium has impaired the accuracy of some of these charts. Perhaps two of the most reliable for use in the field of petroleum distillation are the Natural Gasoline Association of America charts prepared by the Fluor Company (13) and the Kellogg Charts prepared by the M. W. Kellogg Company (1). Each system recognizes the composition effect on individual component vapor-liquid equilibrium and attempts to present a generalized technique for a fairly rapid and accurate determination of these constants.

In the Fluor method, composition affects are taken into account by means of the convergence pressure principle. This principle is rigorous for binary mixtures. For multicomponent mixtures the assumption is made that the convergence pressure is that pressure determined by the system temperature and the critical locus curve between the critical temperature of the lightest component and the critical of the remaining mixture. Several charts for the rapid estimation of convergence pressure for different mixtures are presented in the N.G.A.A. Tables.

The Kellogg charts are based on a reduction of an empirical relationship of the thermodynamic properties of light hydrocarbon mixtures. Their use is somewhat more difficult than the Fluor charts. It is necessary to assume a molal average boiling point of the stream to be calculated in order to determine the equilibrium constants and the molal average boiling point of the calculated stream must approximate the assumption. Thus in a flash vaporization calculation, a multiple trial and error, V/L and molal average boiling point, type of calculation is encountered.

The selection of sources for enthalpy and physical property data is an arbitrary choice. In the choice the source of enthalpy values, because of the nature of the problem, it was necessary to select a source having the correction for pressure variation as an integral part of the data. Because of this requirement, the Western Supply Charts (12) were used as a source of enthalpy data. The Natural Gasoline Supply Men's Association Manual (14) was used as a source of molecular weights and vapor pressures.

CHAPTER III

EXPERIMENTAL PROCEDURE

Plate to Plate Runs. The relative volatility of each component was calculated at three equally spaced temperatures for three different pressures. The vapor-liquid equilibrium constants were read from the 5,000 psia convergence pressure charts of the N.G.A.A. Charts. The latent heats of vaporization of the individual components were calculated from enthalpy data read at the same temperatures and pressures as the equilibrium constants. The specific heat of the individual components was read at these same temperatures. The enthalpy data and relative volatilities were then fitted to a curve of the type $A + BT + CT^2$ by a least squares technique.

The feed composition, control constants, and allied data were assembled with machine calculated constants for the relative volatility and enthalpy equations of a particular pressure. This constituted the primary data deck. Appropriate cards containing the desired V/F and D/F ratios and number of stages were added to the primary data deck. This final data deck was then placed in the plate to plate deck. The console switches were set to their proper positions and the program deck with data was read into the machine. During the computing cycle, the trial summary cards were checked to determine if the problem was converging properly on the feed plate composition and to make sure that the specified allowable number of trials would not be exceeded. In instances of poor convergence and subsequent approach to the allowable number of trials, it was necessary to force the computer to accept the current trial and go to the plate punch out phase of the program. At the completion of a problem, the number of stages was changed and the program deck with modified data read into the machine again. When all of the changes of the number of stages for a particular V/F and D/F ratio had been covered, the V/F and D/F ratio was changed and process of changing the number of stages repeated. This process was repeated for three different pressures.

The collator was used to merge the result cards with the appropriate alphabetic deck. The merged deck was then printed on the tabulator.

Flash Runs. A flash calculation deck was written to handle the experimentally determined "K'" values. The deck consisted of a flash calculation loop, vapor pressure calculation loop and the necessary punch loops.

The "K'" values were obtained from Figure 41. It may be noticed that in all instances the "K'" values for methane is considered to be infinity. Since the computer is limited to ten digit numbers, it was assumed that any "K'" value going to infinity could be approximated by a value of 1,000.

Cards containing the feed composition, control constants, and appropriate vapor liquid equilibrium data were assembled with the flash deck. This deck was read into the machine which performed the mechanics of flash calculation.

The result cards were merged with a listing deck and printed on the tabulator.

Fenske Runs. The relative volatility data were recalculated using the same vapor-liquid equilibrium constants but a different reference component. The resulting values were curve-fitted to the equation $A + BT + CT^2$, with any component equation having either a maximum or minimum being adjusted to eliminate this defect.

Cards containing the feed composition, control constants, and relative volatility data were assembled with the Fenske Deck. This deck was read into the machine which performed the Fenske calculation at the specified conditions.

The result cards were merged with the Fenske Listing Deck and printed on the tabulator.

CHAPTER IV

RESULTS

The following pages contain the data obtained in this investigation. The data is presented in both tabular and graphical form. Graphical data includes the temperature profiles and tray concentration diagrams of the plate to plate calculations, curves relating the effect of pressure on recovery, pseudo vapor-liquid equilibrium constants calculated from the plate to plate results, and curves relating minimum number of stages, pressure, and pseudo equilibrium constant curve number. Tabulated data includes final material and heat balances of the plate to plate calculation, flash results using the calculated pseudo equilibrium values, and product distribution with the minimum number of stages of the Fenske calculations.

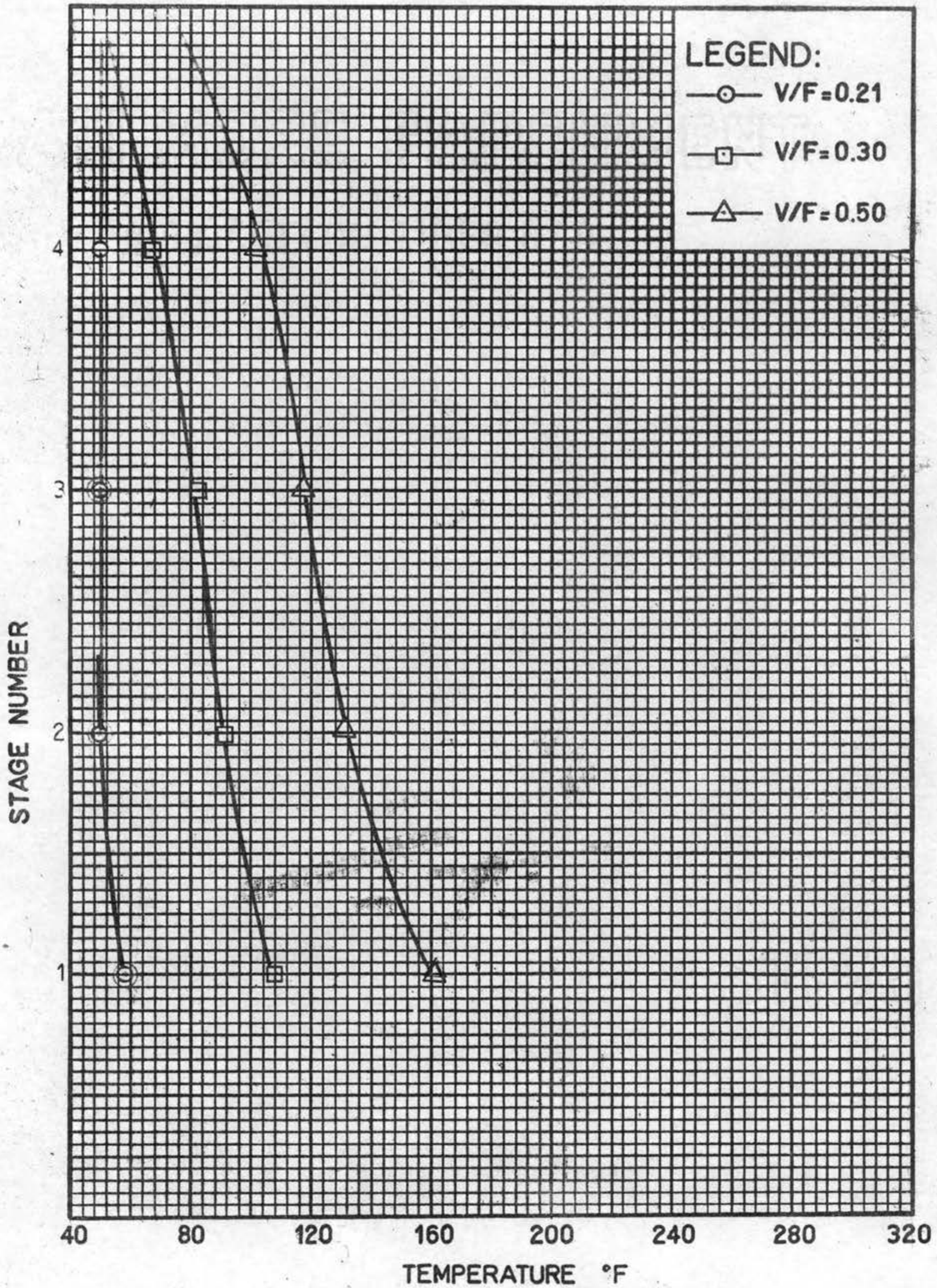


FIGURE 2

TEMPERATURE PROFILE, 5 STAGES & 75 PSIA

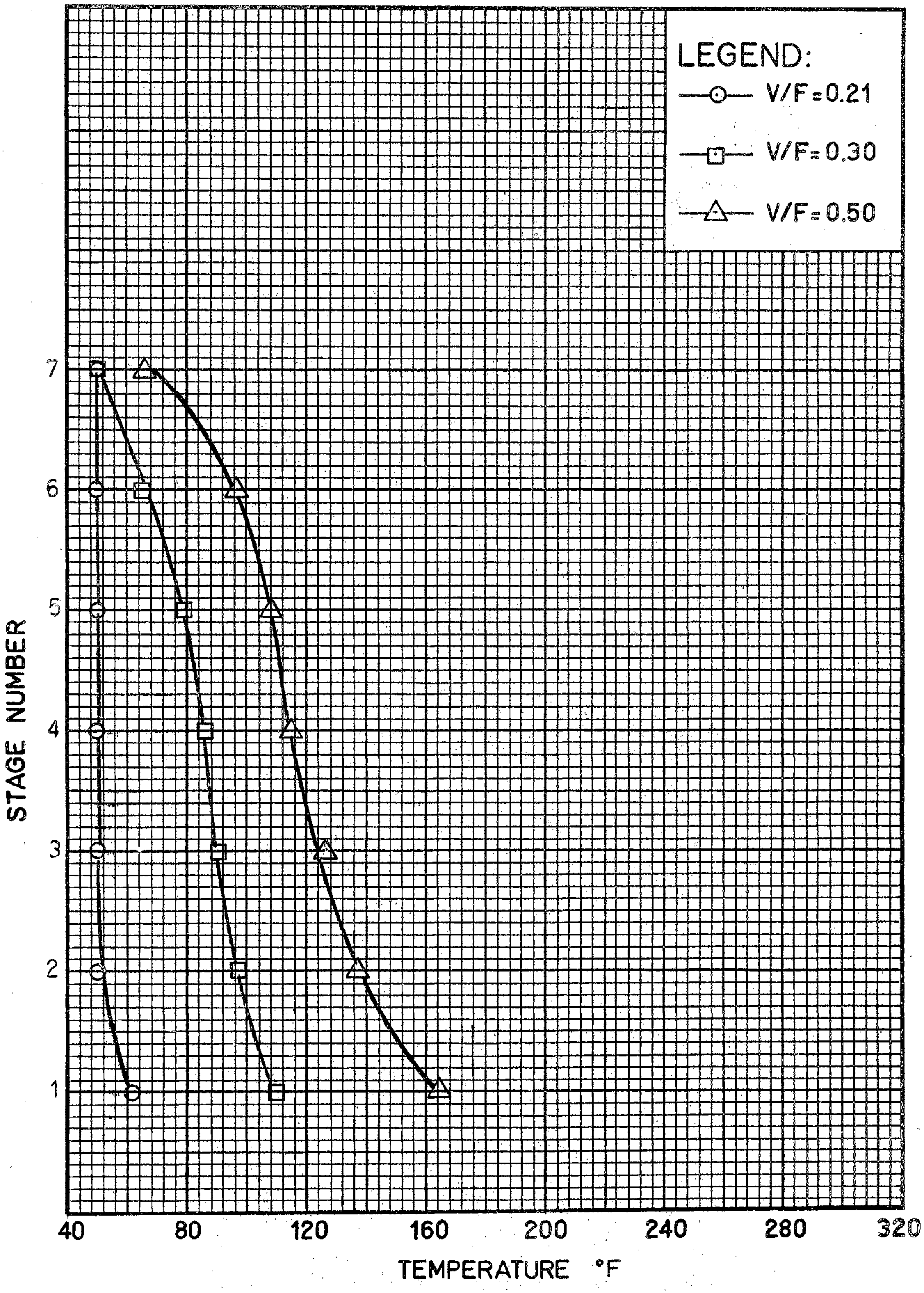


FIGURE 3

TEMPERATURE PROFILE, 7 STAGES & 75 PSIA

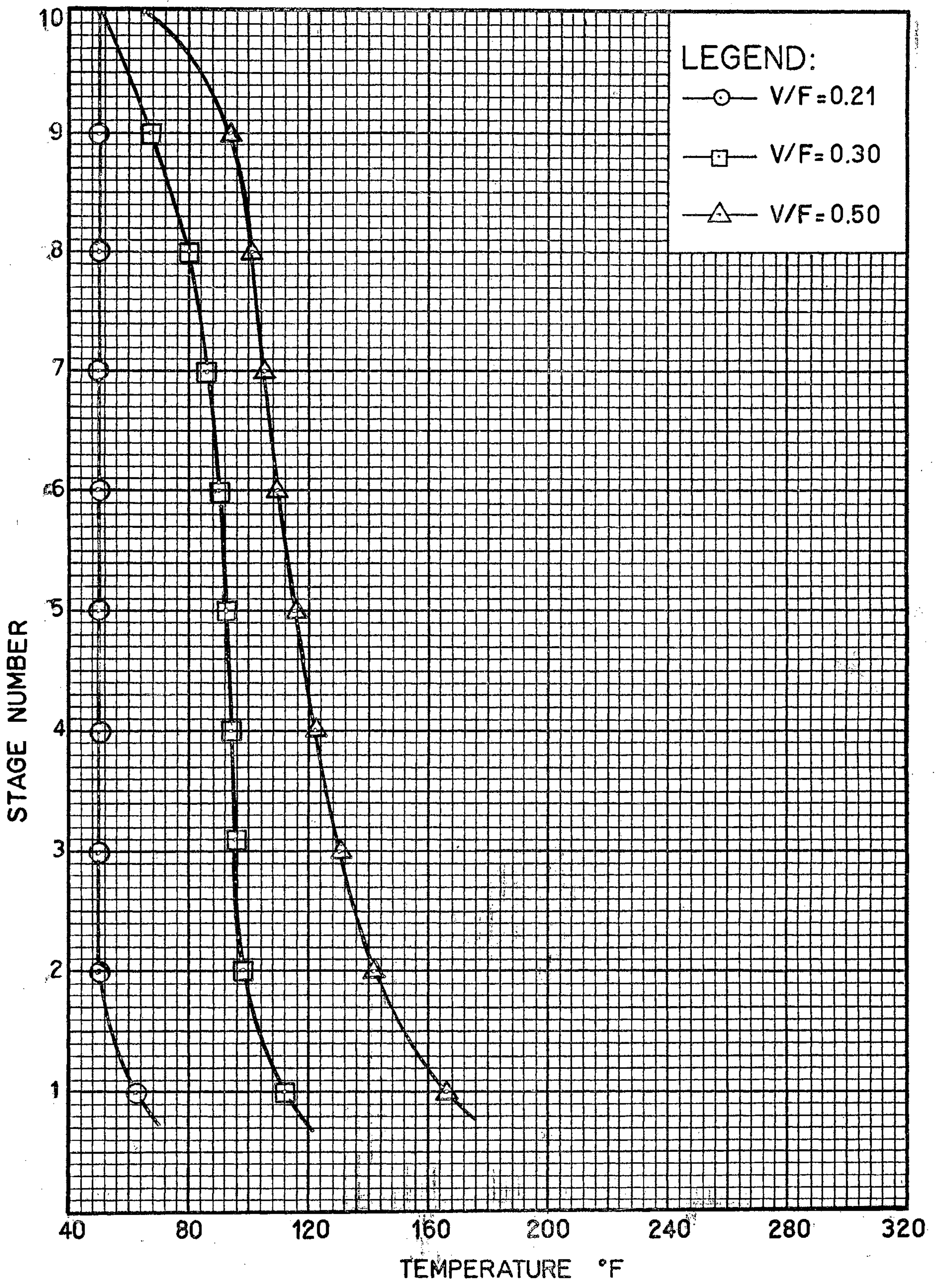


FIGURE 4

TEMPERATURE PROFILE, 10 STAGES & 75 PSIA

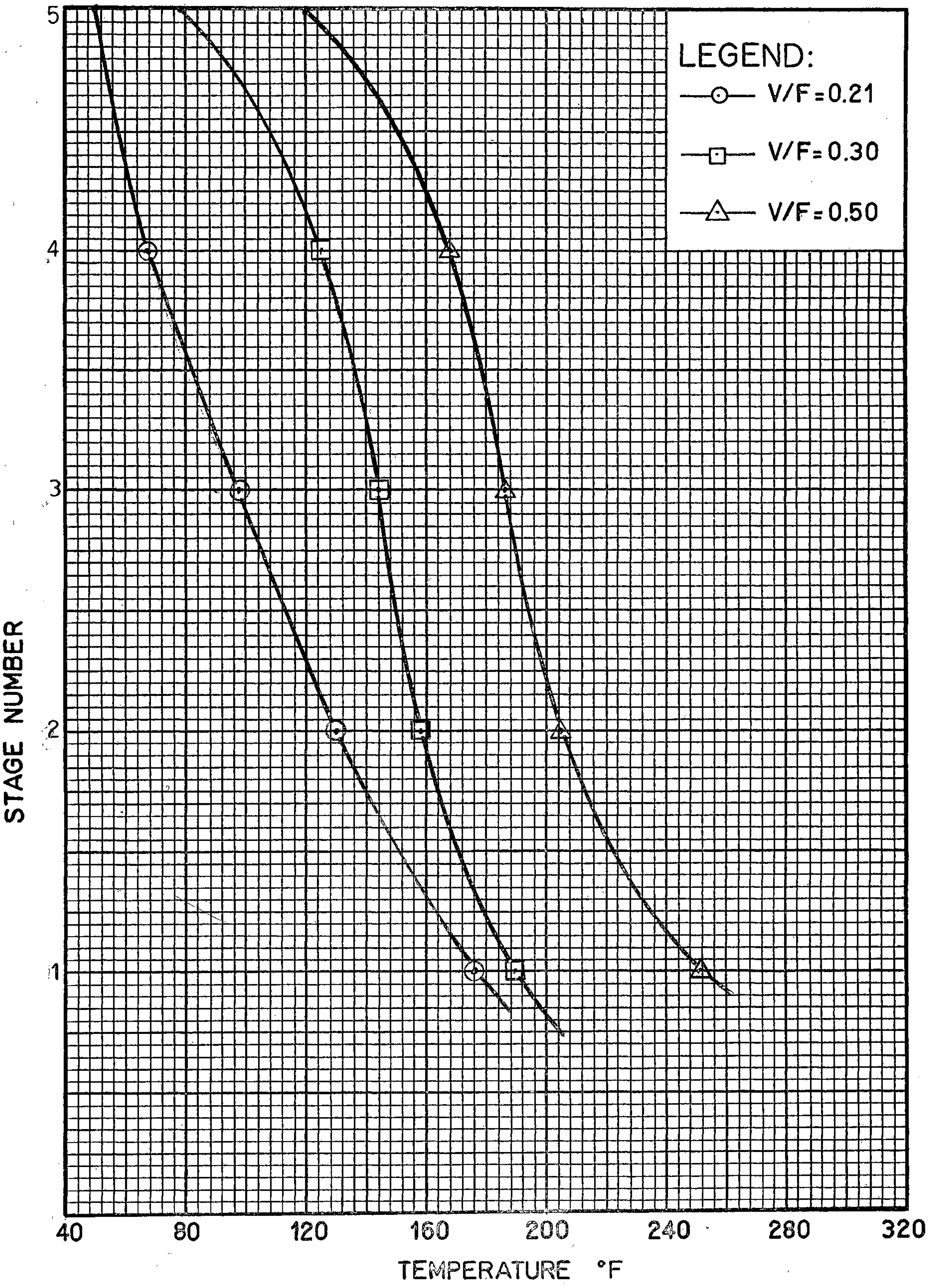


FIGURE 5

TEMPERATURE PROFILE, 5 STAGES & 200 PSIA

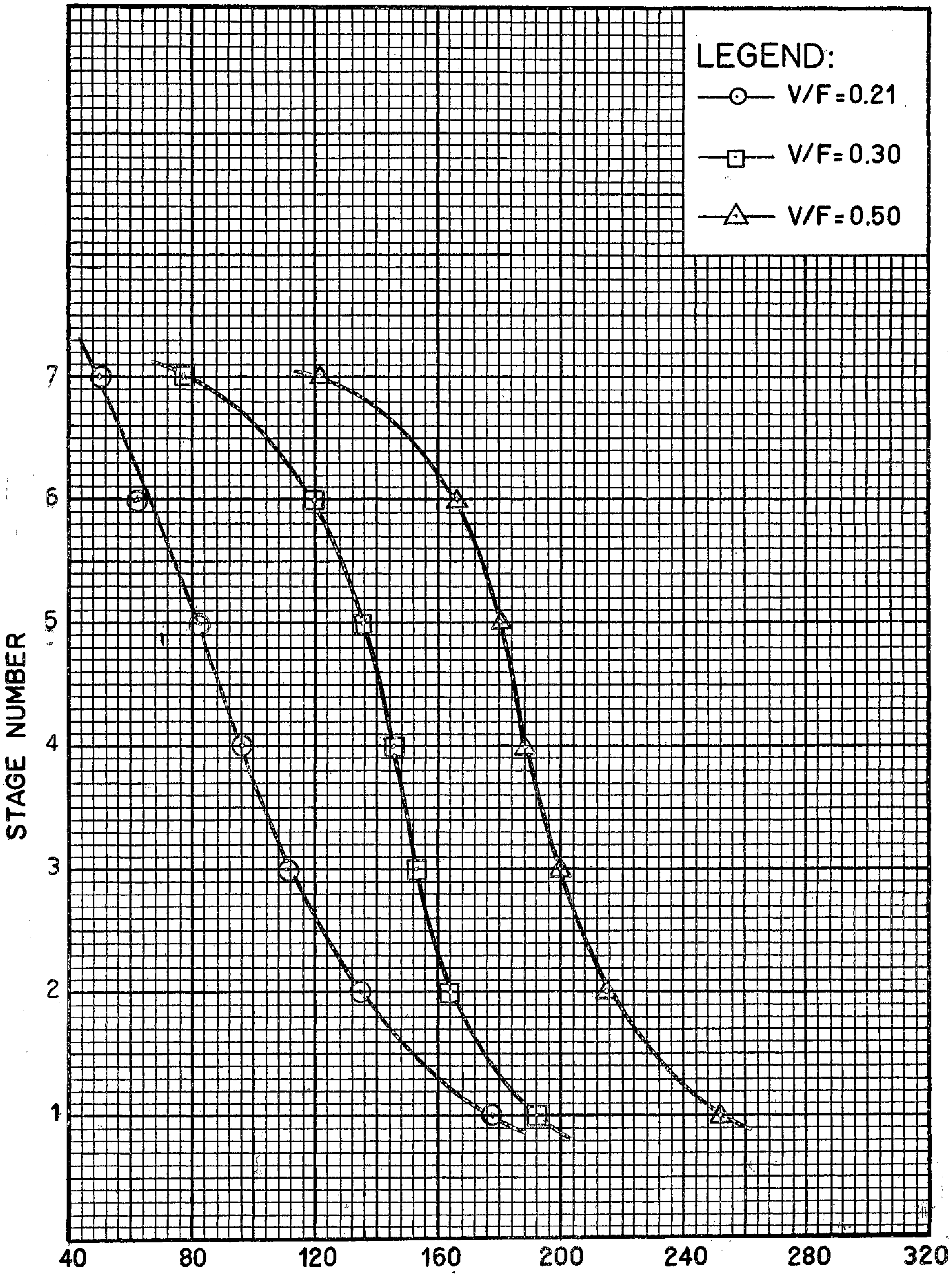


FIGURE 6

TEMPERATURE PROFILE, 7 STAGES & 200 PSIA

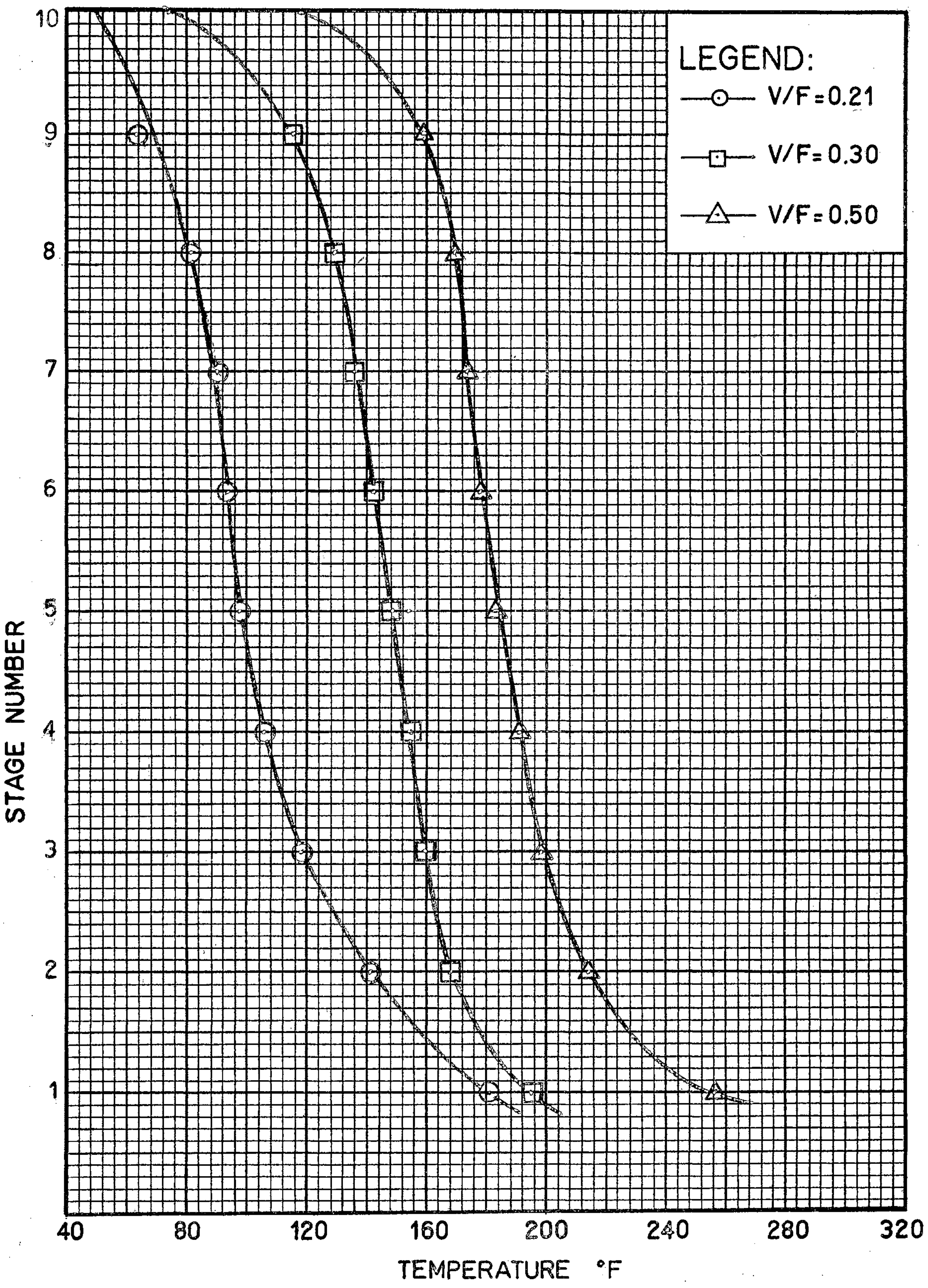


FIGURE 7

TEMPERATURE PROFILE, 10 STAGES & 200 PSIA

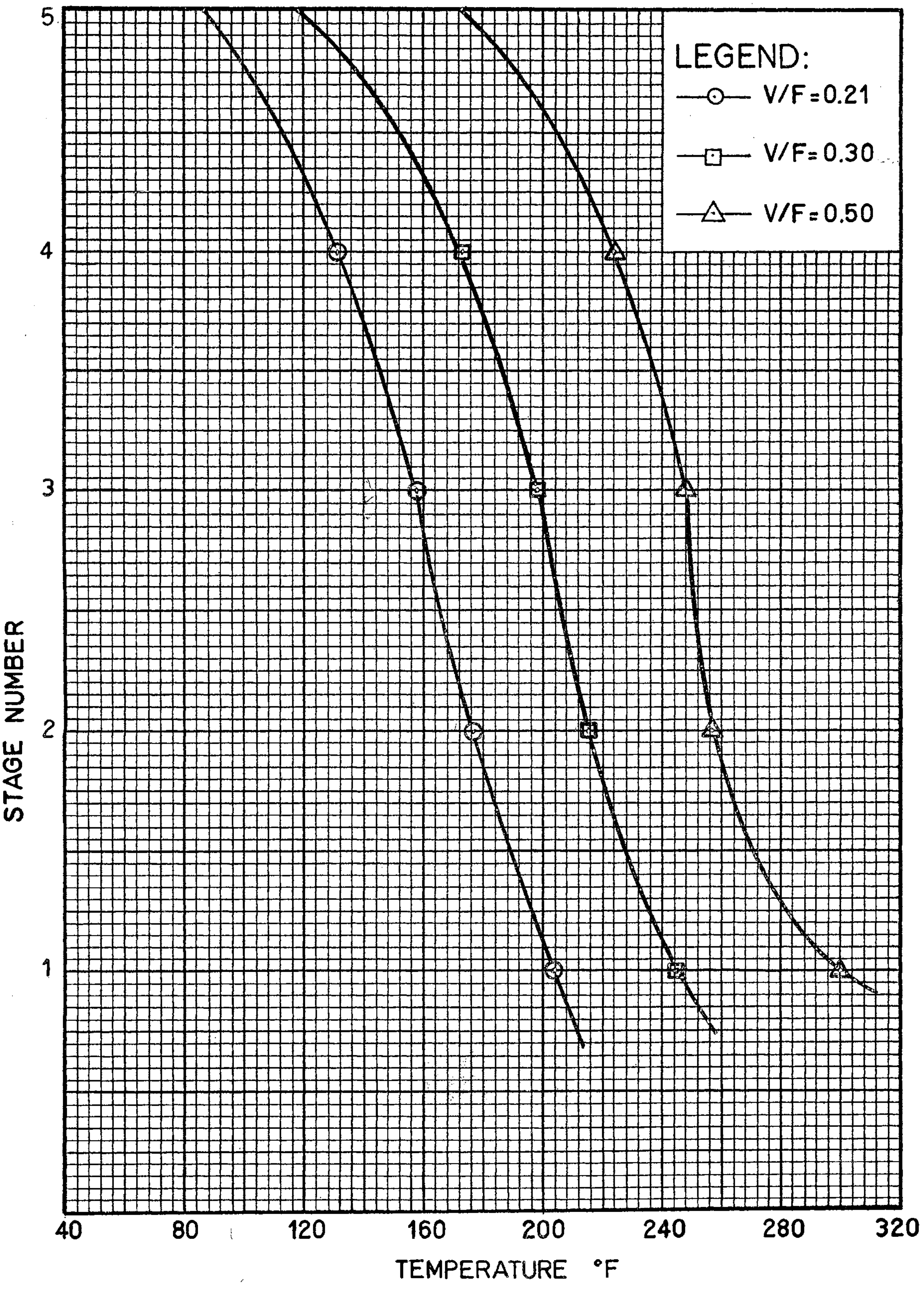


FIGURE 8

TEMPERATURE PROFILE, 5 STAGES & 300 PSIA

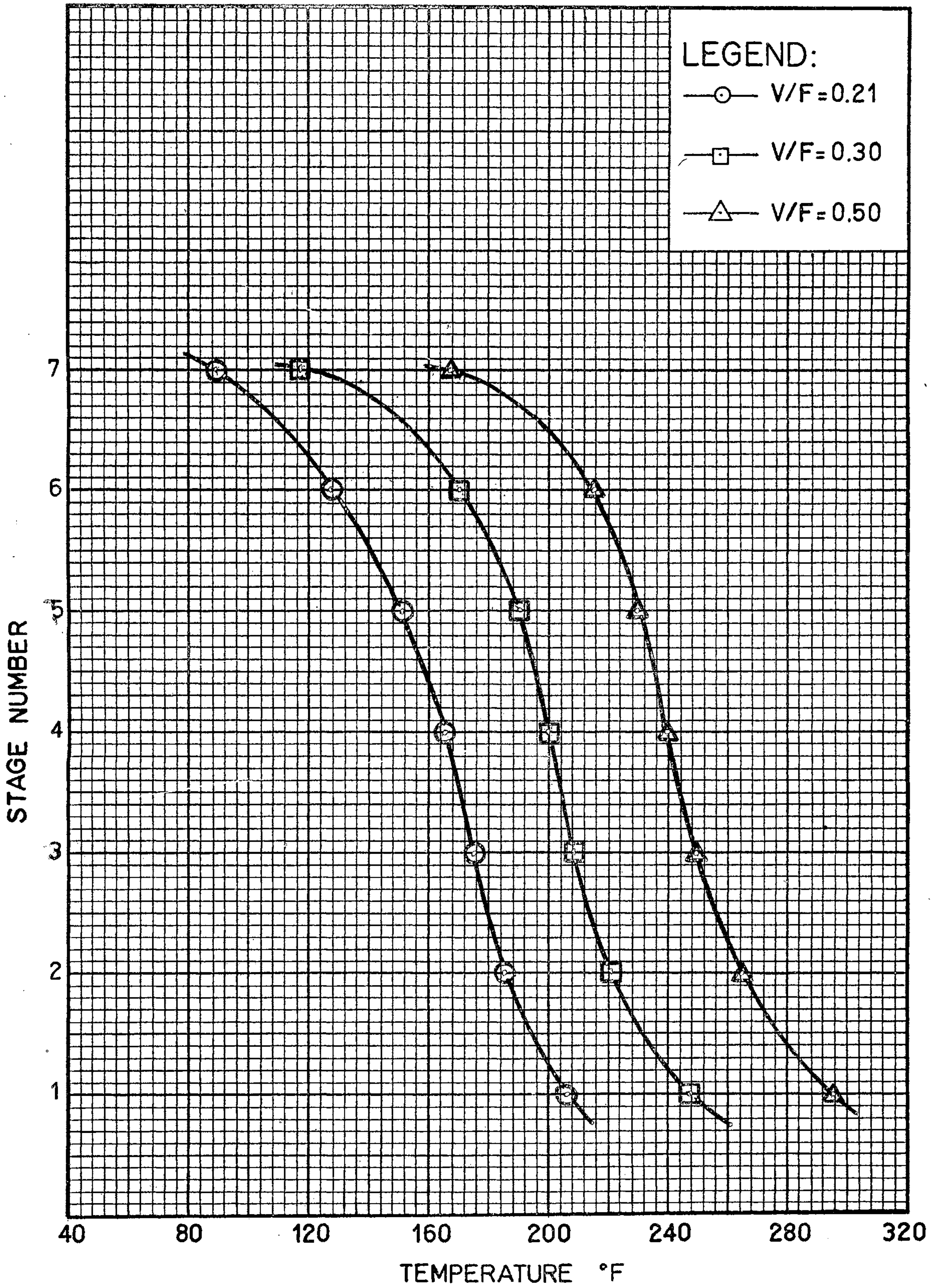


FIGURE 9

TEMPERATURE PROFILE, 7 STAGES & 300 PSIA

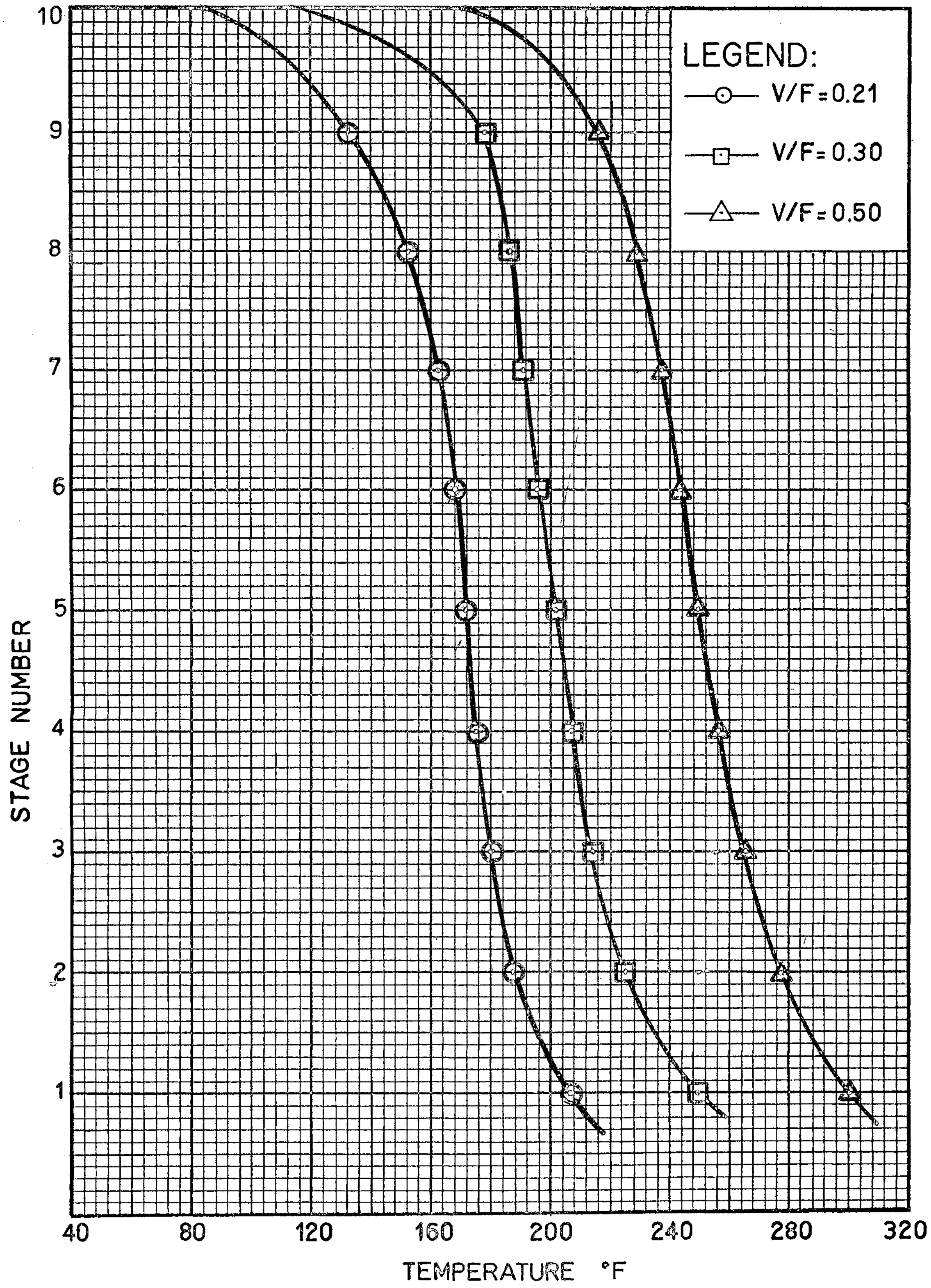


FIGURE 10

TEMPERATURE PROFILE, 10 STAGES & 300 PSIA

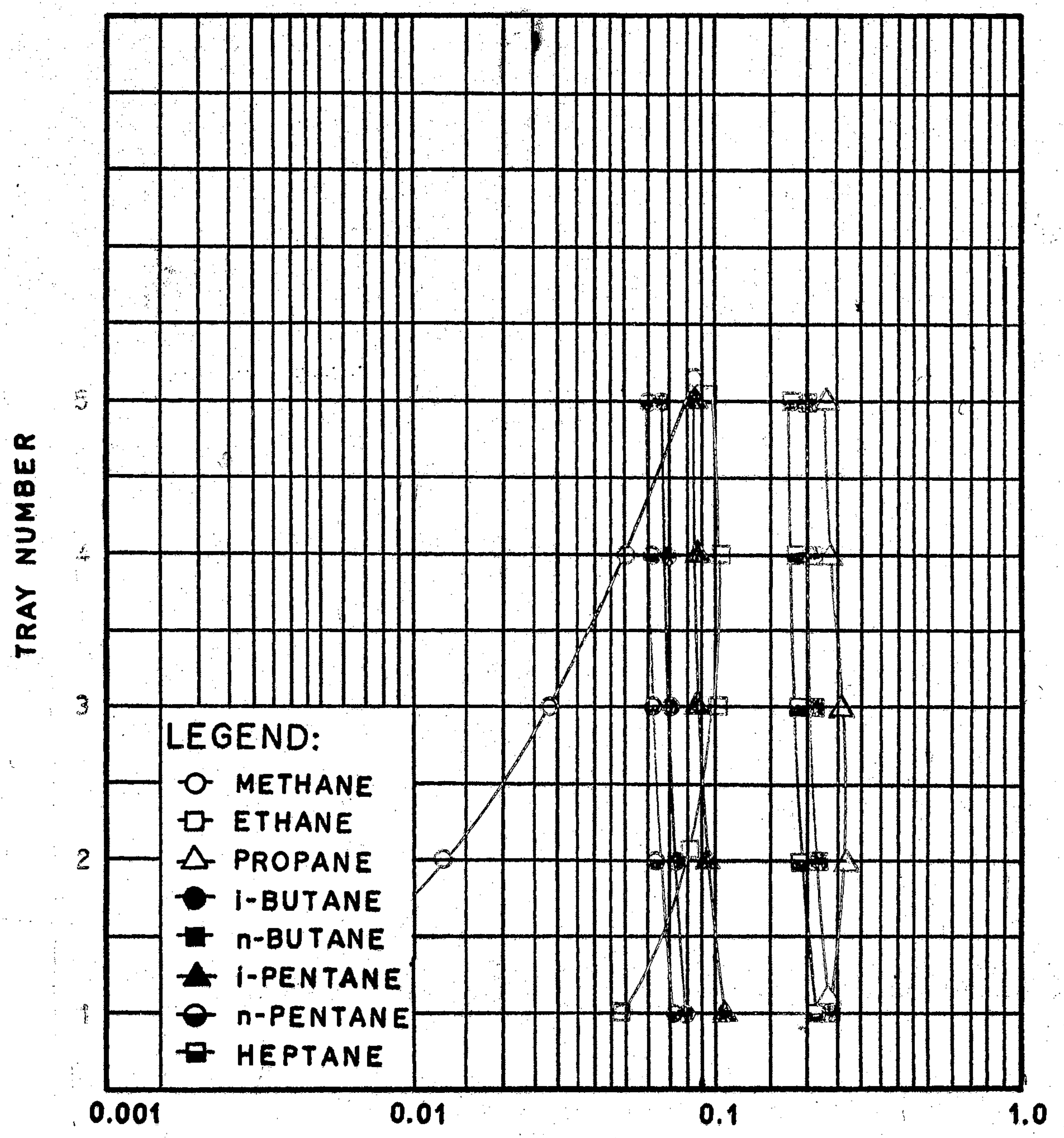


FIGURE 11

TRAY CONCENTRATION DIAGRAM

V/F = 0.21, 5 TRAYS, & 75 PSIA

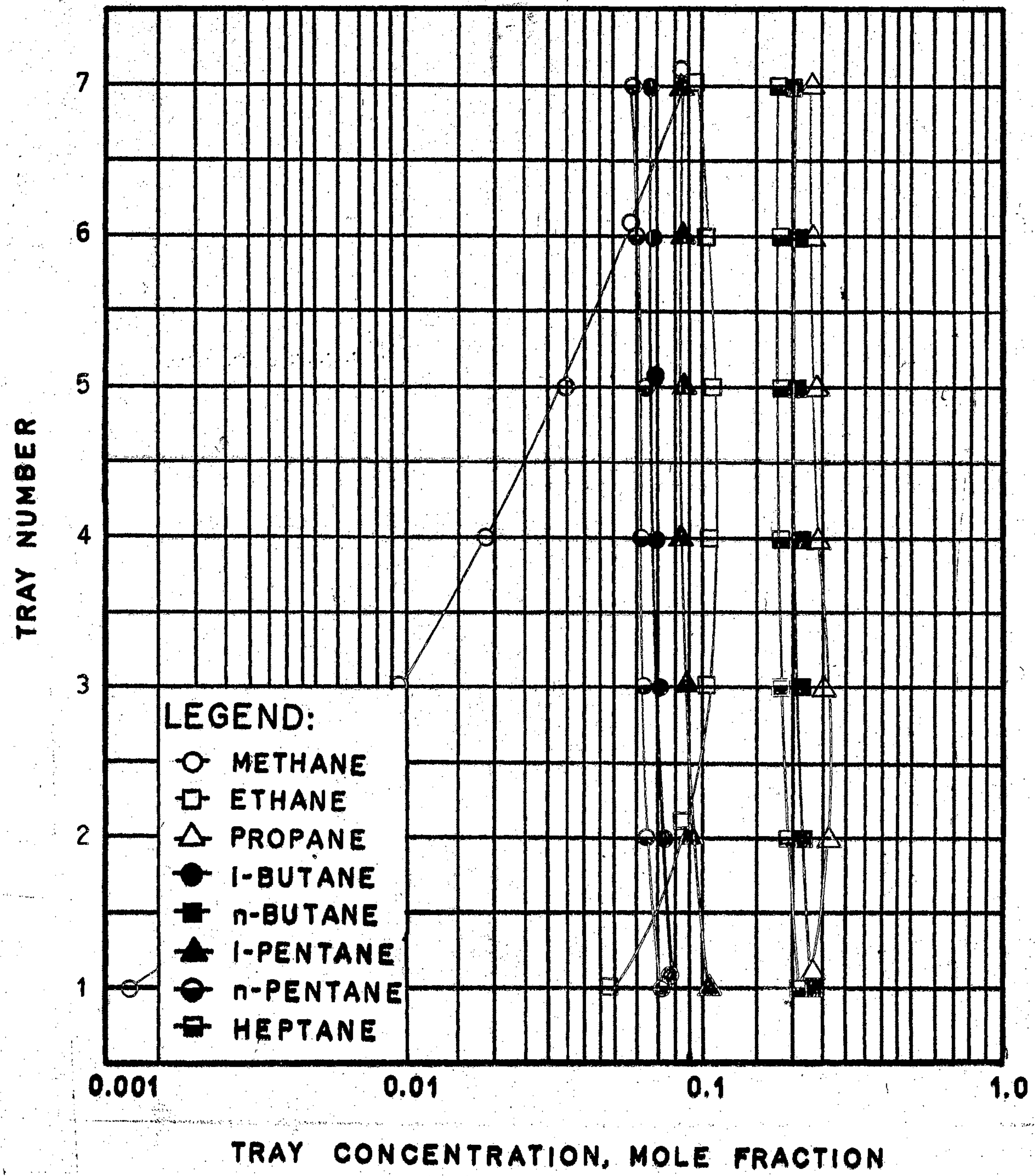


FIGURE 12

TRAY CONCENTRATION DIAGRAM

V/F = 0.21, 7 TRAYS, & 75 PSIA

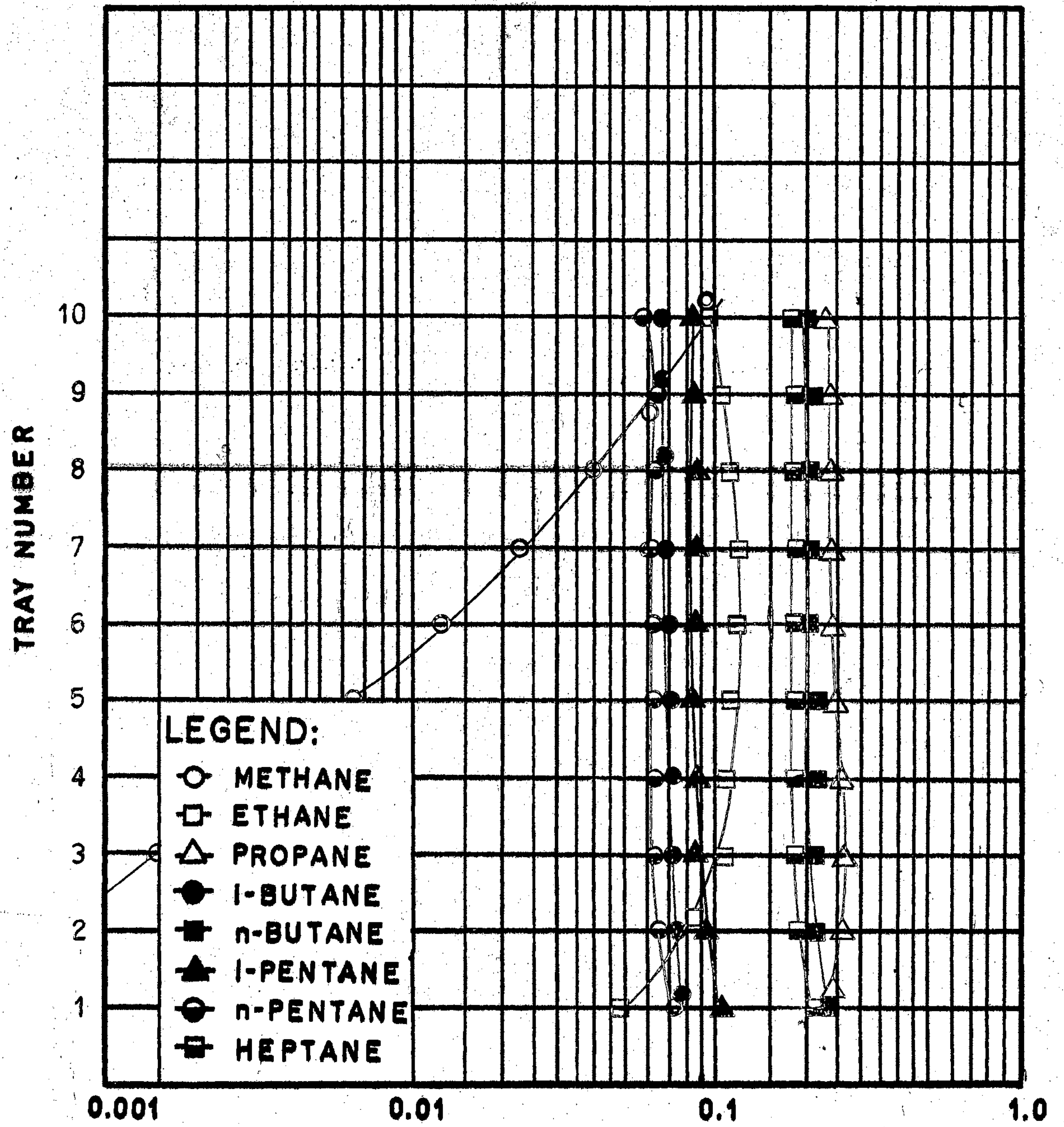
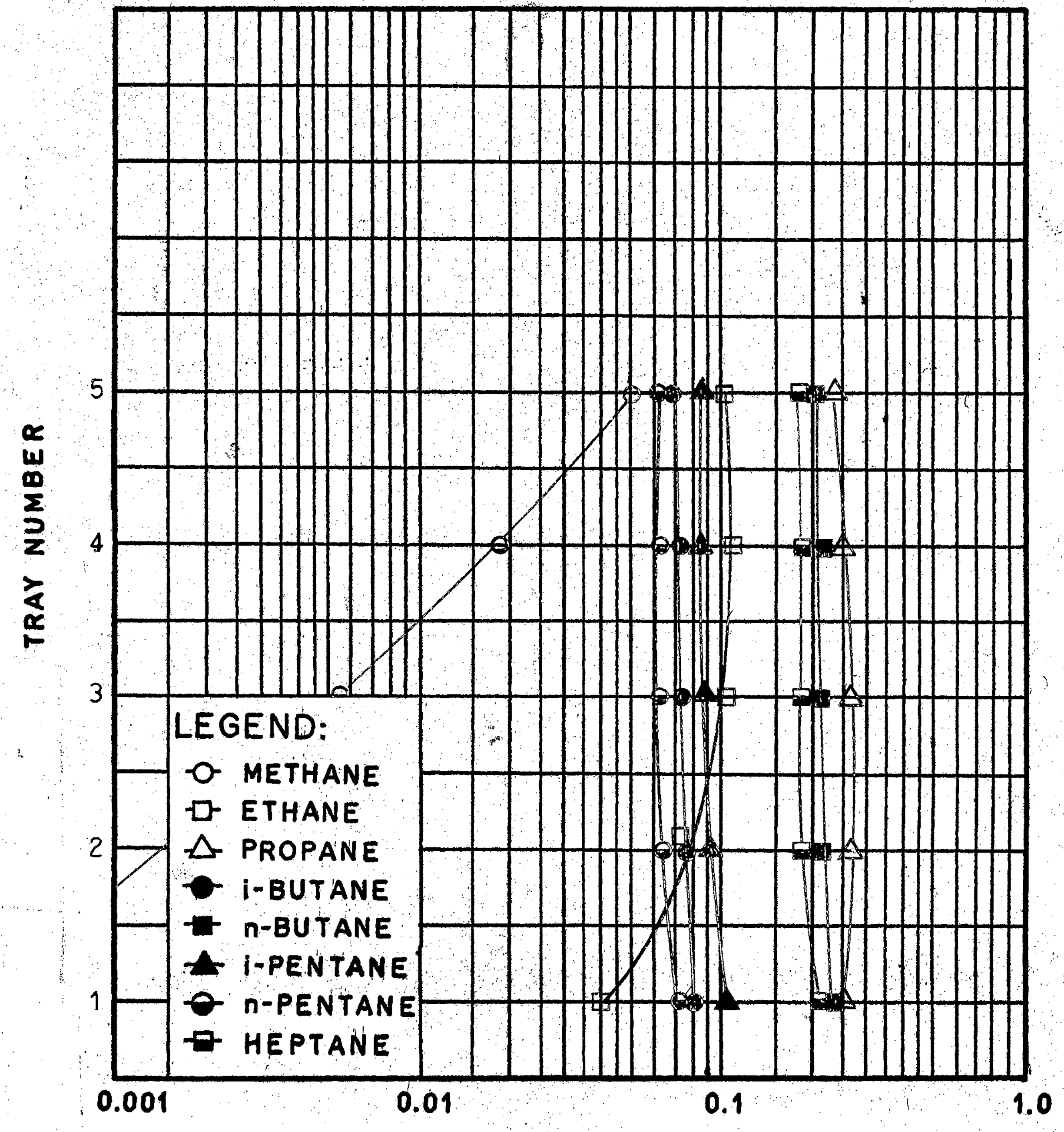


FIGURE 13

TRAY CONCENTRATION DIAGRAM

V/F = 0.21, 10 TRAYS, & 75 PSIA



TRAY CONCENTRATION, MOLE FRACTION

FIGURE 14

TRAY CONCENTRATION DIAGRAM

V/F = 0.21, 5 TRAYS, & 200 PSIA

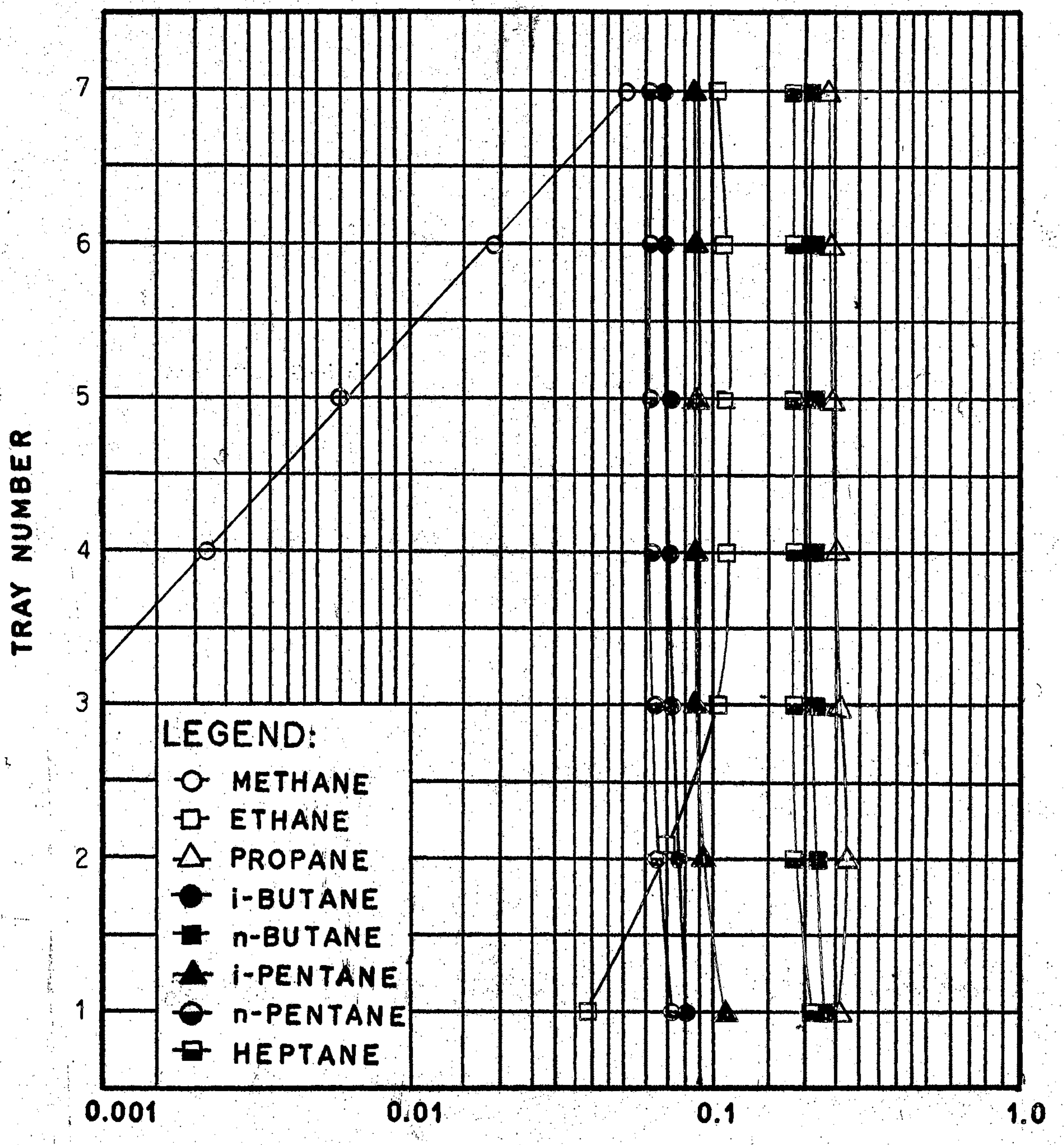


FIGURE 15

TRAY CONCENTRATION DIAGRAM

V/F = 0.21 , 7 TRAYS, & 200 PSIA

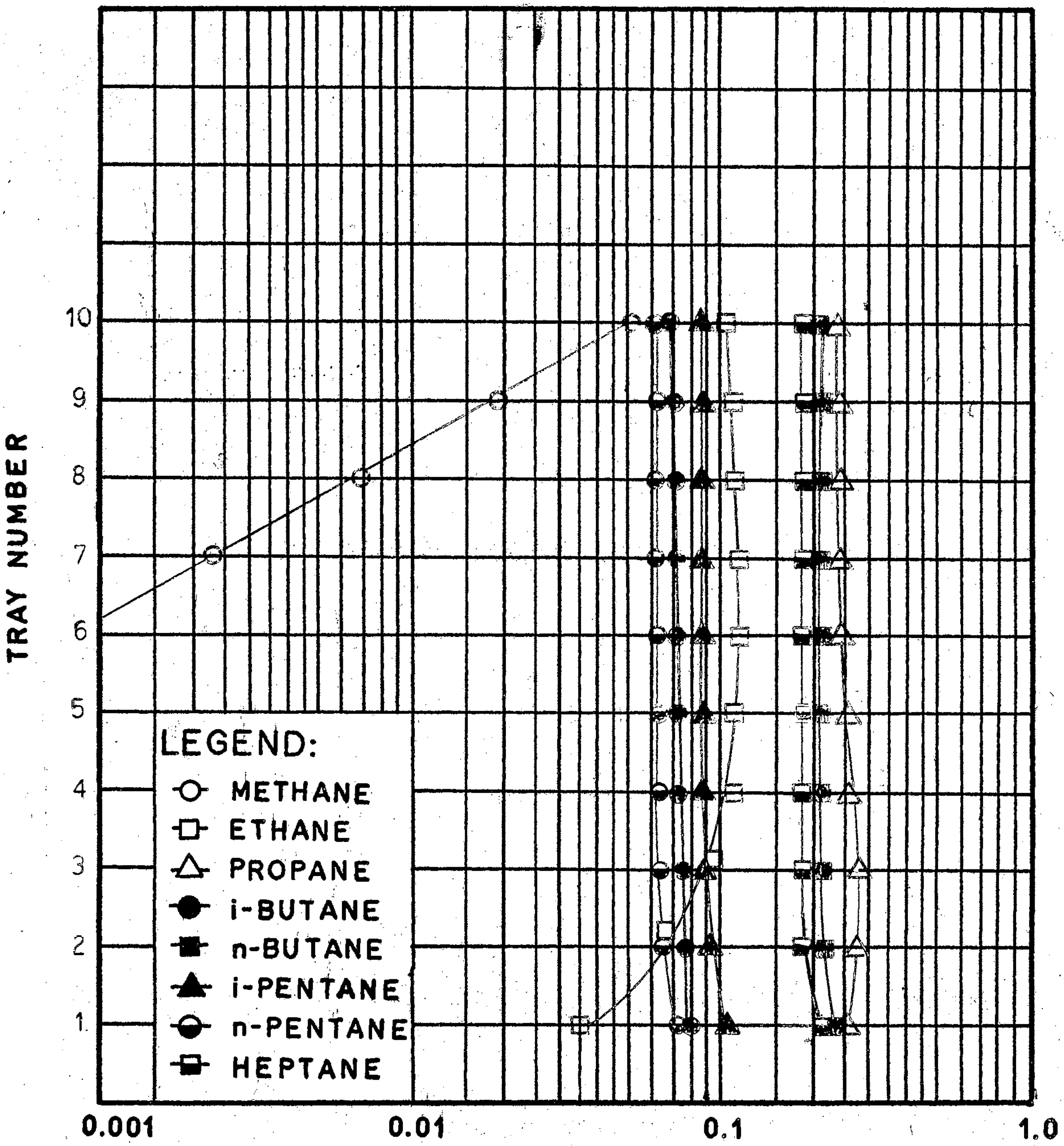
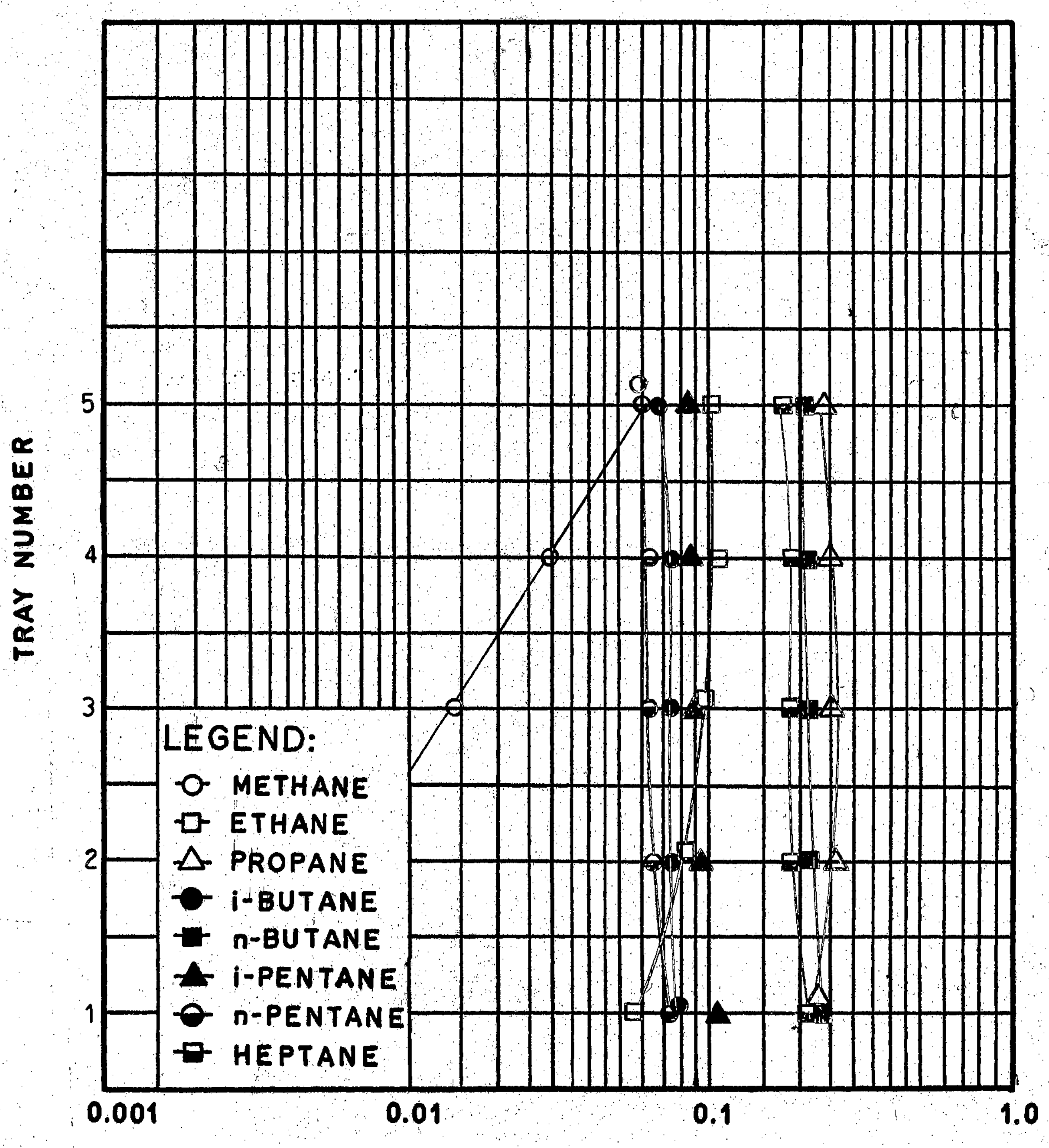


FIGURE 16

TRAY CONCENTRATION DIAGRAM

V/F = 0.21, 10 TRAYS, & 200 PSIA



TRAY CONCENTRATION, MOLE FRACTION

FIGURE 17

TRAY CONCENTRATION DIAGRAM

V/F = 0.21 , 5 TRAYS, & 300 PSIA

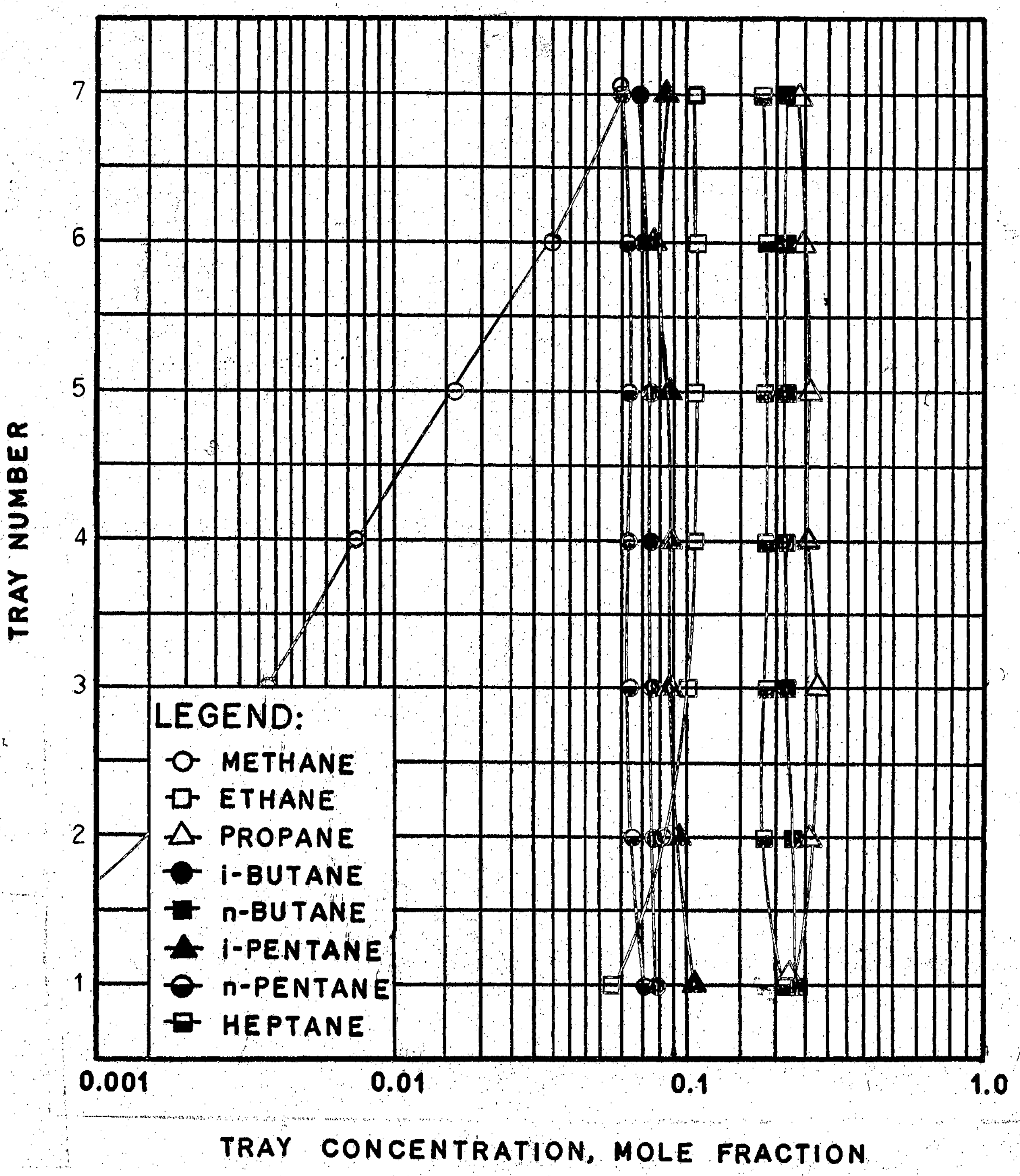


FIGURE 18

TRAY CONCENTRATION DIAGRAM

V/F = 0.21 , 7 TRAYS, & 300 PSIA

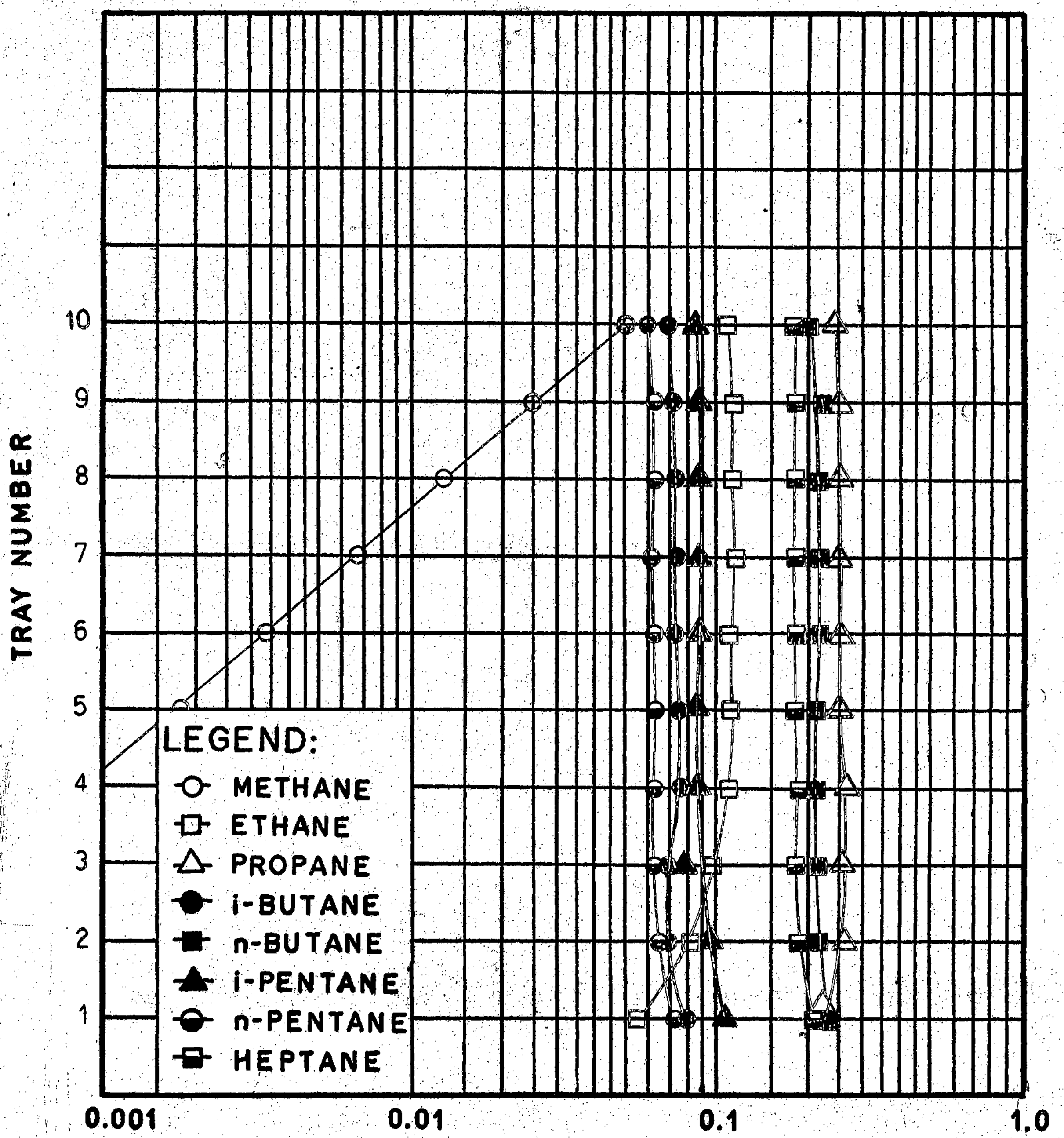


FIGURE 19

TRAY CONCENTRATION DIAGRAM

V/F = 0.21, 10 TRAYS, & 300 PSIA

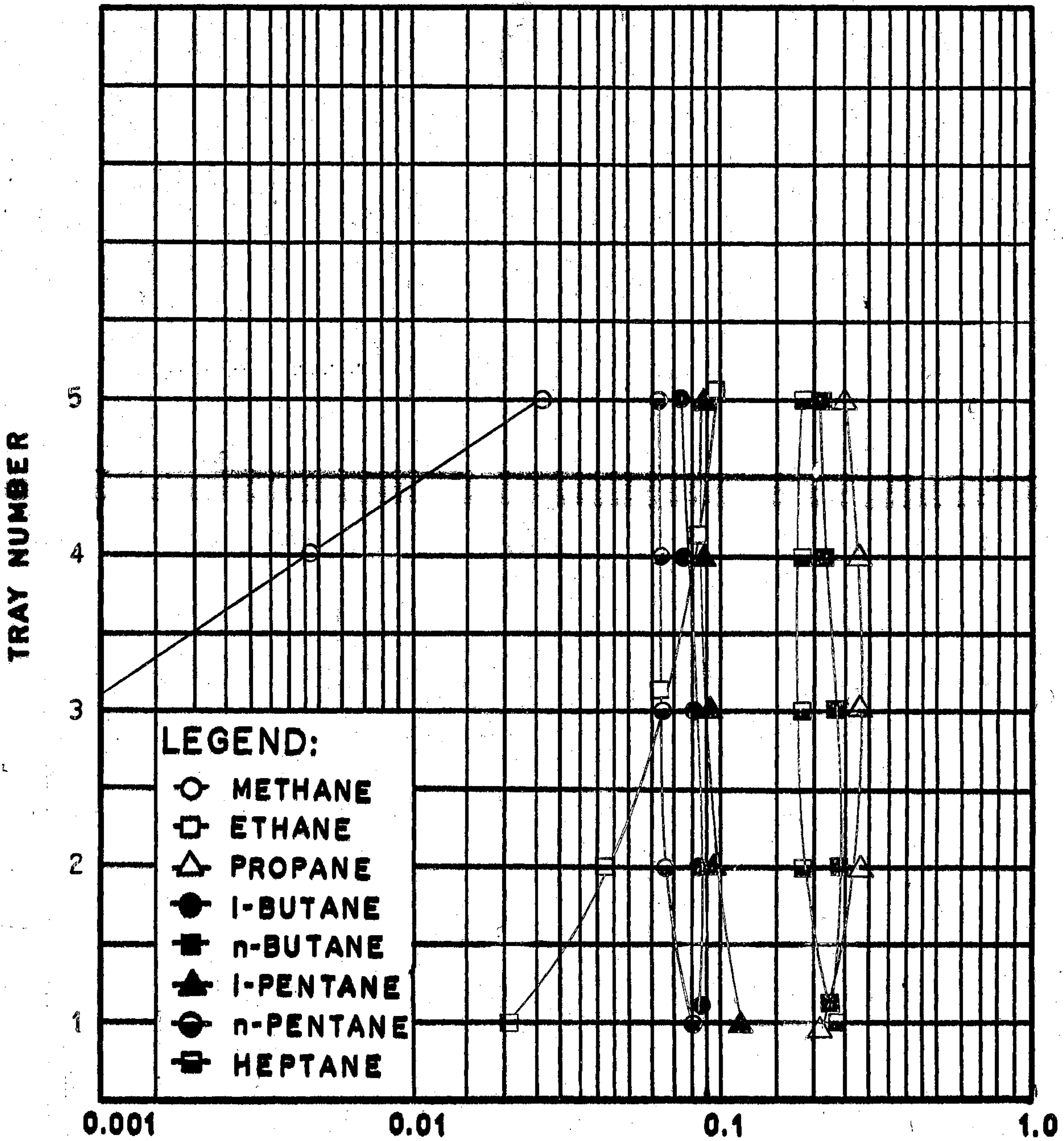


FIGURE 20

TRAY CONCENTRATION DIAGRAM

V/F = 0.30, 5 TRAYS, & 75 PSIA

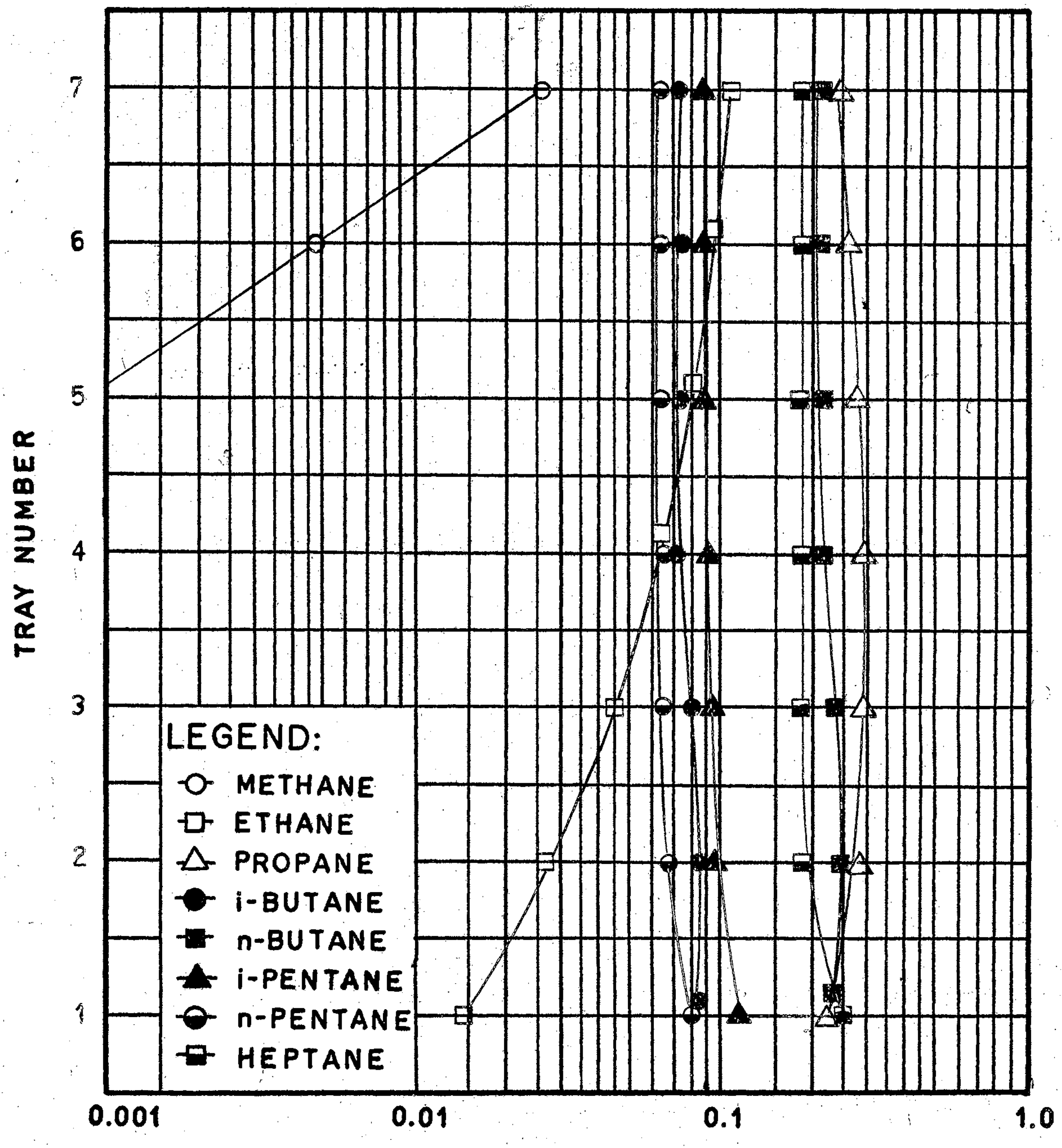
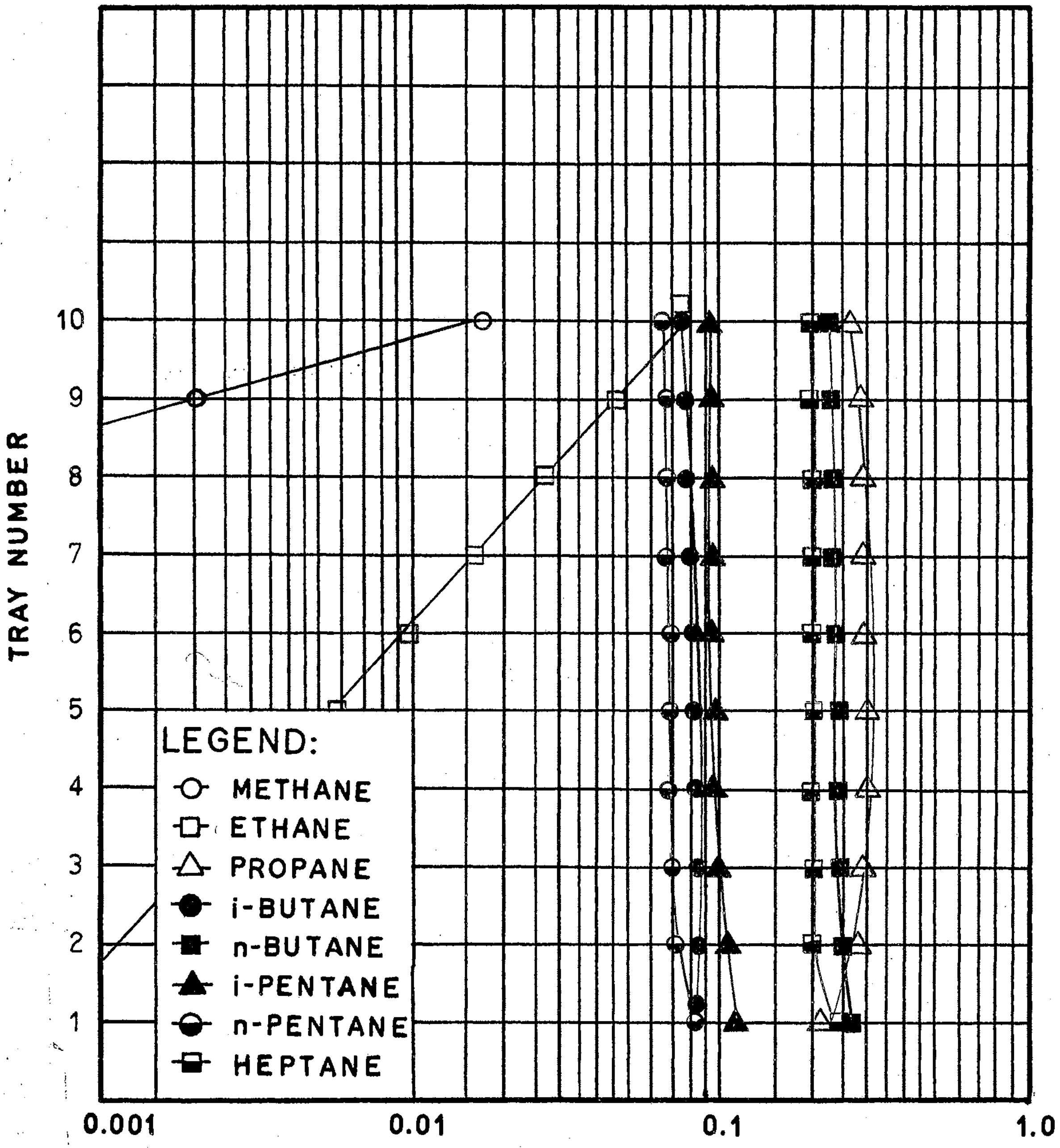


FIGURE 21

TRAY CONCENTRATION DIAGRAM

V/F = 0.30, 7 TRAYS, & 75 PSIA



TRAY CONCENTRATION, MOLE FRACTION

FIGURE 22

TRAY CONCENTRATION DIAGRAM

V/F = 0.30, 10 TRAYS, & 75 PSIA

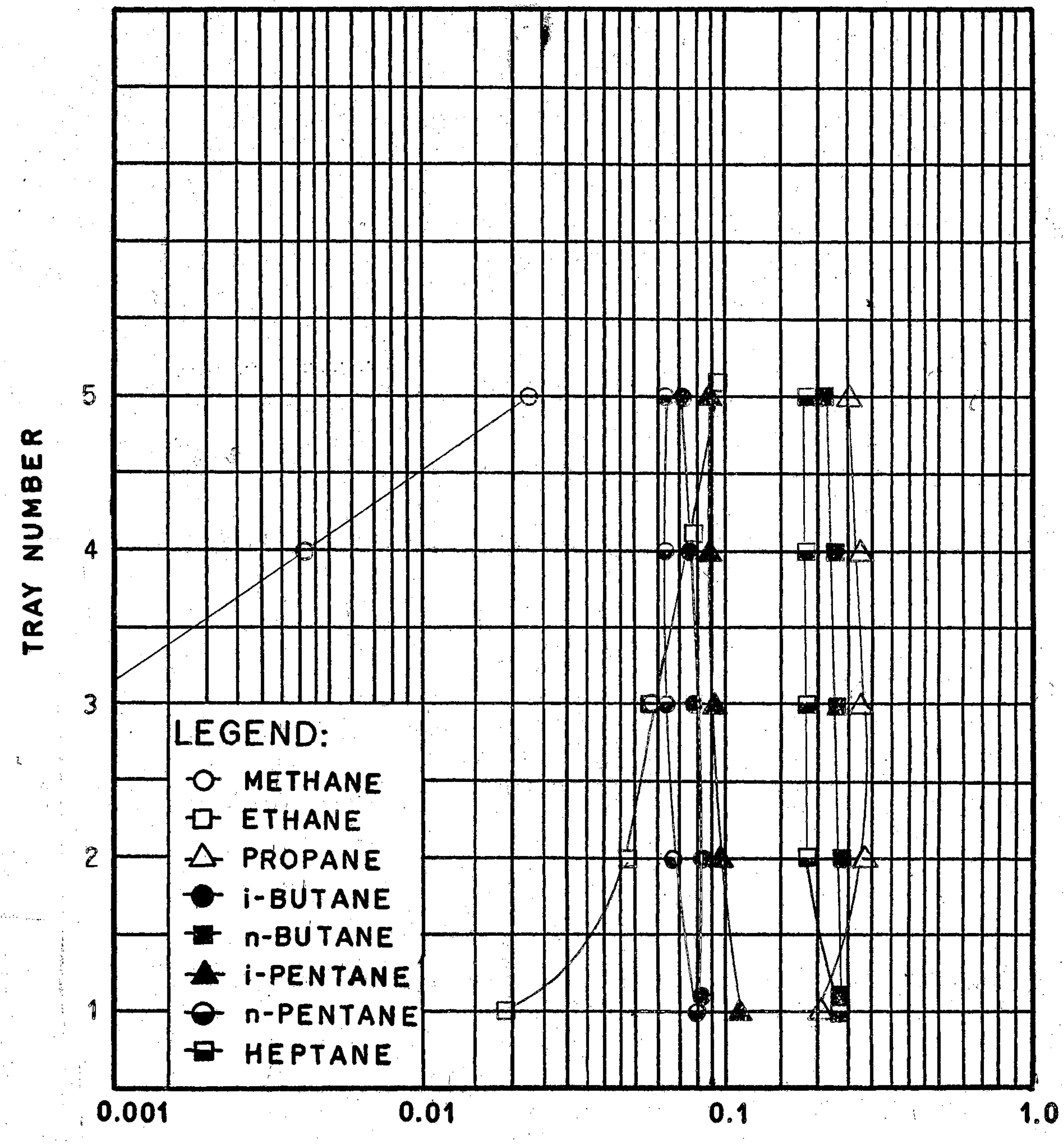


FIGURE 23

TRAY CONCENTRATION DIAGRAM

V/F = 0.30, 5 TRAYS, & 200 PSIA

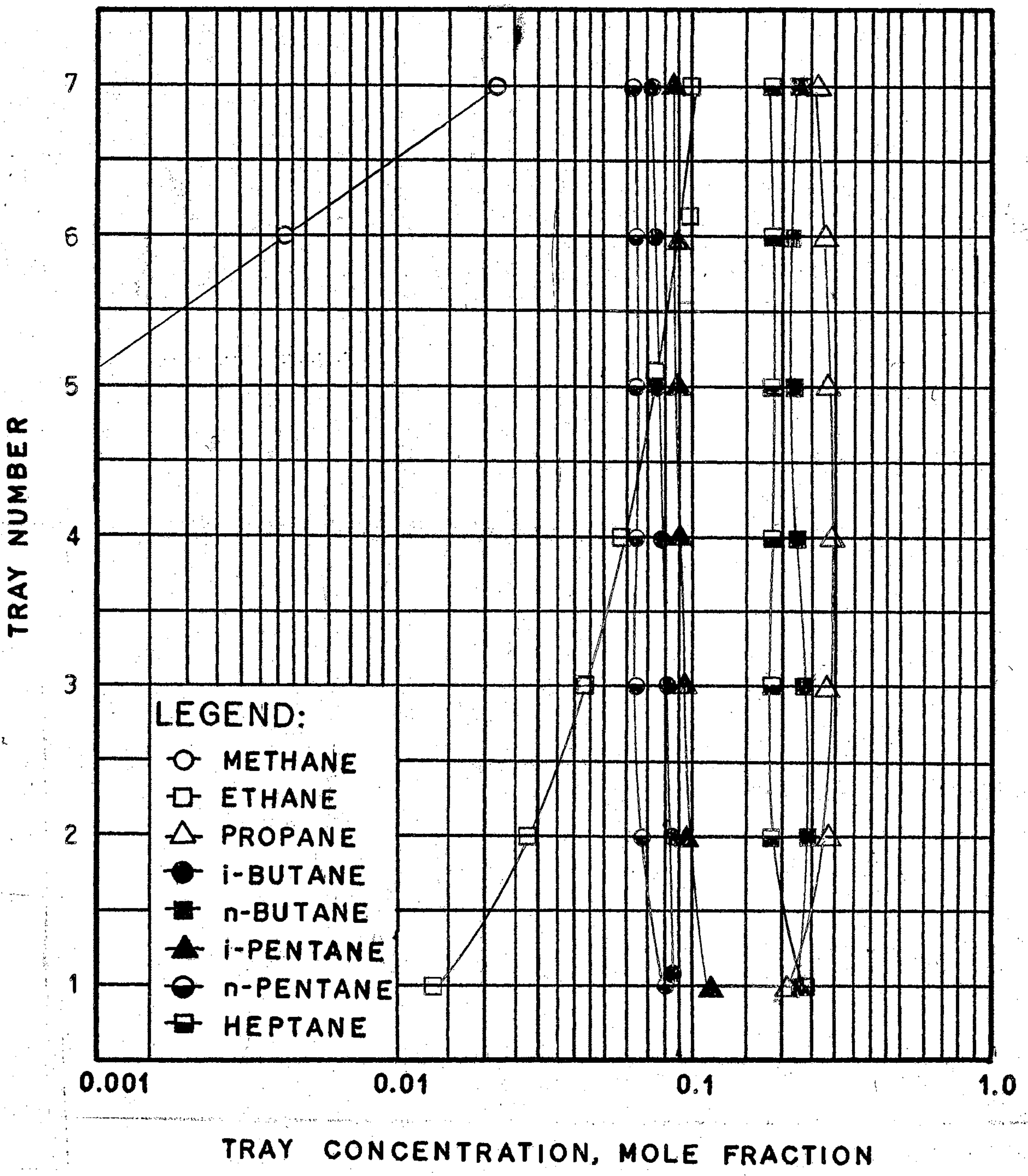


FIGURE 24

TRAY CONCENTRATION DIAGRAM

V/F = 0.30 , 7 TRAYS, & 200PSIA

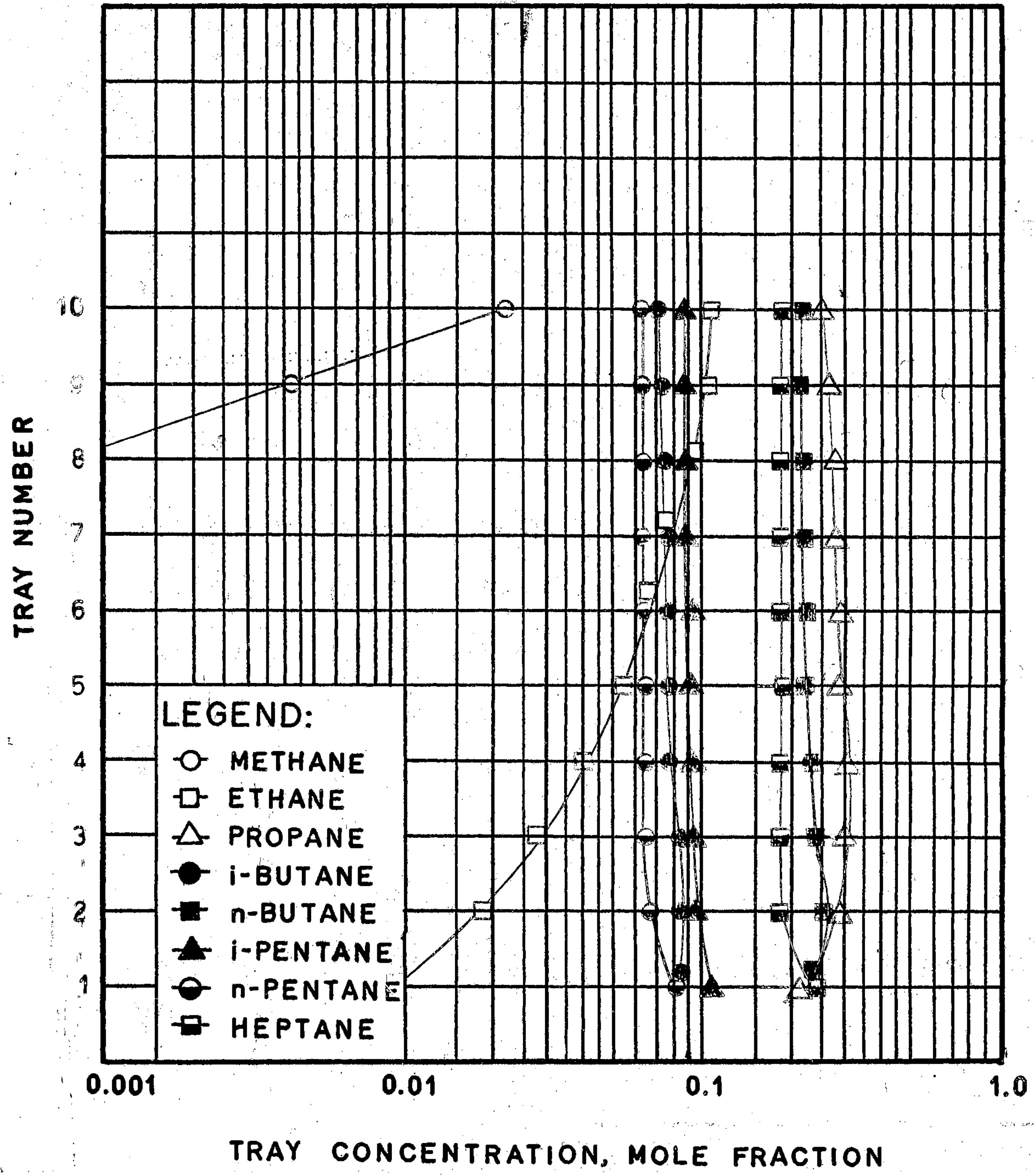


FIGURE 25

TRAY CONCENTRATION DIAGRAM

V/F = 0.30, 10 TRAYS, & 200 PSIA

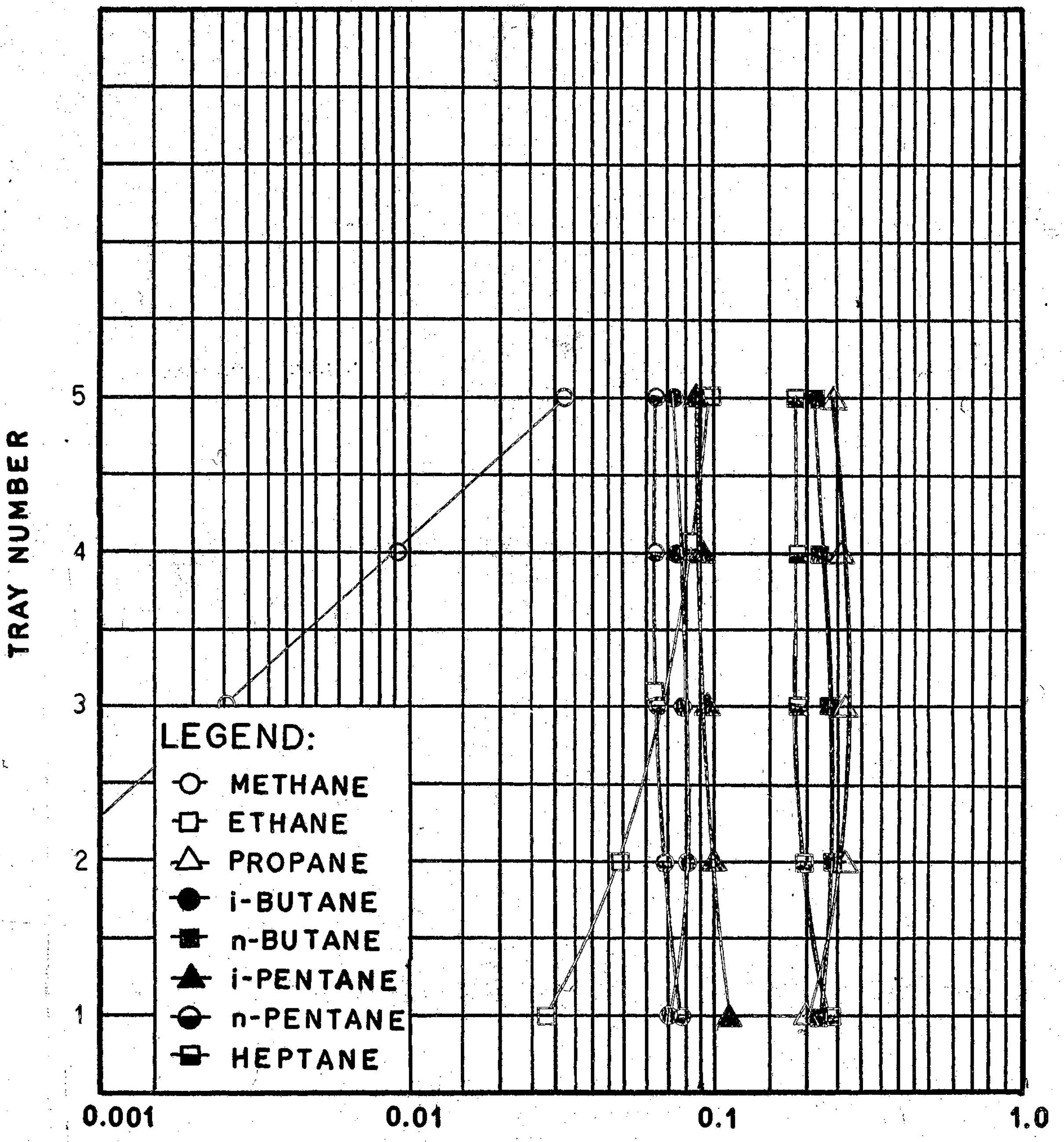


FIGURE 26

TRAY CONCENTRATION DIAGRAM

V/F = 0.30, 5 TRAYS, & 300 PSIA

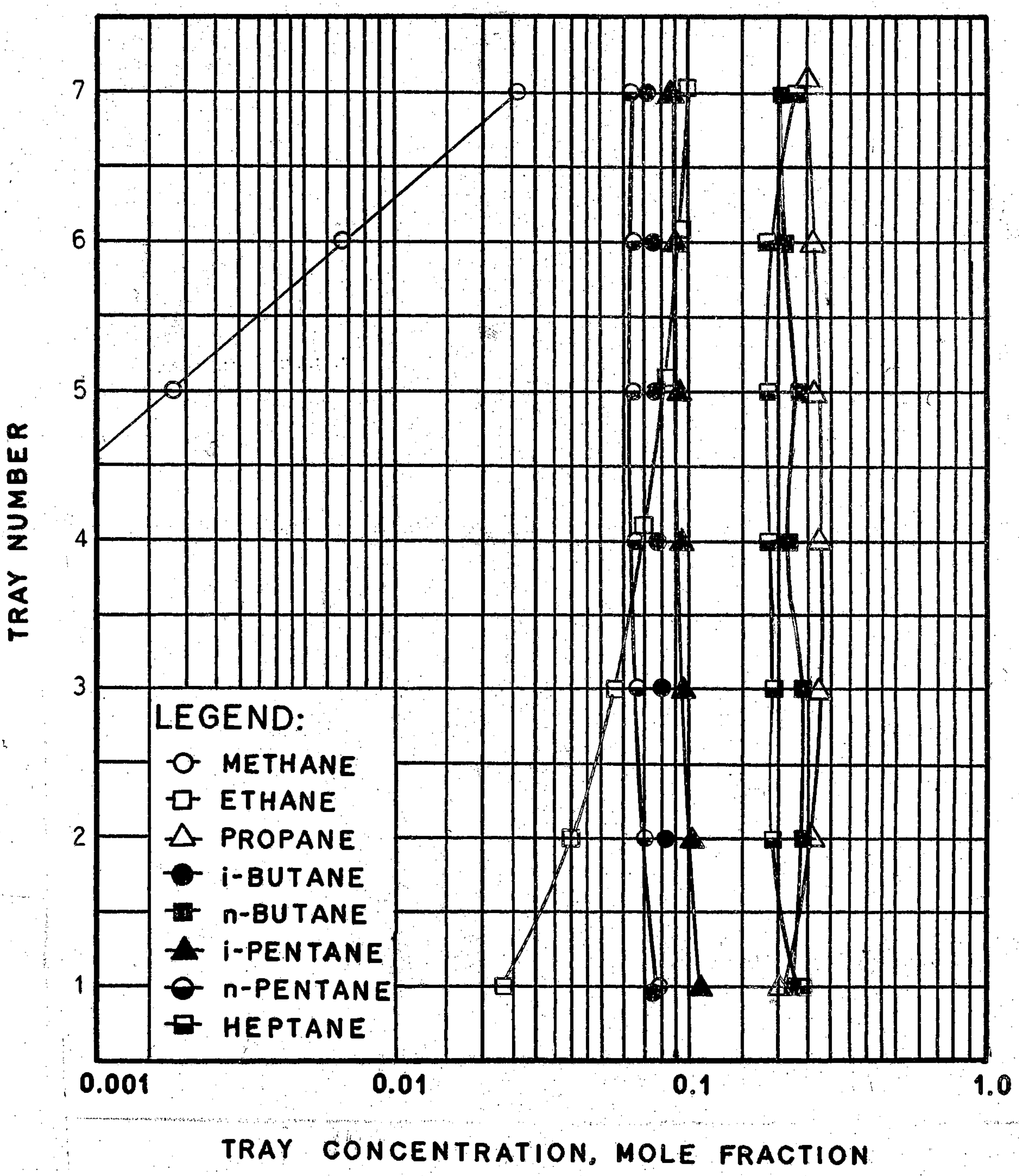


FIGURE 27

TRAY CONCENTRATION DIAGRAM

V/F = 0.30, 7 TRAYS, & 300 PSIA

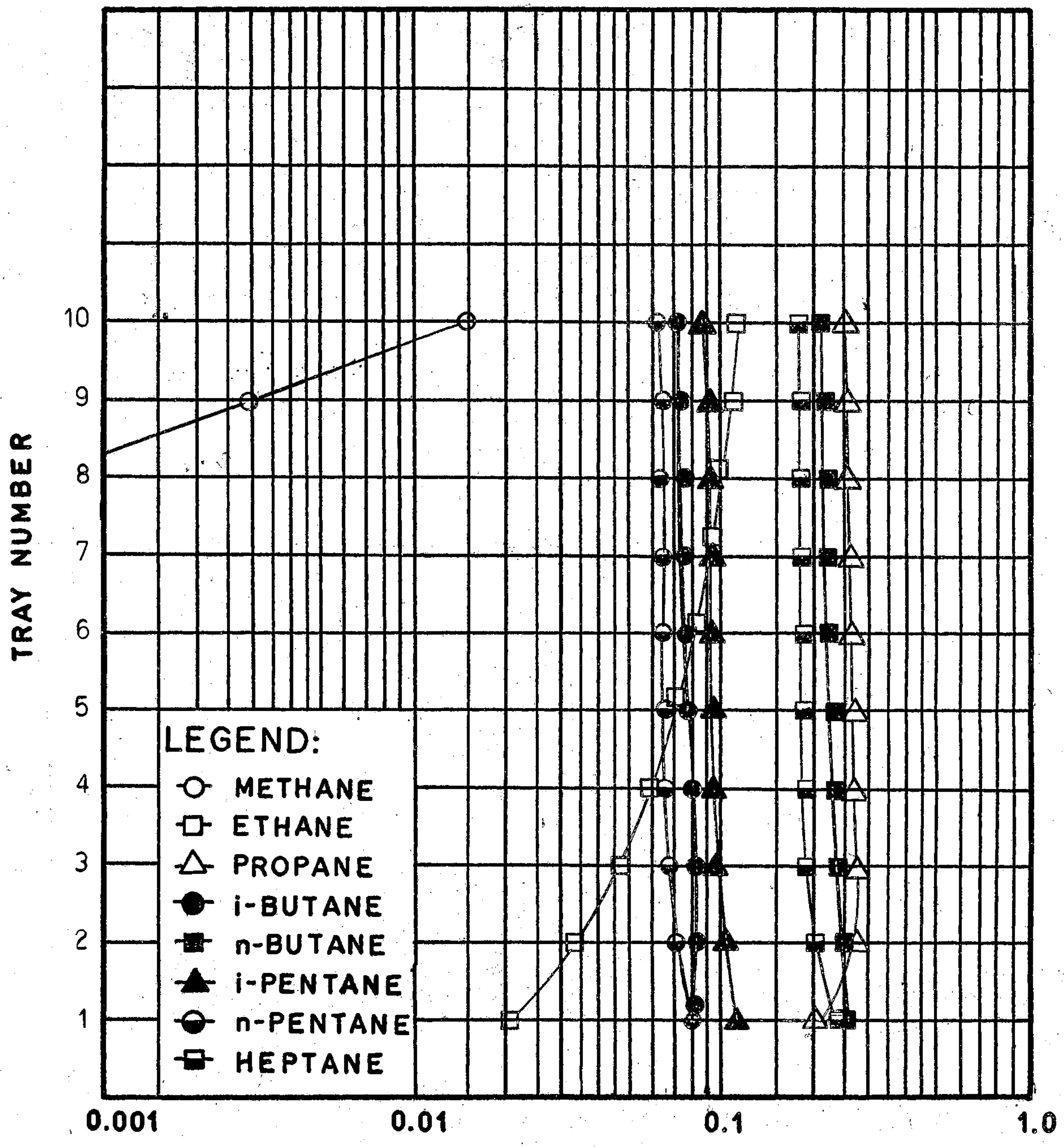


FIGURE 28

TRAY CONCENTRATION DIAGRAM

V/F = 0.30, 10 TRAYS, & 300 PSIA

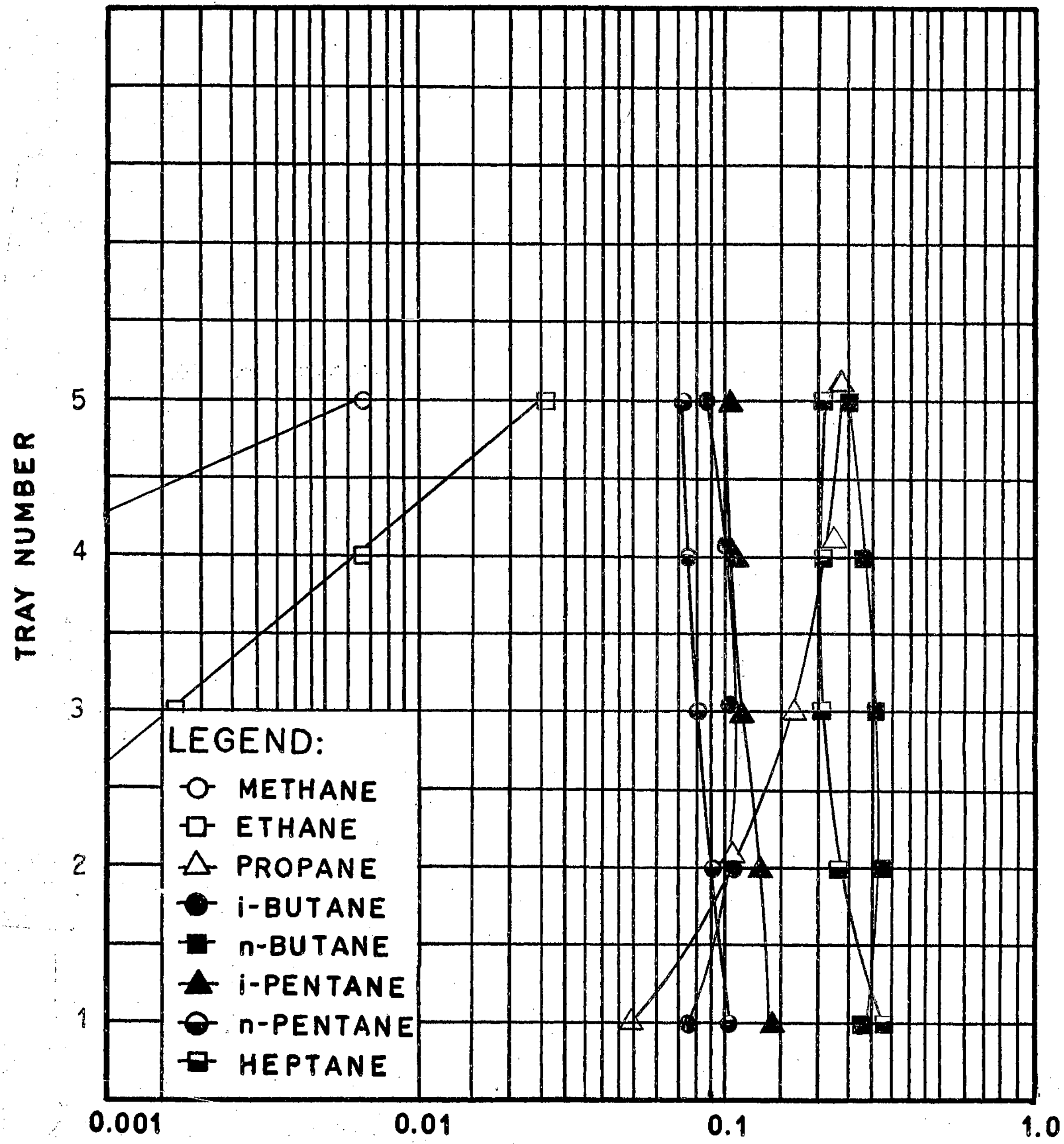


FIGURE 29

TRAY CONCENTRATION DIAGRAM

V/F = 0.50, 5 TRAYS, & 75 PSIA

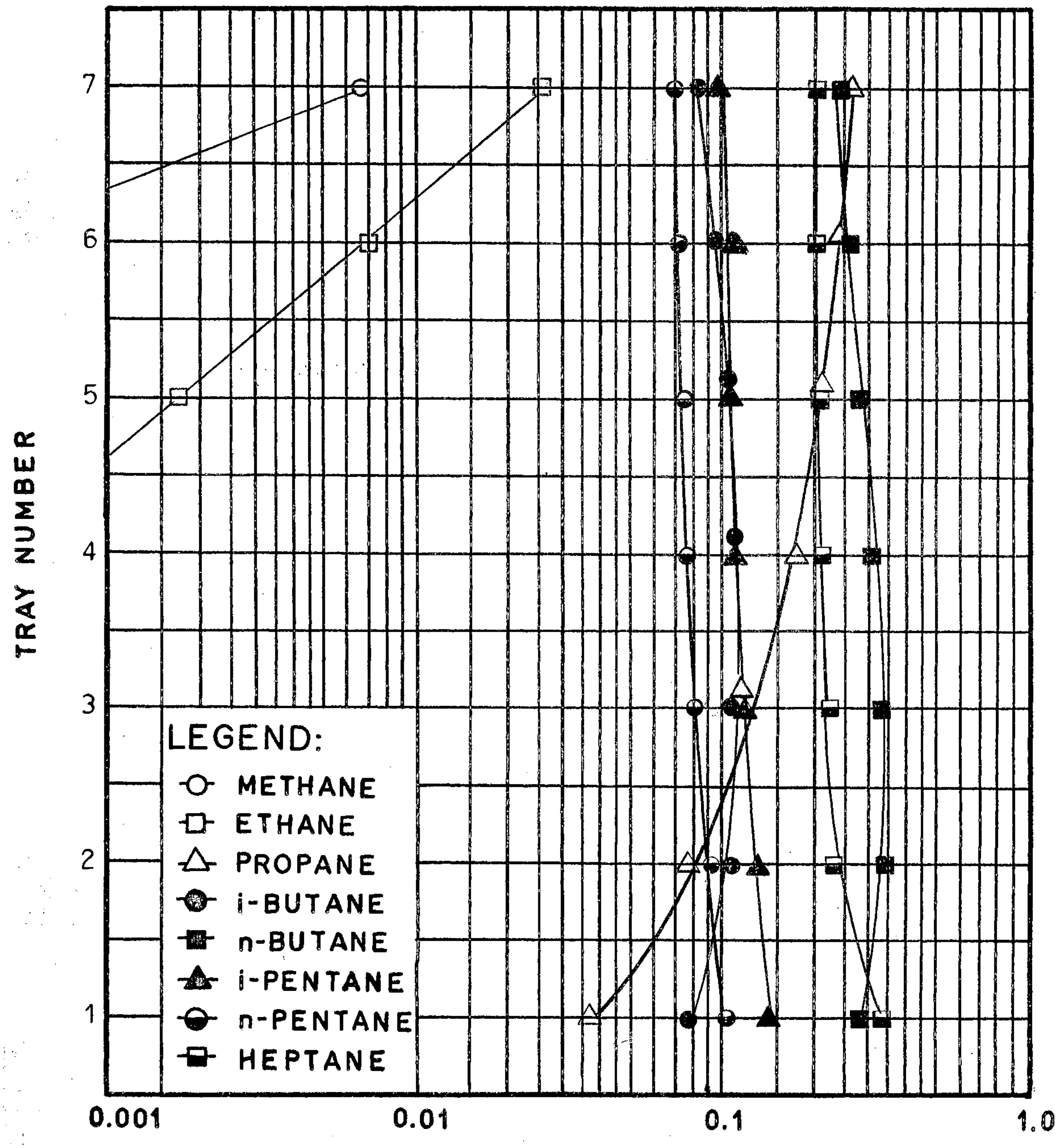


FIGURE 30

TRAY CONCENTRATION DIAGRAM

V/F = 0.50, 7 TRAYS, & 75 PSIA

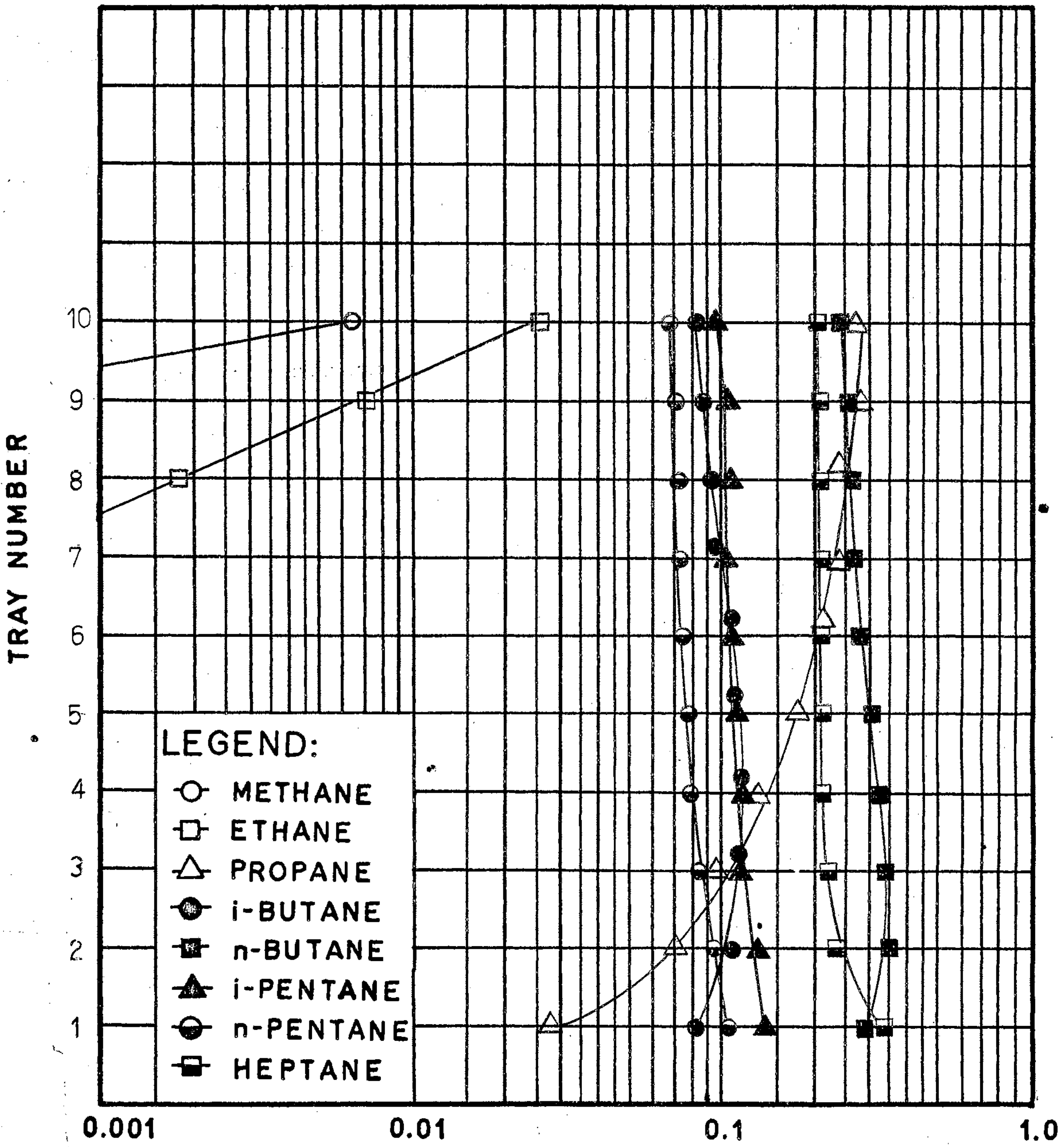


FIGURE 31

TRAY CONCENTRATION DIAGRAM

V/F = 0.50, 10 TRAYS, & 75 PSIA

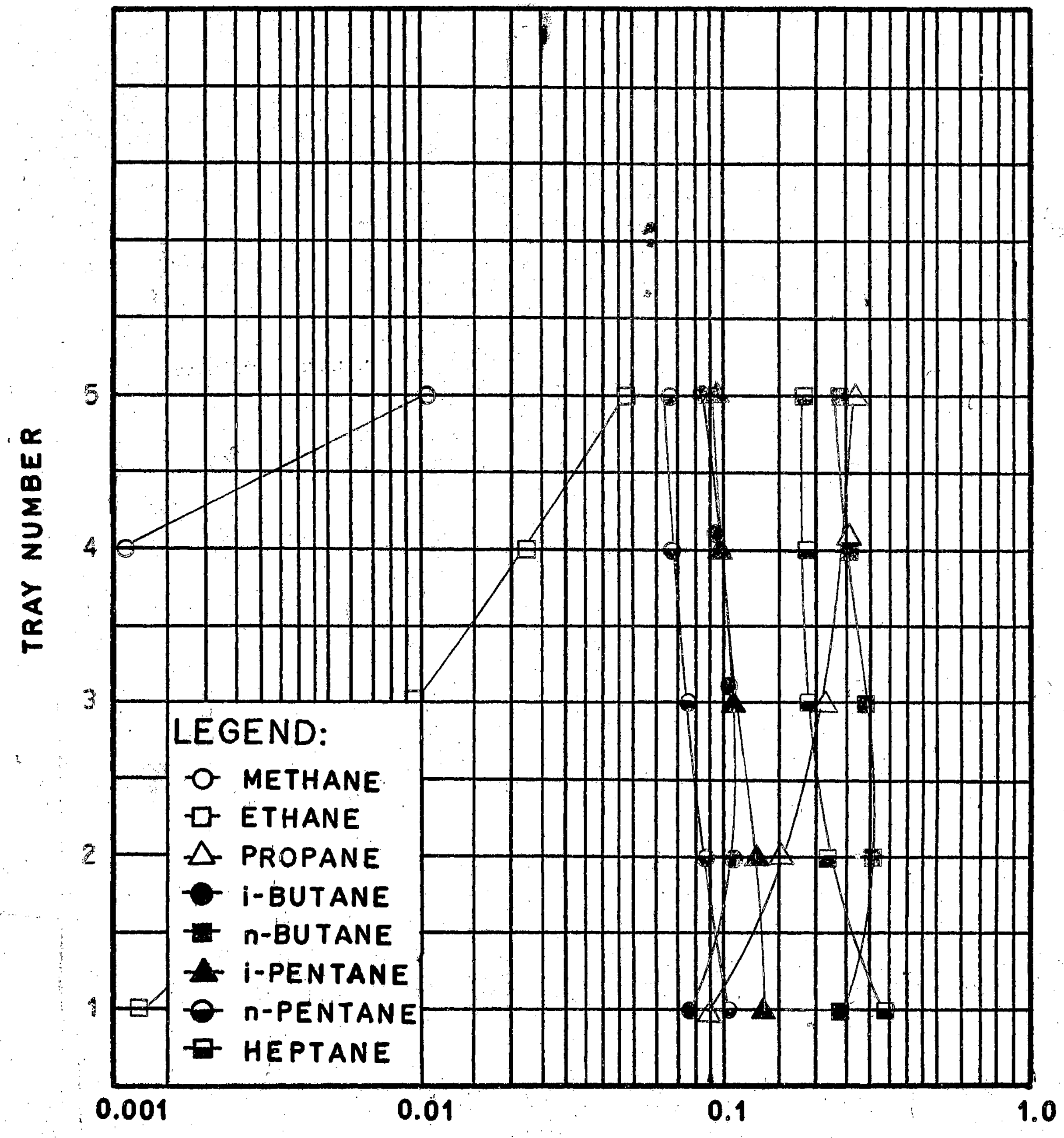


FIGURE 32

TRAY CONCENTRATION DIAGRAM

V/F = 0.50, 5 TRAYS, & 200 PSIA

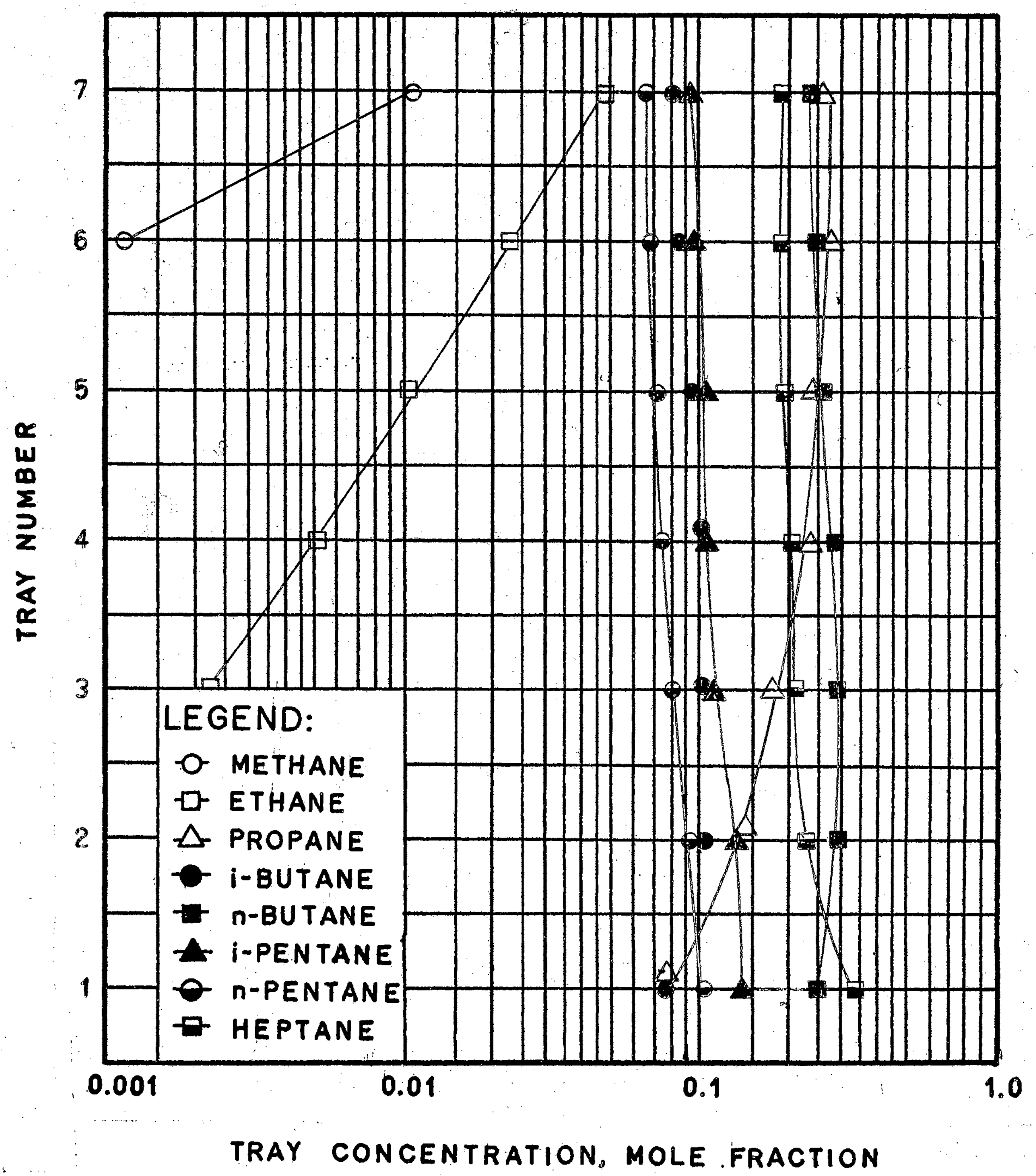


FIGURE 33

TRAY CONCENTRATION DIAGRAM

V/F = 0.50, 7 TRAYS, & 200 PSIA

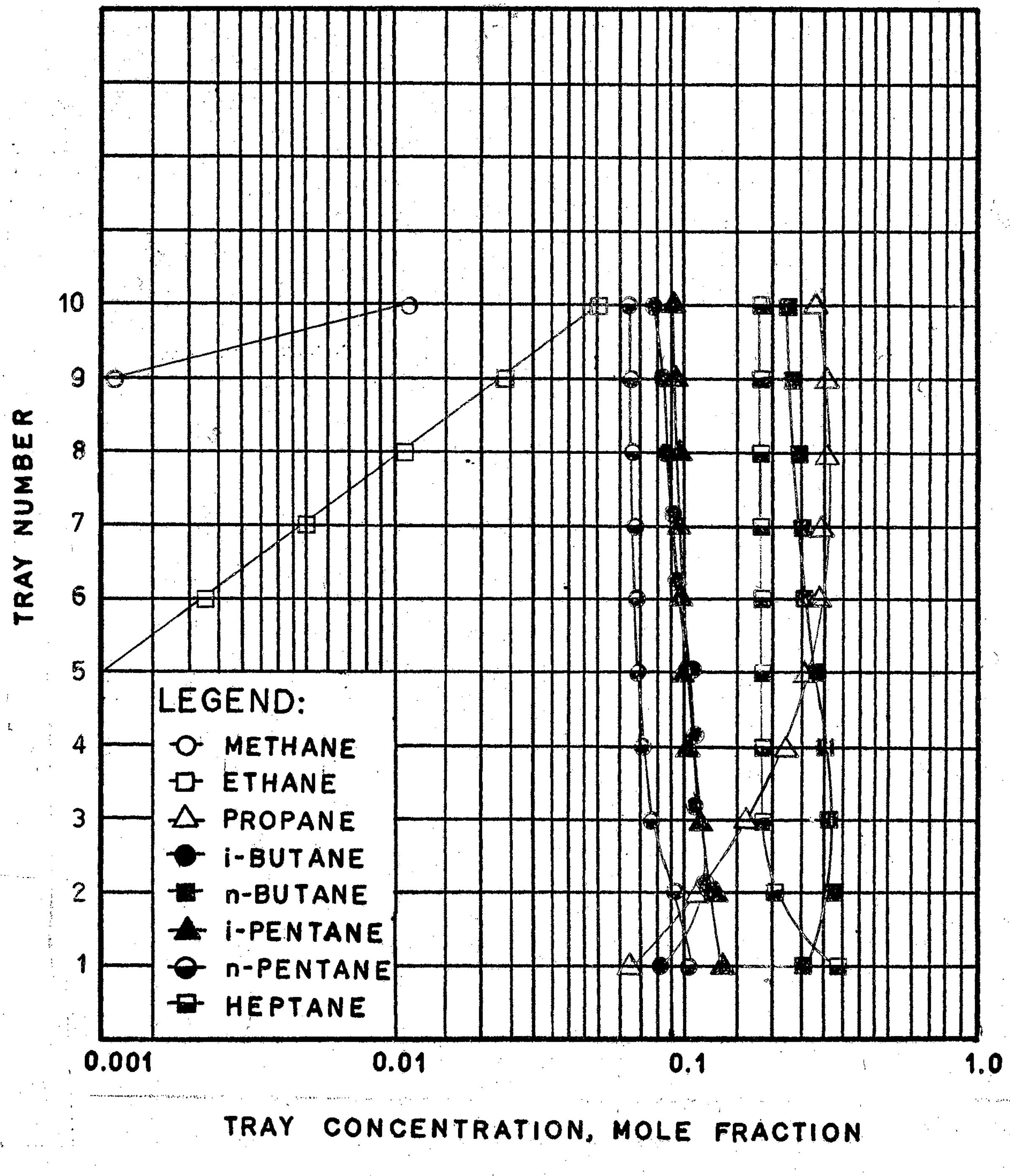


FIGURE 34

TRAY CONCENTRATION DIAGRAM

V/F = 0.50, 10 TRAYS, & 200 PSIA

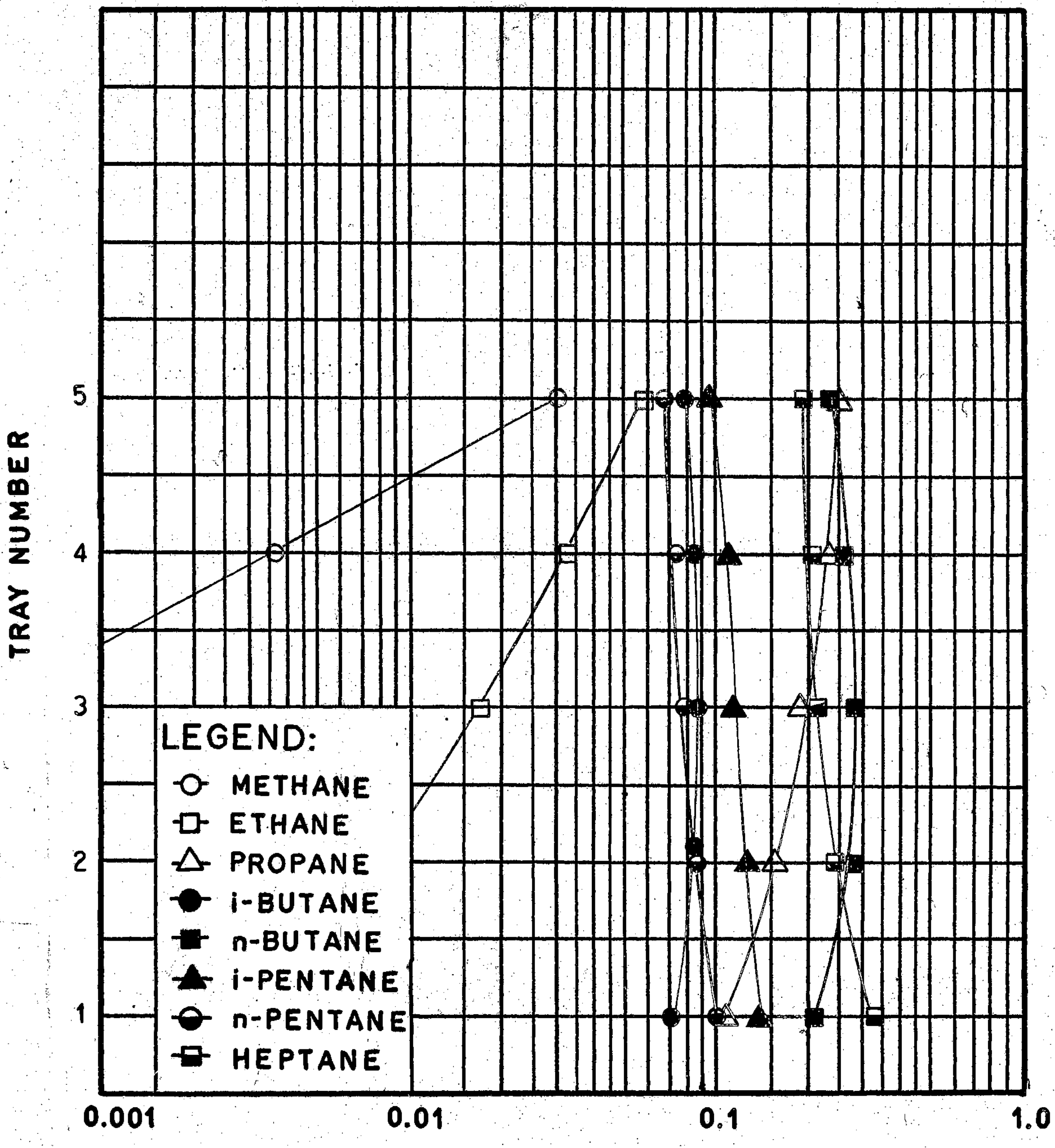


FIGURE 35

TRAY CONCENTRATION DIAGRAM

V/F = 0.50, 5 TRAYS, & 300 PSIA

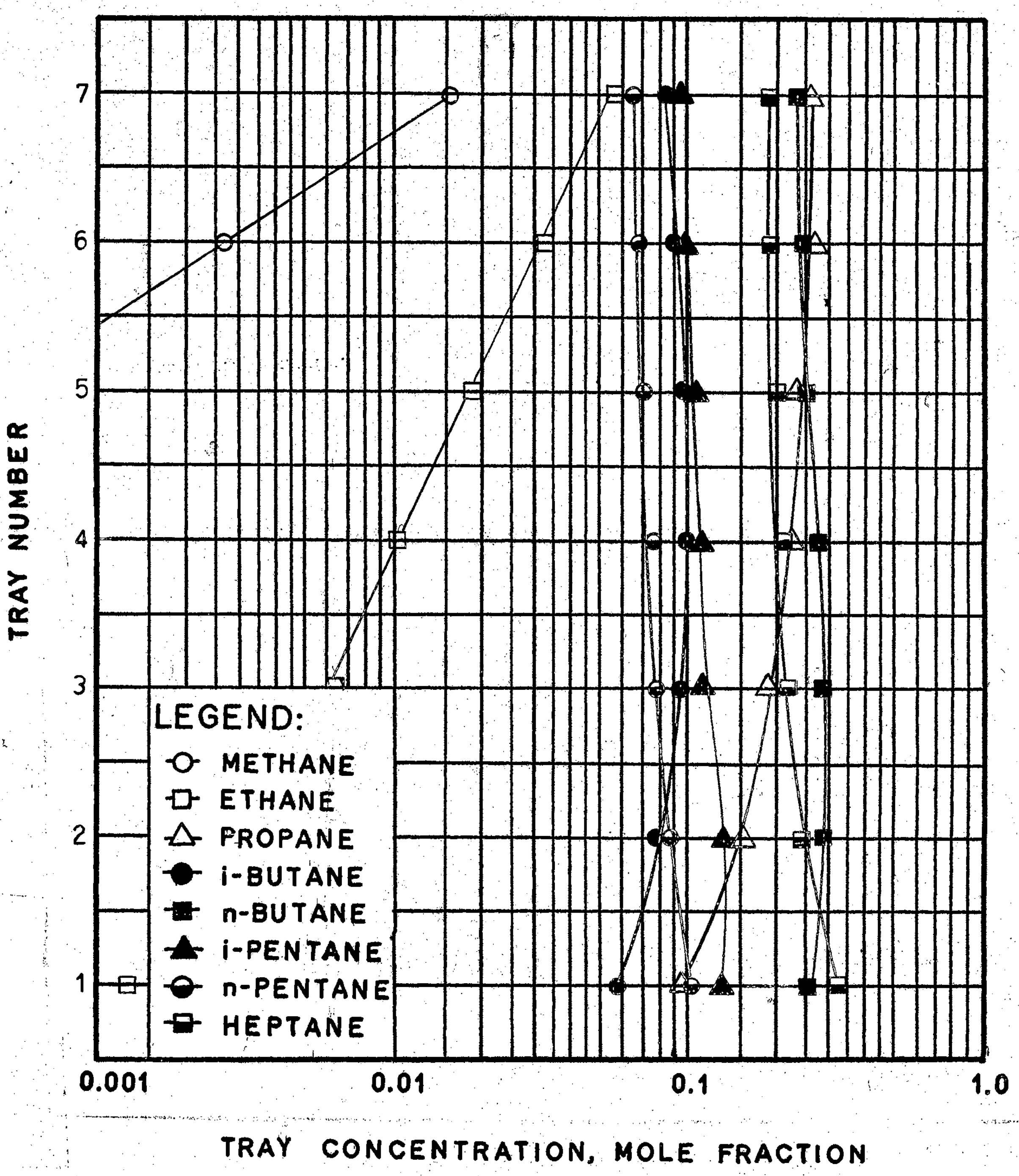


FIGURE 36

TRAY CONCENTRATION DIAGRAM

V/F = 0.50, 7 TRAYS, & 300 PSIA

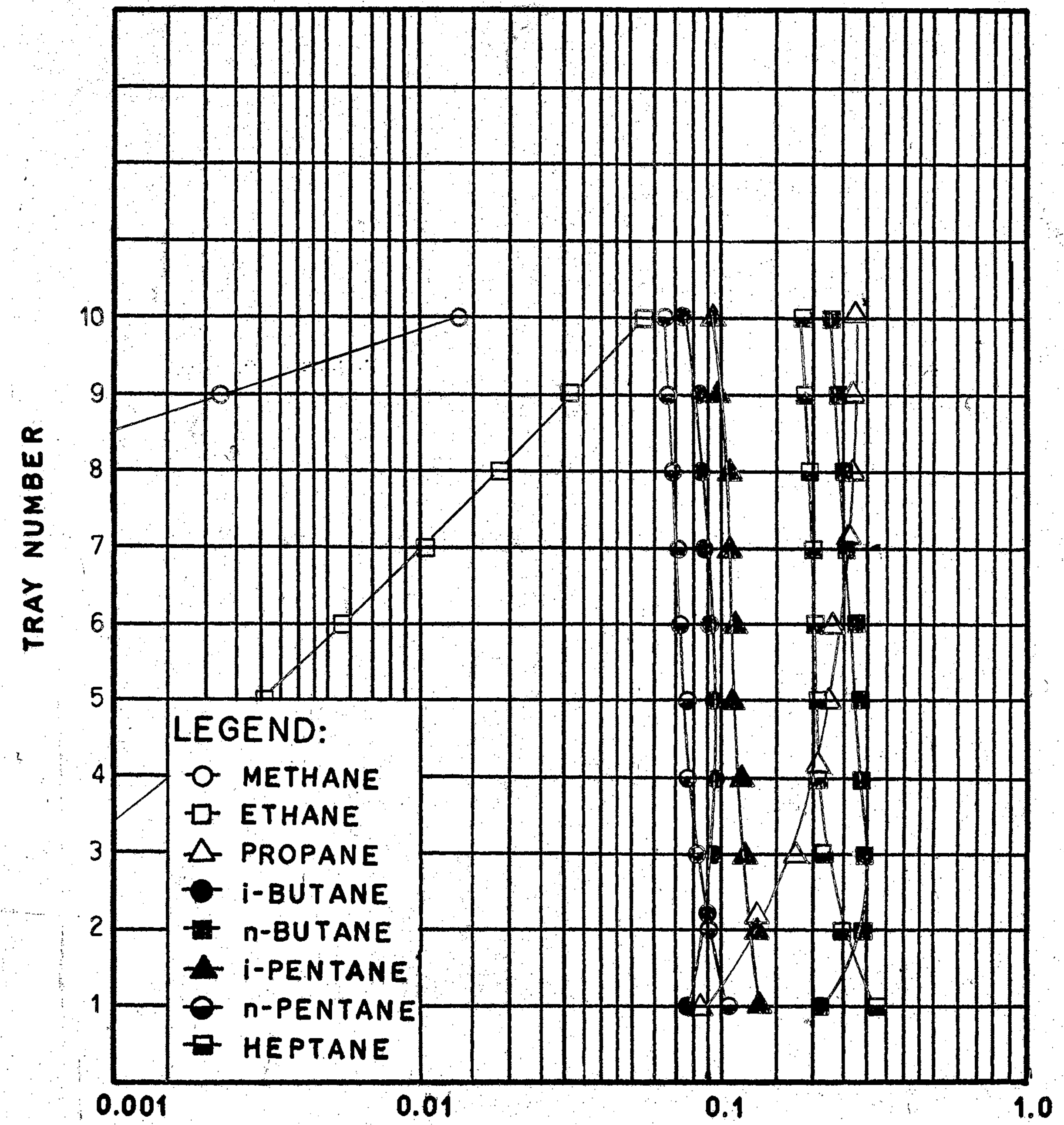


FIGURE 37

TRAY CONCENTRATION DIAGRAM

V/F = 0.50, 10 TRAYS, & 300 PSIA

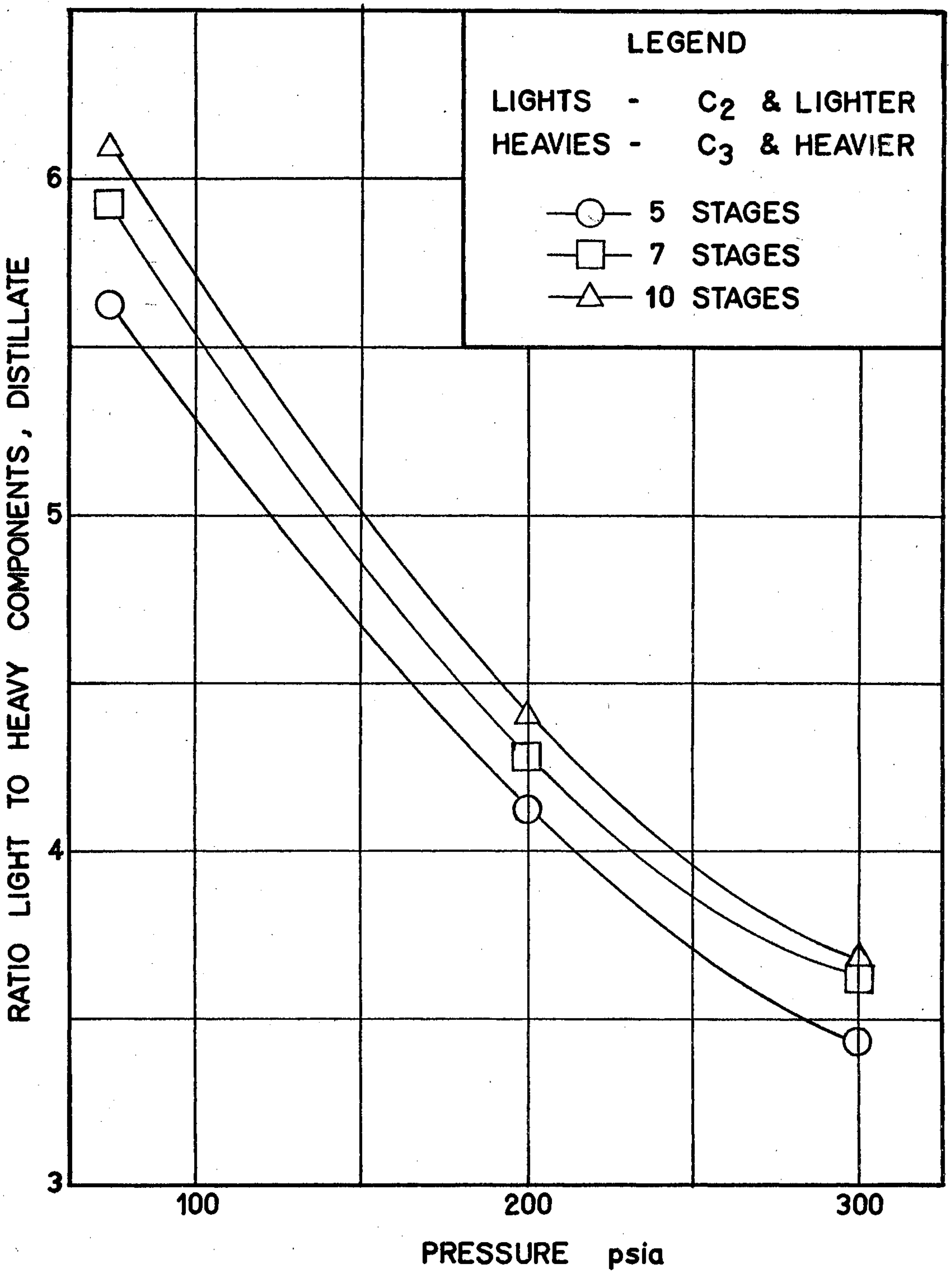


FIGURE 38

RATIO LIGHT TO HEAVY COMPONENTS vs PRESSURE

D/F = 0.21

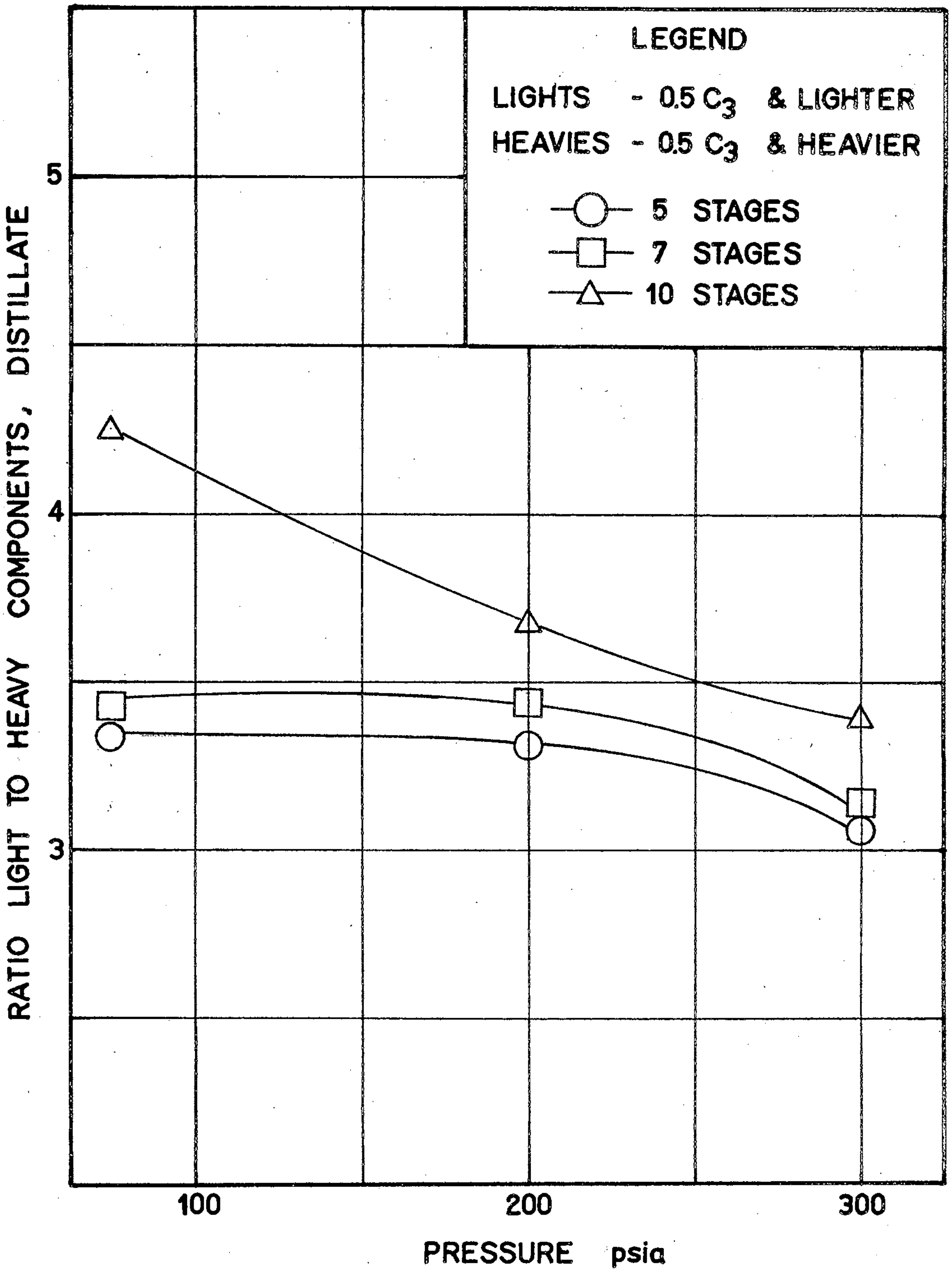


FIGURE 39

RATIO LIGHT TO HEAVY COMPONENTS vs PRESSURE

D/F = 0.30

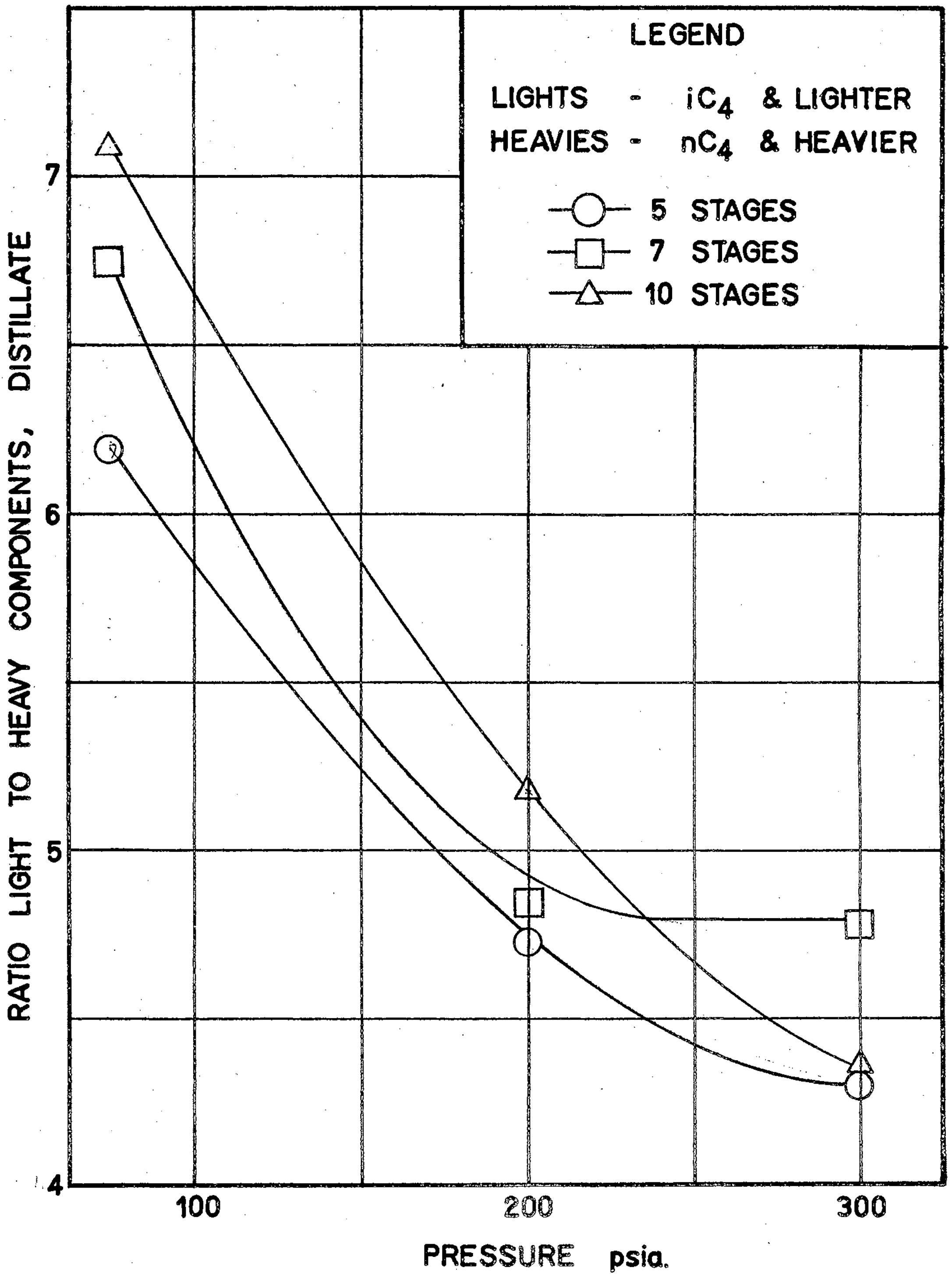


FIGURE 40

RATIO LIGHT TO HEAVY COMPONENTS vs PRESSURE

D/F = 0.50

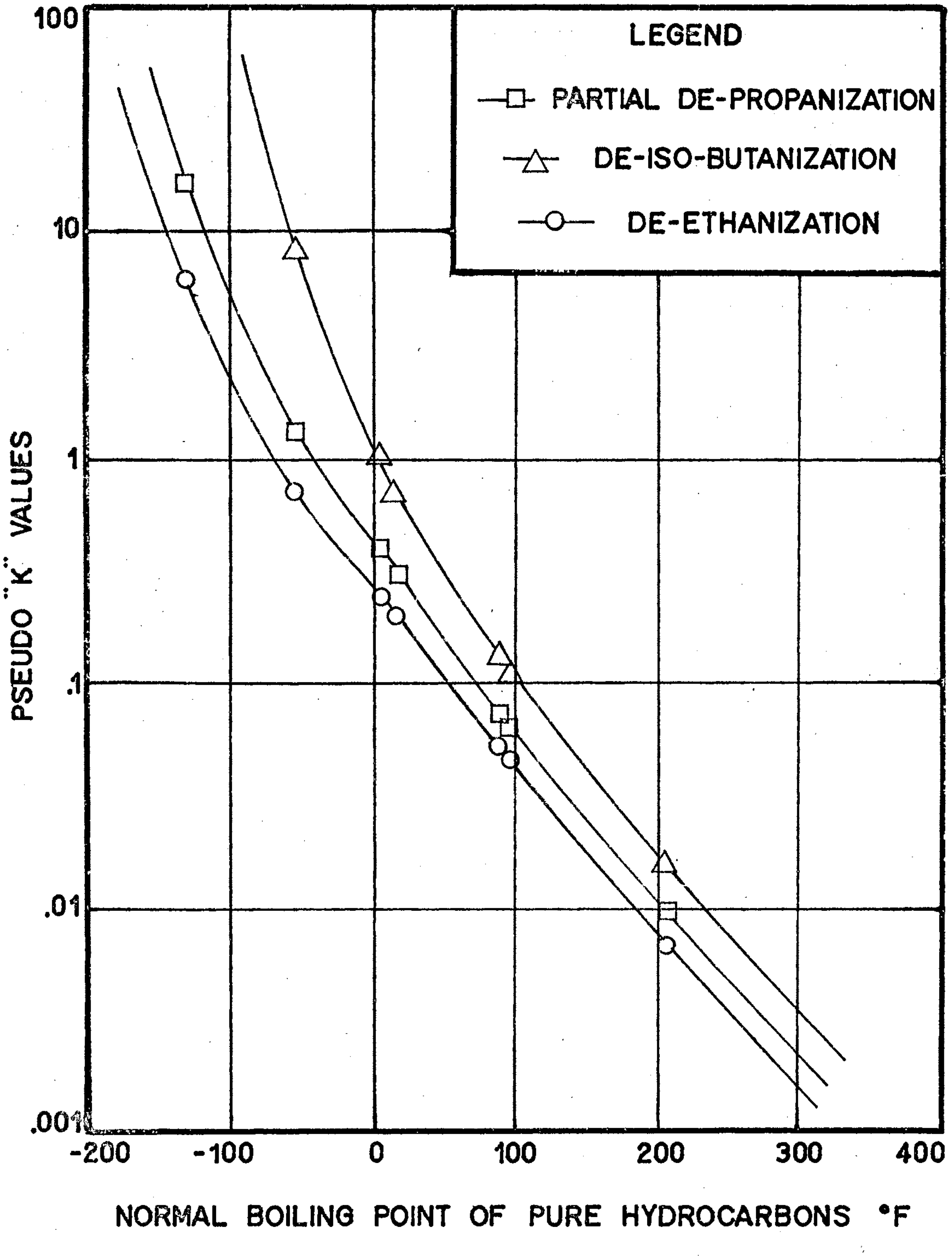


FIGURE 41

PSEUDO "K" VALUES

TABLE II

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 75-.21-5

NUMBER OF STAGES = 5.00

PRESSURE = 75 psia.

V/F AND D/F = 0.21

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.14561	0.54191	0.48465	0.00385
C_2	8.87368	0.26501	6.04615	0.04800
C_3	4.58171	0.13683	30.11803	0.23910
iC_4	0.47714	0.01425	9.82340	0.07799
nC_4	1.08793	0.03249	29.78179	0.23643
iC_5	0.17206	0.00514	12.86785	0.10215
nC_5	0.10015	0.00299	9.17990	0.07288
C_{6+} (C_7)	0.04643	0.00139	27.66297	0.21961
Totals	33.48472	1.00000	125.96473	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.66227 BTU/hr.
Reboiler	1.59049 BTU/hr.	Condenser	-0.00057 BTU/hr.
		Bottoms	0.21699 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE III

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 75-.21-7

NUMBER OF STAGES = 7.00

PRESSURE = 75 psia.

V/F AND D/F = 0.21

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.47923	0.55187	0.15103	0.00120
C_2	8.65364	0.25844	6.26619	0.04975
C_3	4.49427	0.13422	30.20548	0.23979
iC_4	0.47026	0.01404	9.83028	0.07804
nC_4	1.07278	0.03204	29.79694	0.23655
iC_5	0.16984	0.00507	12.87008	0.10217
nC_5	0.09886	0.00295	9.18119	0.07289
C_{6+} (C_7)	0.04586	0.00137	27.66354	0.21961
Totals	33.48472	1.00000	125.96473	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.65320 BTU/hr.
Reboiler	1.58898 BTU/hr.	Condenser	-0.00057 BTU/hr.
		Bottoms	0.22455 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE IV

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 75-.21-10

NUMBER OF STAGES = 10.00

PRESSURE = 75 psia.

V/F AND D/F = 0.21

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.60606	0.55566	0.02420	0.00019
C_2	8.67561	0.25909	6.24422	0.04957
C_3	4.38610	0.13099	30.31364	0.24065
iC_4	0.45984	0.01373	9.84070	0.07812
nC_4	1.04927	0.03134	29.82045	0.23674
iC_5	0.16619	0.00496	12.87372	0.10220
nC_5	0.09675	0.00289	9.18330	0.07290
C_{6+} (C_7)	0.04490	0.00134	27.66451	0.21962
Totals	33.48472	1.00000	125.96473	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.64278 BTU/hr.
Reboiler	1.58250 BTU/hr.	Condenser	-0.00057 BTU/hr.
		Bottoms	0.22849 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE V

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 200-.21-5

NUMBER OF STAGES = 5.00

PRESSURE = 200 psia.

V/F AND D/F = 0.21

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.48220	0.55439	0.04806	0.00038
C_2	9.88025	0.29477	5.03958	0.04002
C_3	3.23513	0.09652	31.46461	0.24986
iC_4	0.47724	0.01424	9.82330	0.07801
nC_4	1.01813	0.03038	29.85159	0.23705
iC_5	0.17130	0.00511	12.86861	0.10219
nC_5	0.09708	0.00290	9.18297	0.07292
C_{6+} (C_7)	0.05708	0.00170	27.65232	0.21958
Totals	33.51842	1.00000	125.93104	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.07221 BTU/hr.
Reboiler	1.52040 BTU/hr.	Condenser	-0.00324 BTU/hr.
		Bottoms	0.73963 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE VI

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 200-.21-7

NUMBER OF STAGES = 7.00

PRESSURE = 200 psia.

V/F AND D/F = 0.21

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.62524	0.55517	0.00502	0.00004
C_2	9.87290	0.29727	4.94693	0.03929
C_3	3.16260	0.09427	31.53714	0.25049
iC_4	0.46683	0.01391	9.83372	0.07811
nC_4	0.99380	0.02962	29.87592	0.23730
iC_5	0.16920	0.00504	12.87071	0.10223
nC_5	0.09621	0.00287	9.18384	0.07294
C_{6+} (C_7)	0.06164	0.00184	27.64776	0.21960
Totals	33.54842	1.00000	125.90103	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.06918 BTU/hr.
Reboiler	1.52506 BTU/hr.	Condenser	-0.00518 BTU/hr.
		Bottoms	0.74926 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE VII

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 200-.21-10

NUMBER OF STAGES = 10.00

PRESSURE = 200 psia.

V/F AND D/F = 0.21

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63011	0.55320	0.00016	0.00000
C_2	10.41683	0.30932	4.50299	0.03580
C_3	2.78803	0.08576	31.81171	0.25293
iC_4	0.45369	0.01347	9.84685	0.07829
nC_4	1.01488	0.03014	29.85484	0.23737
iC_5	0.20464	0.00608	12.83527	0.10205
nC_5	0.12617	0.00375	9.15388	0.07278
C_{6+} (C_7)	-0.05728	-0.00170	27.76668	0.22077
Totals	33.67707	1.00000	125.77238	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.06078 BTU/hr.
Reboiler	1.52798 BTU/hr.	Condenser	-0.01056 BTU/hr.
		Bottoms	0.76596 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE VIII

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 300-.21-5

NUMBER OF STAGES = 5.00

PRESSURE = 300 psia.

V/F AND D/F = 0.21

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.31335	0.54692	0.31691	0.00252
C_2	7.62590	0.22774	7.29393	0.05790
C_3	4.95299	0.14792	29.74675	0.23615
iC_4	0.68567	0.02048	9.61487	0.07633
nC_4	1.43928	0.04298	29.43044	0.23364
iC_5	0.26114	0.00780	12.77877	0.10145
nC_5	0.14408	0.00430	9.13597	0.07253
C_{6+} (C_7)	0.06231	0.00186	27.64709	0.21948
Totals	33.48472	1.00000	125.96473	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.65345 BTU/hr.
Reboiler	2.23991 BTU/hr.	Condenser	-0.00079 BTU/hr.
		Bottoms	0.87545 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE IX

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 300-.21-7

NUMBER OF STAGES = 7.00

PRESSURE = 300 psia.

V/F AND D/F = 0.21

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.54893	0.55395	0.08133	0.00065
C_2	7.70594	0.23013	7.21389	0.05727
C_3	4.73787	0.14149	29.96187	0.23786
iC_4	0.65579	0.01958	9.64475	0.07657
nC_4	1.38543	0.04137	29.48429	0.23407
iC_5	0.25205	0.00753	12.78786	0.10152
nC_5	0.13918	0.00416	9.14087	0.07257
C_{6+} (C_7)	0.05953	0.00178	27.64987	0.21950
Totals	33.48472	1.00000	125.96473	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.63618 BTU/hr.
Reboiler	2.23941 BTU/hr.	Condenser	-0.00098 BTU/hr.
		Bottoms	0.89241 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE X

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 300-.21-10

NUMBER OF STAGES = 10.00

PRESSURE = 300 psia.

V/F AND D/F = 0.21

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.62123	0.55611	0.00903	0.00007
C_2	7.68188	0.22941	7.23795	0.05746
C_3	4.70980	0.14066	29.98994	0.23808
iC_4	0.64929	0.01939	9.65125	0.07662
nC_4	1.37641	0.04111	29.49331	0.23414
iC_5	0.25004	0.00747	12.78987	0.10154
nC_5	0.13807	0.00412	9.14199	0.07258
C_{6+} (C_7)	0.05801	0.00173	27.65139	0.21952
Totals	33.48472	1.00000	125.96473	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.63246 BTU/hr.
Reboiler	2.24029 BTU/hr.	Condenser	-0.00093 BTU/hr.
		Bottoms	0.89696 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XI

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 75-.3-5

NUMBER OF STAGES = 5.00

PRESSURE = 75 psia.

V/F AND D/F = 0.30

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.62758	0.38941	0.00268	0.00002
C_2	12.65651	0.26459	2.26332	0.02028
C_3	10.95091	0.22893	23.74883	0.21278
iC_4	1.24063	0.02594	9.05991	0.08117
nC_4	3.53101	0.07382	27.33871	0.24494
iC_5	0.47320	0.00989	12.56671	0.11259
nC_5	0.25207	0.000527	9.02798	0.08089
C_{6+} (C_7)	0.10341	0.000216	27.60599	0.24733
Totals	47.83532	1.00000	111.61413	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.93817 BTU/hr.
Reboiler	2.38407 BTU/hr.	Condenser	-0.00041 BTU/hr.
		Bottoms	0.73451 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XII

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 75-.3-7

NUMBER OF STAGES = 7.00

PRESSURE = 75 psia.

V/F AND D/F = 0.30

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63017	0.38946	0.00009	0.00000
C_2	13.29408	0.27791	1.62575	0.01457
C_3	10.48278	0.21914	24.21697	0.21697
iC_4	1.19991	0.02508	9.10063	0.08154
nC_4	3.42245	0.07155	27.44727	0.24591
iC_5	0.45958	0.00961	12.58033	0.11271
nC_5	0.24517	0.00513	9.03488	0.08095
C_{6+} (C_7)	0.10118	0.00212	27.60822	0.24735
Totals	47.83532	1.00000	111.61413	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.92388 BTU/hr.
Reboiler	2.38366 BTU/hr.	Condenser	-0.00029 BTU/hr.
		Bottoms	0.74827 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XIII

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 75-.3-10

NUMBER OF STAGES = 10.00

PRESSURE = 75 psia.

V/F AND D/F = 0.30

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63026	0.38947	0.00000	0.00000
C_2	14.87886	0.31104	0.04097	0.00037
C_3	10.46511	0.21877	24.23463	0.21713
iC_4	1.08330	0.02265	9.21724	0.08258
nC_4	2.14527	0.04485	28.72445	0.25735
iC_5	0.34374	0.00719	12.69617	0.11375
nC_5	0.20537	0.00429	9.07468	0.08130
C_{6+} (C_7)	0.08341	0.00174	27.62599	0.24751
Totals	47.83532	1.00000	111.61413	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	2.74012 BTU/hr.
Reboiler	2.85328 BTU/hr.	Condenser	0.00140 BTU/hr.
		Bottoms	0.39996 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XIV

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 200-.3-5

NUMBER OF STAGES = 5.00

PRESSURE = 200 psia.

V/F AND D/F = 0.30

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.62789	0.38956	0.00237	0.00002
C_2	12.70710	0.26741	2.13273	0.01911
C_3	10.73865	0.22457	23.96110	0.21464
iC_4	1.25457	0.02624	9.04597	0.08103
nC_4	3.56760	0.07461	27.30212	0.24457
iC_5	0.48071	0.01005	12.55920	0.11251
nC_5	0.25633	0.00536	9.02372	0.08083
C_{6+} (C_7)	0.10496	0.00219	27.60444	0.24728
Totals	47.81781	1.00000	111.63165	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.93449 BTU/hr.
Reboiler	2.38236 BTU/hr.	Condenser	-0.00042 BTU/hr.
		Bottoms	0.73649 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XV

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 200-.3-7

NUMBER OF STAGES = 7.00

PRESSURE = 200 psia.

V/F AND D/F = 0.30

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63018	0.38950	0.00008	0.00000
C_2	13.38828	0.27991	1.53155	0.01372
C_3	10.35136	0.21641	24.34839	0.21814
iC_4	1.20597	0.02521	9.09457	0.08148
nC_4	3.44053	0.07193	27.42919	0.24574
iC_5	0.46287	0.00918	12.57704	0.11268
nC_5	0.24702	0.00516	9.03303	0.08093
C_{6+} (C_7)	0.10508	0.00220	27.60432	0.24731
Totals	47.83129	1.00000	111.61816	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.92029 BTU/hr.
Reboiler	2.38159 BTU/hr.	Condenser	-0.00022 BTU/hr.
		Bottoms	0.74972 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XVI

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 200-.3-10

NUMBER OF STAGES = 10.00

PRESSURE = 200 psia.

V/F AND D/F = 0.30

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63026	0.38939	0.00000	0.00000
C_2	13.91487	0.29083	1.00496	0.00900
C_3	9.99605	0.20893	24.70369	0.22135
iC_4	1.17071	0.02447	9.12983	0.08181
nC_4	3.34414	0.06990	27.52558	0.24663
iC_5	0.44836	0.00937	12.59155	0.11282
nC_5	0.23840	0.00498	9.04165	0.08102
C_{6+} (C_7)	0.10207	0.00213	27.60733	0.24737
Totals	47.84487	1.00000	111.60458	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	1.91009 BTU/hr.
Reboiler	2.38300 BTU/hr.	Condenser	0.00006 BTU/hr.
		Bottoms	0.76105 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XVII

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 300-.3-5

NUMBER OF STAGES = 5.00

PRESSURE = 300 psia.

V/F AND D/F = 0.30

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.60464	0.38893	0.02562	0.00023
C_2	11.68746	0.24433	3.23237	0.02896
C_3	11.44026	0.23916	23.25948	0.20839
iC_4	1.54744	0.03235	8.75310	0.07842
nC_4	3.35610	0.07016	27.51362	0.24651
iC_5	0.66685	0.01394	12.37306	0.11086
nC_5	0.37566	0.00785	8.90439	0.07978
C_{6+} (C_7)	0.15691	0.00328	27.55250	0.24685
Totals	47.83532	1.00000	111.61413	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	2.78354 BTU/hr.
Reboiler	3.49921 BTU/hr.	Condenser	0.00328 BTU/hr.
		Bottoms	1.00059 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XVIII

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 300-.3-7

NUMBER OF STAGES = 7.00

PRESSURE = 300 psia.

V/F AND D/F = 0.30

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.62865	0.38943	0.00161	0.00001
C_2	12.25036	0.25609	2.66947	0.02392
C_3	10.97965	0.22953	23.72009	0.21252
iC_4	1.50827	0.03153	8.79227	0.07877
nC_4	3.28502	0.06867	27.58470	0.24714
iC_5	0.65776	0.01375	12.38215	0.11094
nC_5	0.36905	0.00771	8.91100	0.07984
C_{6+} (C_7)	0.15657	0.00327	27.55283	0.24686
Totals	47.83532	1.00000	111.61413	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	2.76288 BTU/hr.
Reboiler	3.49009 BTU/hr.	Condenser	0.00162 BTU/hr.
		Bottoms	1.01379 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XIX

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 300-.3-10

NUMBER OF STAGES = 10.00

PRESSURE = 300 psia.

V/F AND D/F = 0.30

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63024	0.38947	0.00002	0.00000
C_2	12.61218	0.26366	2.30765	0.02068
C_3	11.38055	0.23791	23.31919	0.20893
iC_4	1.31253	0.02744	8.98801	0.08053
nC_4	2.90401	0.06071	27.96571	0.25056
iC_5	0.57775	0.01208	12.46216	0.11165
nC_5	0.29342	0.00613	8.98663	0.08052
C_{6+} (C_7)	0.12465	0.00261	27.58475	0.24714
Totals	47.83532	1.00000	111.61413	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	2.68817 BTU/hr.
Reboiler	3.42721 BTU/hr.	Condenser	-0.00054 BTU/hr.
		Bottoms	1.02778 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XX

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 75-.5-5

NUMBER OF STAGES = 5.00

PRESSURE = 75 psia.

V/F AND D/F = 0.50

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63026	0.23356	0.00000	0.00000
C_2	14.91357	0.18697	0.00626	0.00008
C_3	30.75591	0.38558	3.94383	0.04949
iC_4	4.37322	0.05483	5.92732	0.07439
nC_4	8.59157	0.10771	22.27815	0.27958
iC_5	1.38725	0.01739	11.65266	0.14624
nC_5	0.81173	0.01018	8.46832	0.10627
C_{6+} (C_7)	0.30171	0.00378	27.40768	0.34395
Totals	79.76523	1.00000	79.68422	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	6.40428 BTU/hr.
Reboiler	6.57807 BTU/hr.	Condenser	-0.00704 BTU/hr.
		Bottoms	0.46903 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XXI

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 75-.5-7

NUMBER OF STAGES = 7.00

PRESSURE = 75 psia.

V/F AND D/F = 0.50

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63026	0.23359	0.00000	0.00000
C_2	14.91942	0.18706	0.00041	0.00001
C_3	31.72919	0.39783	2.97055	0.03728
iC_4	4.08339	0.05120	6.21715	0.07801
nC_4	8.02220	0.10058	22.84752	0.28669
iC_5	1.31112	0.01644	11.72879	0.14718
nC_5	0.77194	0.00968	8.50811	0.10676
C_{6+} (C_7)	0.28911	0.00362	27.42029	0.34407
Totals	79.75663	1.00000	79.69282	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	6.35686 BTU/hr.
Reboiler	6.54325 BTU/hr.	Condenser	-0.00663 BTU/hr.
		Bottoms	0.48122 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XXII

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 75-.5-10

NUMBER OF STAGES = 10.00

PRESSURE = 75 psia.

V/F AND D/F = 0.50

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63026	0.23386	0.00000	0.00000
C_2	14.91982	0.18729	0.00001	0.00000
C_3	32.43491	0.40715	2.26483	0.02839
iC_4	3.81541	0.04789	6.48513	0.08128
nC_4	7.59188	0.09530	23.27784	0.29175
iC_5	1.25329	0.01573	11.78662	0.14773
nC_5	0.74009	0.00929	8.53996	0.10704
C_{6+} (C_7)	0.27796	0.00349	24.43144	0.34381
Totals	79.66363	1.00000	79.78582	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	6.30986 BTU/hr.
Reboiler	6.52215 BTU/hr.	Condenser	0.01059 BTU/hr.
		Bottoms	0.48990 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XXIII

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 200-.5-5

NUMBER OF STAGES = 5.00

PRESSURE = 200 psia.

V/F AND D/F = 0.50

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63014	0.23384	0.00012	0.00000
C_2	14.83220	0.18617	0.08763	0.00110
C_3	28.13648	0.35315	6.56326	0.08227
iC_4	4.11840	0.05169	6.18214	0.07749
nC_4	11.14454	0.13988	19.72519	0.24725
iC_5	1.59269	0.01999	11.44722	0.14349
nC_5	0.85772	0.01077	8.42233	0.10557
C_{6+} (C_7)	0.35970	0.00451	27.34970	0.34282
Totals	79.67187	1.00000	79.77759	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	3.91355 BTU/hr.
Reboiler	4.42148 BTU/hr.	Condenser	-0.00050 BTU/hr.
		Bottoms	0.79663 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XXIV

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 200-.5-7

NUMBER OF STAGES = 7.00

PRESSURE = 200 psia.

V/F AND D/F = 0.50

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63026	0.23368	0.00000	0.00000
C_2	14.89472	0.18683	0.02511	0.00031
C_3	28.44103	0.35674	6.25872	0.07850
iC_4	4.00527	0.05024	6.29527	0.07896
nC_4	10.86629	0.13630	20.00343	0.25091
iC_5	1.62520	0.02038	11.41471	0.14318
nC_5	0.88044	0.01104	8.39961	0.10536
C_{6+} (C_7)	0.38233	0.00480	27.32707	0.34277
Totals	79.72553	1.00000	79.72392	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	4.06187 BTU/hr.
Reboiler	4.57442 BTU/hr.	Condenser	-0.00087 BTU/hr.
		Bottoms	0.80162 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XXV

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 200-.5-10

NUMBER OF STAGES = 10.00

PRESSURE = 200 psia.

V/F AND D/F = 0.50

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63026	0.23360	0.00000	0.00000
C_2	14.91828	0.18705	0.00156	0.00002
C_3	29.52984	0.37026	5.16990	0.06487
iC_4	3.75574	0.04709	6.54480	0.08212
nC_4	10.24178	0.12842	20.62794	0.25883
iC_5	1.51222	0.01896	11.52769	0.14465
nC_5	0.81447	0.01021	8.46558	0.10622
C_{6+} (C_7)	0.35106	0.00440	27.35835	0.34328
Totals	79.75365	1.00000	79.69580	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	3.90169 BTU/hr.
Reboiler	4.42714 BTU/hr.	Condenser	-0.00370 BTU/hr.
		Bottoms	0.81735 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XXVI

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 300-.5-5

NUMBER OF STAGES = 5.00

PRESSURE = 300 psia.

V/F AND D/F = 0.50

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.62534	0.23361	0.00492	0.00006
C_2	14.62492	0.18343	0.29491	0.00370
C_3	26.71448	0.33507	7.98526	0.10017
iC_4	4.70970	0.05907	5.59084	0.07013
nC_4	10.68714	0.13404	20.18258	0.25317
iC_5	2.34466	0.02941	10.69525	0.13416
nC_5	1.30690	0.01639	7.97315	0.10001
C_{6+} (C_7)	0.71571	0.00898	26.99369	0.33860
Totals	79.72885	1.00000	79.72060	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	6.03869 BTU/hr.
Reboiler	6.73845 BTU/hr.	Condenser	-0.00175 BTU/hr.
		Bottoms	0.98971 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XXVII

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 300-.5-7

NUMBER OF STAGES = 7.00

PRESSURE = 300 psia.

V/F AND D/F = 0.50

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63012	0.23368	0.00014	0.00000
C_2	14.82069	0.18590	0.09914	0.00124
C_3	26.82143	0.33642	7.87831	0.09882
iC_4	5.75117	0.07214	4.54937	0.05706
nC_4	9.76032	0.12242	21.10940	0.26473
iC_5	2.11560	0.02654	10.92431	0.13703
nC_5	1.18723	0.01489	8.09282	0.10151
C_{6+} (C_7)	0.63898	0.00801	27.07042	0.33955
Totals	79.72553	1.00000	79.72392	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	5.95778 BTU/hr.
Reboiler	6.63712 BTU/hr.	Condenser	-0.00351 BTU/hr.
		Bottoms	0.97105 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XXVIII

FINAL MATERIAL AND HEAT BALANCES

PLATE TO PLATE RUN 300-.5-10

NUMBER OF STAGES = 10.00

PRESSURE = 300 psia.

V/F AND D/F = 0.50

MATERIAL BALANCE:

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.63025	0.23368	0.00001	0.00000
C_2	14.55842	0.18261	0.36141	0.00453
C_3	26.31134	0.33002	8.38840	0.10522
iC_4	5.27788	0.06620	5.02266	0.06300
nC_4	10.48182	0.13147	20.38790	0.25573
iC_5	2.36362	0.02965	10.67629	0.13392
nC_5	1.33215	0.01671	7.94790	0.09969
C_{6+} (C_7)	0.77005	0.00966	26.93935	0.33791
Totals	79.72553	1.00000	79.72392	1.00000

HEAT BALANCES:

Heat In:		Heat Out:	
Feed	0.28820 BTU/hr.	Distillate	6.09267 BTU/hr.
Reboiler	6.77990 BTU/hr.	Condenser	-0.00719 BTU/hr.
		Bottoms	0.98262 BTU/hr.

Note:

All enthalpies are scaled by a factor of 100,000.

TABLE XXIX

FLASH CALCULATION RESULTS

DE-ETHANIZATION

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.62313	0.54751	0.00687	0.00005
C_2	9.07097	0.26668	5.84903	0.04663
C_3	4.21648	0.12396	30.48352	0.24302
iC_4	0.54335	0.01597	9.75665	0.07778
nC_4	1.18217	0.03475	29.68783	0.23668
iC_5	0.20573	0.00605	12.83427	0.10232
nC_5	0.11737	0.00345	9.16563	0.07305
C_{6+} (C_7)	0.05586	0.00164	27.65414	0.22047
Totals	34.01506	1.00000	125.43494	1.00000

TABLE XXX

FLASH CALCULATION RESULTS

PARTIAL DEPROPANIZATION

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.62566	0.38916	0.00534	0.00004
C_2	13.05677	0.27281	1.86329	0.01670
C_3	10.76665	0.22496	23.93335	0.21448
iC_4	1.28342	0.02682	9.01658	0.08080
nC_4	3.23001	0.06749	27.63999	0.24768
iC_5	0.50648	0.01058	12.53352	0.11232
nC_5	0.27624	0.00577	9.00376	0.08069
C_{6+} (C_7)	0.11535	0.00241	27.59465	0.24729
Totals	47.86052	1.00000	111.58948	1.00000

TABLE XXXI

FLASH CALCULATION RESULTS

PARTIAL DEBUTANIZATION

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.62823	0.22758	0.00577	0.00002
C_2	14.91859	0.18226	0.00141	0.00002
C_3	30.19818	0.36892	4.50182	0.05802
iC_4	4.64802	0.05678	5.65198	0.07284
nC_4	10.14019	0.12388	20.72981	0.26715
iC_5	1.82106	0.02225	11.21894	0.14458
nC_5	1.02054	0.01247	8.25946	0.10644
C_{6+} (C_7)	0.47980	0.00586	27.23020	0.35093
Totals	81.85461	1.00000	77.59539	1.00000

TABLE XXXII

PRODUCT DISTRIBUTION AND MINIMUM STAGES

FENSKE CALCULATION - 75-.21

LIGHT KEY SPLIT - 0.05569

HEAVY KEY SPLIT - 25.11299

PRESSURE - 75 psia

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	12.08471	0.52373	6.54529	0.04799
C_2	4.80289	0.20815	10.11711	0.07419
C_3	3.99801	0.17327	30.70199	0.22513
iC_4	0.54335	0.02355	9.75665	0.07154
nC_4	1.18217	0.05123	29.68783	0.21769
iC_5	0.22742	0.00986	12.81258	0.09395
nC_5	0.14713	0.00638	9.13287	0.06697
C_{6+} (C_7)	0.00866	0.00384	27.62134	0.20254
Totals	23.07434	1.00000	136.37566	1.00000

Minimum Number of Stages - 0.81496

TABLE XXXIII

PRODUCT DISTRIBUTION AND MINIMUM STAGES

FENSKE CALCULATION - 75-.30

LIGHT KEY SPLIT - 0.14234

HEAVY KEY SPLIT - 8.55724

PRESSURE - 75 psia

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	9.82586	0.35101	8.80414	0.06697
C_2	4.98645	0.17813	9.93355	0.07557
C_3	6.47141	0.23118	28.22859	0.21474
iC_4	1.28342	0.04585	9.01658	0.06859
nC_4	3.23001	0.11539	27.63999	0.21026
iC_5	0.89832	0.03209	12.14164	0.09236
nC_5	0.59322	0.02119	8.68678	0.06608
C_{6+} (C_7)	0.70431	0.02516	27.00569	0.20543
Totals	27.99304	1.00000	131.45696	1.00000

Minimum Number of Stages - 0.49596

TABLE XXXIV

PRODUCT DISTRIBUTION AND MINIMUM STAGES

FENSKE CALCULATION - 75-0.5

LIGHT KEY SPLIT - 0.82237

HEAVY KEY SPLIT - 2.04432

PRESSURE - 75 psia

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.53085	0.24392	0.09915	0.00119
C_2	14.29335	0.18814	0.62665	0.00751
C_3	25.41137	0.33448	9.28863	0.11127
iC_4	4.64802	0.06118	5.65198	0.06771
nC_4	10.14019	0.13347	20.72981	0.24833
iC_5	1.71977	0.02264	11.32023	0.13561
nC_5	0.97620	0.01285	8.30380	0.09947
C_{6+} (C_7)	0.25206	0.00332	27.45794	0.32892
Totals	75.97181	1.00000	83.47819	1.00000

Minimum Number of Stages - 1.35184

TABLE XXXV

PRODUCT DISTRIBUTION AND MINIMUM STAGES

FENSKE CALCULATION - 200-.21

LIGHT KEY SPLIT - 0.05569

HEAVY KEY SPLIT - 25.11299

PRESSURE - 200 psia

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	14.16364	0.55173	4.46636	0.03339
C_2	5.02754	0.19584	9.89246	0.07395
C_3	4.36147	0.16990	30.33853	0.22678
iC_4	0.54335	0.02117	9.75665	0.07293
nC_4	1.18217	0.04605	29.68783	0.22192
iC_5	0.21671	0.00844	12.82329	0.09585
nC_5	0.11774	0.00459	9.16226	0.06849
C_{6+} (C_7)	0.05890	0.00229	27.65110	0.20669
Totals	25.67152	1.00000	133.77848	1.00000

Minimum Number of Stages - 0.97526

TABLE XXXVI

PRODUCT DISTRIBUTION AND MINIMUM STAGES

FENSKE CALCULATION - 200-.3

LIGHT KEY SPLIT - 0.14234

HEAVY KEY SPLIT - 8.55724

PRESSURE - 200 psia

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	11.44092	0.38331	7.18908	0.05547
C_2	5.17397	0.17334	9.74603	0.07520
C_3	6.78210	0.22722	27.91790	0.21541
iC_4	1.28342	0.04300	9.01658	0.06957
nC_4	3.23001	0.10822	27.63999	0.21327
iC_5	0.86469	0.02897	12.17531	0.09394
nC_5	0.52611	0.01763	8.75389	0.06754
C_{6+} (C_7)	0.54680	0.01832	27.16320	0.20959
Totals	29.84802	1.00000	129.60198	1.00000

Minimum Number of Stages - 0.61171

TABLE XXXVII

PRODUCT DISTRIBUTION AND MINIMUM STAGES

FENSKE CALCULATION - 200-.50

LIGHT KEY SPLIT - 0.82237

HEAVY KEY SPLIT - 2.04432

PRESSURE - 200 psia

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.59482	0.24465	0.03518	0.00042
C_2	14.39906	0.18944	0.52094	0.00624
C_3	25.82044	0.33971	8.87956	0.10641
iC_4	4.64802	0.06115	5.65198	0.06773
nC_4	10.14019	0.13341	20.72981	0.24843
iC_5	1.54271	0.02030	11.49729	0.13779
nC_5	0.74219	0.00976	8.53781	0.10232
C_{6+} (C_7)	0.11974	0.00158	27.59026	0.33065
Totals	76.00717	1.00000	83.44283	1.00000

Minimum Number of Stages - 1.91323

TABLE XXXVIII

PRODUCT DISTRIBUTION AND MINIMUM STAGES

FENSKE CALCULATION - 300-.21

LIGHT KEY SPLIT - 0.05569

HEAVY KEY SPLIT - 25.11299

PRESSURE - 300 psia

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	13.25985	0.56656	5.37015	0.03947
C_2	4.17513	0.17839	10.74487	0.07898
C_3	3.82674	0.16351	30.87326	0.22693
iC_4	0.54335	0.02322	9.75665	0.07172
nC_4	1.18217	0.05051	29.68783	0.21822
iC_5	0.23195	0.00991	12.80805	0.09415
nC_5	0.12849	0.00549	9.15151	0.06727
C_{6+} (C_7)	0.05641	0.00241	27.65359	0.20327
Totals	23.40409	1.00000	136.04591	1.00000

Minimum Number of Stages - 1.00509

TABLE XXXIX

PRODUCT DISTRIBUTION AND MINIMUM STAGES

FENSKE CALCULATION - 300-.30

LIGHT KEY SPLIT - 0.14234

HEAVY KEY SPLIT - 8.55724

PRESSURE - 300 psia

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	10.61737	0.37957	8.01263	0.06094
C_2	4.47453	0.15996	10.44547	0.07945
C_3	6.34894	0.22697	28.35106	0.21563
iC_4	1.28342	0.04588	9.01658	0.06858
nC_4	3.23001	0.11547	27.63999	0.21023
iC_5	0.90503	0.03235	12.13497	0.09230
nC_5	0.56237	0.02010	8.71763	0.06630
C_{6+} (C_7)	0.55052	0.01968	27.15948	0.20657
Totals	27.97219	1.00000	131.47781	1.00000

Minimum Number of Stages - 0.63060

TABLE XL

PRODUCT DISTRIBUTION AND MINIMUM STAGES

FENSKE CALCULATION - 300-0.5

LIGHT KEY SPLIT - 0.82237

HEAVY KEY SPLIT - 2.04432

PRESSURE - 300 psia

Component	Distillate Mols/hour	Mol Fraction	Bottoms Mols/hour	Mol Fraction
C_1	18.57230	0.24748	0.05770	0.00068
C_2	13.85554	0.18463	1.06446	0.01261
C_3	25.22112	0.33608	9.47888	0.11230
iC_4	4.64802	0.06194	5.65198	0.06696
nC_4	10.14019	0.13512	20.72981	0.24560
iC_5	1.70639	0.02274	11.33361	0.13428
nC_5	0.78759	0.01049	8.49241	0.10062
C_{6+} (C_7)	0.11452	0.00153	27.59548	0.32694
Totals	75.04567	1.00000	84.40433	1.00000

Minimum Number of Stages - 2.08576

CHAPTER V

DISCUSSION OF RESULTS

Certain of the relative volatility values were modified to avoid the presence of maximum or minimum points in these equations. The changes of relative volatility were made at the temperature or temperatures where a variation from the actual value would have the least effect on the system. They were changed only enough to eliminate the presence of the maximum or minimum. Indiscriminate changes can seriously impair the value of the results of the plate to plate calculation.

The temperature profiles of the plate to plate calculations are shown in Figures 2 through 10. In most instantances they approximate the typical temperature profile found in the stripping section of a distillation column. The distorted temperature profiles found in Figures 2, 3, 4, 5, 6, and 7 are a result of limitations placed on the maximum and minimum acceptable temperatures. When a calculated tray temperature is found to be outside the acceptable range either the specified maximum or minimum temperature is used rather than the calculated value. Such limitations are necessary to prevent the computer from going beyond the acceptable range of relative volatility values. The use of the specified temperatures will cause little error if actual tray temperature is not too different from the maximum or minimum.

The tray concentration diagrams are shown in Figures 11 through 37. These agree quite well with those for the stripping section of a distillation column found in the literature (9). The implications drawn from these plots is, that with the exception of propane and lighter components, little effective fractionation is occuring within the column. This idea is given further emphasis by the temperature profile plots especially those having sections that are vertical or nearly so. It might be noted that had a larger V/F and D/F fraction been chosen the component showing the greatest change from top to bottom would be one of the higher boiling components, iso-butane or perhaps n-butane.

The effect of pressure and number of stages on the ratio of lights to heavy components in the vapor and composition of the product streams is shown by Tables II through XXVIII and Figures 38 through 40. As expected, an increase in pressure with a constant number of stages, V/F and D/F caused a slightly poorer separation of the feed and a decrease in the ratio of light to heavy components in the vapor. An increase in the number of stages, holding pressure, V/F, and D/F constant caused a small increase in the ratio of light to heavy components and improved the quality of separation slightly.

The reasons for these behaviors may be best explained by reference to the Fenske equation (Equation 1). Let us first consider the effect of pressure. As pressure increases the temperature must also increase to maintain the fractionation process within the column. At a constant pressure an increase in temperature will cause the relative volatility of the various components to approach unity, ie, the relative volatility of the components lighter than the reference component tend to decrease and those heavier than the reference component tend to increase. This effect is magnified slightly by pressure. Now assume either of the key splits and the minimum number of stages constant, and let the relative volatility change as outlined for pressure increases. Equation (1) shows the other key split must decrease, causing a poorer separation.

Now let us consider the effect of the number of stages. To do this assume that the relative volatility and one of the specified key splits are constant. At a constant reflux ratio the number of stages becomes a function of the minumum number of stages. It is then possible to substitute, S, the number of stages for, S_m, the minimum number of stages. It is readily apparent then that for an increase in S the split not held constant must increase, thereby giving a better separation.

It will be noticed by reference to Figures 38, 39, and 40 that while the same general trend is noted in all curves, there is appreciable difference in the shapes of the curves.

This is brought about by two factors. The first being the size of the feed plate mismatch error and the manner in which the plate to plate routine adjusts this error. The second factor affecting the shape of the curves is the use of altered relative volatility data.

It may be concluded by reference to Figures 38 through 40 and Tables II through XXVIII that the effect of increasing pressure has a small but definite influence upon the quality of separation. They also show that an increase of the number of stages gives a small increase in the quality of separation.

The plate to plate routine assumed the presence of an additional stage which was in reality a partial condenser. The assumption of the condenser was not called for by the input data, but was due to the calculation sequence of the program. The assumption of the additional stage or partial condenser was not considered to be deterimental to the solution of the problem. The amount of reflux returned to the column in all cases was a negligible amount, usually less than 1.0% of the feed plate overflow. Furthermore the composition of the reflux very closely approximated the feed plate composition.

It will be noticed by reference to Tables II through XXVIII that in certain instances a negative condenser load was calculated. This was caused by the calculation of negative amount of reflux which resulted from the manner in which the program handles a partial condenser. The partial condenser is assumed to be an additional stage and is treated as such. Due to an accumulation of rounding errors the amount of distillate produced may be slightly larger than the amount of vapor from the top tray. This then would result in a negative quantity of reflux.

A series of pseudo vapor-liquid equilibrium constants were calculated from the data of Tables II through XXVIII. The resulting values were grouped according to D/F ratio and averaged. The plotted results appear in Figure 41. A curious feature of these curves is the "hump" appearing at the approximate boiling point of the butanes. Other investigators (10) have noticed the appearance of a similar phemonenon. The reasons for the occurence of this "hump" are not fully understood at this time.

Flash calculations were performed on the feed stream using the calculated pseudo equilibrium values. The results are shown in Tables XXIX through XXXI. A comparison of these values with the corresponding plate to plate results shows a good agreement between the two. This is an indication that the pseudo equilibrium constants are of definite value for use as a preliminary estimate in the determination of the approximate fractional overhead cut. It should be noted that had the pseudo equilibrium constants from the individual plate to plate results been used rather than the averaged values, the plate to plate results would have been reproduced.

In order to determine the minimum number of stages necessary to accomplish the separation predicted by the flash equation, a series of Fenske calculations were made using the splits of the flash calculation results. The results with the product distributions appear in Tables XXXII through XL. A comparison of the product distributions with the corresponding flash calculation results shows in most cases a poor agreement of both composition and total moles of overhead and bottoms products. However as the split components approach the chosen key components, iso- and n-butane, the agreement of data becomes better. This may be best explained by a comparison of conventional relative volatility data and relative volatility data derived from the pseudo equilibrium constants. In the case of the lighter components the relative volatilities derived from the pseudo constants are a great deal larger than the conventional data. The relative volatilities of the heavier components tend to be of the same magnitude. This means then the Fenske relationship will not predict as sharp a split on the lighter components. This is especially true when the chosen key components are somewhat heavier than the components being split.

A comparison of the minimum number stages resulting from the Fenske calculation shows the number of stages required to give the separation for both the 0.21 and 0.30 cut to be less than unity in most instances. This says that a flash separation would give a better separation of the chosen key components. However had the components being separated been chosen as keys the minimum number of stages would have been greater than unity. This may be shown by a comparison of the conventional relative volatility data and the relative volatilities derived from the pseudo constants.

A further comparison of the minimum number of stages presents an anomaly, ie, that the second fractional cut (0.30) requires fewer stages than does the 0.21 cut. In a system such as the one under consideration, it is usually found that the minimum number of stages varies as a direct function of the fractional cut. The explanation for this anomaly relies upon the pseudo constants. It will be noticed by reference to Figure 38 a "hump" appears in the curve depicting the 0.30 cut at the iso-normal butane boiling point. The pseudo equilibrium constants for iso- and normal butane are very nearly equal at this point. This then indicates a poor separation for the entire system if these components are chosen as key components.

The conclusions drawn from the Fenske results are that the Fenske calculation will give a fair estimate of the product distribution only if the correct key components are chosen. The best choices for key components are those components between which the separation is being made. The Fenske calculations may also be quite valuable for the determination of the minimum number of stages. The determination of minimum number of stages will at least eliminate a portion of the possible areas of tower operation.

CHAPTER VI

CONCLUSIONS AND RECOMMENDATIONS

The purposes of this investigation were to determine the feasibility of using the Bonner Plate to Plate Routine to provide data for the design non-refluxed distillation columns and to study the effect of variations of pressure and number of stages on the quality of separation. To this end a series of plate to plate calculations, at varying conditions, were performed using a single feed composition.

The data resulting from the Bonner Plate to Plate Routine was found to agree quite well with data on refluxed towers found in the literature. The routine assumed the presence of a partial condenser on the tower, but in view of the small amount of liquid formed in the condenser, this assumption was not deterimental to the value of the results. In all respects the plate to plate routine was found to be most satisfactory for providing data to be used in the design of non-refluxed towers.

An analysis of the plate to plate results indicated that an increase in the number of stages caused only a slight improvement in the quality of feed separation. A further analysis indicated that increases of pressure caused a slight but noticeable decrease in the quality of separation.

A technique for the determination of the tower parameters to be used with the plate to plate routine has been developed. It consists of the following:

(1) Perform a flash calculation on the feed separation using the pseudo equilibrium constants to approximate the desired split.

(2) Using the key component splits determined from the flash calculation in conjunction with the Fenske relationship calculate minimum number of stages necessary to accomplish the separation.

The use of such a technique prior to running the plate to plate routine will provide a better estimation of the tower conditions necessary to accomplish the separation. This in turn should cause a significant reduction in the calculation time.

Recommendations for future work are:

(1) An extension of the pseudo equilibrium curves from plate to plate data. This would entail a series of plate to plate calculations at varying tower conditions and fractional cuts.

(2) An investigation to determine the existence of a generalized correlation between the minimum number of stages and the actual number of theoretical stages required to accomplish a given separation.

NOMENCLATURE

D/F = Distillate rate expressed as a fraction of the total feed.

K' = Pseudo vapor liquid equilibrium constants.

L = Total moles of liquid.

S_m = Minimum number of stages.

V = Total moles of vapor.

V/F = Maximum vapor rate in a distillation column expressed as a fraction of the total feed.

X = Moles of any component in distillate or bottoms per 100 moles of feed.

α = Relative volatility of any component.

Subscripts

D = Distillate.

HK = Heavy key.

LK = Light key.

W = Bottoms.

A SELECTED BIBLIOGRAPHY

1. Benedict, Webb, Rubin and Friend. Chem. Engr. Prog. 47, December, 1951.
2. Bonner, J. S. "Plate to Plate Distillation Calculations on the 650 Computer." (unpublished memorandum of Humble Oil and Refining Company, distributed by International Business Machines, Inc.), 1956.
3. Bonner, J. S. "Solution of Multicomponent Distillation Problems on a Stored-Program Computer." (unpublished paper presented at 21st Midyear Meeting of American Petroleum Institute's Division of Refining), Montreal, Canada, 1956.
4. Fenske, M. R. Ind. Eng. Chem. 24, May, 1932.
5. Gilliland, E. R. Ind. Eng. Chem. 32, August, 1940.
6. Lewis, W. K. and Matheson, G. L. Ind. Eng. Chem. 24, May, 1932.
7. Maxwell, J. B. Data Book on Hydrocarbons. D. Van Nostrand Company, Inc. New York City, New York, 1950.
8. Perry, J. H. Ed. Chemical Engineers Handbook. 3rd Ed. McGraw-Hill Book Company, Inc. New York City, New York, 1950.
9. Robinson, C. S. and Gilliland, E. R. Elements of Fractional Distillation. 4th Ed. McGraw-Hill Book Company, Inc. New York City, New York, 1950.
10. Schierman, W. S. Private Communication with Black, Sivalls and Bryson, Inc.
11. Treybal, R. E. Mass Transfer Operations. McGraw-Hill Book Company, Inc. New York City, New York, 1955.
12. ______________ Hydrocarbon Enthalpy Charts. Western Supply Company.
13. ______________ Natural Gasoline Association of America "K" Value Charts. The Fluor Corporation, Ltd. Los Angeles, California, 1953.

14. ________________ Natural Gasoline Supply Men's Association Manual. 6th Ed. Natural Gasoline Supply Men's Association. Tulsa, Oklahoma, 1951.

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