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UNIVERSITY OF LOUISVILLE

A CORRELATION OF PLATE EFFICIENCIES
//
IN FRACTIONATING COLUMNS

A Thesis

Submitted to the Faculty
of the Graduate School
of the University of Louisville
in Partial Fulfillment
of the Requirements
for the Degree of

MASTER OF CHEMICAL ENGINEERING

Department of Chemical Engineering

Earl K. Stigger
///

March 1947

**A CORRELATION OF PLATE EFFICIENCIES
IN FRACTIONATING COLUMNS**

Earl K. Stigger

Approved by the Examining Committee

Director	G. C. Williams
	R. C. Ernst
	Guy Stevenson

March 1947

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ABSTRACT

A variety of binary systems were used in obtaining fractionation data in the same rectification column. The column was large enough to permit the use of a correlation for predicting plate efficiencies in commercial fractionators, and yielded results well within the limits of mechanical design. Established correlations were shown adequately to represent the results over a wide variation in both relative volatility and viscosity.

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INTRODUCTION

In the design of commercial fractionating columns, one of the major items is the number of plates or trays to be used in order to produce a given separation. Numerous investigators (1, 2, 3, 4) have presented various methods by which the required number of theoretical plates may be calculated. The theoretical plate concept assumes that equilibrium is attained between the liquid and vapor phases; but since, in practice, equilibrium is seldom, if ever, reached, this theoretical treatment yields results which may be as low as twenty percent of the actual number of plates required. The variance between the actual and the theoretical number of plates has been grouped in a coefficient called plate efficiency, and attributed to various mechanical characteristics of the equipment and the materials undergoing separation.

The mechanical features and their effect have been discussed in numerous articles (5, 6, 7, 8, 9, 10). The first significant treatment of the effect of the physical properties of the materials being separated was presented by Walter and Sherwood (11) who indicated that the efficiency of tower absorbers was a function of the viscosity. Later workers (12, 13) showed that viscosity also affected the plate efficiency of fractionating columns. O'Connell (14) expanded the work of other investigators to show a correlation of plate efficiency in fractionators as a

function of viscosity and the relative volatility of the key components. Systems of materials which seemed to disagree with a correlation of plate efficiency as a function of viscosity alone were brought into closer agreement when correlated on the basis of viscosity and the relative volatility of the key components. The work of O'Connell was based on data published by other investigators using different columns.

The purpose of the material presented in this paper is to evaluate and expand the correlation of plate efficiency as a function of viscosity and relative volatility. The data presented for all these systems were obtained on the same column, and the systems were chosen to cover definite ranges of the physical properties. Duplication of system types was avoided, where possible, in order to broaden the scope of the final correlation.

HISTORICAL

Although the practice of fractionation was developed centuries ago, for many years there was no real engineering design of equipment. The first worthwhile attempt to rationalize column design was presented by Serel (4) at the end of the nineteenth century. Serel worked from equilibrium relationships between liquid and vapor mixtures. If it were possible to obtain vapor-liquid equilibrium, Serel would have solved the question of column design from the standpoint of the number of plates required for a given separation. In practically every case, however, the attainment of equilibrium is impossible, and a field was opened for investigators to find a definite basis for the difference between the theoretical and the actual number of necessary plates.

The discrepancy between actual and theoretical plates, termed efficiency, required a mathematical description. This expression took three general forms: (1) point efficiency, (2) Murphree plate efficiency, and (3) overall efficiency.

Point efficiency is described as the efficiency at one point on any plate in a column, and is the ratio of the enrichment obtained to the enrichment under equilibrium conditions at the designated point.

Murphree plate efficiency is essentially the same as point efficiency except that the enrichment is based on the average composition of vapor entering and leaving a plate,

and the liquid leaving the plate.

Overall efficiency is the ratio of the theoretical number of plates to the actual number of plates required for a given separation.

For engineering calculations, overall efficiency values are the most consistent and in most cases the least difficult to obtain.

Early workers, noting that different columns gave different operating characteristics for similar materials, made an effort to relate the column efficiency to mechanical design aspects of the column.

Peavy and Baker (5) investigated the effect of vapor velocity on plate efficiency, and reported that plate efficiency increases with increasing velocity, becoming fairly constant over a considerable range at moderate velocities. At high velocities, efficiency falls off as a result of entrainment and liquid carry-over to the next plate. In order to determine the optimum velocity on the basis of maximum efficiency, a method was developed by Brown and Souders (6) and expanded by Brown and Lockhart (7).

Carey et al (8) determined the effect of slot submergence and slot velocity on plate efficiency. They concluded that the efficiency increased as slot submergence increased; and, for normal operating velocities, the efficiency was essentially independent of slot velocity.

Kirkbride (9) presented a method for calculating the downcomer area required to prevent column flooding and the deleterious effect of flooding on stable column operation is well known.

In order to insure stable plate operation, the liquid load must be maintained below the value which would produce a greater liquid head at the entrance to the plate than at the overflow from the plate. Allowable liquid loads are predictable according to the method discussed by Good et al (10).

O'Connell (14) has stated: "The diameter of the column does not affect plate efficiency for diameters less than five feet. However, for columns larger than seven feet, the length of the liquid path (Cross flow effect) must be taken into account."

The effect of mechanical design factors on plate efficiency has been confirmed in practice. Yet, when a column is designed for optimum efficiency, based on these mechanical factors, a wide range of efficiencies may be obtained when the column is operated with different systems. This variation suggested the possibility that physical properties of the process materials might affect the efficiency of the separation.

One of the first physical properties investigated with regard to its effect on plate efficiency was viscosity.

Walter and Sherwood (11) had shown that in the absorption of various materials the absorption efficiency increased with decreasing viscosity of the absorbant, while Keyes and Byman (12) showed evidence of viscosity affecting plate efficiency in laboratory distillations of ethanol-water mixtures.

If viscosity affects absorption efficiency, as has been shown by Walter and Sherwood, then it follows that there should be a similar effect in fractionation, since both operations involve diffusional processes. This conclusion is based on the postulation that the major portion of the enrichment during fractionation is a result of selective counter-diffusion of the key components in the froth above the plate.

Most of the work on viscosity involved the use of laboratory data and was thereby limited as to commercial application. Drickamer and Bradford (13) presented a definite correlation of plate efficiency as a function of viscosity (Figure 1) wherein the efficiency was shown to increase with decreasing viscosity. Although the data were obtained from commercial operations, the use of the relation is limited to systems of hydrocarbons having low relative volatility between the key components.

Gerster, Keffelt, and Withrow (15) indicated that the actual number of plates required for a given separation

FIG. 1 — EFFECT OF VISCOSITY ON PLATE EFFICIENCY
(CORRELATION OF DRICKAMER AND BRADFORD)

was a function of the slope of the equilibrium curve. From this, O'Connell (14) deduced that the limitations of the correlation of Drickamer and Bradford could be reduced by correlating plate efficiency as affected by viscosity and the relative volatility of the key components. The correlation (Figure 2) incorporated data for several non-hydrocarbon systems, and for systems of relatively high volatilities. Although the correlation was not specific, it was considered well within the limits of fractionation calculations in general and better commercial adaptation resulted. The correlation was not fitted with equations.

FIG. 2 - EFFECT OF RELATIVE VOLATILITY AND VISCOSITY ON PLATE EFFICIENCY
(CORRELATION OF O'CONNELL)

THEORETICAL

A. Effect of Viscosity on Plate Efficiency

The theoretical background for the effect of viscosity on plate efficiency was summarized by Drickamer and Bradford (13).

First, the Stokes-Einstein equation for diffusivity,

$$k = \frac{RT}{6\pi\eta Zr}$$

presents diffusivity, k , as an inverse function of viscosity, Z .

The equation for mass transfer developed by Chilton and Colburn (16) indicates that mass transfer under turbulent conditions, K , is an inverse function of viscosity, Z .

$$K = \frac{G_j}{M} \left(\frac{\rho k}{Z} \right)^{0.67}$$

Further evidence of note is the von Karmen (17) equation for film thickness:

$$s = \frac{(\text{constant}) (Z)}{G \left(\frac{r}{Z} \right)}$$

where the film thickness, s , is shown to be a direct function of the viscosity, Z .

Sherwood and Walter (11) developed an equation for liquid mass transfer,

$$K = \frac{3.4}{(Z)^{0.58}}$$

which indicated an inverse relationship between the liquid

mass transfer coefficient, K and viscosity, Z .

Although none of these equations indicates a definite mathematical relationship between viscosity and plate efficiency, they all show that viscosity enters into the value of numerous factors which do affect the degree of separation attainable through fractionation.

B. Effect of Relative Volatility on Plate Efficiency

In the work of Gerster, Koffelt, and Withrow (5) it was pointed out that the slope of the equilibrium curve had a direct bearing on the actual number of plates required per transfer unit. If the number of transfer units is equal to, or has a direct relation to, the number of theoretical plates required, then the plate efficiency may also be considered a function of the slope of the equilibrium curve. If the relative volatility is a function of the slope of the equilibrium curve, or, more precisely, a mathematical representation of the slope, then it follows that there should be a connection between plate efficiency and relative volatility.

Although no theoretical explanation for the relationship is available, O'Connell (14) presents the following in support of such a conclusion.

"The effect of the relative volatility was shown by the results of a test on an ethylene dichloride stabilizer. In this column, the feed contained

both water and ethyl chloride which had to be reduced to low concentrations in the ethylene dichloride bottoms. The plate efficiency based on water as the low boiling key component (ethylene dichloride the high boiling key) was 29% and the relative volatility between water and ethylene dichloride was 16. The plate efficiency based on ethyl chloride as the low boiling key component was 57%, and the relative volatility between ethyl chloride and ethylene dichloride was 3.1."

The presentation of plate efficiency as a function of the product of the viscosity and the relative volatility has little theoretical background. A correlation resulted, however, which indicated that the efficiency increased as the viscosity-volatility product decreased. (See Figure 2, Page 11.)

C. Theoretical Plate Calculations and Equilibrium Data

The theoretical number of plates corresponding to a given separation may be calculated by any of numerous methods reported in the literature. For the purpose of the correlation in question, it was considered that the graphical McCabe-Thiele method would suffice. This selection was based on simplicity and general acceptance. For binary systems, the McCabe-Thiele method is simple, and the results are within the normal limits of engineering calculations.

Vapor-liquid equilibrium data were calculated by the method of Chilton and Colburn (16). In the equilibrium

data, Tables I, II, III, IV, and V,

x_A is the mol fraction of the more volatile component
in the liquid

y_A is the mol fraction of the more volatile component
in the vapor

t_b is the boiling temperature, degrees Centigrade

α is the relative volatility, $\frac{\frac{y_A}{1 - y_A}}{\frac{x_A}{1 - x_A}}$

D. Viscosity Data

Viscosity values were obtained from standard references and the values used in the correlation are the 'psuedo molal viscosities' of Drickamer and Bradford (13). The psuede viscosity assumes that viscosities of mixtures may be determined by adding the viscosities of the components at the given temperature in proportion to the mol fraction of each present:

$$\mu = x_A \mu_A + x_B \mu_B$$

where, μ is the viscosity in centipoise

A denotes component A

B denotes component B

x is mol fraction

Table I: Vapor-Liquid Equilibria for Ethanol-Water

x_A	y_A	t_b	α
0.00	0.000	100.0	----
0.005	0.052	98.6	10.9
0.010	0.100	97.3	11.0
0.015	0.145	96.1	11.0
0.020	0.182	95.0	10.9
0.030	0.250	93.2	10.8
0.05	0.337	90.4	9.65
0.07	0.389	88.5	8.45
0.10	0.443	86.4	7.15
0.15	0.497	84.4	5.60
0.20	0.530	83.2	4.50
0.30	0.574	81.7	3.14
0.40	0.613	80.7	2.37
0.50	0.653	80.0	1.880
0.60	0.699	79.4	1.548
0.70	0.752	78.9	1.300
0.80	0.817	78.4	1.117
0.95	0.946	78.2	0.940
1.00	1.000	78.3	-----

Calculated at 760 mm total pressure by the method of
Chilton and Colburn (16)

Table II: Vapor-Liquid Equilibria for Methanol-Water

x_A	y_A	t_b	α
0.00	0.000	100.0	----
0.05	0.277	92.6	7.27
0.10	0.420	87.7	6.52
0.20	0.576	82.0	5.45
0.30	0.665	78.5	4.64
0.40	0.729	75.8	4.04
0.50	0.781	73.7	3.57
0.60	0.831	71.6	3.29
0.70	0.876	69.7	3.02
0.80	0.918	68.0	2.79
0.90	0.959	66.4	2.57
0.95	0.979	65.6	2.50
1.00	1.000	64.6	----

Calculated at 760 mm total pressure by the method of
Chilton and Colburn (16)

Table III: Vapor-Liquid Equilibria for Benzene-Toluene

x_A	y_A	t_b	α
0.00	0.000	110.7	2.315
0.05	0.113	108.5	2.330
0.10	0.207	106.3	2.345
0.20	0.373	102.2	2.374
0.30	0.507	98.6	2.404
0.40	0.618	95.5	2.433
0.50	0.712	92.5	2.463
0.60	0.790	89.8	2.492
0.70	0.855	87.2	2.522
0.80	0.912	84.6	2.551
0.90	0.959	82.3	2.581
0.95	0.981	81.2	2.595
1.00	1.000	80.1	2.610

Calculated at 760 mm total pressure by the method of
Chilton and Colburn (16)

Table IV: Vapor-Liquid Equilibria for
Carbon Tetrachloride-Benzene

x_A	y_A	t_b	α
0.00	0.000	80.2	1.260
0.05	0.0615	79.8	1.245
0.10	0.120	79.4	1.229
0.20	0.230	78.7	1.195
0.30	0.333	78.1	1.164
0.40	0.432	77.7	1.139
0.50	0.526	77.3	1.113
0.60	0.620	77.0	1.090
0.70	0.714	76.7	1.068
0.80	0.806	76.6	1.043
0.90	0.902	76.5	1.025
0.95	0.951	76.4	1.015
1.00	1.000	76.4	1.007

Calculated at 760 mm total pressure by the method of
Chilton and Colburn (16)

Table V: Vapor-Liquid Equilibria for Acetaldehyde-Water

x_A	y_A	t_b	α
0.00	0.000	100.0	54.6
0.02	0.589	77.4	70.5
0.05	0.816	59.4	85.0
0.10	0.914	45.0	96.2
0.20	0.956	33.6	88.0
0.30	0.968	29.0	71.2
0.40	0.974	26.7	55.2
0.50	0.976	25.7	40.5
0.60	0.977	25.0	28.8
0.70	0.980	24.2	20.7
0.80	0.983	23.2	14.6
0.90	0.989	21.8	10.5
0.95	0.994	20.9	8.74
1.00	1.000	19.8	7.43

Calculated at 760 mm total pressure by the method of
Chilton and Colburn (16)

EXPERIMENTAL

EQUIPMENT

All fractionations were accomplished in an experimental laboratory column having the following specifications.

Number of plates	20 (plus reboiler)
Diameter	10 inches
Plate spacing	8 inches
Vapor risers	1-5/8 inch dia. (4 per plate)
Downcomers	1/2 inch dia. (2 per plate)
Weirs	6 in. X 1 in. (2 per plate)
Bubble caps	2-1/2 in. dia. X 1-3/4 in. high (4 per plate)
Slots	5/8 in. X 1/8 in. (31 per cap)
Calandria (reboiler)	12 in. dia. X 12 in. long (shell) 24 1 in. tubes; with auxiliary open steam connection
Feed arrangement	Liquid feed inlets on plates 1, 3, 5, ----- 15, 17 (from top) Vapor feed inlets on plates 2, 4, 6, ----- 18, 20 (from top) Liquid feed delivered by gear pump through a rotameter
Feed pot	60 gallon capacity

Overheads	Two condensers (tube and shell) Splitter box arranged to deliver any desired portion of overheads as reflux to top plate Product and reflux flow by gravity through rotameters
Sampling	Sample cocks arranged to produce samples of the liquid off plates 1, 3, 5, --- 17, 19 Samples of feed, waste, and overheads easily obtainable
Temperatures	Plate temperatures obtained by copper-constantan thermocouples in the liquid on plates 1 thru 20

The column was designed for a maximum feed rate of twenty gallons per hour, and for satisfactory operation on the systems ethanol-water, benzene-toluene, and acetic acid-water at vapor velocities between 0.75 and 1.25 feet per second. (A sketch of the column and details of plate construction are shown in Figures 14 and 15, Appendix.)

SYSTEMS INVESTIGATED

The first system studied was ethanol-water, chosen because the operation of the column with this system had

been studied and proved dependable. The system yielded viscosity-velocity products in the central portion of the range to be investigated; and it afforded excellent opportunities to become familiar with the operating variables of the column. Added features with respect to the correlation were the inclusion of a system producing a constant boiling mixture, and the ease in analysis. The samples were analyzed by Westphal Balance density measurements and density-composition data for ethanol-water mixtures in standard references. The precision for the analyses was 0.2 mol per cent.

The system methanol-water was chosen in order to expand the correlation in the central portion of the viscosity-velocity product range. The viscosity-velocity product for methanol-water was essentially the same as that for ethanol-water; however, the product was obtained by lower viscosity and higher relative volatility. Again, the ease in analysis was a contributing factor. Methanol-water mixtures were analyzed by Westphal Balance density measurements in conjunction with density-composition data in standard references. The precision was 0.2 mol per cent.

Benzene-toluene was chosen as the third system to be studied for one major reason - - this system, widely reported in the literature, offered a means of comparing

the operating characteristics of the equipment to the characteristics of equipment used by other investigators. The system, in addition, extended the range in viscosity-volatility product beyond that covered by the two previous systems. Difficulty in analysis was overcome by the following procedure: For those samples low in toluene content, freezing point determinations led to analyses with a precision of 1.0 mol per cent. Satisfactory results for the low benzene samples were obtained by vapor pressure measurements as outlined by Matelson and Zuckermann (18). Through careful use of this method, a precision of 0.5 mol per cent was obtained. The precision of either method was unsatisfactory for samples of liquid in the central portion of the column, and, rather than include results of doubtful validity, column efficiencies only were reported for benzene-toluene fractionations.

The system carbon tetrachloride-benzene offered the inclusion of a relative volatility which was very low, filling in the lower portion of the viscosity-volatility product range. The added feature of significance was the fact that the material of higher density was being driven toward the top of the column. Westphal Balance density measurements were sufficient to yield analyses with a precision of 0.2 mol per cent.

Acetaldehyde-water, the final system studied

was chosen for its high relative volatility to complete the desired range in viscosity-volatility product. It also added one more type of material to the correlation. Analyses were by Westphal Balance density measurements. Since the density-composition curve contains a maximum value, it was necessary, in cases of ambiguity, to use the plate temperature in conjunction with the density as measured. Precision for these analyses was excellent in the high aldehyde samples, but was of the order of 2 to 3 mol per cent in the low aldehyde range. Little data were obtained for this system, since decomposition, sufficient to produce an immiscible layer, occurred after three or four operations.

OPERATING PROCEDURE

Material from the feed pot was pumped to the top portion of the column and allowed to flow down the column. When the liquid level in the reboiler was sufficient, steam was admitted to the calandria and feed was discontinued. Partial distribution of the components was attained as vapors rose in the column, heating the entire column to operating temperature. All of the distillate was returned to the column as liquid reflux and the column was allowed to approach equilibrium operating conditions. At this time, distribution in the column was adjusted by the addition of

feed and the removal of either distillate or waste. Plate compositions were estimated by the liquid temperature on each plate. When the desired distribution was reached, the column was operated at total reflux for sufficient time to establish equilibrium operating conditions. The approach to constant conditions was followed by observing plate temperatures at ten to fifteen minute intervals. When operation was essentially constant, liquid samples were withdrawn from alternate plates (1, 3, 5, --- 17, 19) and from the distillate and reboiler. Plate temperatures were recorded as well as other data required for calculations. A sample data sheet for total reflux operation is presented in Figure 3.

Continuous feed was introduced at a metered rate onto the plate most nearly containing the same mixture as the feed mixture. The distillate was split into reflux and product at a product rate representing 70 to 95 per cent of the low boiling component entering in the feed. The bottoms draw-off was adjusted to maintain a constant liquid level in the reboiler. Continuous feed, product, and bottoms rates were carefully controlled for a period sufficient to produce constant operation as indicated by constant plate temperatures. Samples and data were obtained as during total reflux, with the addition of a feed sample and feed and product flow rates. The

bottoms rate was not metered; but it was assumed that maintaining constant feed and product rates and a constant liquid level in the reboiler would cause the bottoms rate to be equal to the difference between the feed rate and the product rate. This adjustment required that column holdup be essentially constant - - which should occur under normal operating conditions. Typical data sheets for continuous operations are presented in Figures 4, 5, 6, and 7.

Two methods were used to effect a change in reflux ratio:

- (1) Feed and product rates were held constant while changing the heat input to the reboiler.
- (2) Heat input to the reboiler was held constant while changing feed and/or product rates.

Samples for total reflux and continuous operations were analyzed by the methods previously described.

DATA AND RESULTS

Typical data sheets for operations are presented in Figures 3, 4, 5, 6, and 7. Figure 3 is the data obtained during one operation at total reflux using the system ethanol-water. The other data presented, Figure 4, 5, 6, and 7, represent one continuous operation for each of the systems methanol-water, benzene-toluene, carbon tetrachloride-benzene, and acetaldehyde-water respectively. Complete data are contained in original research notes on file with the Institute of Industrial Research, University of Louisville.

Results are presented in tabular form for each system investigated (Tables VI, VII, VIII, IX, and X).

An explanation of the tables follows:

Run number

The number assigned to any particular operation in the original notes

Number of Plates

The number of actual plates in the section of the column considered over which the data were of sufficient accuracy to use in calculations - - not necessarily the number of plates between feed and product or feed and waste (this approach eliminates including a mis-matched feed section in calculations)

Vapor Velocity

The superficial vapor velocity in the column, based on overhead measurements and corrected for temperature

**System: Ethanol-Water
(Total Reflux)**

Run No. 7

Date: 7 June, 1946

Feed to plate 15 @ 10:00 A.M.

Steam to calandria @ 10:15 A.M.

Reflux over @ 11:00 A.M.

Temperatures indicate too little ethanol in column

Added 1 gallon ethanol to reflux @ 11:15 A.M.

Samples and temperatures @ 12:10 P.M.

Sample	Sp. Gr.	Mol % ethanol	Plate	Temp. °C.
Reflux	0.8106	77.4	1	78
Plate 1	0.8127	75.9	2	78
" 3	0.8188	71.7	3	78
" 5	0.8256	67.5	4	78
" 7	0.8415	58.0	5	78.5
" 9	0.8981	31.5	6	79
" 11	0.9896	1.8	7	79.5
" 13	0.9962	0.3	8	80.5
" 15	0.9965	---	9	83
" 17	0.9965	---	10	90
" 19	0.9965	---	11	96
Bottoms	0.9965	---	12	98
			13	99.5
			14	100
			15	100
			16	100
			17	100
Steam pressure: 8 psi ga.			18	100
			19	100
			20	100

Reflux rotameter: 55 mm @ 35°C. Rate: 120 lbs./hr.

Figure 3: Typical Data Sheet for Ethanol-Water

**System: Methanol-Water
(Continuous Operation)**

Run No. 11

Date: 18 June, 1946

Column brought to equilibrium in run no. 10

Feed to plate 13 @ 11:10 A.M.

Product and bottoms over @ 11:15 A.M.

Samples and temperatures @ 12:15 P.M.

Sample	Sp. Gr.	Mol % methanol	Plate	Temp. °C.
Product	0.7970	94.8	1	66
Reflux	0.7970	94.8	2	67
Plate 1	0.7968	94.8	3	68
" 3	0.8127	85.3	4	69
" 5	0.8236	80.1	5	71
" 7	0.8432	70.7	6	73
" 9	0.8803	52.6	7	74
" 11	0.9302	28.6	8	74
" 13	0.9551	16.5	9	78
" 15	0.9730	8.5	10	81
" 17	0.9878	2.7	11	83
" 19	0.9942	0.6	12	84
Bottoms	0.9962	0.1	13	85
Feed	0.9503	19.0	14	88
			15	92
			16	92
			17	97
Steam pressure: 12.5 psi ga.			18	98
			19	99
			20	99

Feed rotameter: 200 mm @ 60°C. Rate: 115 lbs./hr.

Product rotameter: 45 mm @ 35°C. Rate: 39 lbs./hr.

Reflux rotameter: 28 mm @ 35°C. Rate: 66 lbs./hr.

Figure 4: Typical Data Sheet for Methanol-Water

**System: Benzene-Toluene
(Continuous Operation)**

Run No. 24

Date: 11 Oct., 1946

Column brought to equilibrium in run no. 23

Feed to plate 13 @ 12:15 P.M.

Product and bottoms over @ 12:20 P.M.

Samples and temperatures @ 1:30 P.M.

Sample	Method of Analysis	Mol % benzene	Plate	Temp. °C.
Product	Freezing point	99	1	80
Reflux	" "	99	2	80
Plate 1	" "	99	3	81
" 3	" "	99	4	82
" 5	" "	96	5	82
" 7	" "	94	6	82
" 9	" "	88	7	82
" 11	Vapor Pressure	65	8	82
" 13	" "	47	9	83
" 15	" "	39	10	84
" 17	" "	28	11	85
" 19	" "	20	12	87
Bottoms	" "	9	13	90
Feed	" "	32	14	91
			15	94
			16	94
			17	97
Steam pressure: 22 psi ga.			18	98
			19	101
			20	103

Feed rotameter: 200 mm @ 65°C. Rate: 100 lbs./hr.

Product rotameter: 42 mm @ 35°C. Rate: 33 lbs./hr.

Reflux rotameter: 120 mm @ 35°C. Rate: 163 lbs./hr.

Figure 5: Typical Data Sheet for Benzene-Toluene

**System: Carbon Tetrachloride-Benzene
(Continuous Operation)**

Run No. 33

Date: 21 Nov. 1946

Column brought to equilibrium in run no. 32

Feed to plate 5 @ 10:30 A.M.

Product and bottoms over @ 10:35 A.M.

Samples and temperatures @ 11:30 A.M.

Sample	Sp. Gr.	Mol % CCl ₄	Plate	Temp. °C.
Product	1.3100	59.5	1	77
Reflux	1.3100	59.5	3	77
Plate 1	1.3008	58	5	78
" 3	1.2849	55.5	7	78
" 5	1.2643	52	9	78
" 7	1.2571	50.5	11	79
" 9	1.2466	49.5	13	79
" 11	1.2343	47.5	15	79
" 13	1.2208	45.5	17	80
" 15	1.2060	43.0	19	80
" 17	1.1880	41.5		
" 19	1.1694	39		
Bottoms	1.1544	37.5		
Feed	1.2465	49		

Steam pressure: 4 psi ga.

Feed rotameter: 175 mm @ 40°C. Rate: 114 lbs./hr.

Product rotameter: 40 mm @ 40°C. Rate: 72 lbs./hr.

Reflux rotameter: 215 mm @ 40°C. Rate: 460 lbs./hr.

Figure 6: Typical Data Sheet for Carbon Tetrachloride-Benzene

**System: Acetaldehyde-Water
(Continuous Operation)**

Run No. 47

Date: 17 Jan. 1947

Column brought to equilibrium in run no. 46

Feed to plate 9 @ 11:50 A.M.

Product and bottoms over @ 12:00 M.

Samples and temperatures @ 12:40 P.M.

Sample	Sp. Gr.	Mol % aldehyde	Plate	Temp. °C.
Product	0.7960	99.5	1	21
Reflux	0.7960	99.5	2	21
Plate 1	0.7985	98.2	3	21
" 3	0.8009	97.5	4	21
" 5	0.8040	96.0	5	21
" 7	0.8087	94.3	6	21
" 9	0.9896	43.5	7	22
" 11	1.0116	34.8	8	24
" 13	1.0012	0.6	9	26
" 15	1.0020	0.8	10	28
" 17	1.0013	0.6	11	33
" 19	1.0015	0.6	12	37
Bottoms	1.0028	1.2	13	75
Feed	1.0212	29.3	14	94
			15	100
			16	100
			17	100
Steam pressure: 4 psi ga.			18	100
			19	100
			20	100

Feed rotameter: 175 mm @ 20°C. Rate: 105 lbs./hr.

Product rotameter: 50 mm @ 20°C. Rate: 42 lbs./hr.

Reflux rotameter: 30 mm @ 20°C. Rate: 63 lbs./hr.

Figure 7: Typical Data Sheet for Acetaldehyde-Water

Table VI: Results for Ethanol-Water Fractionations

Run No.	Section	No. of Plates	Vapor Velocity	L/V	Xm	Xn	Theo. Plates	Overall Eff. %	Avg. t Deg. C.	μ	α	$\mu\alpha$
1	Column	8	0.60	1.00	.013	.693	3.63	45.4	84	.60	2.4	1.44
2	Column	8	1.04	1.00	.011	.676	3.50	43.8	86	.60	3.0	1.80
3	Column	4	0.58	1.00	.022	.498	1.90	47.6	90	.60	3.1	1.86
6	Column	8	0.85	1.00	.008	.700	3.88	48.6	87	.60	3.0	1.80
7	Column	10	0.71	1.00	.018	.759	4.72	47.2	83	.59	1.9	1.11
8	Column	10	1.11	1.00	.002	.740	5.25	52.5	86	.60	2.6	1.56
9	Rect.	6	1.09	0.91	.046	.675	3.05	50.8	84	.60	2.5	1.50

Table VII: Results for Methanol-Water Fractionations

Run No.	Section	No. of Plates	Vapor Velocity	L/V	Xm	Xn	Theo. Plates	Overall Eff. %	Avg. t Deg. C.	μ	α	$\mu\alpha$
10	Column	8	1.07	1.00	.075	.955	4.20	52.5	73	.37	3.2	1.18
11	Column	18	0.76	0.63	.006	.948	9.15	50.9	81	.35	3.8	1.33
	Rect.	10	----	----	.286	.948	4.67	46.7	73	.36	3.1	1.12
	Strip.	6	----	----	.006	.165	3.52	58.6	93	.29	6.9	2.00
12	Column	14	0.73	0.81	.009	.916	5.90	42.1	84	.34	4.2	1.43
	Rect.	6	----	----	.350	.916	2.95	49.2	73	.36	3.1	1.12
	Strip.	2	----	-----	.009	.053	1.16	57.8	96	.30	7.7	2.31
14	Column	12	1.15	0.82	.006	.885	5.61	46.8	84	.35	4.5	1.58
	Rect.	6	----	----	.163	.885	3.22	53.6	75	.37	3.4	1.26
	Strip.	2	----	----	.006	.035	1.15	57.8	97	.30	8.0	2.40
15	Rect.	4	1.16	0.87	.198	.783	2.09	52.2	68	.39	3.6	1.40
17	Column	10	0.63	0.85	.005	.875	5.23	52.3	82	.29	4.2	1.22
	Rect.	6	----	----	.143	.875	2.97	49.4	75	.36	3.5	1.26
	Strip.	4	----	----	.005	.143	2.26	56.5	95	.30	7.6	2.28

Table VIII: Results for Benzene-Toluene Fractionations

Run No.	Section	No. of Plates	Vapor Velocity	L/V	Xm	Xn	Theo. Plates	Overall Eff. %	Avg. t Deg. C.	μ	α	$\mu\alpha$
18(a)	Column	18	0.50	1.00	.005	.98	10.5	58	91	.29	2.46	.715
21	Column	18	0.21	1.00	.005	.98	10.5	58	92	.29	2.46	.715
22	Column	18	0.70	0.78	.07	.95	10.6	59	90	.29	2.46	.715
23	Column	14	0.49	1.00	.09	.99	7.8	56	90	.28	2.49	.695
24	Column	14	0.60	0.83	.20	.96	7.0	50	89	.28	2.49	.695
25	Column	14	0.56	1.00	.12	.98	6.5	46	90	.28	2.49	.695
26	Column	14	0.48	0.86	.11	.94	7.0	50	92	.29	2.46	.715

Table IX: Results for Carbon Tetrachloride-Benzene Fractionations

Run No.	Section	No. of Plates	Vapor Velocity	L/V	Xm	Xn	Theo. Plates	Overall Eff. %	Avg. t Deg. C.	μ	α	$\mu\alpha$
27	Column	18	0.91	1.00	.42	.645	9.5	53	78	.39	1.10	0.43
29	Column	18	0.98	0.84	.33	.52	11.5	64	78	.36	1.13	0.40
	Rect.	8	----	----	.43	.52	5.2	65	77	.37	1.12	0.42
	Strip.	8	----	----	.33	.42	5.0	62	79	.34	1.14	0.39
31	Column	18	0.98	1.00	.43	.66	10.1	56	78	.39	1.10	0.43
32	Column	18	0.83	1.00	.39	.64	10.6	59	78	.39	1.10	0.43
33	Column	18	1.00	0.86	.38	.58	10.8	60	78	.38	1.10	0.42
	Rect.	6	----	----	.49	.58	3.5	59	77	.39	1.10	0.43
	Strip.	8	----	----	.39	.48	4.0	61	79	.36	1.13	0.41
34	Column	18	1.02	0.93	.31	.55	9.4	52	78	.36	1.13	0.41
	Rect.	4	----	----	.49	.55	2.6	66	77	.38	1.10	0.42
	Strip.	12	----	----	.31	.47	5.7	48	79	.35	1.14	0.40
35	Column	18	1.02	0.92	.30	.54	9.0	50	78	.36	1.13	0.41
	Rect.	14	----	----	.36	.54	7.1	51	77	.37	1.13	0.42
	Strip.	4	----	----	.30	.36	1.9	47	79	.34	1.17	0.40
37	Column	18	0.95	1.00	.48	.68	9.8	55	78	.40	1.08	0.43
39	Column	18	0.99	0.86	.32	.52	10.1	56	78	.36	1.13	0.41
	Rect.	8	----	----	.42	.52	5.1	64	77	.37	1.12	0.41
	Strip.	8	----	----	.32	.40	4.0	50	79	.35	1.15	0.40
40	Column	18	1.09	0.93	.28	.52	9.0	50	78	.36	1.14	0.41
	Rect.	8	----	----	.41	.52	4.3	54	77	.37	1.12	0.41
41	Column	18	0.74	1.00	.42	.64	9.9	55	78	.39	1.11	0.43
42	Column	18	0.77	0.86	.42	.58	12.0	67	78	.38	1.11	0.42
	Rect.	8	----	----	.49	.58	5.5	69	77	.39	1.10	0.43
	Strip.	4	----	----	.42	.46	2.7	67	79	.36	1.13	0.41

Table X: Results for Acetaldehyde-Water Fractionations

Run No.	Section	No. of Plates	Vapor Velocity	L/V	Xm	Xn	Theo. Plates	Overall Eff. %	Avg. t Deg. C.	μ	α	$\mu\alpha$
45	Strip.	4	----	----	.005	.305	1.9	48	72	.36	97	35.0
46	Column	8	0.48	1.00	.006	.980	2.63	33	32	.36	19	6.8
47	Column	8	0.41	0.60	.006	.960	2.87	35	32	.46	35	16.0
	Rect.	4	----	----	.435	.960	1.00	25	27	.34	15.5	5.3
	Strip.	4	----	----	.006	.435	1.67	42	40	.52	77	40.0
48	Column	4	0.46	1.00	.521	.985	1.50	38	22	.37	18	6.8

L/V

The reflux ratio calculated from overhead measurements (sample calculation in Appendix)

X_m

Mol fraction of the low boiling component in the liquid off the bottom plate of the section under consideration

X_n

Mol fraction of the low boiling component in the liquid into the top of the section under consideration

Theoretical Plates

Number of theoretical (equilibrium) plates required to enrich from X_m to X_n (sample calculation in Appendix)

Overall Efficiency Per Cent

The ratio of theoretical to actual number of plates required to enrich from X_m to X_n (sample calculation in Appendix)

Average Temperature Degrees C.

The average temperature of the liquid on the plates being considered

μ

The psuede molal viscosity of the liquid in the section of the column under consideration at the average temperature and composition (sample calculation in Appendix)

α

The relative volatility at the average composition of the liquid in the section

 $\mu\alpha$

The product of the viscosity and the relative volatility (viscosity-volatility product)

Complete methods used for obtaining calculated results are presented in the Appendix. The reported efficiencies are considered a representation of the best data obtained in any given section of the column, disregarding all data apparently in error due to operational or analytical discrepancies.

The overall efficiencies are presented as a function of the viscosity-volatility product in Figures 8, 9, 10, 11, and 12. The correlation proposed by O'Connell (14) is included with the data obtained since the original problem was an evaluation of this correlation. Figure 8 presents efficiencies obtained for all systems during total reflux operation. Figures 9, 10, and 11 present column, rectifier, and stripper efficiencies, respectively, for all systems during continuous operation. In Table XII, the section and column efficiencies were averaged for each system in order to minimize the effect of errors due to individual operational discrepancies. The average results are plotted in Figure 12.

FIG. 8 - EFFECT OF RELATIVE VOLATILITY AND
VISCOSITY ON OVERALL EFFICIENCY
(TOTAL REFLUX OPERATIONS)

FIG. 9 - EFFECT OF RELATIVE VOLATILITY AND
VISCOSITY ON OVERALL EFFICIENCY
(CONTINUOUS OPERATION - COLUMN)

FIG. 10 - EFFECT OF RELATIVE VOLATILITY AND
VISCOSITY ON OVERALL EFFICIENCY
(CONTINUOUS OPERATION - RECTIFIER)

FIG. 11 - EFFECT OF RELATIVE VOLATILITY AND VISCOSITY ON OVERALL EFFICIENCY
(CONTINUOUS OPERATION - STRIPPER)

Table XI: Average Results by Systems

(Data for Figure 12)

System	Total Reflux Column		-----Continuous-----					
	Eff.%	$\mu\alpha$	Column	Eff.%	$\mu\alpha$	Rectifier	Eff.%	Stripper $\mu\alpha$
Ethanol-Water	47	1.60	----	----	51	1.50	----	----
Methanol-Water	52	1.18	48	1.39	50	1.23	58	2.25
Benzene-Toluene	54	0.70	53	0.71	--	----	--	----
Carbon tet.-Benzene	56	0.43	57	0.41	61	0.42	54	0.40
Acetaldehyde-Water	36	6.45	35	16.0	25	5.3	42	40.0

FIG. 12 - EFFECT OF RELATIVE VOLATILITY AND VISCOSITY ON OVERALL EFFICIENCY
(AVERAGE RESULTS BY SYSTEMS)

DISCUSSION OF RESULTS

For the purpose of discussion, deviations from the proposed correlation have been calculated and presented in Tables XII, XIII, XIV, and XV.

Overall efficiencies based on total reflux operations are in favorable agreement with the correlation proposed by O'Connell (Figure 8). The average deviation from the correlation is 0.5 % absolute, or an error of 3 %. The maximum deviation was 6 % absolute, or an error of 19 %, obtained from data for acetaldehyde-water fractionations. This maximum deviation occurred under conditions whereby small errors in analyses or calculations are exaggerated in the results. This effect enters through the extreme character of the equilibrium existing between liquid and vapor (relative volatility is 10 to 90). Consistent deviations of from 8 to 10 % were obtained using carbon tetrachloride-benzene. The relative volatility for carbon tetrachloride-benzene, 1.10 to 1.15, indicates that the equilibrium curve is very close to the x equal y line in the McCabe-Thiele diagram, decreasing the accuracy of theoretical plate calculations.

The positive nature of the deviation was expected since the efficiency increases with increasing reflux ratio as pointed out for methanol-water fractionations (Figure 13).

Overall column efficiencies obtained during continuous operations (Figure 9) show excellent agreement

Table XII: Deviation of Results for Total Reflux Operations

System	Absolute Deviation % Efficiency	Relative Deviation %
Ethanol-Water	0	0
	1	2
	2	5
	6	14
	7	16
	8	18
Methanol-Water	6	13
Benzene-Toluene	2	4
	2	4
	4	8
	-7	-13
Carbon tetrachloride-Benzene	-2	-3
	-5	-8
	-6	-10
	-6	-10
	-8	-13
Acetaldehyde-Water	1	3
	6	19
AVERAGE	0.5	3

Table XIII: Deviation of Column Efficiencies for
Continuous Operation

System	Absolute Deviation % Efficiency	Relative Deviation %
Ethanol-Water	---	---
Methanol-Water	-1	-2
	4	9
	6	13
	6	13
Benzene-Toluene	-3	-6
	-3	-6
	6	11
Carbon tetrachloride-Benzene	-11	-18
	-10	-16
	-6	-10
	-2	-3
	2	3
	5	8
Acetaldehyde-Water	7	25
AVERAGE	0	2

Table XIV: Deviation of Rectifier Efficiencies for
Continuous Operation

System	Absolute Deviation % Efficiency	Relative Deviation %
Ethanol-Water	7	16
Methanol-Water	0	0
	2	4
	3	7
	7	15
	7	15
Carbon tetrachloride-Benzene	-11	-18
	-9	-14
	-3	-5
	2	3
	3	5
	3	5
	7	11
Acetaldehyde-Water	2	6
AVERAGE	2	4

Table XV: Deviation of Stripper Efficiencies for
Continuous Operation

System	Absolute Deviation % Efficiency	Relative Deviation %
Methanol-Water	17	44
	18	44
	19	49
	19	49
Carbon tetrachloride-Benzene	5	8
	0	0
	0	0
	-13	-21
	-15	-24
	-16	-25
AVERAGE	3	12

FIG.13 - EFFECT OF REFLUX RATIO
ON OVERALL EFFICIENCY OF
RECTIFYING SECTION
SYSTEM: METHANOL-WATER

with the proposed correlation. The average deviation was less than 1 % absolute, or an error of 2 %. The maximum deviation was 7 % absolute, or an error of 25 %, obtained during acetaldehyde-water fractionations. Again, the analyses and calculations for this system were the least consistent. Excluding the deviation for acetaldehyde-water, the maximum deviation indicated an error of 18 %. Deviations up to 15 % were not considered excessive since the operations included various reflux ratios, and the effect of varying reflux ratio on plate efficiency has been presented (Figure 13).

Results obtained for overall rectifier efficiencies (Figure 10) showed very favorable agreement with the proposed correlation. The absolute average deviation for this case was 2 %, or an error of 4 %, while the maximum absolute deviation of 11 %, or 18 % error, occurred during rectification of carbon tetrachloride-benzene. The relative volatility during the operation leading to the 18 % deviation was 1.13. If the number of theoretical plates were recalculated on the basis of a relative volatility of 1.10 (3 % decrease), the deviation would then be 4 %. In view of this, the deviations for carbon tetrachloride-benzene seem unimportant.

Stripper operations led to efficiencies which deviated from the correlation for central values of the

viscosity-volatility product, but were in fair agreement for low values of the viscosity-volatility product.

(See Figure 11.) In the low range, the absolute deviation was 6 %, or an error of 10 %, while in the central range of viscosity-volatility product the average absolute deviation was 18 %, or an error of 46 %. This high deviation was obtained during methanol-water fractionation where the analytical procedure and the calculations were most explicit.

The average deviation of the results by systems (Figure 12) were computed and are presented in Table XVI. These results show excellent agreement with the proposed correlation with the exception of stripper operations at viscosity-volatility products greater than 1.0.

Table XVI: Deviation of the Average Results from the
Proposed Correlation

System	Operation Involved	Section of Column	Relative Deviation %
Ethanol- Water	Total Reflux	Column	9
" "	Continuous	Column	---
" "	"	Rectifier	16
" "	"	Stripper	---
Methanol- Water	Total Reflux	Column	13
" "	Continuous	Column	8
" "	"	Rectifier	8
" "	"	Stripper	46
Benzene- Toluene	Total Reflux	Column	2
" "	Continuous	Column	0
Carbon tetrachloride- Benzene	Total Reflux	Column	8
" "	Continuous	Column	6
" "	"	Rectifier	0
" "	"	Stripper	10
Acetaldehyde- Water	Total Reflux	Column	12
" "	Continuous	Column	25
" "	"	Rectifier	4

SUMMARY AND CONCLUSIONS

Overall efficiencies for fractionations involving five different binary systems have been presented. The systems included alcohol-water mixtures, aromatic hydrocarbon-chlorinated aliphatic hydrocarbon, two aromatic hydrocarbons, and aldehyde-water. Rectifier, stripper, and column overall efficiencies during total reflux and continuous operation were correlated on the basis of the viscosity-volatility product, calculated at the average temperature and composition in the respective portion of the column involved.

The correlation, although semi-empirical, may be used to predict overall efficiencies in commercial fractionators under certain limitations. For viscosity-volatility products between the values 0.1 and 10, overall column and rectifier section efficiencies may be predicted within the limits of mechanical design variables. Stripper efficiencies for materials yielding viscosity-volatility products between 0.1 and 1.0 may be quite accurately predicted, but predictions in the higher ranges of viscosity-volatility product should be used with discretion.

For use of the correlation under circumstances not specified in this work, reference is made to an article by O'Connell (14).

The correlation is in substantial agreement with the correlation of O'Connell, and the types of systems have

been expanded. Although the entire investigation involved the use of binary systems, the expansion to include multicomponent fractionations seems possible in view of the agreement between this correlation and the correlation of O'Connell which included multicomponent separations.

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APPENDIX

Figure 14: Experimental Fractionating Column

MATT. CORCORAN & CO.	
LOUISVILLE, KY. U. S. A.	
FOR SPECIAL SPECIFICATIONS - SEE DRAWING	
TITLE: FRACTIONATING COLUMN	
DATE: 1/1/28	DESIGNED BY: MATT. CORCORAN
ORDER NO. 1000	APPROVED BY: MATT. CORCORAN
	DRAWING NO. 448

FIG.15- DETAILS OF PLATE CONSTRUCTION

SAMPLE CALCULATIONS

These calculations are for Run No. 11,
methanol-water.

Vapor Velocity

$$\text{Vap. Vel.} = \frac{W_d (359)(t_1)(4)}{3600 (M)(273) \pi D^2}$$

where, W_d = Pounds vapor/hour
 M = Molecular weight of product
 t_1 = Temperature, plate (degrees K.)
 D = Column diameter, feet

$$\text{Vap. Vel.} = \frac{(39 + 66)(359)(339)(144)(4)}{3600(.948 \times 32 + .052 \times 18)(273)(\pi)(10)^2}$$

$$\text{Vap. Vel.} = 0.76 \text{ ft./sec.}$$

Reflux Ratio

$$L/V = \frac{\text{Pounds liquid reflux / unit time}}{\text{Pounds vapor / unit time}}$$

$$L/V = \frac{66}{66 + 39}$$

$$L/V = 0.63$$

Material Balance

Methanol input in feed = $W_f(\text{wt. \% methanol})$

$$\text{" " " " } = 115(0.294)$$

$$\text{" " " " } = 33.8 \text{ pounds / hour}$$

Methanol in product = $W_p(\text{wt. \% methanol})$

$$\text{" " " " } = 39(0.972)$$

$$\text{" " " " } = 37.8 \text{ pounds / hour}$$

Methanol in waste = Negligible

Material Balance (Cont.)

$$\text{Balance} = \frac{100(\text{output})}{(\text{input})}$$

$$\text{Balance} = \frac{100(37.8)}{33.8}$$

$$\text{Balance} = 112 \%$$

(Material balances within 80 to 120 % were considered within the limits of the experimental data.)

$$\text{Slope of } q \text{ Line} \left(\frac{q}{q-1} \right)$$

$$q = \frac{\text{latent heat of feed at feed temperature}}{\text{latent heat of feed at boiling point}}$$

$$q = 1 + \frac{\text{sensible heat (tb to tf)}}{\text{latent heat at tb}}$$

$$q = 1 + \frac{(85 - 60)(.81 \times 18 + .19 \times 22.5)}{.81(9720) + .19(8410)}$$

$$q = 1.05$$

$$\text{Slope of } q \text{ line} = 21$$

Theoretical Plates

The number of theoretical plates was calculated by the method of McCabe-Thiele. The equilibrium data were plotted in Figure 16, and the rectifier operating line drawn from X_p at a slope of L/V . From the X equal Y line at X_f , the q line was drawn with a slope of 21 to intersect the rectifier operating line. From this intersection to X_w , the stripper operating line was drawn. By a stepwise calculation from X_{19} to X_1 , the number of theoretical plates was found to be 9.15.

FIG.16 - SAMPLE CALCULATION OF
THEORETICAL PLATES
(METHANOL-WATER FRACTIONATION)

Overall Efficiency

$$\text{Efficiency} = \frac{(100) \text{ theoretical plates required}}{\text{actual plates used}}$$

$$\text{Efficiency} = \frac{100(9.15)}{18}$$

$$\text{Efficiency} = 50.9 \% \text{ (Overall column efficiency)}$$

Stripper and rectifier efficiencies were calculated in the same manner, using different end values for X.

Relative Volatility (α)

From a plot of α vs. composition, determine α at the average composition in the section. $\alpha = 3.8$.

Pseudo Molal Viscosity (μ)

$$\mu = X_A \mu_A + X_B \mu_B$$

$$\mu = .44(.33) + .56(.36)$$

$$\mu = 0.35$$

Viscosity-Volatility Product ($\mu\alpha$)

$$\mu\alpha = 0.35(3.8)$$

$$\mu\alpha = 1.33$$

NOMENCLATURE

C _p	Metal heat capacity of liquid
D	Column diameter, feet
G	Mass flow rate
K	Mass transfer coefficient
L	Liquid flow in column, mols
M	Mean molecular weight
N	Number of mols of substance transferred
R	Gas constant
T	Absolute temperature, degrees Kelvin
V	Vapor flow in column, mols
W	Flow rate, pounds per hour
X, x	Mol fraction of a component in liquid
Y, y	Mol fraction of a component in vapor
Z	Viscosity, pounds per foot-hour
f	Fanning friction factor
j	Mass transfer factor
k	Diffusion coefficient, square feet per hour
q	Factor determining slope of line joining feed composition and operating line intersection in McCabe-Thiele calculation of theoretical plates
δ	Film thickness
t	Temperature, degrees Centigrade

NOMENCLATURE (Cont.)

- α Relative volatility
 μ Pseudo molal viscosity
 λ Latent heat of vaporization
 ρ Density, pounds per cubic foot

Subscripts

- A More volatile component
B Less volatile component
b Boiling point
d Distillate
f Feed
p Product
w Bottoms, waste
1, 2, 3, Plate number from top (top plate No. 1)

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VITA

Earl Kenneth Stigger, son of Clifford T. and Elnora (Knauer) Stigger, was born in Louisville, Kentucky, June 12, 1920. After completing grammar school in Louisville, he entered duPont Manual Training High School and Graduated Valedictorian in 1938. Entering the Speed Scientific School of the University of Louisville in 1938, he pursued a course in Chemical Engineering, receiving the Degree B. Ch. E. in June, 1942. He was presented the A. I. Ch. E. Student Chapter award at the close of the sophomore year, and won three awards in recognition of his scholarship in 1942.

After two years industrial experience with Monsanto Chemical Co. (St. Louis) and two years service in the U. S. N. R., he re-entered the Speed Scientific School in 1946, receiving the Degree M. Ch. E. in March, 1947.