AN O'CONNELL TYPE CORRELATION FOR PREDICTION OF OVERALL EFFICIENCY OF VALVE TRAY COLUMNS

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CHAPTER I

INTRODUCTION

Distillation, king in separation, the most important and the most visible separation technology (Kister, 1992), is believed to have originated around 800 BC (Resetarits and Lockett, 2002). In the United States, there are over 40,000 distillation columns (Weiland and Resetarits, 2002).

Distillation accounts for nearly 3 % of the total U. S. annual energy consumption, and represents approximately 30 % of the energy used in chemical plants and petroleum refineries in the United States (Resetarits and Lockett, 2002). Seader and Henley (1998) reported that in early 1991 an average of 13 million bbl/day of crude oil was processed by petroleum refineries in the United States. The massive scale of operation implies significant potential impact with small improvement in distillation efficiency.

A distillation column differs from a simple pressure vessel due to its internal vapor-liquid contacting devices (Fair, 1965). These internals can be broadly classified into packing and trays. The selection of distillation column internals is a part of the economic optimization of the overall column design. The basis for selection of column internals along with their advantages and disadvantages are discussed elsewhere (Lockett, 1986;

Kister, 1992; Fair, 1965). This study focuses on distillation columns that use trays.

The most common types of trays are bubble-cap, sieve, valve, and dual flow (Kister, 1992). Dual flow trays do not have downcomers. Vapor and liquid in dual flow trays exhibit counter-current flow through holes. These trays are limited in services such as handling dirty liquids (Fair, 1983). Bubble-cap trays were favored during the first half of the 20th century, but by the mid-1960's, sieve and valve trays became the most commonly used tray types (Fair, 1983). Since 1975, bubble-cap trays have generally been specified only for specialized services. In the mid-1950's, simple, relatively easy to model, and inexpensive sieve trays became prevalent (Fair, 1983). Valve trays, an intermediate cost alternative, were developed early in the 1950-1975 period to overcome the disadvantages of complicated tray geometry and high cost of bubble-cap trays as well as the limited turndown of sieve trays (Fair, 1983).

The selection of tray type for a distillation column is important for effective separation. The primary factors affecting selection are capacity, efficiency, cost, and pressure drop (Fair, 1965). This study focuses only on efficiency.

The number of actual trays is calculated from the number of theoretical trays using the overall column efficiency (E_O). The overall column efficiency (E_O) directly affects capital cost, and hence is an important parameter in the design of a distillation column. Seader and Henley (1998) showed that for a propane / propylene splitter column with 150 theoretical trays, a decrease in E_O from 70 to 60 % causes an increase of 34 actual trays in the column. These additional 34 trays increase the height of the distillation column, thereby increasing the capital cost of the column. Therefore, there is an economic incentive to predict E_O as accurately as possible.

The preferred method to predict efficiency is the use of field performance data from a similar column or system. However, such data are seldom available in the open literature (Klemola and Ilme, 1996). Therefore, design engineers rely on efficiency models. This work is an attempt to help design engineers generate an improved estimate of overall column efficiency.

There are two distinct types of efficiency models. The first type is the theoretical model, which is based on first principles. However, consideration of theoretical models is outside the scope of this thesis.

The second type is the empirical model. Empirical models to predict tray efficiencies have been developed by Drickamer and Bradford (1943), O’Connell (1946), Chaiyavech and Van Winkle (1961), English and Van Winkle (1963), Onda, et al. (1971), Tarat, et al. (1974), and MacFarland, et al. (1972). Of the available empirical models, the O’Connell correlation is the most popular and widely recommended correlation (Kister, 1992).

The O’Connell correlation was developed in 1946 using bubble-cap and sieve tray data to estimate the overall column efficiency. Lockhart and Leggett (1958) attempted to improve the correlation using bubble-cap tray column data.

The O’Connell correlation should be used only for columns with similar tray design due to its empirical nature. This implies that the O’Connell correlation should be limited to prediction of E_O for columns with bubble-cap and sieve trays. There had been no attempt to improve the correlation using valve tray data until 1998. In 1998, Seader and Henley augmented Lockett and Leggett's updated version of the O'Connell correlation by incorporating FRI valve tray data for a single valve tray design (Glitsch Ballast tray). The reliability of an empirical model improves with increase in the amount of quality data with a wide range of variations. Therefore, the O'Connell correlation needs to be modified using more than one valve tray design for it to be reliable.

The basis of this work is that, an efficiency correlation having the same form as the O'Connell correlation, based on more than one valve tray design data, can be developed to provide more reliable efficiency predictions for valve tray column.

Thus, the objective of this work is to develop an efficiency correlation that will correlate E_O as a function of product of liquid viscosity and relative volatility using valve tray data with various tray designs.

1.1 Contribution of this work

The main contribution of this work is development of a new O'Connell type correlation to predict overall column efficiency using FRI valve tray data from eleven valve tray designs. An augmented O'Connell correlation using O'Connell (1946), Williams, et al. (1950), and additional FRI valve tray data is also presented.

The intended impact of this research is better and more economical design of trayed distillation columns from use of the proposed overall efficiency correlation.

1.2 Thesis outline

Chapter 2 describes the background of the O’Connell correlation. This includes discussion of the evolution, previous updates, theoretical implications, and literature observations of the O’Connell correlation. The impact of system property parameters on overall column efficiency is also explained.

Chapter 3 presents the new FRI valve tray data along with the estimation of the parameters required for the study. A new correlation for prediction of valve tray column efficiency is reported along with an augmented O’Connell correlation developed from previous bubble-cap and sieve tray data and new FRI valve tray data.

Chapter 4 analyzes the composite database used in this study. These datasets include O’Connell (1946), Williams, et al. (1950), and FRI valve tray data. Variations in pressure, diameter, and tray types are discussed in the context of potential impact on overall column efficiency.

Chapter 5 presents conclusions, recommendations and directions for future work.

CHAPTER II

BACKGROUND AND LITERATURE SURVEY

The O'Connell correlation to predict overall column efficiency was reported in 1946. This chapter presents a review of the O'Connell correlation. This review includes the development, updates, theoretical significance, literature observations, applications, limitations, and implications of the correlation. The agreement of this correlation to experimental data shown by researchers is also documented. Finally, an analysis of the needs and gaps in the research is presented.

2.1 Efficiency definitions

This section of the chapter summarizes the definitions of the most common types of efficiencies.

1. **Point efficiency.** It is defined as

$$E_{OG} = \frac{y_{n,local} - y_{n-1,local}}{y^{*}_{n,local} - y_{n-1,local}} \tag{2.1}$$

where,

E_{OG} is the point efficiency (fraction),

$y_{n,local}$ is the composition of vapor leaving the point,

$y_{n-1,local}$ is the composition of vapor approaching the point, and

$y^*_{n,local}$ is the vapor composition in equilibrium with the composition of the liquid at the point $x_{n,local}$

The liquid composition of the vertical element of fluid above the "local point" is uniform, while the vapor composition varies with elevation above the tray deck.

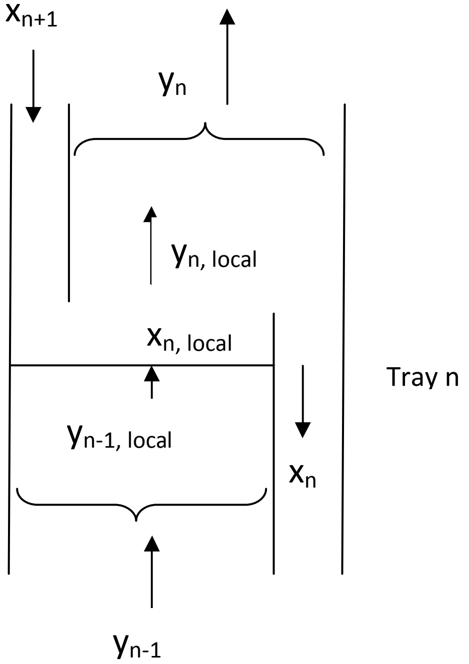

Figure 2.1 Tray efficiency

2. **Murphree vapor phase tray efficiency.** It is defined as

$$E_{MV} = \frac{y_n - y_{n-1}}{y_n^* - y_{n-1}} \tag{2.2}$$

where,

E_{MV} is the Murphree vapor phase tray efficiency (fraction),

y_n is the composition of the perfectly mixed (uniform composition) vapor leaving the tray,

y_{n-1} is the composition of the perfectly mixed (uniform composition) vapor approaching the tray, and

y^*_n is the vapor composition in equilibrium with the composition of the perfectly mixed (uniform composition) liquid leaving the tray (x_n)

3. **Overall column efficiency.** It is defined as

$$E_O = \frac{Number\ of\ theoretical\ or\ ideal\ stages\ in\ the\ column}{Number\ of\ actual\ trays\ in\ the\ column} \tag{2.3}$$

where,

E_O is the overall column efficiency (fraction).

Lockett (1986) stated that the use of more rigorous and acceptable definitions of efficiency is practically difficult. Overall efficiency is the most commonly used efficiency for the design of distillation columns (Lockett, 1986). Therefore, this study considers only overall column efficiency (E_O).

A number of theoretical and empirical models for efficiency prediction are available in the literature. Theoretical models have been developed by Syeda, et al. (2007), Garcia and Fair (2000), Scheffe and Weiland (1987), Chan and Fair (1984), Lockett and Ahmed (1983), Zuiderweg (1982), Garrett, et al. (1977), Bolles (1976), Todd and Van Winkle (1972), Kastanek (1970), Bakowski (1952), AIChE (1958), and Prado and Fair (1990).

Empirical models to predict tray efficiencies have been developed by Drickamer and Bradford (1943), O'Connell (1946), Chaiyavech and Van Winkle (1961), English and Van Winkle (1963), Onda, et al. (1971), Tarat, et al. (1974), and MacFarland, et al. (1972). The aforementioned empirical correlations are summarized and presented below.

In 1943, Drickamer and Bradford developed a correlation to predict the overall column efficiency (E_O) of fractionators and absorbers using bubble-cap and sieve tray data. The form of the correlation is:

$$E_o = 0.17 - 0.616 \log_{10} \sum_{i=1}^{c} Z_i \, \mu_L \tag{2.4}$$

where,

E_O is the overall column efficiency (fraction),

Z_i is the feed liquid mole fraction,

c is the number of components in the feed, and

$\mu_{L,i}$ is the liquid viscosity of component "i" (cP).

In 1946, O'Connell included a relative volatility term in Drickamer and Bradford's (1943) correlation and reported a correlation to predict the overall column

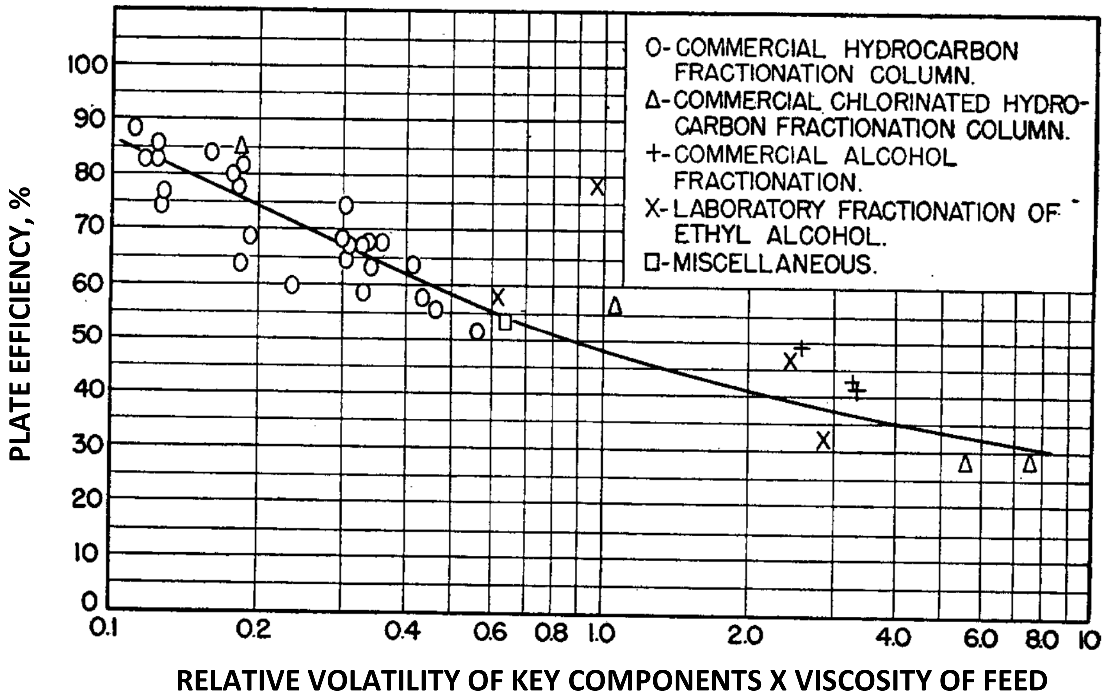

Figure 2.2: O'Connell correlation (Adapted from: O'Connell, H. E. (1946). Plate efficiency of fractionating columns and absorbers. *Transactions of the American Institute of Chemical Engineers, 42*(4), 741-755)

efficiency for fractionating columns using bubble-cap and sieve tray data. The O'Connell correlation is presented Figure 2.2.

In 1961, Chaiyavech and Van Winkle developed a correlation to predict the Murphree plate efficiency of perforated trays. The equation is:

$$E_{MV} = C\left(\frac{\sigma}{\mu_L V_g}\right)^{0.64}\left(\frac{\mu_L}{\rho_L D_L}\right)^{0.19}(\alpha)^{0.056} \tag{2.5}$$

where,

C is the constant required to determine for each set of design,

σ is the liquid surface tension (dyne/cm),

μ_L is the liquid viscosity (cP),

V_g is the superficial vapor velocity based on column cross-sectional area (ft/sec),

ρ_L is the liquid density (grams/cm^3)

D_L is the molecular diffusion coefficient for the liquid phase (cm^2/sec), and

α is the relative volatility.

In 1963, English and Van Winkle attempted to predict the Murphree vapor efficiency of bubble-cap and perforated trays by developing the correlation:

$$E_{MV} = 10.84(FA)^{-0.28}\left(\frac{L}{V}\right)^{0.024}(h_w)^{0.241}(G)^{-0.013} \\ X\left(\frac{\sigma}{\mu_L V_g}\right)^{0.044}\left(\frac{\mu_L}{\rho_L D_L}\right)^{0.137}(\alpha)^{-0.028} \tag{2.6}$$

where,

FA is the free area fraction in column cross-section (ft^2),

L/V is the reflux ratio,

h_w is the weir height (inch),

σ is the surface tension (dyne/cm),

μ_L is the liquid viscosity of the mixture (poise),

ρ_L is the liquid density of the mixture ($grams/cm^3$),

D_L is liquid molecular diffusion coefficient (cm^2/sec),

G is the superficial mass vapor velocity based on column cross-section ($lb/hr.ft^2$), and

V_g is the superficial vapor velocity based on column cross-section (cm/sec),

Onda, et al.'s (1971) modified Murphree vapor efficiency correlation was developed based on bubble-cap tray data. Onda, et al. (1971) defined the modified Murphree vapor efficiency as:

$$(E_{MV})'_{\frac{1}{2}} = \frac{y''_n - y''_{n-1}}{(y^o_n) - y''_{n-1}} \tag{2.7}$$

where,

$(E_{MV})'_{1/2}$ is the modified plate efficiency based on vapor in equilibrium with the mean of the liquid compositions at the inlet and the outlet of the plate,

y_n'' is the vapor composition from the nth plate,

y''_{n-1} is the vapor composition approaching tray n from the n-1th plate,

(y_n^o) is the vapor composition in equilibrium with x_n', and

x_n', is the mean of the liquid compositions at the inlet and the outlet of the tray.

Onda, et al.'s correlation is:

$$(E_{MV})'_{\frac{1}{2}} = 0.1359(N_{ReV})^{-0.192}(N_{ReL})^{0.242}(N_{ScL})^{0.095}$$
$$X\ (N_{Ca})^{0.085}(\alpha)^{-0.267}\left(\frac{Z.h_w}{S_a}\right)^{0.076} \tag{2.8}$$

where, N_{ReV}, N_{ReL}, N_{ScL}, and N_{Ca} are the vapor Reynold's number, liquid Reynold's number, liquid Schmidt's number, and capillary number respectively. These numbers are defined as follows:

$$N_{ReL} = \frac{4\Gamma}{\mu_L} \tag{2.9}$$

$$N_{ReV} = \frac{u_V d_{s1} \rho_V}{\mu_V} \tag{2.10}$$

$$N_{ScL} = \frac{\mu_L}{\rho_L D_L} \tag{2.11}$$

$$N_{Ca} = \frac{\mu_V u_V}{\sigma} \tag{2.12}$$

where,

u_V is the vapor velocity at slot (cm/sec),

d_{s1} is the equivalent diameter of slot (cm),

Γ is the liquid flow rate per mean unit width of the plate, (g/cm.sec),

σ is the surface tension (dyne/cm),

z is the distance from inlet weir to outlet weir (cm),

h_w is the weir height (cm),

S_a is the sum of area of all slots on one plate (cm^2),

μ_L is the liquid viscosity (grams/cm.sec),

ρ_V is the vapor density ($grams/cm^3$),

μ_V is the vapor viscosity (grams/cm.sec),

ρ_L is the liquid density ($grams/cm^3$),

D_L is the diffusivity of the liquid (cm^2/sec), and

α is the relative volatility.

In 1972, MacFarland, et al. presented two equations for the predicting Murphree vapor plate efficiency of bubble-cap and perforated trays in the form of the dimensionless numbers as follows:

$$E_{MV} = 7.0(N_{Dg})^{0.14}(N_{SC})^{0.1}(N_{Re})^{0.08} \tag{2.13}$$

$$E_{MV} = 6.8(N_{Dg}N_{SC})^{0.115}(N_{Re}N_{SC})^{0.1} \tag{2.14}$$

where, N_{Dg}, N_{SC}, and N_{Re} are the surface tension number, Schmidt number, and Reynolds number respectively and defined as follows:

$$N_{Dg} = \frac{\sigma_L}{\mu_L U_V} \tag{2.15}$$

$$N_{SC} = \frac{\mu_L}{\rho_L D_{LK}} \tag{2.16}$$

$$N_{Re} = \frac{h_w U_V \rho_V}{\mu_L FA} \tag{2.17}$$

where,

E_{MV} is the Murphree vapor plate efficiency (%),

σ_L is the liquid surface tension (dyne/cm),

μ_L is the liquid viscosity (lb/hr-ft),

U_V is the superficial vapor velocity (ft/hr),

ρ_L is the liquid density (lb/ft^3),

D_{LK} is the molecular diffusion coefficient of the liquid light key (ft^2/hr),

h_W is the weir height (inches),

ρ_V is the vapor density (lb/ft^3), and

FA is the fractional free area.

The advantage of Equation 2.14 over Equation 2.13 is that it avoids the need for liquid viscosity (μ_L).

Tarat, et al. (1974) was the first one to develop mass transfer correlations for valve tray data, which is given as:

For the vapor phase:

$$N_G = 0.28(Re_V)^{0.1}\left(\frac{h_v}{a_1}\right)^{0.47}\left(F_f\right)^{-0.7}\left(\frac{A_s}{4A_O}\right)^{1.4} \tag{2.18}$$

For the liquid phase:

$$N_L = 0.32(Re_V)^{0.18}\left(\frac{h_v}{a_1}\right)^{0.47}\left(F_f\right)^{-0.8}\left(\frac{A_s}{4A_O}\right)^{2.4} \tag{2.19}$$

where,

N_G is the number of gas phase transfer units,

N_L is the number of liquid phase transfer units,

Re_V is the Reynolds number for vapor flow through tower-cross section,

h_V is the froth height (m),

a_l is the maximum valve lift (m),

F_f is the fractional hole area,

A_S is the slot area (m^2), and

A_O is the orifice area (m^2).

Of the available empirical models, the O'Connell correlation is the most popular and widely recommended correlation (Kister, 1992). The O'Connell correlation is the most simple, reliable, and accurate correlation available in literature (Kister, 1992).

Therefore, this study adopts the form of the O'Connell correlation as the basis for a new overall efficiency model developed explicitly for valve tray columns.

2.2 Development and updates of the O'Connell correlation

The O'Connell correlation was developed in 1946. Numerous researchers have modified this correlation over the last 64 years. The development and modifications of the O'Connell correlation are summarized in Figure 2.3 and the details are reported in this section of the chapter.

In 1943, Drickamer and Bradford developed an efficiency correlation for fractionating columns and absorbers. They correlated E_O as a function of molal average liquid viscosity (μ_L) of the feed at the average tower temperature. The data used for the development of the Drickamer and Bradford correlation are presented in Appedix A. These data were obtained from refinery columns including stabilizers, debutanizers, butane depropanizers, deisobutanizers, and naphtha fractionators. All of the columns utilized bubble-cap trays. One of the columns used a combination of bubble-cap and sieve trays. The three data points associated with the combination tray type column are classified as sieve tray data points for purposes of this study.

The graphical correlation for fractionating columns and absorbers as presented by Drickamer and Bradford (1943) is reproduced and shown in Figure 2.4. The equation form of this correlation is given by Equation 2.4.

1943 **Drickamer and Bradford correlation**

- Correlated efficiency as a function of liquid viscosity for fractionating columns and absorbers
- Developed using bubble-cap and sieve tray data

1946 **O'Connell correlation**

- Correlated efficiency as a function of product of liquid viscosity and relative volatility for fractionating columns
- Individual correlations are presented for distillation and absorption
- Developed using bubble-cap and sieve tray data

1958 **Lockhart and Leggett's augmented O'Connell correlation**

- Augmented the O'Connell correlation by adding Williams, et al.'s (1950) bubble-cap tray data (presented a single efficiency correlation for distillation and absorption)
- Based on bubble-cap and sieve tray data

40 years

1998 **Seader and Henley's augmented O'Connell correlation**

- Augmented the Lockhart and Leggett's updated O'Connell correlation by adding FRI valve tray data from a single valve tray design
- Based on bubble-cap and sieve tray data and **a single valve tray design data**

2011 **OSU-FRI valve tray correlation**

- Presented the O'Connell type correlation for valve tray data only using **11 valve tray designs data**
- Presented the augmented O'Connell correlation by adding FRI valve tray and Williams, et al.'s (1950) bubble-cap tray data to the O'Connell data (1946)

Figure 2.3: Historical timeline of the development and updates of the O'Connell correlation

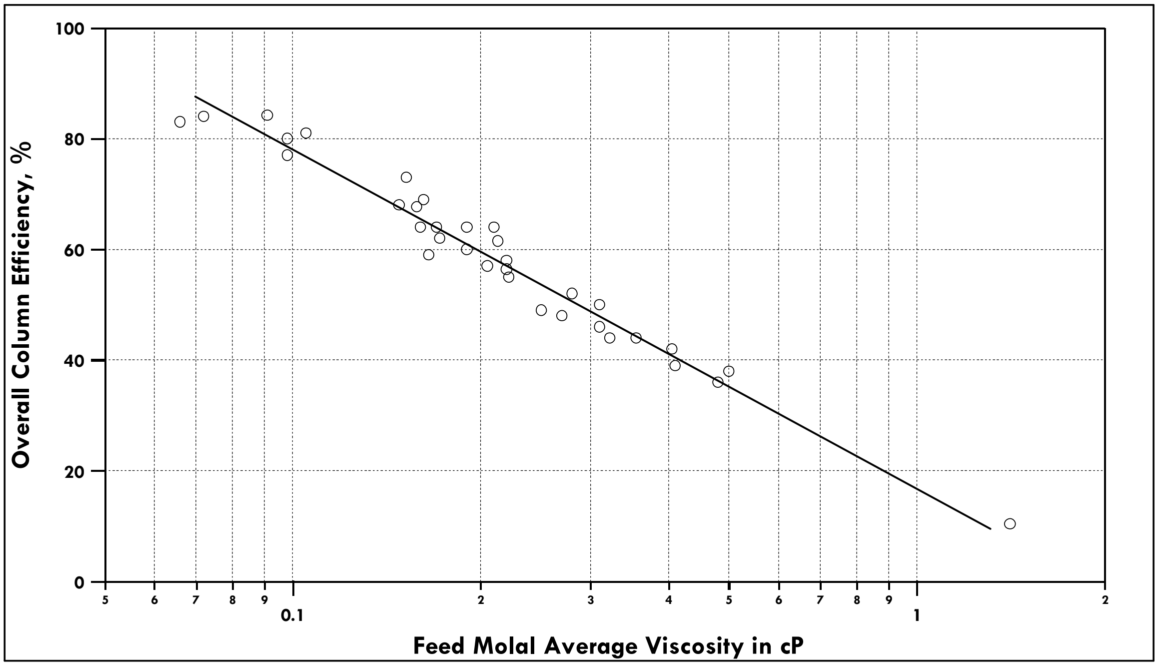

Figure 2.4: Drickamer and Bradford correlation (Reproduced from: Drickamer, H. G., and Bradford, J. R. (1943). Overall plate efficiency of commercial hydrocarbon fractionating columns as a function of viscosity. *Transaction of American Institute of Chemical Engineers, 39*, 319-360.)

The mean absolute relative error between the Drickamer and Bradford correlation predicted E_O and the measured E_O for the data used for the development of the correlation as calculated in this report was 4.9 %.

In 1946, O'Connell pointed out that Drickamer and Bradford's correlation is suitable only for hydrocarbon systems with low relative volatility. To overcome this limitation, O'Connell correlated E_O as a function of the product of the liquid viscosity and relative volatility between the key components of the column feed at the average column temperature and pressure. O'Connell also presented a correlation for absorbers, but this study focuses only on the correlation for fractionating columns.

The O'Connell correlation was developed using data with relative volatilities between 1.16 and 20.51. Data used for the development of the correlation were collected from commercial as well as laboratory columns and included hydrocarbons, chlorinated hydrocarbons, alcohol-water, and trichloroethylene-toluene-water systems (O'Connell, 1946). The data used to develop the correlation are shown in Table 2.1 and the O'Connell correlation plot is shown in Figure 2.2. The data used were obtained from bubble-cap and sieve tray columns. Bubble-cap tray columns contributed 32 data points, and sieve tray columns contributed six data points. The average deviation between the O'Connell correlation predicted E_O and the measured E_O was ±10 %.

2.2.1 Equation forms of the O'Connell correlation

The O'Connell correlation for fractionating columns was presented in the form of a plot, which is shown in Figure 2.2. O'Connell did not provide a correlation equation

O'CONNELL DATA (1946)

Table 2.1: O'Connell data (1946)

System	**Lit. Ref. No.**	**Average Column**		**Types of trays**	**Components present in system**	**Key Components**		**D, inch**	**N**	**$\alpha\mu_L$**	**Meas. Plate E_O, %**
		Temp., F	**Press., psia**			**High Volatility**	**Low Volatility**				
Gasoline Fractionator	1,0	197	215	BC	C_2-C_7	Iso-Butane	Butane	72	30	0.14	74.0
Gasoline Fractionator	1,0	217	265	BC	C_2-C_7	Iso-Butane	Butane	72	30	0.12	88.0
Gasoline Fractionator	1,0	200	218	BC	C_2-C_7	Iso-Butane	Butane	72	30	0.13	86.0
Gasoline Fractionator	1,0	205	223	BC	C_2-C_7	Iso-Butane	Butane	72	30	0.13	83.0
Naphtha Fractionator Test 1	2,0	240	135	BC	C_2, C_3, nC_4, iC_4, nC_5, C_6	Iso-Pentane	Pentane	42	30	0.18	63.0
Naphtha Fractionator Test 2	2,0	236	125	BC	C_2, C_3, nC_4, iC_4, nC_5, C_6	Iso-Pentane	Pentane	42	30	0.19	69.0
Naphtha Fractionator Test 3	2,0	214	125	BC	C_2, C_3, nC_4, iC_4, nC_5, C_6	Butane	Pentane	42	30	0.33	67.4
Naphtha Fractionator Test 4	2,0	157	100	BC	C_2, C_3, nC_4, iC_4, nC_5, C_6	Butane	Pentane	42	30	0.57	51.0
Cracking Unit Stabilizer	3,0	315	347	BC	C_1-C_6	Propane	Butane	72		0.13	84.0
Cracking Unit Stabilizer	3,0,	309	363	BC	C_1-C_6	Butane	Pentane			0.18	77.0
Cracking Unit Stabilizer	3,0	312	357	BC	C_1-C_6	Propane	Butane	72		0.12	83.0
Cracking Unit Stabilizer	3,0	319	363	BC	C_1-C_6	Propane	Butane	72		0.19	81.0
Cracking Unit Stabilizer	3,0	313	366	BC	C_1-C_6	Propane	Butane	72		0.16	84.2
Cracking Unit Stabilizer	3,0	314	350	BC	C_1-C_6	Propane	Butane	72		0.18	80.0
Poly Plant Stabilizer	3,0	198	362	BC		Propane	Butane			0.47	55.0
Poly Plant Stabilizer	3,0	228	365	BC		Propane	Butane			0.44	58.0
Butane Depropanizer	3,0	162	235	BC	C_3, C_3=, C_4. C_4=, C_5, C_5=,C_6	Propane	Butane	54	26	0.36	68.0
Butane Depropanizer	3,0	161	235	BC	C_3, C_3=, C_4. C_4=, C_5, C_5=,C_6	Propane	Butane			0.42	64.0
Debutanizer$^{\#}$	3,0	275	117	Sieve	C_3,C_4, C_5, iC_5, nC_5, C_6+	Butane	Pentane	72	27	0.33	59.0
Deisopentanizer$^{\#}$	3,0	280	116	Sieve	C_3, C_4, C_5=, iC_5, nC_5, C_6+	Pentane	Hexane			0.30	64.0
Deisopentanizer$^{\#}$	3,0	270	116	Sieve	C_3, C_4, C_5=, iC_5, nC_5, C_6+	Pentane	Hexane			0.34	62.0

#Bottom 11 trays perforated in tower

0	O'Connell (1946)	1	Brown and Lockhart (1943)
2	McGiffin (1942)	3	Drickamer and Bradford (1943)

(Table 2.1 Continued)

System	Lit. Ref. No.	Average Column		Types of trays	Components present in system	Key Components		D, inch	N	$\alpha\mu_L$	Meas. Plate E_O, %
		Temp., F	Press., psia			High Volatility	Low Volatility				
Deisopentanizer	3,0	262	116		C_3, C_4, C_5=, iC_5, nC_5, C_6+	Pentane	Hexane				
Deisopentanizer	3,0	264	115	BC	C_3, C_4, C_5=, iC_5, nC_5, C_6+	Pentane	Hexane			0.30	69.0
Deisopentanizer	3,0	260	117	BC	C_3, C_4, C_5=, iC_5, nC_5, C_6+	Pentane	Hexane	48	30	0.30	73.0
Deisobutanizer	3,0	304	367	BC	C_1, C_2=, C_2, C_3=, C_3,iC_4, C_4,nC_4, iC_5,C_5=,nC_5, C_6+	Iso-Butane	Butane			0.13	76.7
Deisopentanizer	3,0	253	98	BC	C_3,iC_4,C_4=,nC_4,iC_5,C_5=, nC_5,C_6+	Iso-Pentane	Pentane			0.24	59.5
Stabilization of Ethylene Dichloride	4,0	206	21	BC	H_2O, $C_2H_4Cl_2$	Water	Ethylene Dichloride	24	7	7.60	29.0
Stabilization of Ethylene Dichloride*	4,0	220	28	BC	C_2,C_2H_5CL, H_2O, $C_2H_4CL_2$,$C_2H_3CL_3$	Water	Ethylene Dichloride	18	7	5.60	29.0
Stabilization of Ethylene Dichloride*	4,0	220	28	BC	C_2,C_2H_5CL, H_2O, $C_2H_4CL_2$,$C_2H_3CL_3$	Ethyl Chloride	Ethylene Dichloride	18	7	1.08	57.0
Stabilization of Ethylene Dichloride	4,0	173	151	BC	iC_4, C_2H_5CL	Iso-Butane	Ethyl Chloride	30	24	0.19	85.0
Alcohol-Water, lab, column	5,0	196	14.7	BC		Ethyl Alcohol	Water			2.89	32.0
Alcohol-Water, lab, column	5,0	188	14.7	BC		Ethyl Alcohol	Water			2.54	47.0
Alcohol-Water, lab, column	5,0	179	14.7	BC		Ethyl Alcohol	Water			0.98	77.0
Alcohol-Water, lab, column	5,0	174	14.7	BC		Ethyl Alcohol	Water			0.57	62.0
Beer Stills (Perforated trays)	6,0	208	16	Sieve	C_2H_5OH, H_2O	Ethyl Alcohol	Water	66		3.46	41.0
Beer Stills (Perforated trays)	6,0	210	17	Sieve	C_2H_5OH, H_2O	Ethyl Alcohol	Water	66		3.46	42.5
Alcohol-Water	7,0,	200	15	Sieve	Alcohol, water	Alcohol	Water	36		2.61	49.0
Trichloroethylene Toluene and Water, lab. Column	8,0	210	15	BC	C_2HCL_3, C_7H_8, H_2O	Trichloroethylene	Toluene	8	15	0.635	53.0

#Bottom 11 trays perforated in tower *Results of one test with plate E_O based on different components

BC	Bubble-cap plate				
Temp.	Temperature, F	Press.	Pressure, psia	D	Diameter of the column, inches
N	Number of trays in the column	α	Relative volatility	μ_L	Liquid viscosity, cP
4	Brown and Souders (1934).	5	Keyes and Byman (1941)	6	Gunness and Baker (1938)
7	Peters (1922)	8	Schoenborn, et al. (1941)		

with the plot (1946). However, various forms of equations for the O'Connell correlation have been reported in the literature. These forms of the correlation are reported in Table 2.2. The details of the performance of the various equation forms of the O'Connell correlation is presented in Table 2.3 and summarized in Table 2.2.

Table 2.2: Various equation forms of the O'Connell correlation

Author	Correlation	Mean absolute relative error*, %	Comments
Economopoulos (1978)	$E_O = 0.485 - 0.129 \ln(\alpha\mu_{Feed}) + 0.018(\ln \alpha\mu_{Feed})^2 + 0.001 (\ln \alpha\mu_{Feed})^3$	9.3	All O'Connell data points are considered
Lockett (1986)	$E_O = 0.492(\alpha\mu_L)^{-0.245}$	9.1	Bubble-cap tray data are considered
Kessler and Wankat (1988)	$E_O = 0.54159 - 0.28531 \log (\alpha\mu_L)$	9.0	All O'Connell data points are considered
OSU (2011)	$E_O = 0.514(\alpha\mu_L)^{-0.23}$	9.0	All O'Connell data points are considered

All O'Connell data consists of sieve and bubble-cap tray data.
*Mean absolute relative error is reported based on all O'Connell data (38 data points)

2.2.2 OSU's equation form of the O'Connell correlation

The equation form of the O'Connell correlation as reported in Table 2.2 has been obtained using the best-fit curve technique of Microsoft Excel. All the O'Connell data (1946) were plotted in Figure 2.5 and the best-fit curve was drawn through it using Microsoft Excel trend line of power function. The OSU's equation form of the O'Connell correlation, which is abbreviated as "A" is:

$$E_o = 0.514(\alpha\mu_L)^{-0.23} \tag{2.20}$$

In this report, Eq. 2.20 will be used to represent the original O'Connell correlation, since it was generated using all of the original O'Connell data.

Table 2.3: Performance of various forms of the O’Connell correlation

System	Lit. Ref. No.	Average Column		$\alpha\mu_L$	Meas. Plate E_O, %	Economopoulos equation form		Lockett’s equation form		Kessler and Wankat’s equation form		OSU’s equation form	
		Temp., F	Press., psia			Pred. E_O, %	% Error	Pred. E_O, %	% Error	Pred. E_O, %	% Error	Pred. E_O, %	% Error
Gasoline Fractionator	1,0	197	215	0.14	74.0	80.1	8.2	79.6	7.6	78.5	6.1	80.8	9.2
Gasoline Fractionator	1,0	217	265	0.12	88.0	83.0	5.7	82.7	6.0	80.4	8.6	83.7	4.9
Gasoline Fractionator	1,0	200	218	0.13	86.0	81.5	5.3	81.1	5.7	79.4	7.6	82.2	4.4
Gasoline Fractionator	1,0	205	223	0.13	83.0	81.5	1.9	81.1	2.3	79.4	4.3	82.2	1.0
Naphtha Fractionator Test 1	2,0	240	135	0.18	63.0	75.4	19.7	74.9	18.9	75.4	19.7	76.3	21.0
Naphtha Fractionator Test 2	2,0	236	125	0.19	69.0	74.4	7.9	73.9	7.1	74.7	8.3	75.3	9.1
Naphtha Fractionator Test 3	2,0	214	125	0.33	67.4	64.9	3.7	64.6	4.2	67.9	0.7	66.3	1.6
Naphtha Fractionator Test 4	2,0	157	100	0.57	51.0	56.3	10.4	56.5	10.7	61.1	19.9	58.5	14.7
Cracking Unit Stabilizer	3,0	315	347	0.13	84.0	81.5	3.0	81.1	3.4	79.4	5.4	82.2	2.2
Cracking Unit Stabilizer	3,0,	309	363	0.18	77.0	75.4	2.1	74.9	2.7	75.4	2.1	76.3	1.0
Cracking Unit Stabilizer	3,0	312	357	0.12	83.0	83.0	0.0	82.7	0.3	80.4	3.1	83.7	0.8
Cracking Unit Stabilizer	3,0	319	363	0.19	81.0	74.4	8.1	73.9	8.8	74.7	7.7	75.3	7.0
Cracking Unit Stabilizer	3,0	313	366	0.16	84.2	77.6	7.9	77.1	8.5	76.9	8.7	78.3	7.0
Cracking Unit Stabilizer	3,0	314	350	0.18	80.0	75.4	5.7	74.9	6.4	75.4	5.7	76.3	4.7
Poly Plant Stabilizer	3,0	198	362	0.47	55.0	59.2	7.7	59.2	7.6	63.5	15.5	61.1	11.2
Poly Plant Stabilizer	3,0	228	365	0.44	58.0	60.2	3.9	60.2	3.7	64.3	10.9	62.1	7.0
Butane Depropanizer	3,0	162	235	0.36	68.0	63.5	6.7	63.2	7.1	66.8	1.7	65.0	4.4
Butane Depropanizer	3,0	161	235	0.42	64.0	61.0	4.7	60.9	4.9	64.9	1.4	62.8	2.0
Debutanizer[#]	3,0	275	117	0.33	59.0	64.9	10.0	64.6	9.4	67.9	15.1	66.3	12.4
Deisopentanizer[#]	3,0	280	116	0.30	64.0	66.5	3.9	66.1	3.3	69.1	7.9	67.8	5.9
Deisopentanizer[#]	3,0	270	116	0.34	62.0	64.4	3.8	64.1	3.4	67.5	8.9	65.9	6.3
Deisopentanizer	3,0	262	116	0.31	67.7	65.9	2.6	65.6	3.2	68.7	1.4	67.3	0.6

#Bottom 11 trays perforated in tower

0	O’Connell (1946).	1	Brown and Lockhart (1943).
2	McGiffin (1942)	3	Drickamer and Bradford (1943)

(Table 2.3 Continued)

System	Lit. Ref. No.	Average Column		$\alpha\mu_L$	Meas. Plate E_O, %	Economopoulos equation form		Lockett's equation form		Kessler and Wankat's equation form		OSU's equation form	
		Temp., F	Press., psia			Pred. E_O, %	% Error	Pred. E_O, %	% Error	Pred. E_O, %	% Error	Pred. E_O, %	% Error
Deisopentanizer	3,0	264	115	0.30	69.0	66.5	3.7	66.1	4.2	69.1	0.1	67.8	1.7
Deisopentanizer	3,0	260	117	0.30	73.0	66.5	9.0	66.1	9.5	69.1	5.4	67.8	7.1
Deisobutanizer	3,0	304	367	0.13	76.7	81.5	6.2	81.1	5.7	79.4	3.6	82.2	7.1
Deisopentanizer	3,0	253	98	0.24	59.5	70.3	18.1	69.8	17.3	71.8	20.7	71.4	19.9
Stabilization of Ethylene Dichloride	4,0	206	21	7.60	29.0	30.6	5.4	29.9	3.2	29.0	0.1	32.2	11.2
Stabilization of Ethylene Dichloride*	4,0	220	28	5.60	29.0	32.1	10.8	32.3	11.2	32.8	13.1	34.6	19.3
Stabilization of Ethylene Dichloride*	4,0	220	28	1.08	57.0	47.5	16.6	48.3	15.3	53.2	6.7	50.5	11.4
Stabilization of Ethylene Dichloride	4,0	173	151	0.19	85.0	74.4	12.4	73.9	13.1	74.7	12.1	75.3	11.4
Alcohol-Water, lab, column	5,0	196	14.7	2.89	32.0	37.0	15.5	37.9	18.5	41.0	28.2	40.3	25.8
Alcohol-Water, lab, column	5,0	188	14.7	2.54	47.0	38.1	18.9	39.2	16.7	42.6	9.3	41.5	11.7
Alcohol-Water, lab, column	5,0	179	14.7	0.98	77.0	48.8	36.7	49.4	35.8	54.4	29.3	51.6	32.9
Alcohol-Water, lab, column	5,0	174	14.7	0.57	62.0	56.3	9.2	56.5	8.9	61.1	1.4	58.5	5.7
Beer Stills (Perforated trays)	6,0	208	16	3.46	41.0	35.5	13.5	36.3	11.5	38.8	5.4	38.6	5.8
Beer Stills (Perforated trays)	6,0	210	17	3.46	42.5	35.5	16.6	36.3	14.6	38.8	8.8	38.6	9.1
Alcohol-Water	7,0,	200	15	2.61	49.0	37.9	22.7	38.9	20.6	42.3	13.7	41.2	15.9
Trichloroethylene Toluene and Water, lab. Column	8,0	210	15	0.635	53.0	54.7	3.2	55.0	3.8	59.8	12.8	57.1	7.7

#Bottom 11 trays perforated in tower *Results of one test with plate E_O based on different components

Temp.	Temperature, F	Press.	Pressure, psia
α	Relative volatility	μ_L	Liquid viscosity, cP
4	Brown and Souders (1934).	5	Keyes and Byman (1941)
6	Gunness and Baker (1938)	7	Peters (1922)
8	Schoenborn, et al. (1941)		

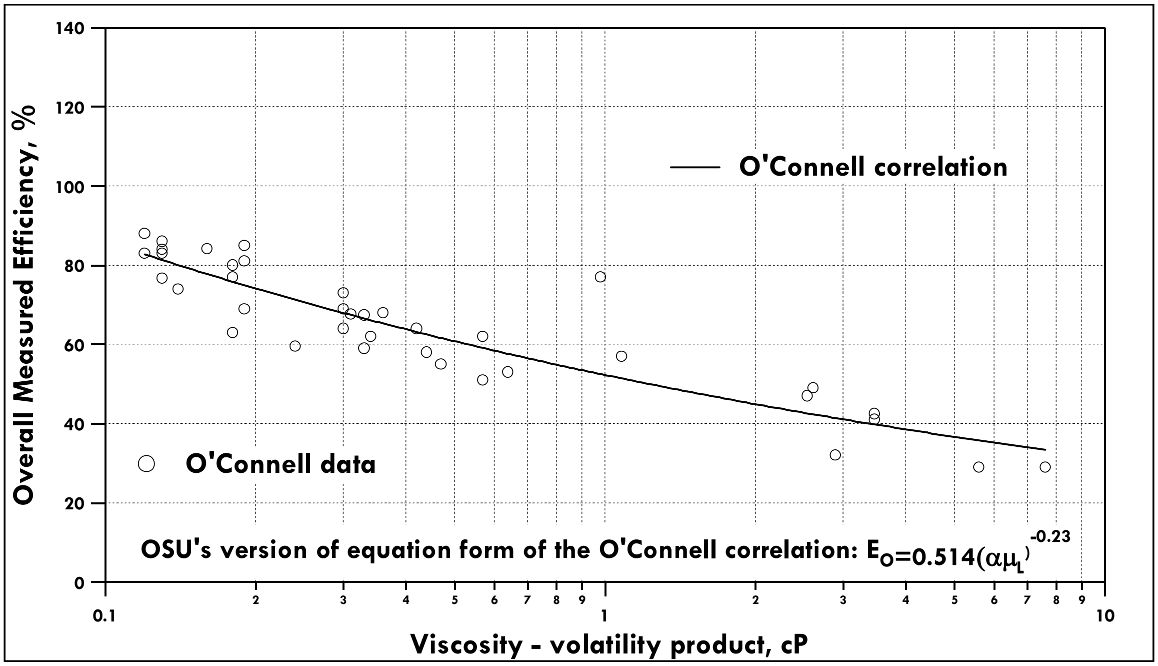

Figure 2.5: OSU curve fit to original O'Connell data (38 data points)

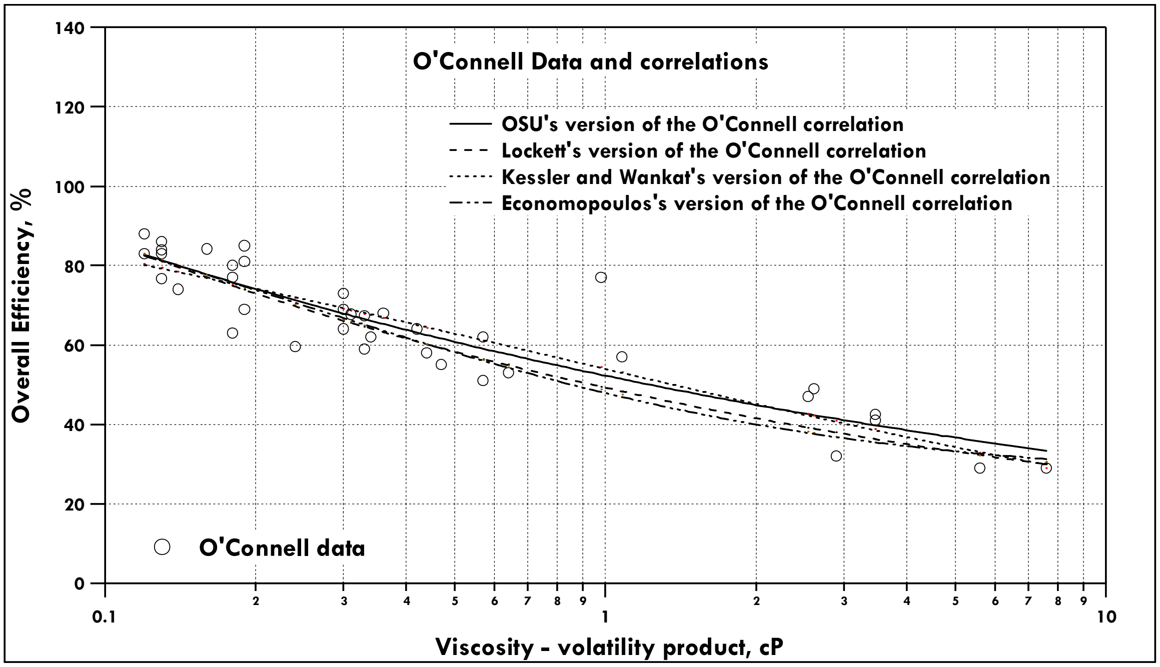

Figure 2.6: Various equation forms of the O'Connell correlation

Many literature sources (Kister, 1992; Weiland and Resetarits, 2002; Wilkinson, et al., 2007) use Lockett's equation form of the O'Connell correlation. All the correlations reported in Table 2.2 are plotted in a single plot in Figure 2.6. It can be seen from Figure 2.6 that these correlations are almost identical.

Although Lockett (1986) and this study used the "power" form of the equation, differences were observed between both equation forms of the correlation. These differences might have occurred because Lockett (1986) reported the correlation only for bubble-cap trays, whereas OSU considered all 38 (bubble-cap and sieve tray) data points.

In 1950, Williams, et al. generated intermediate data points for the O'Connell plot, thereby extending the range of the relative volatility-liquid viscosity product in the O'Connell plot. They utilized systems different from those used to develop the original O'Connell plot. Efficiency data using eight binary systems, having a wide choice of chemical structures and physical properties were reported (Williams, et al., 1950). These data were obtained using a 10-inch diameter column with bubble-cap trays. The Williams, et al. (1950) data were later used by Lockhart and Leggett (1958) to update the O'Connell correlation.

In 1958, Lockhart and Leggett reported a single efficiency correlation for fractionating columns and absorbers using O'Connell's distillation and absorption column (1946), and Williams, et al.'s (1950) distillation data, which is shown in Figure 2.7 They correlated E_O as a function of the $\alpha\mu_L$ product for both fractionating columns and absorbers. In the case of hydrocarbon absorbers, the viscosity-volatility product is represented by the product of the molal average liquid viscosity of the rich oil and ten times the equilibrium constant of the key components (Lockhart and Leggett, 1958). All the distillation data used to produce the Lockhart and Leggett correlation correspond to bubble-cap and sieve trays.

In 1998, Seader and Henley incorporated a small amount of FRI valve tray efficiency data into Lockhart and Leggett's updated O'Connell efficiency correlation plot. The details of the FRI valve tray data used by Seader and Henley (1998) were not reported. This was the first attempt to modify the O'Connell correlation using efficiency data from a valve tray column. However, all the FRI data were obtained from a single valve tray design (Glitsch Ballast tray). The FRI valve tray data correspond to cyclohexane / n-heptane and iso-butene / n-butane systems. Seader and Henley reported the equation form of the O'Connell correlation for distillation data with liquid viscosities between 0.1 to 10 cP. The equation is:

$$E_o = 0.503(\alpha\mu_L)^{-0.226} \tag{2.21}$$

Seader and Henley (1998) reported that the actual efficiencies associated with the FRI data are 10 to 20 % higher than those predicted by their augmented correlation (Eq. 2.21). They observed that the correlation is conservative for well-designed trays. O'Connell (in Williams, et al., 1950) reported that there is much variation in tray design. Therefore, O'Connell recommended using the correlation for similar column designs on which it was developed (Williams, et al., 1950). This implies that the original O'Connell correlation should be limited to prediction of E_O for columns with bubble-cap and sieve trays.

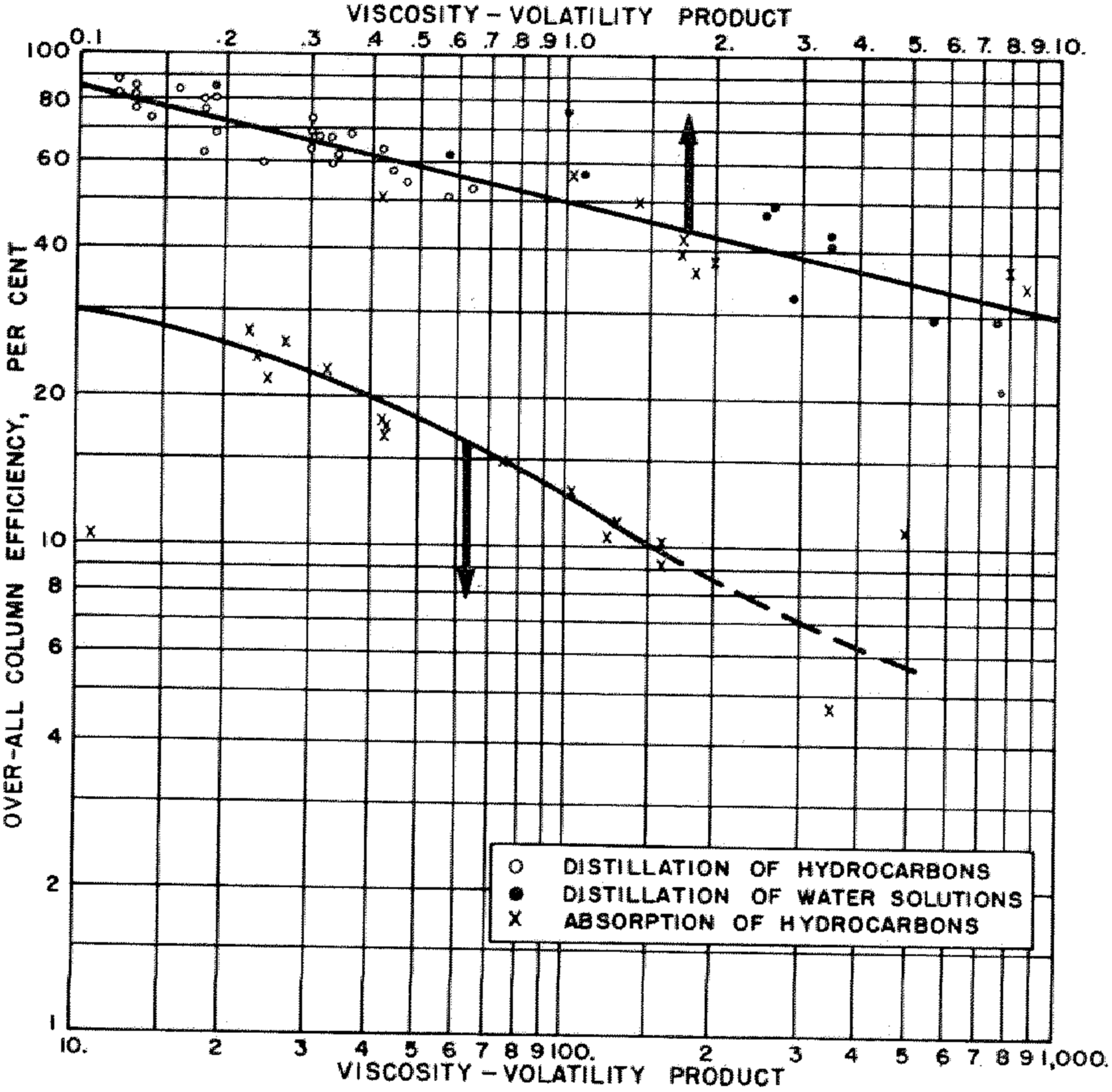

Figure 2.7: Lockhart and Leggett's augmented O'Connell correlation (Adapted from: Lockhart, F. J., and Leggett C. W. . (1958). *Advances in Petroleum Chemistry and Refining, 1*, 323-326)

2.3 Theoretical significance of the O'Connell correlation

Although the O'Connell correlation is an empirical correlation, its connection to first principles has been established by Chen and Chuang (1995) and the analysis of the Chen and Chuang (1993) model by Yang and Chuang (1995).

Chen and Chuang (1995) showed that E_O can be correlated to the $\alpha\mu_L$ product, based on the two-film theory using a steady state mass balance over an elemental strip of the froth shown in Figure 2.8.

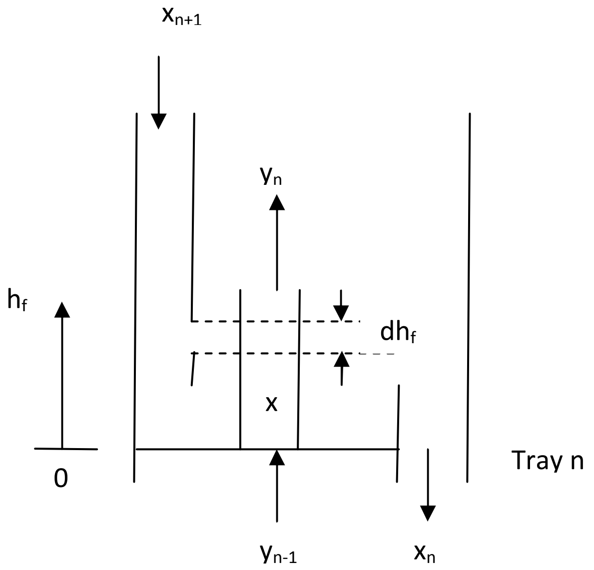

Figure 2.8: Vapor-liquid froth on tray n

The Chen and Chuang (1995) derivation of the O'Connell correlation are described as follows:

A mass balance is performed over an elemental strip of the froth (Figure 2.8) using the following assumptions:

- The gas is in vertical plug flow through the froth
- The liquid is completely mixed in the vertical direction

The equation obtained from this mass balance is:

$$u_s(dA_b)(dy) = K_{OG}(y^* - y)a(dA_b)(dh_f) \tag{2.22}$$

where,

u_s is the gas velocity based on the bubbling area (ft/s),

A_b is the tray bubbling area (ft^2),

y is the mole fraction of the more volatile component in the gas phase,

K_{OG} is the overall gas-phase mass-transfer coefficient (ft/s),

y* is the mole fraction of the more volatile component in the gas phase in equilibrium with the liquid, x,

a is the effective interfacial area (ft^2/ft^3), and

h_f is the froth height (ft).

Rearranging Equation 2.22 gives

$$\frac{K_{OG}a}{u_s}dh_f = \frac{dy}{y^* - y} \tag{2.23}$$

Integrating Equation 2.23 yields

$$\int_0^{h_f} \frac{K_{OG}a}{u_s} dh_f = \int_{y_{n-1}}^{y_n} \frac{dy}{y^* - y} = -\ln\left(\frac{y^* - y_n}{y^* - y_{n-1}}\right) \quad (2.24)$$

If N_{OG} is defined as

$$N_{OG} = \int_0^{h_f} \frac{K_{OG}a}{u_s}\, dh_f \quad (2.25)$$

where N_{OG} is the number of overall gas-phase mass-transfer units.

Assuming K_{OG}/u_S is independent of h_f, Equation 2.25 becomes

$$N_{OG} = K_{OG}a\frac{h_f}{u_s} = K_{OG}at_G \quad (2.26)$$

where,

$$t_G = \frac{h_f}{u_s} \quad (2.27)$$

Combining Equations 2.24 and 2.25 gives

$$N_{OG} = -\ln\left(\frac{y^* - y_n}{y^* - y_{n-1}}\right) \quad (2.28)$$

From Equation 2.28,

$$N_{OG} = -\ln\left(\frac{y^* - y_n + y_{n-1} - y_{n-1}}{y^* - y_{n-1}}\right)$$

$$exp\,(-N_{OG}) = -\left(\frac{y_n - y_{n-1}}{y^* - y_{n-1}}\right) + \left(\frac{y^* - y_{n-1}}{y^* - y_{n-1}}\right) \quad (2.29)$$

But, E_{OG} (gas-phase point efficiency) can be defined as:

$$E_{OG} = \frac{y_n - y_{n-1}}{y^* - y_{n-1}} \quad (2.30)$$

Substituting Equation 2.30 into Equation 2.29 gives

$$exp\ (-N_{OG}) = -E_{OG} + 1$$

and $$E_{OG} = 1 - exp\ (-N_{OG}) \tag{2.31}$$

Using the two-film theory,

$$\frac{1}{K_{OG}} = \frac{1}{k_G} + \frac{m\rho_G M_L}{k_L \rho_L M_G} \tag{2.32}$$

where,

k_G is the gas-phase mass-transfer coefficient (ft/s),

k_L is the liquid-phase mass-transfer coefficient (ft/s),

m is the slope of the vapor-liquid equilibrium line,

ρ_G is the gas density (lbm/ft^3),

ρ_L is the liquid density (lbm/ft^3),

M_L is the liquid molecular weight (lbm/lbmol), and

M_G is the gas molecular weight (lbm/lbmol).

Dividing Equation 2.32 by (at_G),

$$\frac{1}{N_{OG}} = \frac{1}{N_G} + \frac{m}{N_L} \tag{2.33}$$

with

$$N_G = k_G a t_G \tag{2.34}$$

$$N_L = k_L a t_G \frac{\rho_L M_G}{\rho_G M_L} \tag{2.35}$$

where,

N_G is the number of gas-phase mass-transfer units, and

N_L is the number of liquid-phase mass-transfer units.

$(1/N_{OG})$, $(1/N_G)$, and (m/N_L) constitute the total mass-transfer resistance, gas-phase mass-transfer resistance, and liquid-phase mass-transfer resistance, respectively. The fraction of liquid phase resistance (LPR) can be given as

$$LPR = \frac{m/N_L}{1/N_{OG}} = \frac{mN_{OG}}{N_L} \tag{2.36}$$

The fraction of gas-phase resistance (GPR) can be given as

$$GPR = \frac{1/N_G}{1/N_{OG}} = \frac{N_{OG}}{N_G} = 1 - LPR \tag{2.37}$$

From Equation 2.33,

$$N_{OG} = \frac{N_L N_G}{N_L + mN_G} \tag{2.38}$$

Substituting N_{OG} value from Equation 2.38 in Equation 2.36 yields,

$$LPR = \frac{m/N_L}{1/N_{OG}} = \frac{m}{N_L}\left[\frac{N_L N_G}{N_L + mN_G}\right] \tag{2.39}$$

Dividing and multiplying the right hand side of Equation 2.39 by $N_L N_G$ yields

$$LPR = \frac{m}{\frac{N_L}{N_G} + m} \tag{2.40}$$

If gas-phase resistance in distillation is assumed to be negligible ($1/N_G = 0$) then Equation 2.33 yields

$$\frac{1}{N_{OG}} = \frac{m}{N_L} \tag{2.41}$$

Substituting the value of N_L from Equation 2.35 into Equation 2.41 yields

$$\frac{1}{N_{OG}} = \frac{k_L}{m} a t_G \frac{\rho_L M_G}{\rho_G M_L} \propto \frac{k_L}{m} \tag{2.42}$$

Zuiderweg (1982) obtained

$$k_L = \frac{0.000026}{{\mu_L}^{0.25}} \propto {\mu_L}^{-0.25} \tag{2.43}$$

where μ_L is the liquid viscosity (N s/m^2).

The mole fraction of the more volatilite component in a binary mixture can be related by

$$y = \frac{\alpha x}{1 + (\alpha - 1)x} \tag{2.44}$$

where,

α is the relative volatility, and

x is the mole fraction of the more volatile component in the liquid phase.

Differentiating Equation 2.44, m can be given as

$$m = \frac{dy}{dx} = \frac{\alpha}{(1 + (\alpha - 1)x)^2} \tag{2.45}$$

Chen and Chuang (1995) showed that when α is in the range from 1 to 2 and x=0.29, then Equation 2.45 is approximately equivalent to

$$m = 1.02\alpha^{0.25} \tag{2.46}$$

Substituting Equations 2.46 and 2.43 into Equation 2.42 gives

$$N_{OG} \propto (\alpha\mu_L)^{-0.25} \tag{2.47}$$

If N_{OG} is assumed to be small, Equation 2.31 can be approximately given as

$$E_{OG} = N_{OG} \tag{2.48}$$

If the liquid on the tray is completely mixed, then the relationship between E_{OG} and E_{MV} can be given as

$$E_{OG} = E_{MV} \tag{2.49}$$

where E_{MV} is the Murphree gas-phase tray efficiency (fraction).

If λ (λ=mG/L) is close to unity, the overall column efficiency can be approximately given by

$$E_O = E_{MV} \tag{2.50}$$

where,

G is the molar gas flow rate,

L is the molar liquid flow rate, and

E_O is the overall column efficiency (fraction).

Combining Equations 2.50, 2.49, 2.48, and 2.47 yields

$$E_O \propto (\alpha\mu_L)^{-0.25} \tag{2.51}$$

Equation 2.51 is almost identical to the O'Connell correlation.

The theoretical significance of the O'Connell correlation can also be realized indirectly from Chen and Chuang (1993). According to the O'Connell correlation, the impact of physical properties on efficiency is more significant than that of liquid and vapor flow rate and tray geometry. Chen and Chuang (1993) developed a semi-empirical model based on physical properties and tray geometry. The analysis of this model by Yang and Chuang (1995) showed that the influence of internal tray design parameters such as weir height, weir length per unit bubbling area, fractional perforated tray area, and pitch of holes are less significant than the physical properties of the system.

2.4 Theoretical significance of the O'Connell correlation variables

Although the O'Connell correlation is empirical, O'Connell found that its variables have theoretical significance, which can be explained as follows (Williams, et al., 1950):

- O'Connell (in Williams, et al., 1950) reported that Chilton and Colburn (1934) showed that mass transfer of vapor and liquid is a function of viscosity.
- O'Connell (in Williams, et al., 1950) also reported that relative volatility is approximately proportional to the slope of the equilibrium curve. Gerster, et al. (1945) showed that relative volatility affects efficiency.

Other researchers who found theoretical significance in the variables of the O'Connell correlation include King (1980), Lockett (1986), and Kister (1992). With an increase in μ_L, liquid diffusivity decreases, which results in an increase in liquid phase resistance, and hence efficiency decreases.

The inverse relationship between liquid viscosity (μ_L) and efficiency can be explained using the hydrodynamical theory for molecular diffusivity. The Stoke-Einstein equation (Welty, et al., 2007) developed from the hydrodynamical theory for molecular diffusivity shows that diffusivity is an inverse function of viscosity.

$$D = \frac{kT}{6\pi r\mu} \tag{2.52}$$

where,

D is the diffusivity (ft^2/hr),

k is the Boltzmann constant (1.38×10^{-16} erg/K),

T is the absolute temperature (K),

r is the solute particle radius (ft), and

μ is the solvent viscosity (lbm/ft s).

O'Connell included a term, α based on a multi-component stabilization test of ethylene dichloride. This test showed that efficiency increased with a decrease in relative volatility using different low boiling components as observed from Table 2.4. O'Connell could not find any theoretical explanation for this relationship.

Table 2.4: Tests of Stabilization of Ethylene Dichloride (O'Connell, 1946)

High-Boiling component	Low-Boiling component	Relative volatility	Efficiency, %
Ethylene Dichloride	Water	16	29
Ethylene Dichloride	Ethyl-Chloride	3.1	57

2.5 Literature observations regarding the O'Connell correlation

The O'Connell correlation was developed in 1946. Over the last 64 years, this correlation has been criticized and evaluated by many literature sources, which are summarized in this section of the chapter.

This correlation has received criticism for being conservative (Chen and Chuang, 1993; Bribika and Biddulph, 1986; Seader and Henley, 1998) and for not having been validated (Chen and Chuang, 1993; Bribika and Biddulph, 1986; Chan and Fair, 1993; Wankat, 1988). In addition, Onda, et al. (1971) reported that it gives satisfactory results only for hydrocarbon systems.

Although the O'Connell correlation is considered conservative by many researchers, Yang and Chuang (1995) considered this criticism as "unfortunate" and further stated that it is one of the best available empirical correlations to predict efficiency. The other researchers who recommended the correlation, as cited by Kister (1992), include Ludwig (1979), Fair, et al. (1984), King (1980), Vital, et al. (1984), Wankat (1988), Hines and Maddox (1985). Kister (1992) acknowledged this correlation as a "standard of the industry" because of the combination of reliability, reasonable accuracy, and simplicity along with the weakness of theoretical tray efficiency correlations. Weiland and Resetarits (2002) added that this is the most successful E_O correlation in the literature.

2.6 Application of the O’Connell correlation to literature data

Several researchers have applied the O’Connell correlation to literature data and showed that predictions of the correlation are in good agreement with the data. Weiland and Resetarits (2002) applied the O’Connell correlation to data from the proprietary ULTRA-FRAC co-current flow tray and showed that measured efficiencies agree “extremely well” with the O’Connell predicted efficiencies. Other researchers who were able to show that predictions of the correlation are in good agreement with experimental data include Wilkinson, et al. (2007) and Marek and Novosad (1955).

Marek and Novosad (1955) showed that the experimental efficiencies of an acetic acid-water system in a 30-plate column with a diameter of 23.6 inches are in satisfactory agreement with the O’Connell correlation predictions, at 5.4, 8.4, and 14.4 psia pressures. This plot is shown in Figure 2.9. These data were collected at total reflux conditions and in a bubble-cap tray column.

Wilkinson, et al. (2007) showed that the O’Connell correlation provides a reasonable estimate for efficiency of super high capacity trays. They summarized the predicted efficiencies of super high capacity trays such as the Shell ConSep, Jaeger COFLO, and Koch-Glitsch ULTRA-FRAC trays and compared predictions with experimental efficiency data. Comparison of the measured and O’Connell predicted efficiencies are shown in Table 2.5.

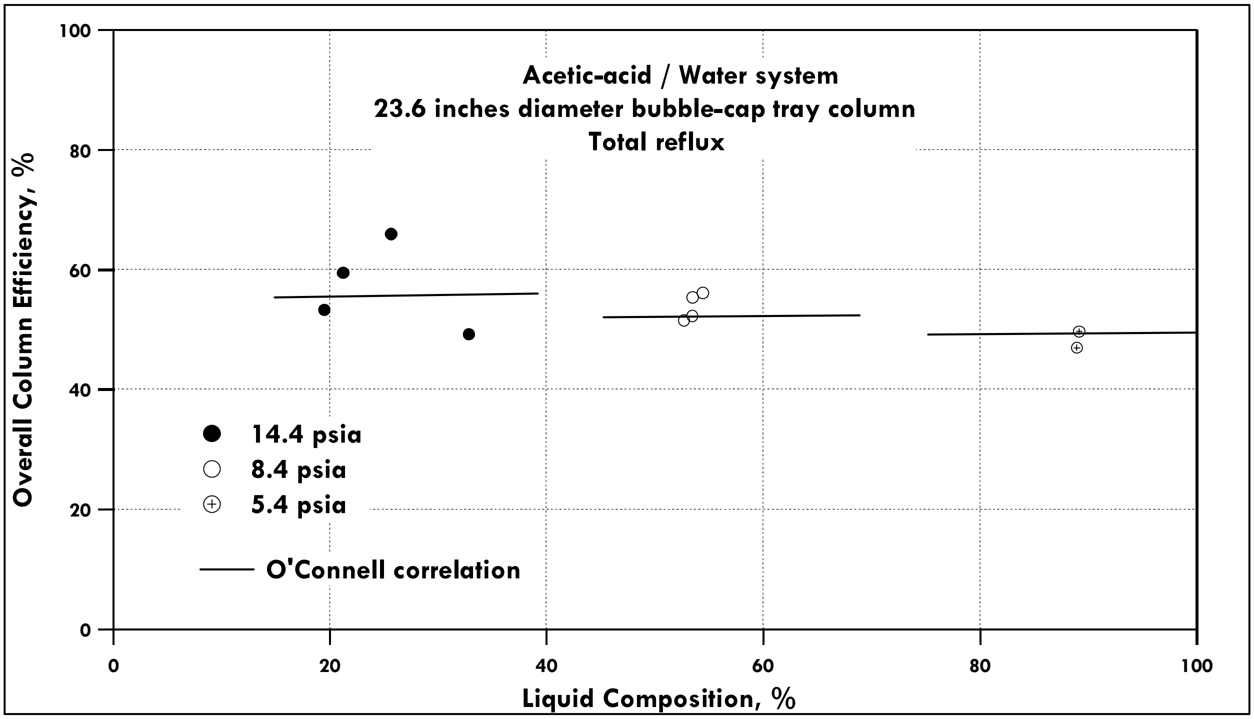

Figure 2.9: Agreement of the experimental efficiencies of acetic-acid / water system with O'Connell correlation (Reproduced from: Marek, J., and Novosad, Z. (1956). Plate efficiencies of an industrial column. *Collection of Czechoslovak Chemical Communications, 21*(4), 795-801.)

Table 2.5: Comparison of observed efficiencies with efficiencies predicted using the O'Connell correlation for super-high capacity trays. (Adapted from: Wilkinson, P., Vos, E., Konijn, G., Kooijman, H., Mosca, G., and Tonon, L., (2007). "Distillation trays that operate beyond the limits of gravity by using centrifugal separation." *Chemical Engineering Research and Design* ***85*****(**A1): 130-135.)

Tray	Application	Efficiency	
		Actual (%)	O'Connell Predicted (%)
ConSep Tray (De Villiers, et al., 2004)	IC4/NC4 at 160 psia	89	82
COFLO tray (Fair, et al., 1999)	C6/C7 at 4.8 psia	60	55
	C6/C7 at 15.4 psia	70	58
	C6/C7 at 24 psia	75	63
ULTRA-FRAC tray	Deethanizer	85	82
	Depropanizer	78-82*	79
	Debutanizer	75-85*	69

*ranges of efficiencies observed by making several measurement on multiple towers

2.7 Limitations and implications of the O'Connell correlation

The original O'Connell correlation has several limitations, as reported in detail in the Williams, et al. (1950) article. This correlation is valid for (Williams, et al., 1950):

- similar tray designs on which the correlation was based,
- liquid flow path length less than 5 feet,
- columns operating near minimum reflux, and
- columns designed with reliable vapor-liquid equilibrium data.

2.8 Analysis of needs and gaps

The O’Connell correlation became the “standard of industry” due to the combination of simplicity, reliability, and accuracy along with weakness of theoretical models for prediction of efficiency (Kister, 1992). However, this correlation was developed using only bubble-cap and sieve tray data. No valve tray data has been used. An effort to update this correlation using valve tray data was made by Seader and Henley (1998). However, they used only one valve tray design (Glitsch Ballast tray). The valve tray data used showed 10 to 20 % higher efficiency than those predicted by their updated O’Connell correlation.

A separate correlation needs to be developed using only valve tray data. Thus, an O’Connell type correlation for valve tray columns is developed in the next chapter.

CHAPTER III

THE OSU-FRI VALVE TRAY EFFICIENCY CORRELATION

In this chapter, a new correlation of the same form as the O'Connell correlation is presented to predict E_O of valve tray columns.

The FRI valve tray data along with the details of the data screening procedure are reported in Section 3.1. The estimation of the values of the physical properties (α and μ_L) required as input for the correlation is documented in Section 3.2. An augmented O'Connell correlation and the OSU-FRI valve tray type valve tray efficiency correlation are presented in Sections 3.3 and 3.4, respectively.

3.1 Valve tray data selection

Although distillation is the one of the oldest methods of separation, distillation efficiency data are not available in abundance in the open literature. The preferred method for predicting overall column efficiencies is the use of field performance data from a similar column or system. However, scarcity of such data in the open literature limits this choice (Klemola and Ilme, 1996). Numerous researchers have attempted to use laboratory columns to estimate efficiencies of commercial columns. Fair, et al. (1983)

stated that reliable point efficiencies could be obtained using Oldershaw columns and these efficiencies can be applied to commercial columns. However, these point efficiency data must be corrected for vapor and liquid mixing effects to obtain overall column efficiency (Chan and Fair, 1984). Furthermore, Kister (2008) reported that laboratory-scale distillation column efficiency data are not applicable for prediction of efficiencies of commercial columns. Therefore, commercial column data are more desirable than laboratory column data to predict the efficiency of commercial columns.

Fractionation Research, Inc. (FRI) was formed in 1952 and its contribution to the distillation field is impressive. FRI has 54 years worth of data (Resetarits and King, 2008). The FRI data is an excellent source of commercial-scale test data as they are obtained from a closely monitored research facility using systems with well-known physical properties (Kister, 2008). All these data are the property of the companies sponsoring the work. A part of FRI data have been released to the public (Resetarits and King, 2008). The data released to the public were obtained during tests of bubble-cap, sieve, and valve trays as well as structured and random packing. The valve tray data have been used for this work. The FRI data that have been released to the public and used in this study are referred to as "FRI public data" throughout this thesis.

3.1.1 Criteria used for FRI valve tray data screening

Data screening procedure used is summarized in Figure 3.1. The numbers in parentheses represent the number of data points. Each FRI test run produces one data point.

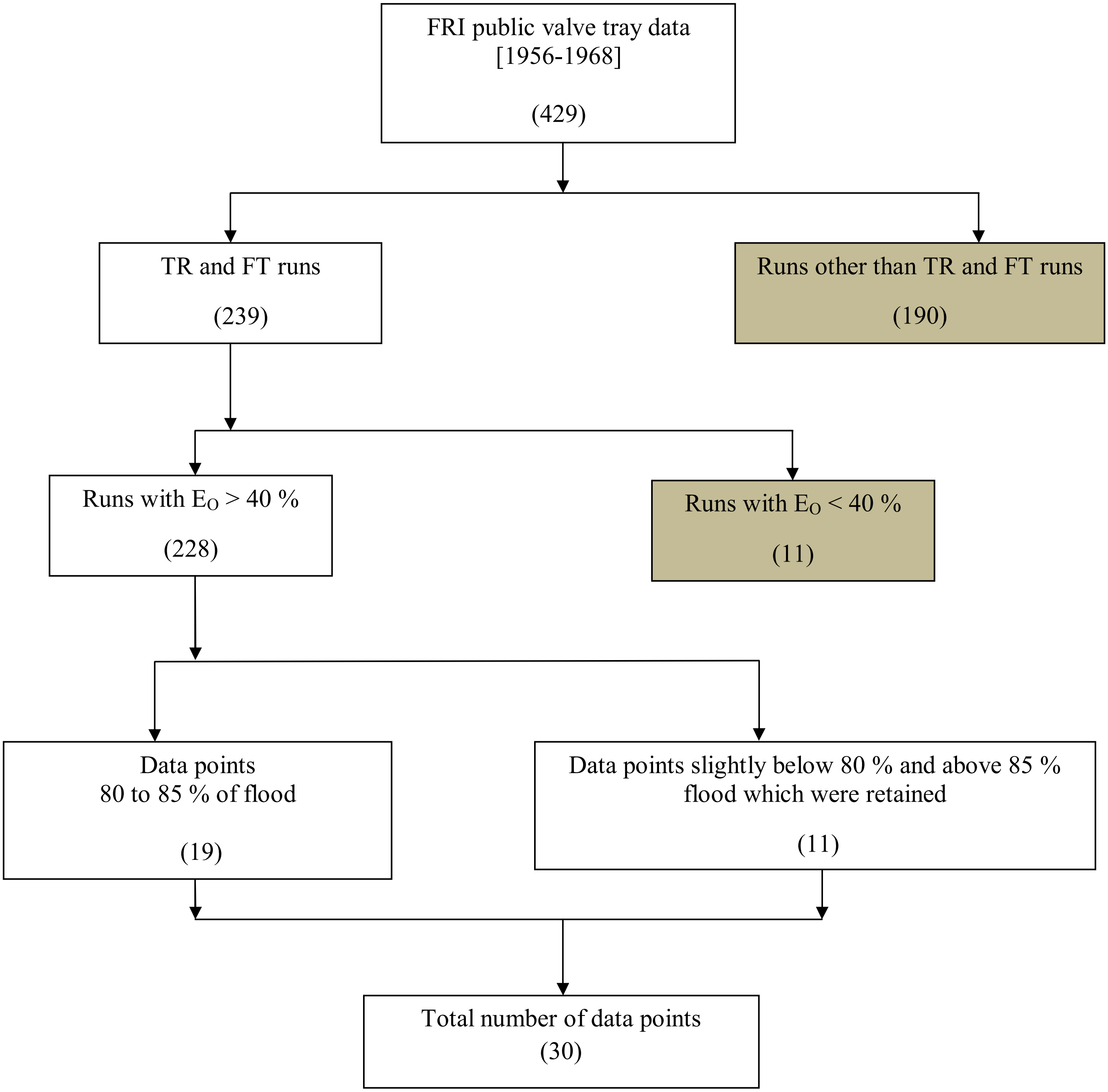

Figure 3.1: Data selection procedure

E_O- Overall column efficiency (%)
TR- Total reflux
FT- Flooding with total reflux

The screening steps shown in Figure 3.1 are explained as follows:

1. A total of 429 FRI valve tray data points were obtained from FRI Topical Reports 10, 15, 19, 27, 31, 40, 41, and 44, which were released to the public.
2. Measured E_O values were available only for total reflux (TR) and flooding with total reflux (FT) data points. Of the 429 data points, 239 such data points were available. The remaining 190 data points with no measured efficiencies could not be considered in this study.
3. The first criterion used to screen the remaining data was that E_O be greater than or equal to 40 %. Of the 239 data points, 228 such data points met this criterion and were retained.
4. The next criterion was to select data points between 80 and 85 % of flood. Only 19 of the 228 retained data points fell within this range. The basis for this criterion can be explained as follows:

    > The original O'Connell correlation was presented as a tool for design engineers to estimate E_O (O'Connell in Williams, et al., 1950). As cited by Kister (1992), it was recommended (Van Winkle, 1967; Chase, 1967; Treybal, 1980) to design a column at 80 to 85 % of flood. Therefore, the new correlation should be based on data collected over this same range.

5. Data points just below 80 % and just above 85 % of flood were re-evaluated from the 209 data points discarded in step 4. Eleven such data points were ultimate retained. Nine of the eleven points are between 78 and 80 % of flood. The other two correspond to 90 and 90.4 % of flood. Inspection of the full dataset for these

two points indicated that the runs did not exhibit any significant loss of efficiency due to entrainment.

Using the above methodology, a total of 30 data points from the original 429 were selected as the basis for a new O'Connell type correlation for valve tray columns. A complete listing of the information collected for each of the 30 data points is provided in Table 3.1.

Inspection of Table 3.1 reveals that 11 valve tray designs are represented in the dataset used to generate the new correlation. The details of the different tray designs are documented in Appendices B, C, and D. Both fixed and moving valve designs are included from four vendors (Glitsch, Koch, Nutter, and Zink).

Appendix B provides photographs of the valve tray designs used in this study. Appendix C defines the differences between trays in a common product line. Appendix D provides the available tray geometry information for the tray designs considered in this study.

The 30 data points cover four test systems as shown in Table 3.1. The four systems are ortho-xylene / para-xylene (O P Xylene), isobutane / normal-butane (IC4/NC4), cyclohexane / n-heptane (C6/C7), and n-octanol / n-decanol (C8/C10).

Table 3.1: Selected FRI (Moving and Fixed) valve tray data (48 inch diameter column$^{\$\$}$, Total reflux operation)

RN	System	TP, psia	BP, psia	Col. Avg. Pres., psia	TT, F	BT, F	Col. Avg. Temp., F	Liquid MVC Comp.*, fraction	Tray Design	TPR	N	E_o, %	% Flood
9414	C8/C10	0.2	0.5	0.3	213.0	266.5	239.8	0.57	Koch Type T Flexitray	TR 44	6	58.7	80.5
9344	OPX	0.4	0.7	0.5	120.7	137.8	129.3	0.64	Koch Type T Flexitray	TR 44	6	74.6	80.1
9375	OPX	0.9	1.3	1.1	148.0	160.3	154.2	0.63	Koch Type T Flexitray	TR 44	6	77.2	80.3
9391	OPX	2.0	2.3	2.1	176.7	184.7	180.7	0.63	Koch Type T Flexitray	TR 44	6	95.0	80.1
1182	C6/C7	4.3	5.2	4.7	114.2	137.1	125.7	0.70	Koch Flexitray	TR 10	10	71.5	79.9
2553	C6/C7	4.9	5.8	5.4	118.4	136.5	127.5	0.77	Glitsch Ballast Tray	TR 15	9	60.0	79.1
3521	C6/C7	4.7	5.8	5.2	117.7	138.5	128.1	0.74	Glitsch Ballast Tray V1	TR 19	9	69.5	80.3
5542	C6/C7	4.7	5.5	5.1	118.3	141.0	129.7	0.62	Nutter Float valve D437	TR 27	10	68.6	79.5
5585	C6/C7	4.7	5.7	5.2	118.0	137.0	127.5	0.70	Nutter Float valve P437	TR 27	10	52.0	82.1
1166	C6/C7	23.5	24.6	24.0	208.1	228.8	218.5	0.63	Koch Flexitray	TR 10	10	80.0	80.3
2535	C6/C7	23.5	24.5	24.0	208.4	226.8	217.6	0.66	Glitsch Ballast Tray	TR 15	9	82.6	80.2
3502	C6/C7	23.6	24.9	24.3	207.2	225.9	216.6	0.74	Glitsch Ballast Tray V1	TR 19	9	82.0	80.5
3530	C6/C7	23.8	24.9	24.4	210.2	230.6	220.4	0.67	Glitsch Ballast Tray V1	TR 19	9	87.1	80.1
5525	C6/C7	23.7	24.6	24.2	208.1	223.7	215.9	0.70	Nutter Float valve D437	TR 27	10	82.2	79.9
5575	C6/C7	23.7	24.7	24.2	209.5	225.9	217.7	0.67	Nutter Float valve P437	TR 27	10	78.6	80.2
7273	C6/C7	23.5	24.7	24.1	209.8	228.2	219.0	0.70	Nutter Type B Float valve(same as B with larger slot area)	TR 31	10	82.0	80.0
9211	C6/C7	23.5	24.0	23.8	209.0	221.5	215.3	0.71	*#Zink Bi-Mix, Tray No. 1	TR 40	7	67.9	81.3
9241	C6/C7	23.5	24.1	23.8	208.7	221.6	215.2	0.75	*#Zink Bi-Mix, Tray No. 2	TR 40	7	71.7	81.1
9255	C6/C7	23.5	24.8	24.2	207.4	227.6	217.5	0.69	*#Glitsch V-O Ballast Tray	TR 41	10	82.4	81.3
2560	C6/C7	49.9	51.1	50.5	264.0	282.6	273.3	0.70	Glitsch Ballast Tray	TR 15	9	85.0	89.9
1147	IC4/NC4	164.3	165.5	164.9	168.2	181.0	174.6	0.44	Koch Flexitray	TR 10	10	112.0	90.4

Continued on next page

(Table 3.1 Continued)

RN	System	TP, psia	BP, psia	Col. Avg. Pres., psia	TT, F	BT, F	Col. Avg. Temp., F	Liquid MVC Comp.*, fraction	Tray Design	TPR	N	E_o, %	% Flood
2573	IC4/NC4	164.5	165.3	164.9	166.6	179.4	173.0	0.44	Glitsch Ballast Tray	TR 15	9	104.0	80.3
3555	IC4/NC4	164.4	165.2	164.8	166.2	178.8	172.5	0.54	Glitsch Ballast Tray V1	TR 19	9	107.0	78.6
3537	IC4/NC4	164.6	165.4	165.0	154.0	176.3	165.2	0.55	Glitsch Ballast Tray V1	TR 19	9	113.0	78.9
5552	IC4/NC4	165.0	165.8	165.4	164.0	178.5	171.3	0.48	Nutter Float valve D437	TR 27	10	91.1	80.2
5563	IC4/NC4	165.0	165.8	165.4	167.0	179.4	173.2	0.47	Nutter Float valve P437	TR 27	10	103.0	81.3
7247	IC4/NC4	165.0	165.9	165.4	162.1	175.7	168.9	0.60	Nutter Type B Float valve	TR 31	10	104.0	80.7
7259	IC4/NC4	165.0	165.9	165.4	162.1	175.8	169.0	0.62	Nutter Type B Float valve(same as B with larger slot area)	TR 31	10	109.0	79.4
9233	IC4/NC4	164.5	164.5	164.5	167.0	176.1	171.6	0.53	*#Zink Bi-Mix, Tray No. 1	TR 40	7	95.3	79.4
9272	IC4/NC4	164.7	165.5	165.1	167.6	180.2	173.9	0.44	*#Glitsch V-O Ballast Tray	TR 41	10	107.0	80.4

*This is the composition (Mol fraction) of the tray having approximately average temperature and average pressure.

*#These are fixed valve trays.

More volatile component for C8/C10, C6/C7, OPX, and IC4/NC4 are C8, C6, PX, and IC4, respectively

$^{\$\$}$Diameter in TR 10 was known through personal communication with FRI.

Avg.	Average	BP	Bottom pressure (psia)
BT	Bottom temperature (F)	Comp.	Composition
C6/C7	cyclohexane / n-heptane	C8/C10	n-octanol/n-decanol
E_O	Overall column efficiency (%)	IC4/NC4	iso-butane / normal-butane
N	Number of trays in column	OPX	ortho-xylene / para-xylene
Pres.	Pressure	RN	Run number
Temp.	Temperature	TP	Top Pressure, (psia)
TPR	Topical report	TT	Top temperature (F)

3.2 Estimation of physical properties

The O'Connell correlation predicts the efficiency using only the relative volatility (α) of the key components and liquid viscosity (μ_L) at the average temperature and pressure of the column. Therefore, α and μ_L need to be estimated.

Estimation of α and μ_L was performed using Aspen Plus V7.0 and the "PURE22" properties databank. The "SRK" property method was used without modification of the default option settings. Values obtained for the α and μ_L are tabulated in Table 3.2 for the 30 FRI valve tray data points.

It can be observed from Table 3.2 that the product $\alpha\mu_L$ is approximately constant for a particular system-pressure combination. As seen in Table 3.2, the number of data points for each system-pressure combination is not the same. System-pressure combinations C8/C10 at 0.2 psia, O P Xylene at 0.4 psia, O P Xylene at 0.9 psia, O P Xylene at 2.0 psia, and C6/C7 at 49.9 psia have only one data point, whereas C6/C7 at 4.7 psia, C6/C7 at 23.6 psia, and IC4/NC4 at 164.7 psia have five, ten, and ten data points, respectively.

Using all 30 data points to generate the correlation would result in undue weightage being given to the system-pressure combinations C6/C7 at 4.7 psia, C6/C7 at 23.6 psia, and IC4/NC4 at 164.7 psia. To avoid this undue weightage, an average E_o was calculated for each system-pressure combination with more than one data point. These single runs are referred to as "Averaged run" in this report. Results are presented in Table 3.3, 3.4, and 3.5.

Table 3.2: Estimated α and μ_L for the selected FRI valve tray data (48 inch diameter column$^{\$\$}$, Total reflux operation)

RN	System	TP, psia	BP, psia	Col. Avg. Pres., psia	TT, F	BT, F	Col. Avg. Temp., F	Liquid MVC Comp.*, fraction	Tray Design	TPR	N	E_o, %	% Flood	α	μ_L, cP	$\alpha\mu_L$, cP
9414	C8/C10	0.2	0.5	0.3	213.0	266.5	239.8	0.57	Koch Type T Flexitray	TR 44	6	58.7	80.5	3.63	0.87	3.14
9344	OPX	0.4	0.7	0.5	120.7	137.8	129.3	0.64	Koch Type T Flexitray	TR 44	6	74.6	80.1	1.24	0.47	0.58
9375	OPX	0.9	1.3	1.1	148.0	160.3	154.2	0.63	Koch Type T Flexitray	TR 44	6	77.2	80.3	1.23	0.41	0.50
9391	OPX	2.0	2.3	2.1	176.7	184.7	180.7	0.63	Koch Type T Flexitray	TR 44	6	95.0	80.1	1.21	0.36	0.44
1182	C6/C7	4.3	5.2	4.7	114.2	137.1	125.7	0.70	Koch Flexitray	TR 10	10	71.5	79.9	1.91	0.45	0.86
2553	C6/C7	4.9	5.8	5.4	118.4	136.5	127.5	0.77	Glitsch Ballast Tray	TR 15	9	60.0	79.1	1.77	0.5	0.86
3521	C6/C7	4.7	5.8	5.2	117.7	138.5	128.1	0.74	Glitsch Ballast Tray V1	TR 19	9	69.5	80.3	1.81	0.47	0.85
5542	C6/C7	4.7	5.5	5.1	118.3	141.0	129.7	0.62	Nutter Float valve D437	TR 27	10	68.6	79.5	1.88	0.43	0.81
5585	C6/C7	4.7	5.7	5.2	118.0	137.0	127.5	0.70	Nutter Float valve P437	TR 27	10	52.0	82.1	1.83	0.47	0.86
1166	C6/C7	23.5	24.6	24.0	208.1	228.8	218.5	0.63	Koch Flexitray	TR 10	10	80.0	80.3	1.56	0.26	0.41
2535	C6/C7	23.5	24.5	24.0	208.4	226.8	217.6	0.66	Glitsch Ballast Tray	TR 15	9	82.6	80.2	1.55	0.27	0.42
3502	C6/C7	23.6	24.9	24.3	207.2	225.9	216.6	0.74	Glitsch Ballast Tray V1	TR 19	9	82.0	80.5	1.53	0.28	0.43
3530	C6/C7	23.8	24.9	24.4	210.2	230.6	220.4	0.67	Glitsch Ballast Tray V1	TR 19	9	87.1	80.1	1.57	0.25	0.40
5525	C6/C7	23.7	24.6	24.2	208.1	223.7	215.9	0.70	Nutter Float valve D437	TR 27	10	82.2	79.9	1.52	0.28	0.43
5575	C6/C7	23.7	24.7	24.2	209.5	225.9	217.7	0.67	Nutter Float valve P437	TR 27	10	78.6	80.2	1.55	0.27	0.42
7273	C6/C7	23.5	24.7	24.1	209.8	228.2	219.0	0.70	Nutter Type B Float valve(same as B with larger slot area)	TR 31	10	82.0	80.0	1.54	0.28	0.43
9211	C6/C7	23.5	24.0	23.8	209.0	221.5	215.3	0.71	*#Zink Bi-Mix, Tray No. 1	TR 40	7	67.9	81.3	1.53	0.28	0.43
9241	C6/C7	23.5	24.1	23.8	208.7	221.6	215.2	0.75	*#Zink Bi-Mix, Tray No. 2	TR 40	7	71.7	81.1	1.53	0.28	0.43
9255	C6/C7	23.5	24.8	24.2	207.4	227.6	217.5	0.69	*#Glitsch V-O Ballast Tray	TR 41	10	82.4	81.3	1.54	0.27	0.42
2560	C6/C7	49.9	51.1	50.5	264.0	282.6	273.3	0.70	Glitsch Ballast Tray	TR 15	9	85.0	89.9	1.42	0.21	0.30
1147	IC4/NC4	164.3	165.5	164.9	168.2	181.0	174.6	0.44	Koch Flexitray	TR 10	10	112.0	90.4	1.23	0.11	0.14

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(Table 3.2 Continued)

RN	**System**	**TP, psia**	**BP, psia**	**Col. Avg. Pres., psia**	**TT, F**	**BT, F**	**Col. Avg. Temp., F**	**Liquid MVC Comp.*, fraction**	**Tray Design**	**Topical Report**	**N**	**E_o, %**	**% Flood**	**α**	**μ_L, cP**	**$\alpha\mu_L$, cP**
2573	IC4/NC4	164.5	165.3	164.9	166.6	179.4	173.0	0.44	Glitsch Ballast Tray	TR 15	9	104.0	80.3	1.23	0.11	0.14
3555	IC4/NC4	164.4	165.2	164.8	166.2	178.8	172.5	0.54	Glitsch Ballast Tray V1	TR 19	9	107.0	78.6	1.25	0.11	0.14
3537	IC4/NC4	164.6	165.4	165.0	154.0	176.3	165.2	0.55	Glitsch Ballast Tray V1	TR 19	9	113.0	78.9	1.25	0.12	0.15
5552	IC4/NC4	165.0	165.8	165.4	164.0	178.5	171.3	0.48	Nutter Float valve D437	TR 27	10	91.1	80.2	1.24	0.12	0.15
5563	IC4/NC4	165.0	165.8	165.4	167.0	179.4	173.2	0.47	Nutter Float valve P437	TR 27	10	103.0	81.3	1.23	0.11	0.14
7247	IC4/NC4	165.0	165.9	165.4	162.1	175.7	168.9	0.60	Nutter Type B Float valve	TR 31	10	104.0	80.7	1.24	0.11	0.14
7259	IC4/NC4	165.0	165.9	165.4	162.1	175.8	169.0	0.62	Nutter Type B Float valve(same as B with larger slot area)	TR 31	10	109.0	79.4	1.24	0.11	0.14
9233	IC4/NC4	164.5	164.5	164.5	167.0	176.1	171.6	0.53	*#Zink Bi-Mix, Tray No.1	TR 40	7	95.3	79.4	1.24	0.11	0.14
9272	IC4/NC4	164.7	165.5	165.1	167.6	180.2	173.9	0.44	*#Glitsch V-O Ballast Tray	TR 41	10	107.0	80.4	1.24	0.11	0.14

*This is the composition (Mol fraction) of the tray having approximately average temperature and average pressure.

*#These are fixed valve trays.

More volatile component for C8/C10, C6/C7, OPX, and IC4/NC4 are C8, C6, PX, and IC4, respectively

[$$]Diameter in TR 10 was known through personal communication with FRI.

Avg.	Average	BP	Bottom pressure (psia)
BT	Bottom temperature (F)	Comp.	Composition
C6/C7	cyclohexane / n-heptane	C8/C10	n-octanol/n-decanol
E_O	Overall column efficiency (%)	IC4/NC4	iso-butane / normal-butane
N	Number of trays in column	OPX	ortho-xylene / para-xylene
Pres.	Pressure	RN	Run number
Temp.	Temperature	TP	Top Pressure, (psia)
TPR	Topical report	TT	Top temperature (F)
α	Relative volatility	μ_L	Liquid viscosity (cP)

Table 3.3: C6/C7 binary system at 4.7 psia pressure with average for all runs (48 inch diameter column[$$], Total reflux)

RN	**System**	**TP, psia**	**BP, psia**	**Col. Avg. Pres., psia**	**TT, F**	**BT, F**	**Col. Avg. Temp., F**	**Liquid MVC Comp.*, fraction**	**Tray Design**	**TPR**	**N**	**E_o, %**	**% Flood**	**α**	**μ_L, cP**	**$\alpha\mu_L$, cP**
1182	C6/C7	4.3	5.2	4.7	114.2	137.1	125.7	0.70	Koch Flexitray	TR 10	10	71.5	79.9	1.91	0.45	0.86
2553	C6/C7	4.9	5.8	5.4	118.4	136.5	127.5	0.77	Glitsch Ballast Tray	TR 15	9	60.0	79.1	1.77	0.5	0.86
3521	C6/C7	4.7	5.8	5.2	117.7	138.5	128.1	0.74	Glitsch Ballast Tray V1	TR 19	9	69.5	80.3	1.81	0.47	0.85
5542	C6/C7	4.7	5.5	5.1	118.3	141.0	129.7	0.62	Nutter Float valve D437	TR 27	10	68.6	79.5	1.88	0.43	0.81
5585	C6/C7	4.7	5.7	5.2	118.0	137.0	127.5	0.70	Nutter Float valve P437	TR 27	10	52.0	82.1	1.83	0.47	0.86
Averaged Run	**C6/C7**	**4.7**	**5.6**	**5.1**	**117.3**	**138.0**	**127.7**	**0.70**				**64.3**		**1.84**[#]	**0.47**[#]	**0.86**

*This is the composition (Mol fraction) of the tray having approximately average temperature and average pressure.
#These values are estimated using average temperature and pressure of the averaged runs.
More volatile component for C6/C7 is C6.
[$$] Diameter in TR 10 was known through personal communication with FRI.

Avg.	Average	BP	Bottom pressure (psia)
BT	Bottom temperature (F)	Comp.	Composition
C6/C7	cyclohexane / n-heptane	E_O	Overall column efficiency (%)
N	Number of trays in column	Pres.	Pressure
RN	Run number	Temp.	Temperature
TP	Top Pressure, (psia)	TPR	Topical report
TT	Top temperature (F)	α	Relative volatility
μ_L	Liquid viscosity (cP)		

Table 3.4: C6/C7 binary system at 23.6 psia pressure with average for all runs (48 inch diameter column[$$], Total reflux)

RN	System	TP, psia	BP, psia	Col. Avg. Pres., psia	TT, F	BT, F	Col. Avg. Temp., F	Liquid MVC Comp.*, fraction	Tray Design	TPR	N	E_o, %	% Flood	α	μ_L, cP	$\alpha\mu_L$, cP
1166	C6/C7	23.5	24.6	24.0	208.1	228.8	218.5	0.63	Koch Flexitray	TR 10	10	80.0	80.3	1.56	0.26	0.41
2535	C6/C7	23.5	24.5	24.0	208.4	226.8	217.6	0.66	Glitsch Ballast Tray	TR 15	9	82.6	80.2	1.55	0.27	0.42
3502	C6/C7	23.6	24.9	24.3	207.2	225.9	216.6	0.74	Glitsch Ballast Tray V1	TR 19	9	82.0	80.5	1.53	0.28	0.43
3530	C6/C7	23.8	24.9	24.4	210.2	230.6	220.4	0.67	Glitsch Ballast Tray V1	TR 19	9	87.1	80.1	1.57	0.25	0.40
5525	C6/C7	23.7	24.6	24.2	208.1	223.7	215.9	0.70	Nutter Float valve D437	TR 27	10	82.2	79.9	1.52	0.28	0.43
5575	C6/C7	23.7	24.7	24.2	209.5	225.9	217.7	0.67	Nutter Float valve P437	TR 27	10	78.6	80.2	1.55	0.27	0.42
7273	C6/C7	23.5	24.7	24.1	209.8	228.2	219.0	0.70	Nutter Type B Float valve(same as B with larger slot area)	TR 31	10	82.0	80.0	1.54	0.28	0.43
9211	C6/C7	23.5	24.0	23.8	209.0	221.5	215.3	0.71	*#Zink Bi-Mix, Tray No. 1	TR 40	7	67.9	81.3	1.53	0.28	0.43
9241	C6/C7	23.5	24.1	23.8	208.7	221.6	215.2	0.75	*#Zink Bi-Mix, Tray No. 2	TR 40	7	71.7	81.1	1.53	0.28	0.43
9255	C6/C7	23.5	24.8	24.2	207.4	227.6	217.5	0.69	*#Glitsch V-O Ballast Tray	TR 41	10	82.4	81.3	1.54	0.27	0.42
Averaged Run	**C6/C7**	**23.6**	**24.6**	**24.1**	**208.6**	**226.1**	**217.5**	**0.69**				**79.7**		**1.55**[#]	**0.27**[#]	**0.42**

*This is the composition (Mol fraction) of the tray having approximately average temperature and average pressure.

*#These are fixed valve trays.

#These values are estimated using average temperature and pressure of the averaged runs.

More volatile component for C6/C7 is C6. [$$]Diameter in TR 10 was known through personal communication with FRI.

Avg.	Average	BP	Bottom pressure (psia)
BT	Bottom temperature (F)	Comp.	Composition
C6/C7	cyclohexane / n-heptane	E_O	Overall column efficiency (%)
N	Number of trays in column	Pres.	Pressure
RN	Run number	Temp.	Temperature
TP	Top Pressure, (psia)	TPR	Topical report
TT	Top temperature (F)	α	Relative volatility
μ_L	Liquid viscosity (cP)		

Table 3.5: IC4/NC4 binary system at 164.7 psia pressure with average for all runs (48 inch diameter column$^{\$\$}$, Total reflux)

RN	**System**	**TP, psia**	**BP, psia**	**Col. Avg. Pres., psia**	**TT, F**	**BT, F**	**Col. Avg. Temp., F**	**Liquid MVC Comp.*, fraction**	**Tray Design**	**TPR**	**N**	**E_o, %**	**% Flood**	**α**	**μ_L, cP**	**$\alpha\mu_L$, cP**
1147	IC4/NC4	164.3	165.5	164.9	168.2	181.0	174.6	0.44	Koch Flexitray	TR 10	10	112.0	90.4	1.23	0.11	0.14
2573	IC4/NC4	164.5	165.3	164.9	166.6	179.4	173.0	0.44	Glitsch Ballast Tray	TR 15	9	104.0	80.3	1.23	0.11	0.14
3555	IC4/NC4	164.4	165.2	164.8	166.2	178.8	172.5	0.54	Glitsch Ballast Tray V1	TR 19	9	107.0	78.6	1.25	0.11	0.14
3537	IC4/NC4	164.6	165.4	165.0	154.0	176.3	165.2	0.55	Glitsch Ballast Tray V1	TR 19	9	113.0	78.9	1.25	0.12	0.15
5552	IC4/NC4	165.0	165.8	165.4	164.0	178.5	171.3	0.48	Nutter Float valve D437	TR 27	10	91.1	80.2	1.24	0.12	0.15
5563	IC4/NC4	165.0	165.8	165.4	167.0	179.4	173.2	0.47	Nutter Float valve P437	TR 27	10	103.0	81.3	1.23	0.11	0.14
7247	IC4/NC4	165.0	165.9	165.4	162.1	175.7	168.9	0.60	Nutter Type B Float valve	TR 31	10	104.0	80.7	1.24	0.11	0.14
7259	IC4/NC4	165.0	165.9	165.4	162.1	175.8	169.0	0.62	Nutter Type B Float valve(same as B with larger slot area)	TR 31	10	109.0	79.4	1.24	0.11	0.14
9233	IC4/NC4	164.5	164.5	164.5	167.0	176.1	171.6	0.53	*#Zink Bi-Mix, Tray No. 1	TR 40	7	95.3	79.4	1.24	0.11	0.14
9272	IC4/NC4	164.7	165.5	165.1	167.6	180.2	173.9	0.44	*#Glitsch V-O Ballast Tray	TR 41	10	107.0	80.4	1.24	0.11	0.14
Averaged Run	**IC4/NC4**	**164.7**	**165.5**	**165.1**	**164.5**	**178.1**	**171.3**	**0.51**				**104.5**		**$1.24^{\#}$**	**$0.11^{\#}$**	**0.14**

*This is the composition (Mol fraction) of the tray having approximately average temperature and average pressure.

*#These are fixed valve trays.

#These values are estimated using average temperature and pressure of the averaged runs.

More volatile component for IC4/NC4 is IC4. $^{\$\$}$Diameter in TR 10 was known through personal communication with FRI.

Avg.	Average	BP	Bottom pressure (psia)	BT	Bottom temperature (F)
Comp.	Composition	IC4/NC4	iso-butane / normal-butane	E_O	Overall column efficiency (%)
N	Number of trays in column	Pres.	Pressure	RN	Run number
Temp.	Temperature	TP	Top Pressure, (psia)	TPR	Topical report
TT	Top temperature (F)	α	Relative volatility	μ_L	Liquid viscosity (cP)

After creating an "averaged run" value for those system-pressure combinations with multiple runs, there are a total of eight data points available to generate a new O'Connell type overall efficiency correlation for valve tray columns. These eight data points are listed in Table 3.6.

Table 3.6: Summary of simulated α and μ_L values used in this study

System	Top Press., psia	α	$\mu_{L,}$ cP	$\alpha\mu_{L,}$ cP	Min. Eo, %	Max. Eo, %	Average E_O, %	# of data points
C8/C10	0.2	3.63	0.87	3.14			58.7	1
O P Xylene	0.4	1.24	0.47	0.58			74.6	1
O P Xylene	0.9	1.23	0.41	0.50			77.2	1
O P Xylene	2.0	1.21	0.36	0.44			95.0	1
C6/C7*	4.7	1.84	0.47	0.86	52.0	71.5	64.3	5
C6/C7*	23.6	1.55	0.27	0.42	67.9	87.1	79.7	10
C6/C7	49.9	1.42	0.21	0.30			85.0	1
IC4/NC4*	164.7	1.24	0.11	0.14	91.1	113.0	104.5	10
								Total = 30

*These are averaged data points

3.3 The augmented O'Connell correlation

Before generating a new correlation specifically for valve trays, we elected to generate an "augmented O'Connell correlation" using O'Connell's original data combined with that of Williams, et al. (1950), and the new FRI valve tray data described in the previous section. The augmented O'Connell correlation is presented in this section.

This study included Williams, et al. (1950) data with total reflux for the augmented O'Connell correlation. Fifteen such data points consisting of five systems were used. These data points are presented in Table 3.7.

Williams, et al. total reflux data were added to the O'Connell correlation plot as presented in Figure 3.2. The original O'Connell correlation fits the Williams, et al. data well. The O'Connell correlation gives a mean absolute relative error of 11.3 % for the 15 of Williams, et al. data points.

A parity plot showing the predictions of the original O'Connell correlation (A) compared to the O'Connell and Williams, et al. data is shown in Figure 3.3. It can be seen from Figure 3.3 that predictions using the O'Connell correlation are in agreement with O'Connell and Williams, et al. data. The mean absolute relative error of the O'Connell correlation when applied to O'Connell and Williams, et al. data together is 9.6 %, which can be observed from Table 3.8

Table 3.7: Williams, et al.(1950) total reflux data. (Diameter of column = 10 inch, Bubble-cap tray column, Number of plates = 20)

System	**MVC**	**LVC**	**Act. N**	**Theor. N**	**α**	**μ_L, cP**	**α μ_L, cP**	**Meas. E_O, %**
Benzene-Toluene	Benzene	Toluene	18	10.5	2.46	0.29	0.72	58
Benzene-Toluene	Benzene	Toluene	18	10.5	2.46	0.29	0.72	58
Benzene-Toluene	Benzene	Toluene	14	7.8	2.49	0.28	0.70	56
Benzene-Toluene	Benzene	Toluene	14	6.5	2.49	0.28	0.70	46
Benzene-Carbon Tetrachloride	Carbon Tetrachloride	Benzene	18	9.5	1.10	0.39	0.43	53
Benzene-Carbon Tetrachloride	Carbon Tetrachloride	Benzene	18	8.2	1.13	0.36	0.40	46
Benzene-Carbon Tetrachloride	Carbon Tetrachloride	Benzene	18	10.1	1.10	0.39	0.43	56
Benzene-Carbon Tetrachloride	Carbon Tetrachloride	Benzene	18	10.6	1.10	0.39	0.43	59
Benzene-Carbon Tetrachloride	Carbon Tetrachloride	Benzene	18	9.8	1.08	0.40	0.43	55
Benzene-Carbon Tetrachloride	Carbon Tetrachloride	Benzene	18	9.9	1.11	0.39	0.43	55
Acetaldehyde-water	Acetaldehyde	Water	8	2.63	19.0	0.36	6.60	33
Acetaldehyde-water	Acetaldehyde	Water	4	1.50	18.5	0.37	6.40	38
Acetone-water	Acetone	Water	8	4.52	4.7	0.33	1.54	57
Pentane-Toluene	Pentane	Toluene	8	3.5	8.65	0.26	2.25	43.8
Pentane-Toluene	Pentane	Toluene	6	2.93	7.8	0.20	1.56	49

MVC	More volatile component
LVC	Less volatile component
Act. N	Actual number of plates
Theor. N	Theoretical number of plates
α	Relative volatility
μ_L	Liquid viscosity (cP)
Meas. E_O	Measured efficiency (%)

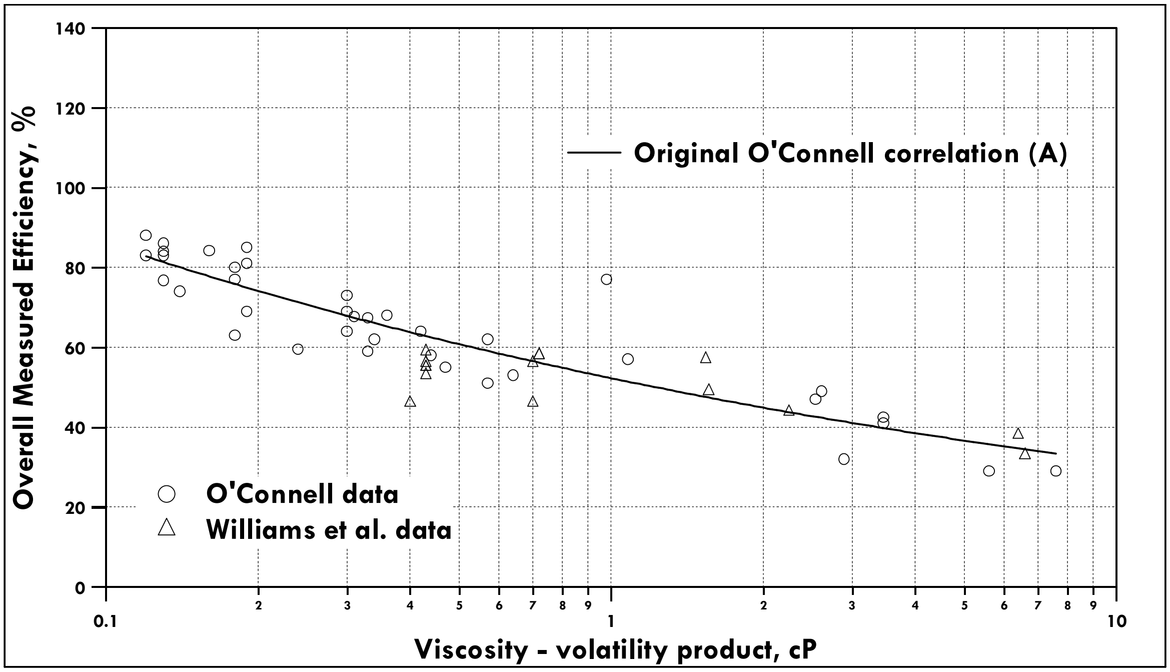

Figure 3.2: Agreement of the O'Connell correlation with Williams, et al.(1950) data.

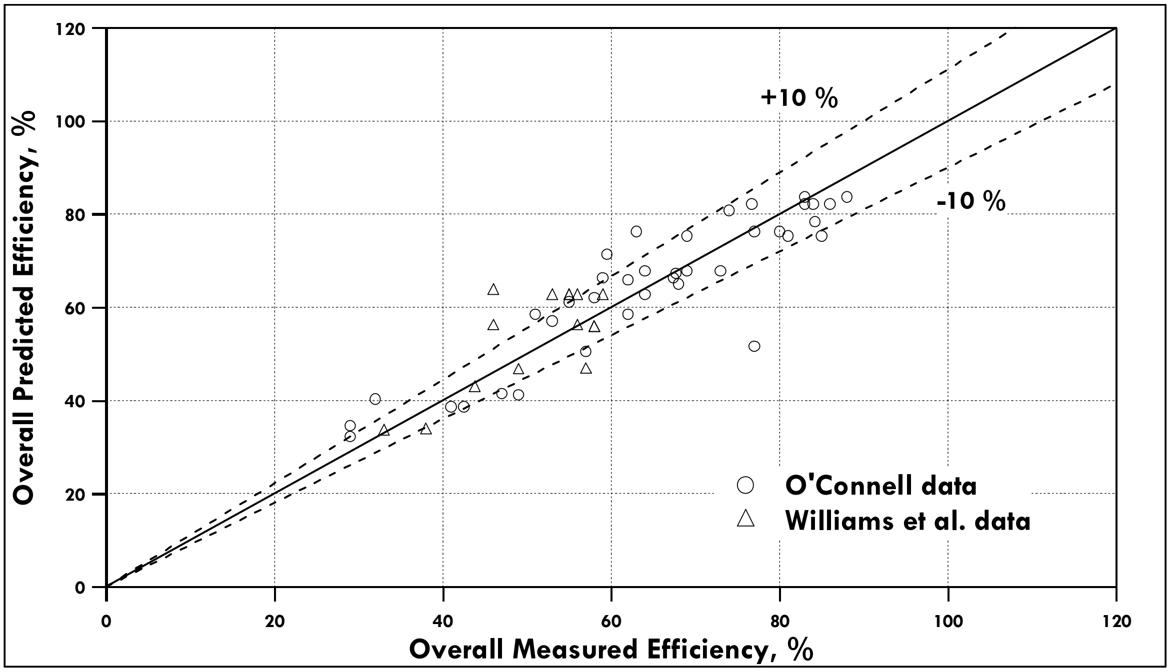

Figure 3.3: Parity plot of the O'Connell correlation predictions compared to O'Connell and Williams, et al. data.

Table 3.8: Comparison of the predicted E_O by OSU’s version of O’Connell correlation with measured E_O for O’Connell and Williams, et al. data

	System	Avg. Press., psia	D, inch	α	μ_L	$\alpha\mu_L$	Measured E_O, %	OSU's ver. of O'Connell correlation	% Error
O' CONNELL DATA	Gasoline Fractionator	215	72	1.36	0.1	0.14	74.0	80.79	9.17
	Gasoline Fractionator	265	72	1.31	0.09	0.12	88.0	83.71	4.88
	Gasoline Fractionator	218	72	1.28	0.1	0.13	86.0	82.18	4.44
	Gasoline Fractionator	223	72	1.28	0.1	0.13	83.0	82.18	0.99
	Naphtha Fractionator Test 1	135	42	1.21	0.15	0.18	63.0	76.25	21.03
	Naphtha Fractionator Test 2	125	42	1.21	0.16	0.19	69.0	75.31	9.14
	Naphtha Fractionator Test 3	125	42	2.2	0.15	0.33	67.4	66.33	1.59
	Naphtha Fractionator Test 4	100	42	2.61	0.22	0.57	51.0	58.49	14.69
	Cracking Unit Stabilizer	347	72	1.76	0.07	0.13	84.0	82.18	2.17
	Cracking Unit Stabilizer	363		1.88	0.1	0.18	77.0	76.25	0.97
	Cracking Unit Stabilizer	357	72	1.77	0.07	0.12	83.0	83.71	0.85
	Cracking Unit Stabilizer	363	72	1.81	0.11	0.19	81.0	75.31	7.03
	Cracking Unit Stabilizer	366	72	1.77	0.09	0.16	84.2	78.35	6.95
	Cracking Unit Stabilizer	350	72	1.83	0.1	0.18	80.0	76.25	4.69
	Poly Plant Stabilizer	362		2.11	0.22	0.47	55.0	61.15	11.18
	Poly Plant Stabilizer	365		2.0	0.22	0.44	58.0	62.08	7.04
	Butane Depropanizer	235	54	2.45	0.15	0.36	68.0	65.01	4.39
	Butane Depropanizer	235		2.45	0.17	0.42	64.0	62.75	1.95
	Debutanizer	117	72	2.0	0.17	0.33	59.0	66.33	12.42
	Deisopentanizer	116		1.90	0.16	0.30	64.0	67.80	5.94
	Deisopentanizer	116		1.96	0.17	0.34	62.0	65.88	6.25
	Deisopentanizer	116		1.98	0.16	0.31	67.7	67.29	0.61
	Deisopentanizer	115		1.87	0.16	0.30	69.0	67.80	1.74
	Deisopentanizer	117	48	1.97	0.15	0.30	73.0	67.80	7.12
	Deisobutanizer	367		1.16	0.11	0.13	76.7	82.18	7.14
	Deisopentanizer	98		1.22	0.19	0.24	59.5	71.37	19.95
	Stabilization of Ethylene Dichloride	21	24	20.51	0.37	7.60	29.0	32.24	11.17
	Stabilization of Ethylene Dichloride*	28	18	16.0	0.35	5.60	29.0	34.58	19.26
	Stabilization of Ethylene Dichloride*	28	18	3.1	0.35	1.08	57.0	50.50	11.41
	Stabilization of Ethylene Dichloride	151	30	1.69	0.11	0.19	85.0	75.31	11.40

Continued on next page

(Table 3.8 Continued)

	System	Avg. Press., psia	D, inch	α	μ_L	$\alpha\mu_L$	Measured E_O, %	OSU's ver. of O'Connell correlation	% Error
O'CONNELL DATA	Alcohol-Water, lab, column	14.7		9.03	0.32	2.89	32.0	40.27	25.84
	Alcohol-Water, lab, column	14.7		7.05	0.36	2.54	47.0	41.48	11.74
	Alcohol-Water, lab, column	14.7		2.34	0.42	0.98	77.0	51.64	32.94
	Alcohol-Water, lab, column	14.7		1.27	0.45	0.57	62.0	58.49	5.65
	Beer Stills (Perforated trays)	16.0	66	10.8	0.32	3.46	41.0	38.63	5.77
	Beer Stills (Perforated trays)	17.0	66	10.8	0.32	3.46	42.5	38.63	9.10
	Alcohol-Water	15.0	36	9.0	0.29	2.61	49.0	41.22	15.87
	Trichloroethylene Toluene and Water, lab. Column	15.0	8	2.12	0.30	0.64	53.0	57.06	7.66
WILLIAMS ET AL. DATA	Benzene-Toluene		10	2.46	0.29	0.715	58.0	55.52	4.27
	Benzene-Toluene		10	2.46	0.29	0.715	58.0	55.52	4.27
	Benzene-Toluene		10	2.49	0.28	0.695	56.0	55.89	0.20
	Benzene-Toluene		10	2.49	0.28	0.695	46.0	55.89	21.49
	Benzene-Carbon Tetrachloride		10	1.10	0.39	0.43	53.0	62.41	17.76
	Benzene-Carbon Tetrachloride		10	1.13	0.36	0.4	46.0	63.46	37.95
	Benzene-Carbon Tetrachloride		10	1.10	0.39	0.43	56.0	62.41	11.45
	Benzene-Carbon Tetrachloride		10	1.10	0.39	0.43	59.0	62.41	5.78
	Benzene-Carbon Tetrachloride		10	1.08	0.4	0.43	55.0	62.41	13.48
	Benzene-Carbon Tetrachloride		10	1.11	0.39	0.43	55.0	62.41	13.48
	Acetaldehyde-water		10	19.0	0.36	6.6	33.0	33.30	0.91
	Acetaldehyde-water		10	18.5	0.37	6.4	38.0	33.54	11.74
	Acetone-water		10	4.7	0.33	1.54	57.0	46.54	18.35
	Pentane-Toluene		10	8.65	0.26	2.25	43.8	42.65	2.62
	Pentane-Toluene		10	7.8	0.20	1.56	49.0	46.40	5.30

Mean absolute error for the O’Connell data = 9.0 %
Mean absolute error for the Williams, et al. data = 11.3 %
Mean absolute error for the O’Connell and Williams, et al. data taken together = 9.6 %

The O'Connell correlation plot with the Williams, et al. (1950) and new FRI valve tray data (shown in Table 3.2) included is shown in Figure 3.4. It is observed from Figure 3.4 that the original O'Connell correlation (A) under predicts the efficiencies of FRI valve tray data and produces large errors for these data. A parity plot showing the comparison of efficiencies predicted using the original O'Connell correlation with measured efficiencies for the O'Connell, Williams, et al., and the selected FRI valve tray data is shown in Figure 3.5. This figure shows that no valve tray data point has less than 10 % error. The correlation gives a mean absolute relative error of 24.0 % for the selected FRI valve tray data. The minimum error was 17.3 % for C6/C7 system at 4.7 psia pressure and the maximum error was 34.5 % for O P Xylene system at 2.0 psia pressure.

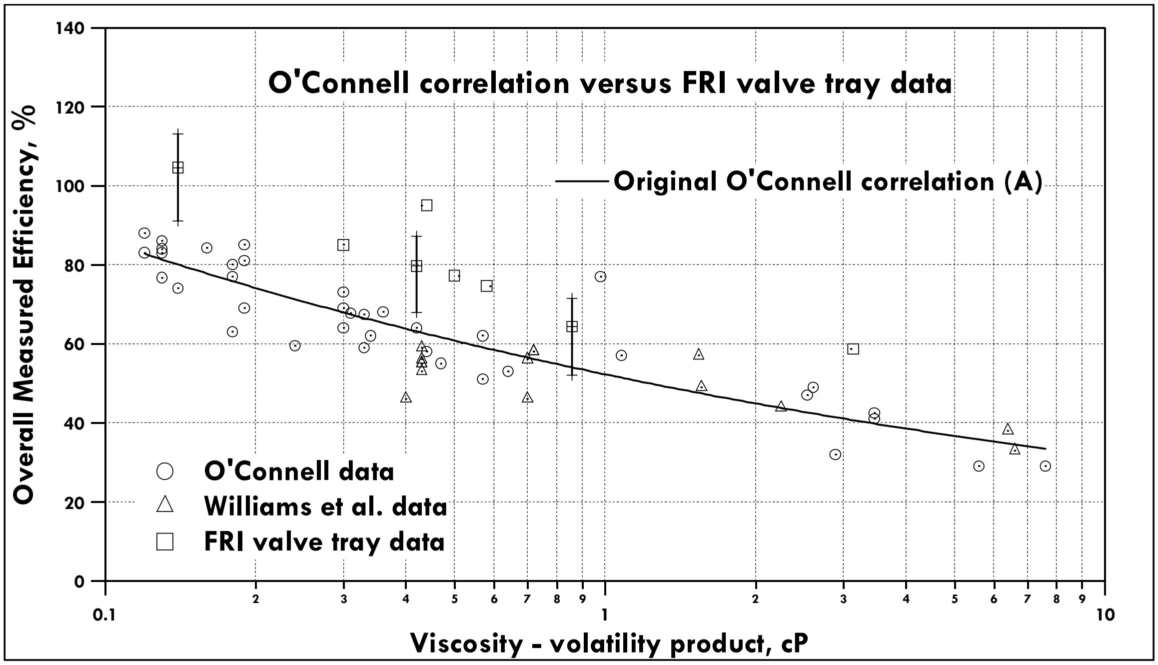

Figure 3.4: The O’Connell correlation versus Williams, et al. and FRI valve tray data. [A line with ‘+’ sign indicate the range of efficiencies. Upper, lower and middle ‘+’ sign indicate higher, lower and average efficiency value respectively]

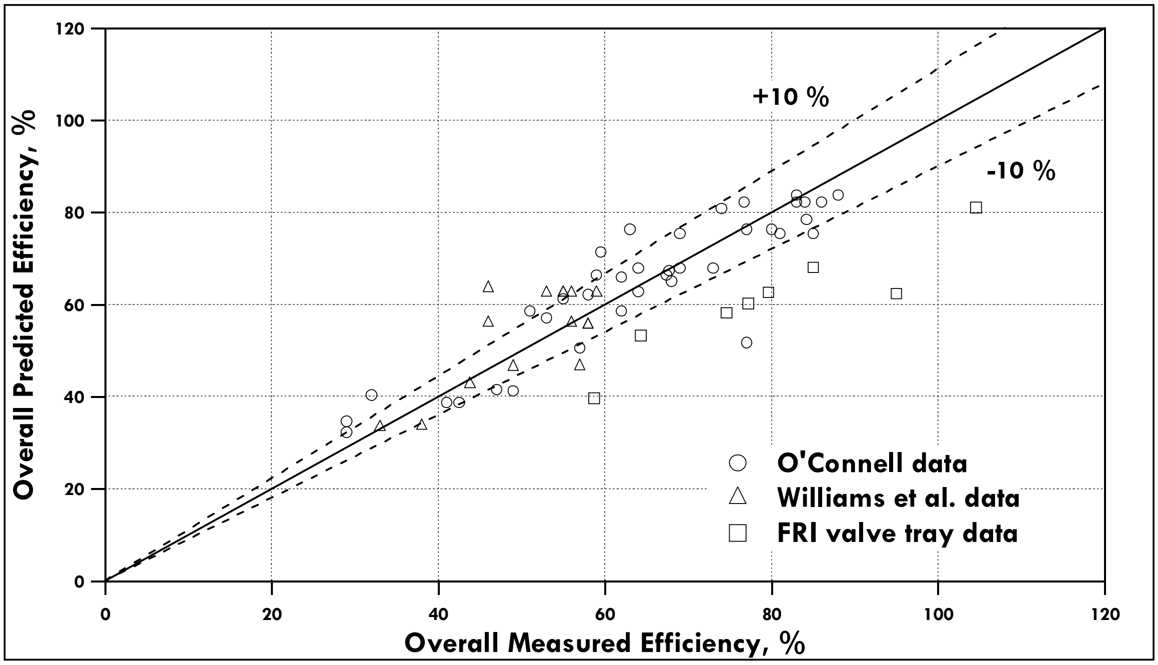

Figure 3.5: A parity plot of the OSU’version of the O’Connell correlation predictions compared to O’Connell (1946), Williams, et al., and FRI valve tray data.

Table 3.9: Comparison of predicted E_O by OSU's version of the O'Connell correlation with measured E_O for FRI valve tray data

System	Avg. Press., psia	α	μ_L	$\alpha\mu_L$	Measured E_O, %	Predicted E_O using original O'Connell correlation	% Error
C8/C10	0.2	3.63	0.87	3.14	58.70	39.50	33.71
O P Xylene	0.4	1.24	0.47	0.58	74.60	58.20	21.99
O P Xylene	0.9	1.23	0.41	0.50	77.20	60.21	22.01
O P Xylene	2.0	1.21	0.36	0.44	95.00	62.23	34.50
C6/C7	4.7	1.84	0.47	0.86	64.32	53.18	17.32
C6/C7	23.6	1.55	0.27	0.42	79.65	62.80	21.15
C6/C7	49.9	1.42	0.21	0.30	85.00	67.91	20.11
IC4/NC4	164.7	1.24	0.11	0.14	104.54	81.32	22.21
						Average absolute error, % =	24.00

As a logical follow-up, we elected to update the original O'Connell correlation using the augmented dataset containing O'Connell (1946), Williams, et al. (1950), and the new FRI valve tray data. The summary of the data used to augment the O'Connell correlation is provided in Table 3.10.

Table 3.10: Summary of the data sets used to augment the O'Connell correlation

O'Connell data	Table 2.1	38 data points
Williams, et al. data	Table 3.7	15 data points
FRI valve tray data	Table 3.2	8 data points

To augment the O'Connell correlation, the values of $\alpha\mu_L$ from the augmented data set were plotted on the x-axis (semi-log scale) and the corresponding E_O values were plotted on the y-axis as shown in Figure 3.6. The augmented O'Connell correlation was obtained using the Trend Line function in Microsoft Excel. The result is shown in Figure

3.6 along with the original O’Connell correlation. The equation form of the augmented O’Connell correlation is:

$$E_o = 0.532(\alpha\mu_L)^{-0.22} \tag{3.1}$$

It is observed from Figure 3.6 that this (Eq. 3.1) correlation appears to fit well for O’Connell (1946) and Williams, et al. data (1950), but not for the FRI valve tray data. The augmented O’Connell correlation predicts higher efficiencies than the original O’Connell correlation by only 1 to 2 %.

A parity plot showing the comparison of efficiencies predicted using the augmented O’Connell correlation with measured efficiencies for the O’Connell (1946), Williams, et al.(1950), and the new FRI valve tray data is shown in Figure 3.7. This figure shows that the augmented O’Connell correlation under-predicts efficiencies of FRI valve tray data. The mean absolute relative error of the augmented correlation is 11.5 % for the augmented data set, 10.0 % for the O’Connell and Williams, et al. data taken together, and 21.9 % for the new FRI valve tray data. The details of the predictions of the augmented O’Connell correlation along with % error are documented in Table 3.11, which shows that no valve tray data point has an error less than 14.0 %.

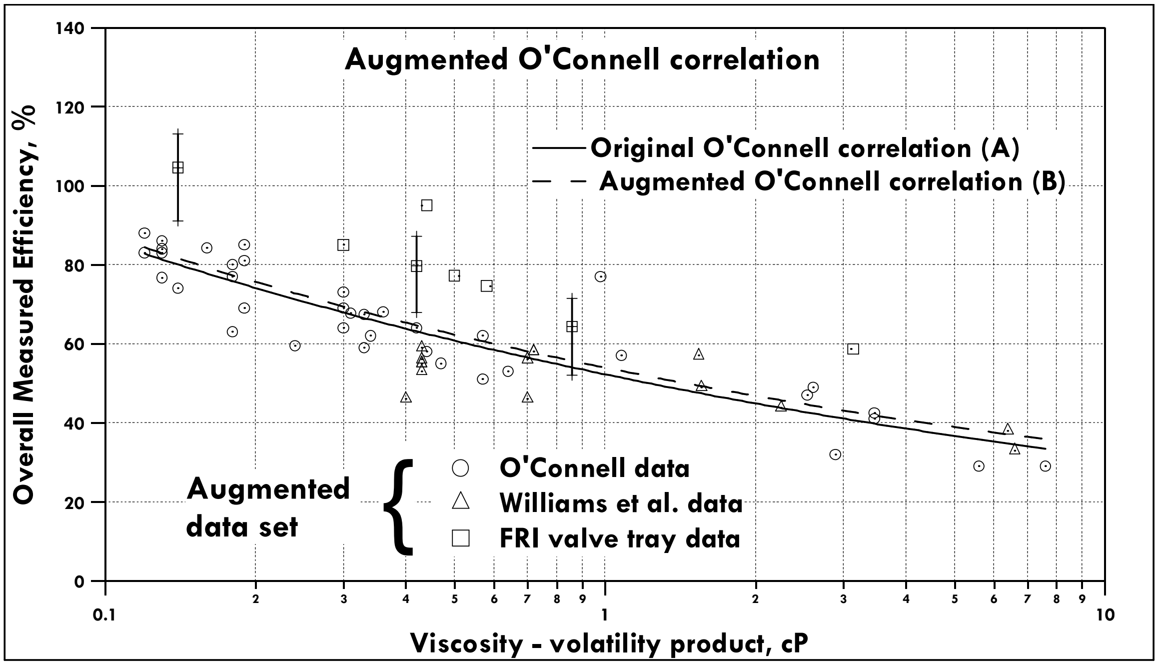

Figure 3.6: The augmented O'Connell correlation. [A line with '+' sign indicate the range of efficiencies. Upper, lower and middle '+' sign indicate higher, lower and average efficiency value respectively]

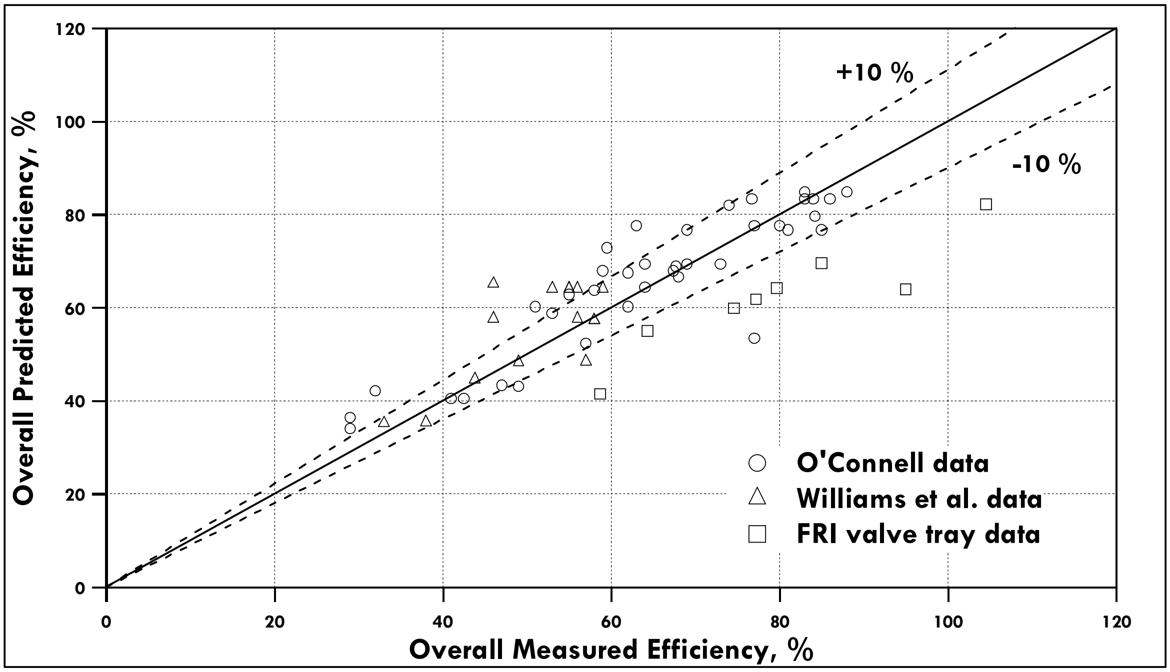

Figure 3.7: A parity plot of the Augmented O'Connell correlation predictions compared to O'Connell, Williams, et al., and FRI data.

Table 3.11: Comparison of predicted E_O by the augmented O'Connell correlation with measured E_O for the augmented data set

	System	Avg. Press., psia	D, inch	α	μ_L	$\alpha\mu_L$	Measured E_O, %	Aug. O'Conn. Corr.	% Error
O'CONNELL DATA	Gasoline Fractionator	215	72	1.36	0.1	0.14	74.0	81.99	10.80
	Gasoline Fractionator	265	72	1.31	0.09	0.12	88.0	84.82	3.62
	Gasoline Fractionator	218	72	1.28	0.1	0.13	86.0	83.34	3.10
	Gasoline Fractionator	223	72	1.28	0.1	0.13	83.0	83.34	0.41
	Naphtha Fractionator Test 1	135	42	1.21	0.15	0.18	63.0	77.58	23.14
	Naphtha Fractionator Test 2	125	42	1.21	0.16	0.19	69.0	76.66	11.11
	Naphtha Fractionator Test 3	125	42	2.2	0.15	0.33	67.4	67.90	0.73
	Naphtha Fractionator Test 4	100	42	2.61	0.22	0.57	51.0	60.20	18.05
	Cracking Unit Stabilizer	347	72	1.76	0.07	0.13	84.0	83.34	0.79
	Cracking Unit Stabilizer	363		1.88	0.1	0.18	77.0	77.58	0.75
	Cracking Unit Stabilizer	357	72	1.77	0.07	0.12	83.0	84.82	2.19
	Cracking Unit Stabilizer	363	72	1.81	0.11	0.19	81.0	76.66	5.35
	Cracking Unit Stabilizer	366	72	1.77	0.09	0.16	84.2	79.62	5.44
	Cracking Unit Stabilizer	350	72	1.83	0.1	0.18	80.0	77.58	3.02
	Poly Plant Stabilizer	362		2.11	0.22	0.47	55.0	62.81	14.21
	Poly Plant Stabilizer	365		2.0	0.22	0.44	58.0	63.73	9.88
	Butane Depropanizer	235	54	2.45	0.15	0.36	68.0	66.61	2.05
	Butane Depropanizer	235		2.45	0.17	0.42	64.0	64.39	0.60
	Debutanizer	117	72	2.0	0.17	0.33	59.0	67.90	15.08
	Deisopentanizer	116		1.9	0.16	0.30	64.0	69.33	8.33
	Deisopentanizer	116		1.96	0.17	0.34	62.0	67.45	8.79
	Deisopentanizer	116		1.98	0.16	0.31	67.7	68.84	1.68
	Deisopentanizer	115		1.87	0.16	0.30	69.0	69.33	0.48
	Deisopentanizer	117	48	1.97	0.15	0.30	73.0	69.33	5.02
	Deisobutanizer	367		1.16	0.11	0.13	76.7	83.34	8.65
	Deisopentanizer	98		1.22	0.19	0.24	59.5	72.82	22.39
	Stabilization of Ethylene Dichloride	21	24	20.51	0.37	7.60	29.0	34.05	17.42
	Stabilization of Ethylene Dichloride*	28	18	16.0	0.35	5.60	29.0	36.42	25.58
	Stabilization of Ethylene Dichloride*	28	18	3.1	0.35	1.08	57.0	52.31	8.23

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(Table 3.11 Continued)

	System	**Avg. Press., psia**	**D, inch**	**α**	**μ_L**	**$\alpha\mu_L$**	**Measured E_O, %**	**Aug. O'Conn. Corr.**	**% Error**
O'CONNELL Data	Stabilization of Ethylene Dichloride	151	30	1.69	0.11	0.19	85.0	76.66	9.81
	Alcohol-Water, lab, column	14.7		9.03	0.32	2.89	32.0	42.12	31.63
	Alcohol-Water, lab, column	14.7		7.05	0.36	2.54	47.0	43.34	7.80
	Alcohol-Water, lab, column	14.7		2.34	0.42	0.98	77.0	53.44	30.60
	Alcohol-Water, lab, column	14.7		1.27	0.45	0.57	62.0	60.20	2.90
	Beer Stills (Perforated trays)	16.0	66	10.8	0.32	3.46	41.0	40.49	1.25
	Beer Stills (Perforated trays)	17.0	66	10.8	0.32	3.46	42.5	40.49	4.74
	Alcohol-Water	15.0	36	9.0	0.29	2.61	49.0	43.08	12.09
	Trichloroethylene Toluene and Water, lab. Column	15.0	8	2.12	0.3	0.64	53.0	58.79	10.92
WILLIAMS et. al. Data	Benzene-Toluene		10	2.46	0.29	0.715	58.0	57.27	1.25
	Benzene-Toluene		10	2.46	0.29	0.715	58.0	57.27	1.25
	Benzene-Toluene		10	2.49	0.28	0.695	56.0	57.63	2.92
	Benzene-Toluene		10	2.49	0.28	0.695	46.0	57.63	25.29
	Benzene-Carbon Tetrachloride		10	1.1	0.39	0.43	53.0	64.05	20.86
	Benzene-Carbon Tetrachloride		10	1.13	0.36	0.4	46.0	65.08	41.48
	Benzene-Carbon Tetrachloride		10	1.1	0.39	0.43	56.0	64.05	14.38
	Benzene-Carbon Tetrachloride		10	1.1	0.39	0.43	59.0	64.05	8.57
	Benzene-Carbon Tetrachloride		10	1.08	0.4	0.43	55.0	64.05	16.46
	Benzene-Carbon Tetrachloride		10	1.11	0.39	0.43	55.0	64.05	16.46
	Acetaldehyde-water		10	19.0	0.36	6.6	33.0	35.12	6.44
	Acetaldehyde-water		10	18.5	0.37	6.4	38.0	35.36	6.94
	Acetone-water		10	4.7	0.33	1.54	57.0	48.38	15.12
	Pentane-Toluene		10	8.65	0.26	2.25	43.8	44.51	1.61
	Pentane-Toluene		10	7.8	0.2	1.56	49.0	48.24	1.55

Continued on next page

(Table 3.11 Continued)

	System	Avg. Press., psia	D, inch	α	μ_L	$\alpha\mu_L$	Measured E_O, %	Aug. O'Conn. Corr.	% Error
FRI Valve Tray Data	C8/C10	0.2	48	3.63	0.87	3.14	58.70	41.36	29.55
	O P Xylene	0.4	48	1.24	0.47	0.58	74.60	59.91	19.69
	O P Xylene	0.9	48	1.23	0.41	0.50	77.20	61.89	19.83
	O P Xylene	2.0	48	1.21	0.36	0.44	95.00	63.87	32.77
	C6/C7	4.7	48	1.84	0.47	0.86	64.32	54.96	14.56
	C6/C7	23.6	48	1.55	0.27	0.42	79.65	64.44	19.10
	C6/C7	49.9	48	1.42	0.21	0.30	85.00	69.44	18.31
	IC4/NC4	164.7	48	1.24	0.11	0.14	104.54	82.51	21.08

Mean absolute error for the O’Connell data = 9.2 %
Mean absolute error for the Williams, et al. data = 12.0 %
Mean absolute error for the FRI valve tray data = 21.9 %
Mean absolute error for the Augmented data set = 11.5 %

Thus, augmenting the O’Connell correlation does not significantly improve the efficiency predictions for FRI valve tray data.

The reason for the under prediction of FRI valve tray efficiencies by the augmented O’Connell correlation is the empirical nature of the correlation. The correlation is influenced by the number of O’Connell and Williams, et al. data points, as it has been developed using 38 O’Connell, 15 Williams, et al., and only 8 FRI valve tray data points. The O’Connell and Williams, et al. data have lower efficiencies than the FRI valve tray data. The FRI valve tray measured efficiencies are uniformly higher than the O’Connell (1946) and Williams, et al. (1950) data at any given $\alpha\mu_L$ value as can be seen from Figure 3.6.

3.4 The OSU-FRI valve tray correlation

A new correlation has been developed using only FRI valve tray data. This newly developed correlation correlates E_O as a function of the product of liquid viscosity and relative volatility between the key components as was done in the O’Connell correlation. The new correlation was developed by plotting E_O versus $\alpha\mu_L$ of FRI valve tray data (Table 3.6), and using the Trend Line function in the Microsoft Excel.

This new correlation will be referred to as the “OSU-FRI valve tray correlation” throughout this thesis, since it has been developed at Oklahoma State University (OSU) using FRI data.

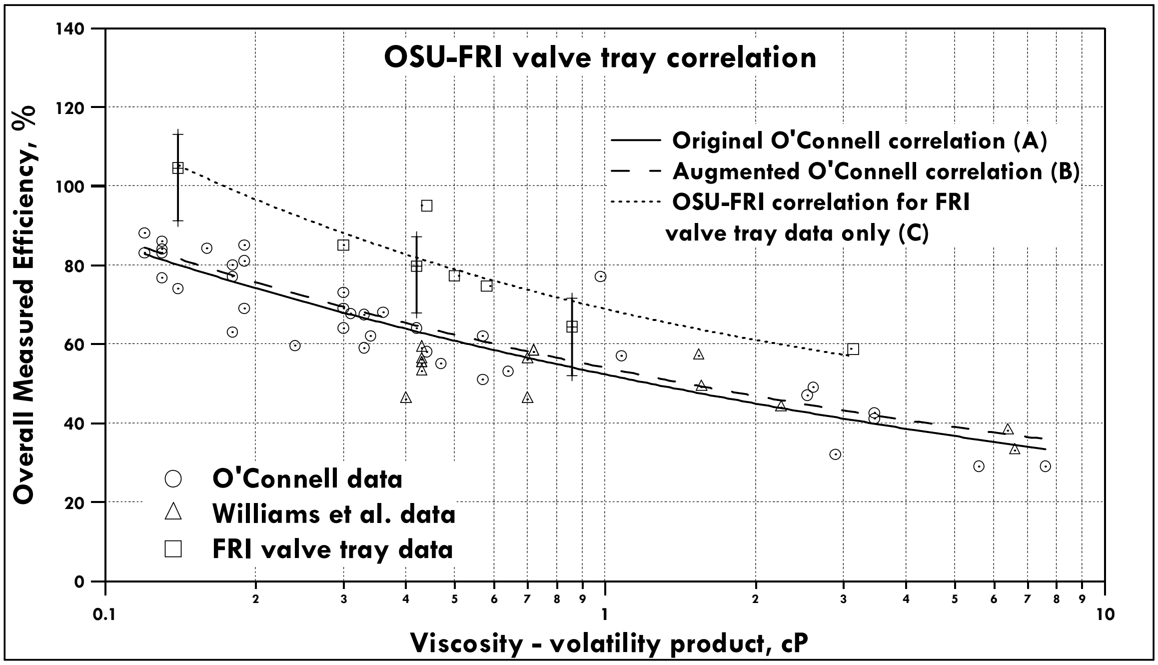

Figure 3.8: The OSU-FRI valve tray correlation for valve tray data. [A line with '+' sign indicate the range of efficiencies. Upper, lower and middle '+' sign indicate higher, lower and average efficiency value respectively]

The equation form of the OSU-FRI valve tray correlation is given as:

$$E_o = 0.695(\alpha\mu_L)^{-0.19} \tag{3.2}$$

The efficiency predictions of the OSU-FRI valve tray correlation (Eq. 3.2) are higher than the predictions of the original O'Connell correlation (Eq. 2.20) by 14.6 to 20.4 % and higher than the augmented O'Connell correlation (Eq. 3.1) by 12.8 to 19.4 %.

Studentized residuals were used to estimate the uncertainty in the coefficient and exponent of the OSU-FRI valve tray correlation (Eq. 3.2). To calculate these uncertainties, the MATLAB^{TM} function nlparci, was used with 95 % confidence limits. The confidence limits of the coefficient are 0.632 to 0.759, while confidence limits of the exponent are -0.12 to -0.28.

The goodness of fit for the new OSU-FRI correlation is shown in Figure 3.9. Efficiency predictions of the OSU-FRI valve tray correlation are in reasonable agreement with the measured efficiencies as inferred from Figure 3.9. The mean absolute relative error for the eight FRI valve tray data points is 5.6 %. Absolute errors shown by six of the data points are less than 5 %, and only two data points have errors over 10 %. These two data points are O P Xylene at 2.0 psia and C6/C7 at 4.7 psia pressure.

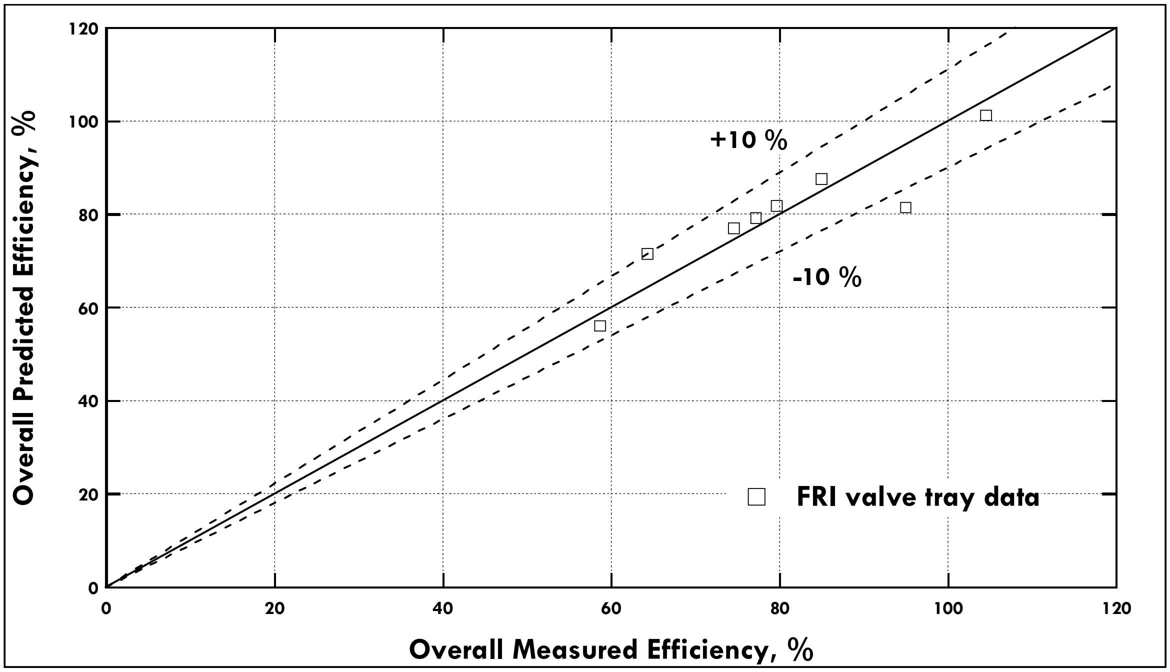

Figure 3.9: A parity plot showing the goodness of the OSU-FRI valve tray correlation for FRI valve tray data.

Table 3.12: Comparison of predicted E_O by the OSU-FRI valve tray correlation with measured E_O for the selected valve tray data

	System	Top Press., psia	α	μL	αμL	Measured EO, %	OSU-FRI O'Conn. Corr.	% Error
FRI valve tray data	C8/C10	0.2	3.63	0.87	3.14	58.70	55.9	4.8
	O P Xylene	0.4	1.24	0.47	0.58	74.60	77.0	3.2
	O P Xylene	0.9	1.23	0.41	0.50	77.20	79.2	2.6
	O P Xylene	2.0	1.21	0.36	0.44	95.00	81.4	14.3
	C6/C7	4.7	1.84	0.47	0.86	64.32	71.5	11.1
	C6/C7	23.6	1.55	0.27	0.42	79.65	82.0	3.0
	C6/C7	49.9	1.42	0.21	0.30	85.00	87.5	2.9
	IC4/NC4	164.7	1.24	0.11	0.14	104.54	101.5	2.9
						Mean absolute % error		5.6

Mean absolute error for the FRI valve tray data = 5.6 %

A residual analysis was also done to validate the goodness of fit of the OSU-FRI valve tray correlation. The residual plot for the correlation is shown in Figure 3.10. The residuals are random as established by performing a six step "Runs test" on the residuals.

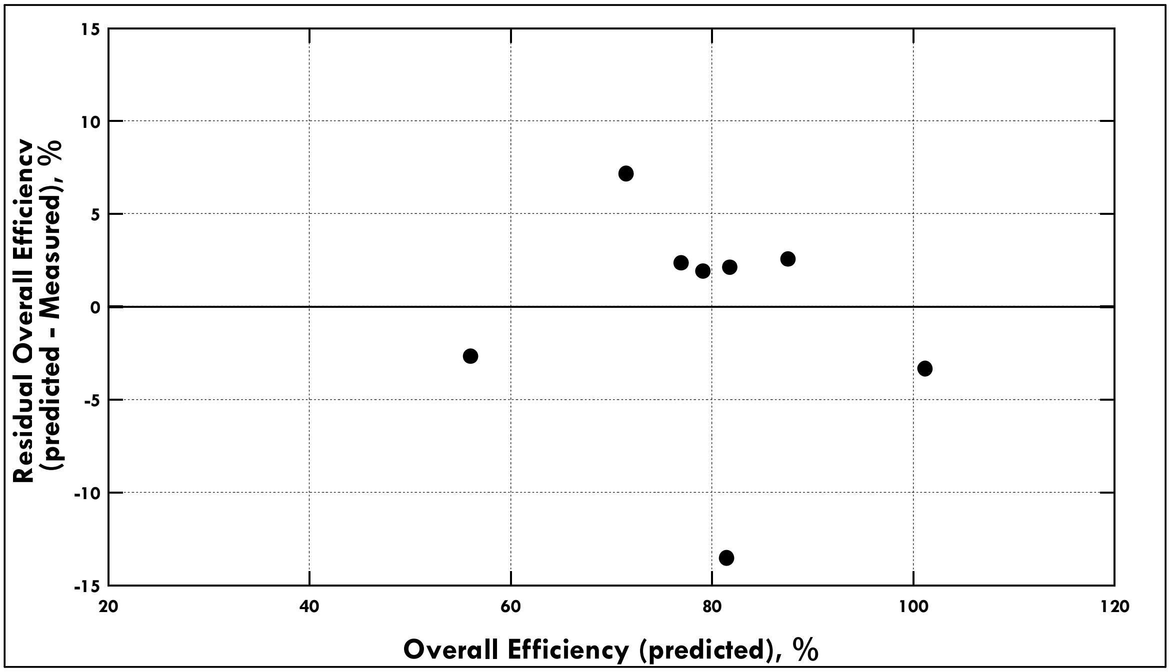

Figure 3.10: A residual plot of the OSU-FRI valve tray correlation with respect to the predicted efficiency values. The residual is calculated as (predicted – measured).

The “Runs test” is performed as follows:

Table 3.13: Performance of the “Runs test” on the residuals

	System	Top Press., psia	$\alpha\mu L$	Measured EO, %	OSU-FRI O'Conn. Corr.	Residuals (Predicted-Measured)	Less/More than Median
FRI valve tray data	C8/C10	0.2	3.14	58.70	55.9	-2.79	B
	C6/C7	4.7	0.86	64.32	71.5	7.16	A
	O P Xylene	0.4	0.58	74.60	77.0	2.41	A
	O P Xylene	0.9	0.50	77.20	79.2	2.00	B
	O P Xylene	2.0	0.44	95.00	81.4	-13.61	B
	C6/C7	23.6	0.42	79.65	82.0	2.36	A
	C6/C7	49.9	0.30	85.00	87.5	2.48	A
	IC4/NC4	164.7	0.14	104.54	101.5	-3.02	B

Step 1: Establish null hypothesis

H_0: Residuals are random.

Step 2: Establish alternative hypothesis

H_1: Residuals are not random.

Step 3: Significance level (α) =0.1

Step 4: Test statistics: V, the total number of runs

Step 5: Computations: For the given sample, the median obtained is 2.18. Replacing each residual measurement by the symbol “A” if it falls above 2.18, and by the symbol “B” if it falls below 2.18, we will obtain the sequence as shown in Table 3.13. This sequence is:

B A A B B A A B

For which, $n_l = 4$, $n_2 = 4$, and $v = 5$. Thefore, using Table A.18 of Walpole and Myers (1989) textbook, the computed p-value is:

$P = 2P\ (V \leq 5 \text{ when } H_0 \text{ is true})$

$P = 2 \text{ X } 0.629$

$P = 1.258 > 0.1.$

Step 6: Decision: Accept the hypothesis that the sequence of residuals varies randomly

The fact that the residuals are random and do not exhibit systematic bias implies that there are no missing independent variables in the correlation.

CHAPTER IV

ANALYSIS OF COMPOSITE DATABASE

Variables such as pressure, diameter, and device (tray) type are commonly identified as factors that influence separation efficiency. This chapter discusses the variations between O'Connell, Williams, et al., and FRI valve tray data used in this study. These variations are summarized in Table 4.1. The purpose of this chapter is to identify trends rather than to perform any type of detailed analysis.

Table 4.1: Variations between O'Connell, Williams, et al., and FRI valve tray data.

	O'Connell data	**Williams, et al. data**	**FRI valve tray data**
Pressure	Atmospheric and above atmospheric	-	Vacuum, atmospheric, and above atmospheric
Diameter	8 to 72 inch[+]	10 inch	48 inch
Tray type	Bubble-cap and Sieve	Bubble-cap	Valve

[+] Diameters for all data points are not known

4.1 Effect of pressure on efficiency

One of the differences between the OSU-FRI valve tray correlation and the original O'Connell correlation is that the former was developed using vacuum and superatmospheric data, while the latter used only superatmospheric data. The points plotted in Figure 4.1 were taken from Table 2.1 (O'Connell) and Table 3.2 (FRI valve tray). The pressure information for the Williams, et al. data was not available.

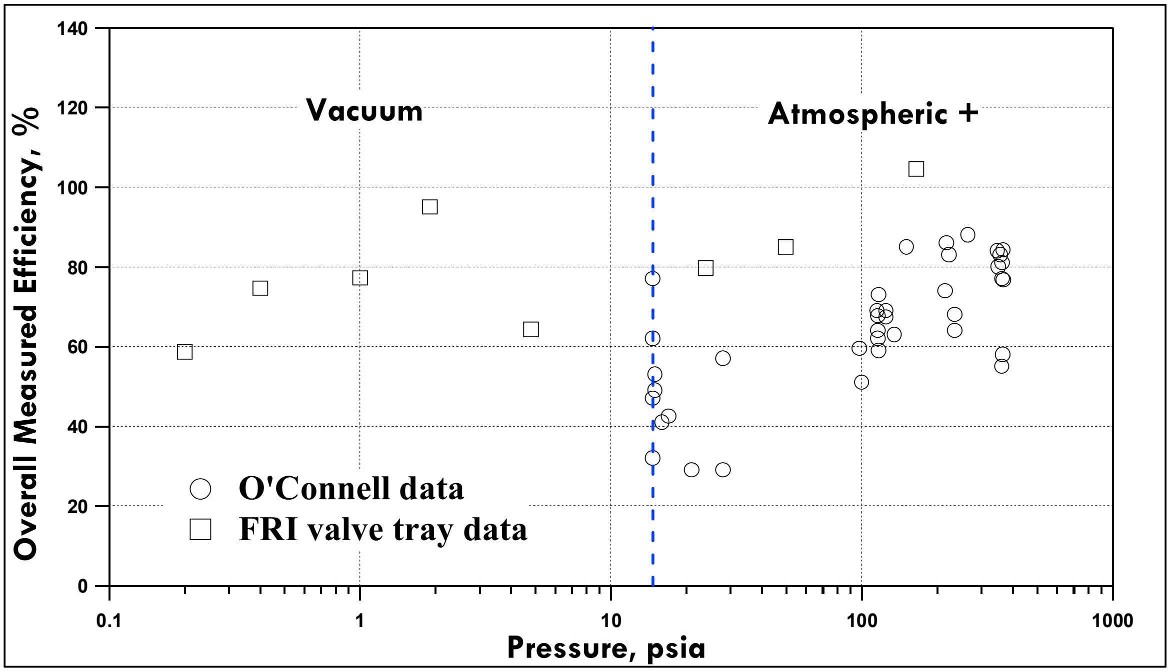

Figure 4.1: Plot of E_O as a function of pressure for the O'Connell and FRI valve tray data.

It is clear from Figure 4.1 that the O'Connell dataset does not include any data points under vacuum conditions. This limitation is seldom mentioned in discussions of the O'Connell correlation in textbooks or the literature.

It is well known that physical properties play an important role in the overall column efficiency. Both the system and pressure affect physical properties. However, the effect of pressure on E_O is easier to identify if done using the same system. Therefore, data in Figure 4.1 are replotted along with system information in Figure 4.2.

Systems for which more than two pressure values were available are plotted in Figure 4.2. Some systems such as IC4/NC4 have data available only at one pressure. Therefore, these systems have not been considered. From visual observation only, Figure 4.2 shows that E_O increases with increase in pressure for O P Xylene and C6/C7 systems at lower pressures (less than 90 psia). However, for pressures higher than 90 psia, it is difficult to identify any trends from Figure 4.2. Therefore, Figure 4.2 has been divided into Figure 4.3 (0.1 to 90 psia) and Figure 4.4 (90 to 500 psia) for detailed observations.

It can be seen from Figure 4.3 that efficiency increases with an increase in pressure for O P Xylene and C6/C7 systems. For stabilization of ethylene dichloride system, there seems to be no change in efficiency with change in pressure. However, at 27.5 psia pressure, two data points of this system show a large difference in efficiency. This is not because of pressure, but because these two efficiencies were calculated using different low boiling components (O'Connell, 1946). No details were available on alcohol-water system data, and hence no explanation can be provided. For beer still data points, an increase in pressure by 1 psia appears to have resulted in an increase in efficiency by 1.5 % as can be seen from Table 2.1.

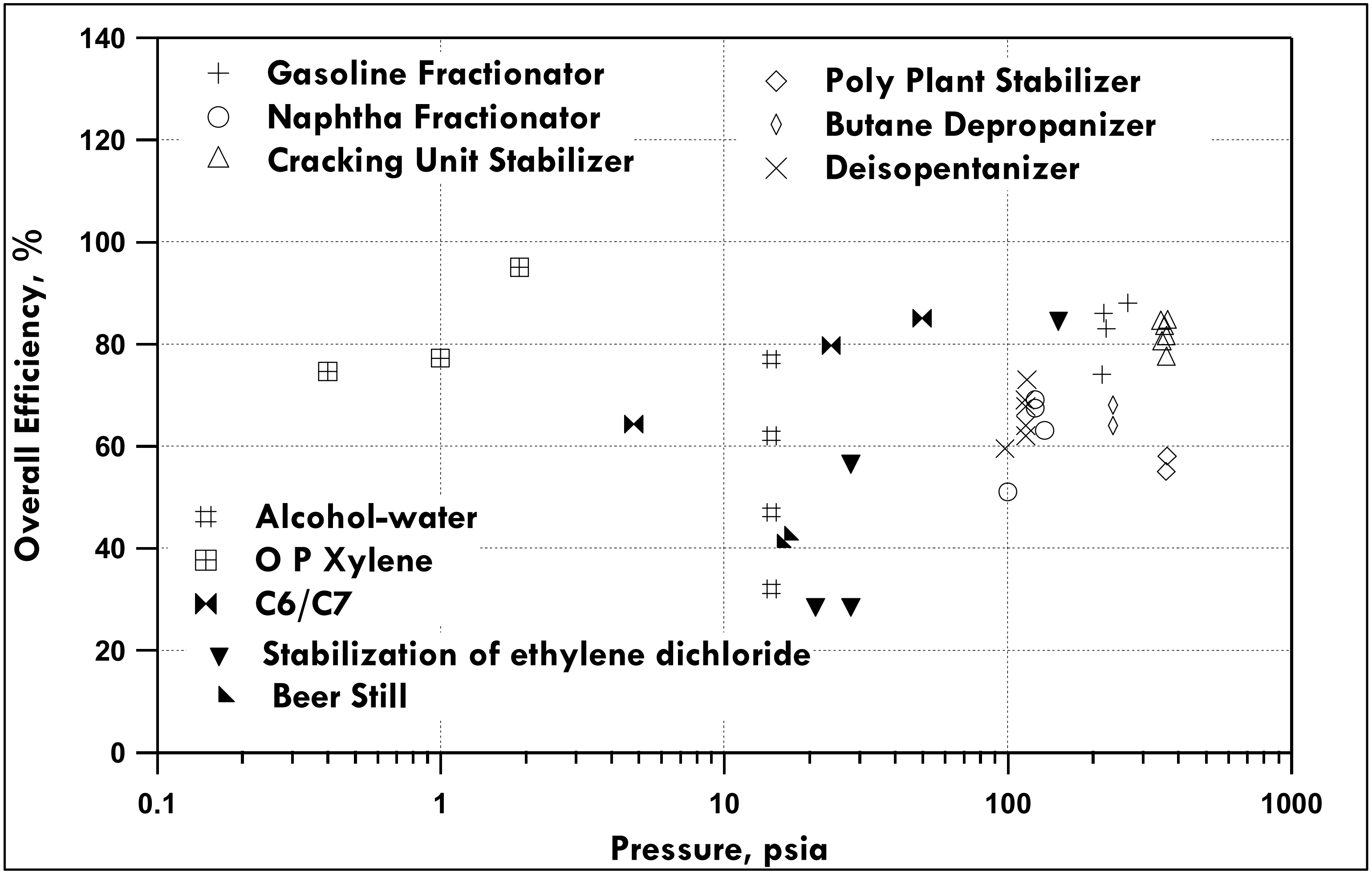

Figure 4.2: Plot of E_O as a function of pressure for the O'Connell and FRI valve tray data with system information.

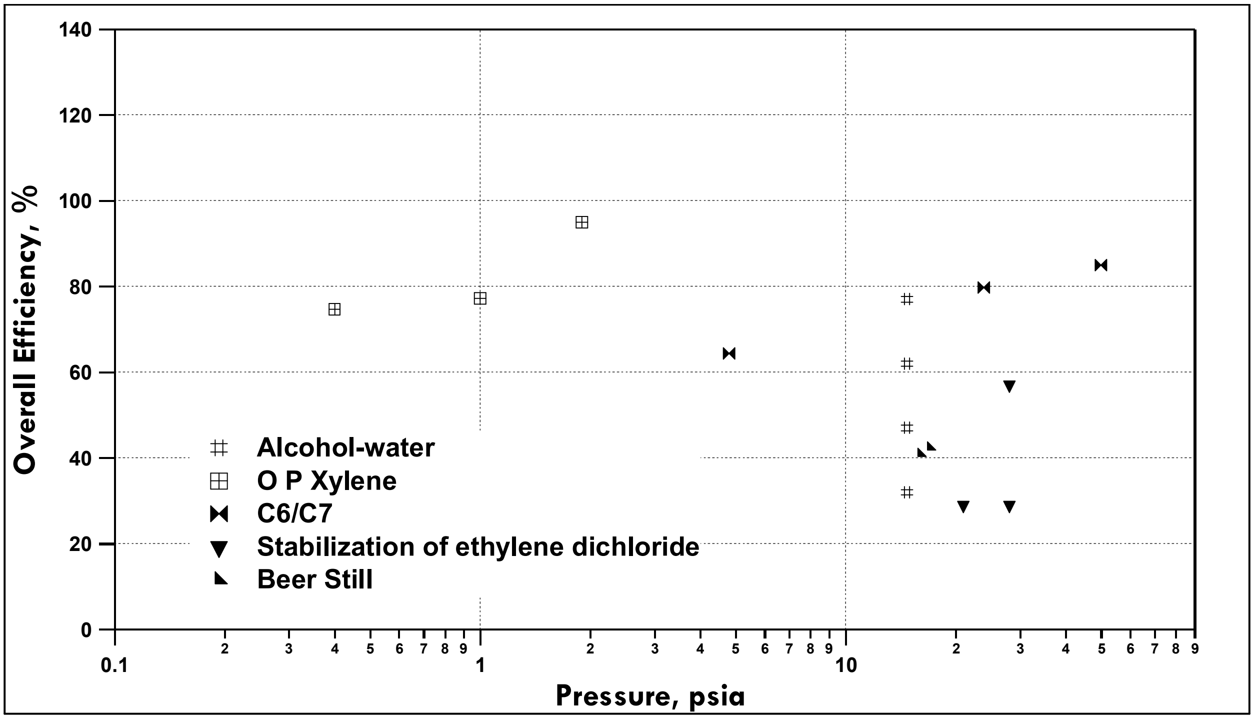

Figure 4.3: Effect of pressure on efficiency for pressure range from 0.1 to 90 psia.

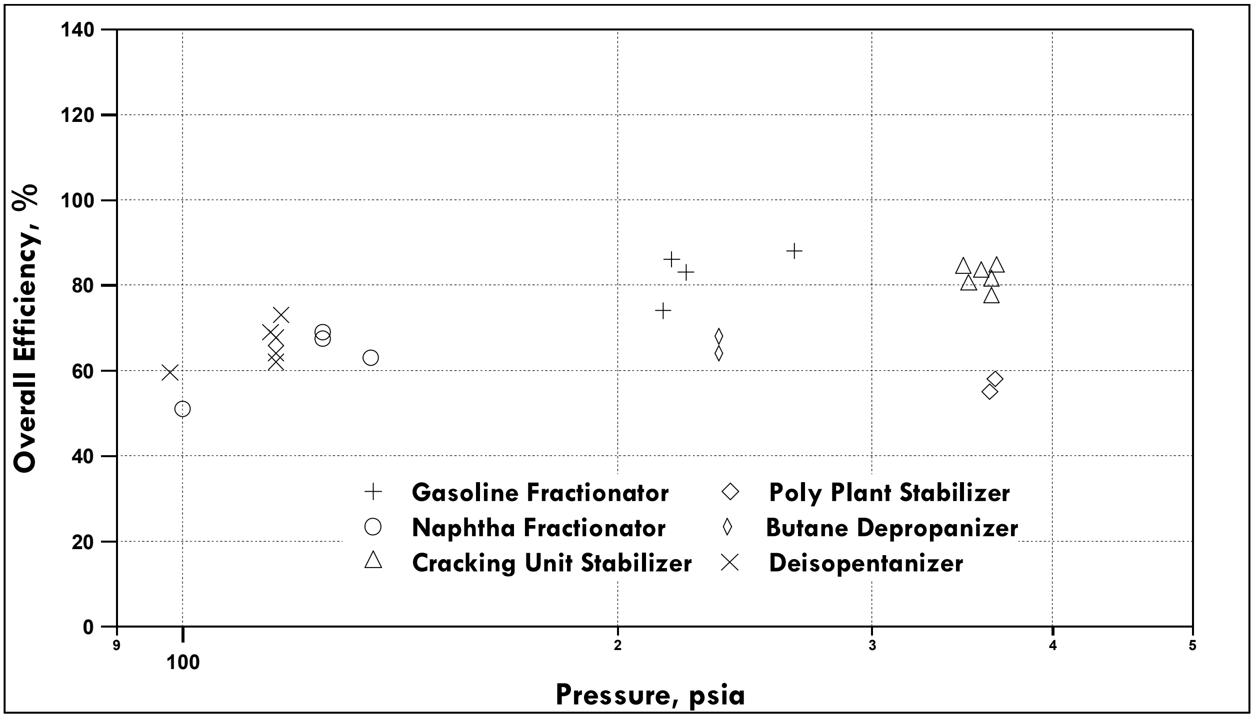

Figure 4.4: Effect of pressure on efficiency for pressure range 90 to 500 psia pressure.

Efficiency appears to increase with an increase in pressure for the deisopentanizer, naphtha and gasoline fractionators as observed from Figure 4.4. There is insufficient variation in the pressure of the butane depropanizer and cracking unit and poly plant stabilizers to observe any trends.

None of the systems included in Figure 4.3 or 4.4 show a trend of reduced efficiency at higher column pressure.

4.2 Effect of column diameter on efficiency

The column diameter directly affects liquid flow path length between downcomers, which is a known variable that influences tray efficiency. Small diameter columns with short flow path lengths lead to back mixing of the liquid, which reduces efficiency. An increase in diameter increases the flow path length, which limits back mixing and directionally increases efficiency. However, a large diameter also increases the potential for stagnant zones where liquid recirculates on the trays. Therefore, column diameter clearly has an impact on column efficiency.

To study the effect of diameter, Willliams, et al.'s 15 data points from a 10 inch diameter column, 8 FRI valve tray data points from a 48 inch diameter column, and 24 of the total 38 O'Connell data points from columns having 8 to 72 inch diameters were used. The column diameters are not known for 14 of the original 38 O'Connell data points. A total of 47 data points were available for the diameter study.

All data for which diameters are known are plotted in Figure 4.5.

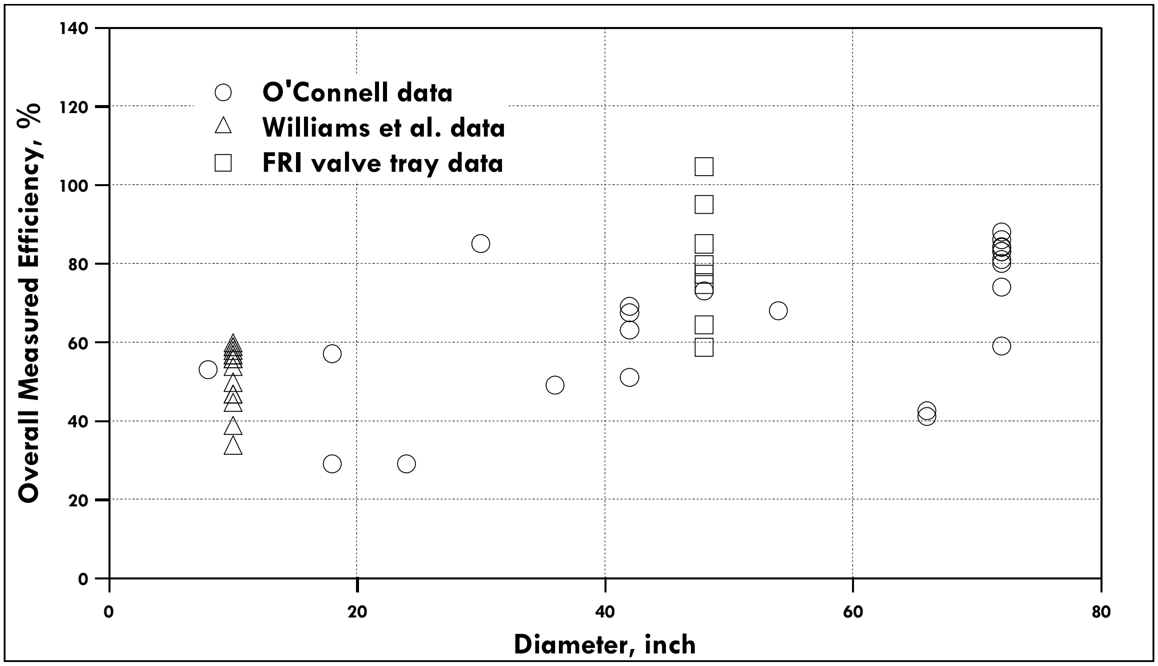

Figure 4.5: Plot of E_O as a function of column diameter

The effect of diameter on E_O shown in Figure 4.5 is not clear because of the number of different systems and operating pressures which are represented. As shown previously, system-pressure effects play an important role in determining overall efficiency (E_O). Therefore, it is useful to consider system-pressure effects along with diameter effects. Hence, measured efficiencies are plotted against viscosity-volatility product for different diameters and shown in Figure 4.6.

Trend lines have been added to Figure 4.6 for the 10, 42, 48, and 72-inch diameter efficiency points. The range of $\alpha\mu_L$ covered by the four column sizes is not the same. However, the 10-inch diameter clearly exhibits the lowest overall efficiencies.

All of the 48-inch diameter data points except a single data point correspond to FRI valve tray data and have the largest overall efficiencies. This result is consistent with

Lockett, et al.'s (1973) conclusion that efficiency increases with column diameter and goes through a maximum at about a diameter of 5 ft. (60 inches). The fact that the 48-inch diameter data were collected at FRI under the best possible conditions may also explain some of the increased efficiencies over the10-inch diameter data.

A much more detailed analysis of the data would be required to draw any further conclusions from a column diameter perspective.

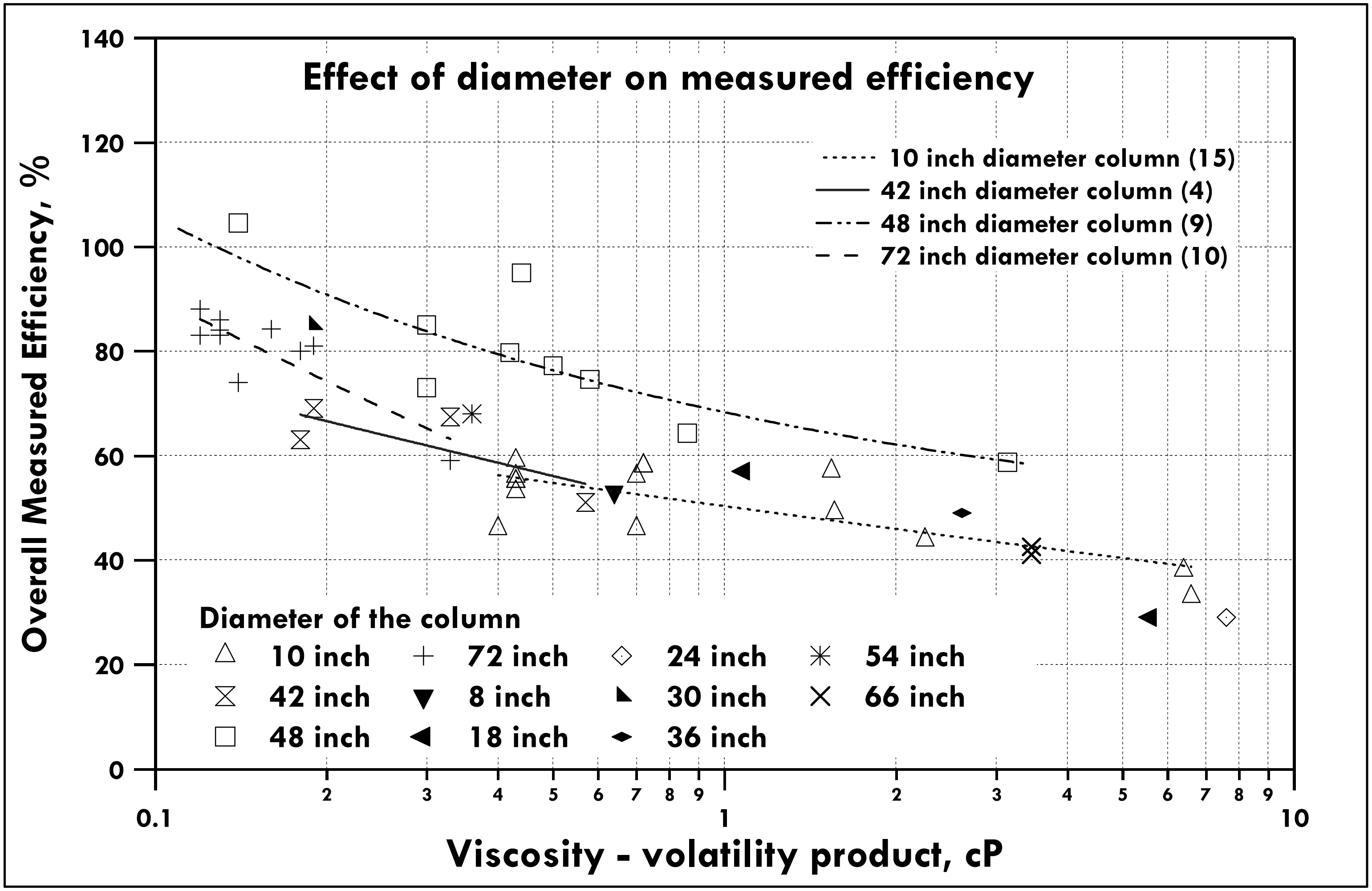

Figure 4.6: Plot of E_O as a function of column diameter with system-pressure effects.

4.3 Tray type considered in data

The OSU-FRI valve tray correlation has been developed using valve tray data, whereas the original O’Connell correlation was developed using bubble-cap and a small amount of sieve tray data. As suggested by O’Connell (in Williams, et al., 1950), the O’Connell correlation should be used for similar column designs. Therefore, the OSU-FRI valve tray correlation is recommended for E_O prediction of valve tray columns and the O’Connell correlation is recommended for E_O prediction of bubble-cap and sieve tray columns. Data used in this study showed that valve tray efficiencies are higher than bubble-cap and sieve tray efficiencies. Some literature sources have also shown that valve tray efficiencies are higher than sieve and bubble-cap tray efficiencies.

- Anderson, et al. (1976) showed that Murphree vapor phase efficiencies of valve trays are higher than those of sieve trays for 1-propanol / toluene system in an 18-inch diameter column. The Anderson, et al. (1976) plot is reproduced and shown in Figure 4.7
- Anderson, et al. (1976) also showed that the E_O of valve trays are higher than E_O of bubble-cap trays for cyclohexane / n- heptane systems in a 48-inch diameter column.
- Bolles (1976) showed that valve tray efficiencies are higher than sieve tray efficiencies for ethylbenzene / styrene system at 1.9 psia pressure in a 20-inch diameter column.

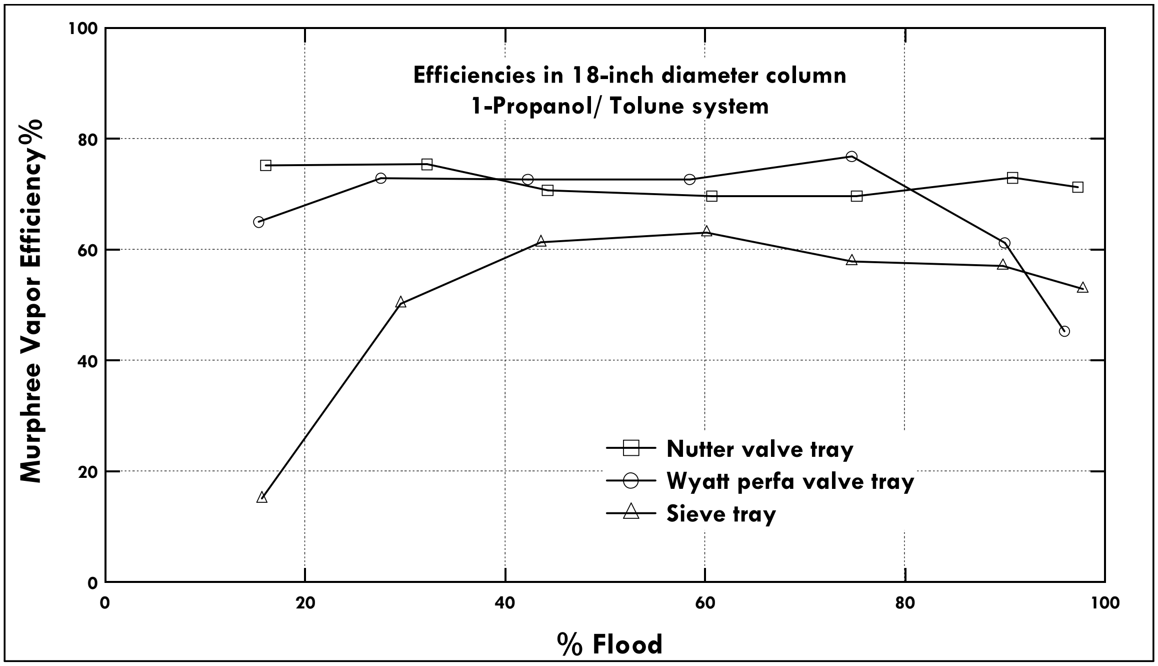

Figure 4.7: Valve tray efficiencies are higher than sieve tray efficiencies (Anderson, et al., 1976; Plot has been replotted against % flood instead of F-factor)

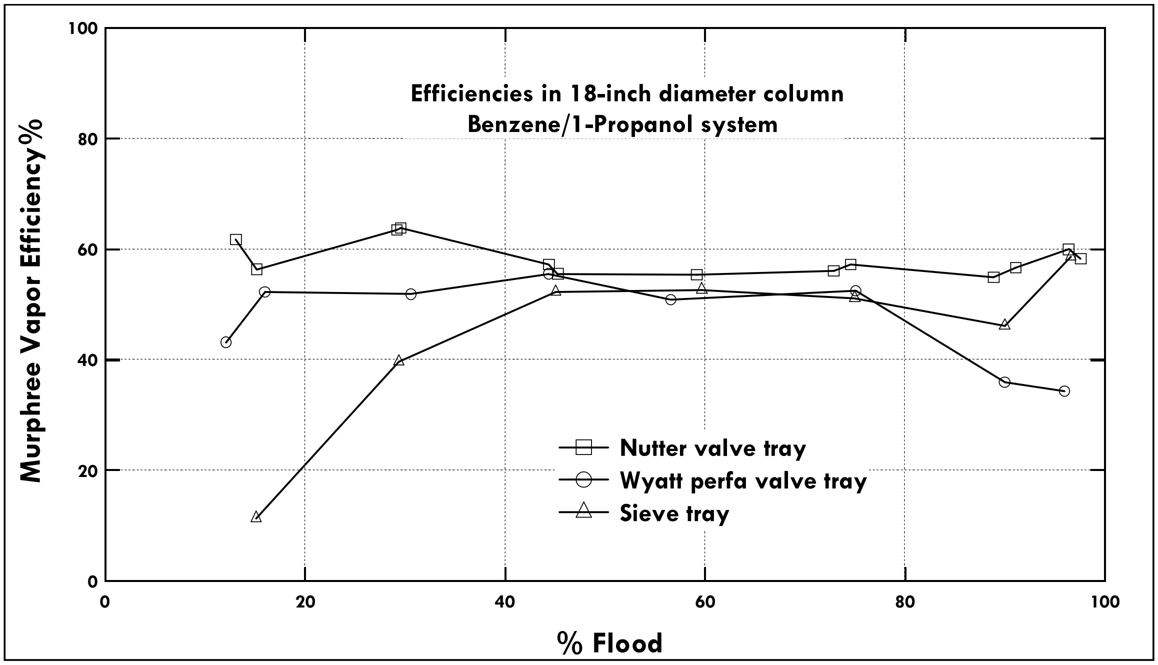

Figure 4.8: Valve tray efficiencies are almost same as sieve tray efficiencies. (Anderson, et al., 1976; Plot has been replotted against % flood instead of F-factor)

However, this might not be true for all systems. Anderson, et al. (1976) showed that for benzene / 1-propanol system, the differences in Murphree efficiencies for sieve and valve trays appear to be insignificant as seen in Figure 4.8. Therefore, it cannot be concluded that the differences in efficiencies are only due to tray type.

Unfortunately, the FRI valve tray and O'Connell data considered in this study do not have any common system for which valve tray efficiencies can be compared with bubble-cap and sieve tray efficiencies. It may be worthwhile to compare FRI sieve and bubble-cap tray efficiencies with FRI valve tray efficiencies for C6/C7, O P Xylene, and IC4/NC4 systems in the future.

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CHAPTER V

CONCLUSIONS AND FUTURE WORK

In this chapter, conclusions and directions for future work in this research are presented.

5.1 Conclusions

The OSU-FRI valve tray correlation has been developed to help design engineers estimate the overall column efficiency (E_O) of valve tray columns. This is the first study that reports a correlation similar to the O'Connell correlation for valve tray columns. The equation form of the OSU-FRI valve tray correlation is given as:

$$E_o = 0.695(\alpha\mu_L)^{-0.19} \tag{5.1}$$

where,

E_O is the overall column efficiency (fraction),

α is the relative volatility between key components, and

μ_L is the liquid viscosity (cP).

The predictions of the OSU-FRI valve tray correlation for valve tray data were higher than those of the O'Connell and augmented correlation by 14.6 to 20.4 % and 12.8 to 19.4 %, respectively, for $\alpha\mu_L$ products in the range 0.1 to 10 cP. This indicates that using the O'Connell and/or augmented O'Connell correlation for valve tray column predictions will result in the under prediction of overall column efficiency by at least 12.0 %. Therefore, the proposed correlation is recommended for E_O prediction of valve tray data.

The mean absolute relative error given by this correlation is 5.6 % for the valve tray data considered in this study. Analysis of the residual plot of the OSU-FRI valve tray correlation shows that the correlation captures the variability in efficiency adequately. In addition, the proposed model does not have systemic bias.

The OSU-FRI valve tray correlation is useful to predict efficiencies of columns with a wide range of pressures at total reflux conditions, as it has been developed using pressures ranging from 0.2 psia to 165 psia. However, note that the presented correlation should be used for efficiency prediction of only valve tray column designs.

5.2 Future Work

Efficiency data is required for a wide range of systems other than hydrocarbon systems to verify the performance of the OSU-FRI valve tray correlation. Surface tension positive data have not been used in this study and the correlation needs to be verified for such data.

It is worthwhile to test the performance of the OSU-FRI valve tray correlation using FRI bubble-cap and sieve tray data at total reflux. Since, these data are similar to the FRI valve tray data used in this study except for the tray type. The similarities between FRI valve tray data and FRI bubble-cap and sieve tray data are:

- available at total reflux conditions,
- obtained from a column of 48 inch diameter, and
- available for binary systems.

Due to the empirical nature of the OSU-FRI valve tray correlation, the reliability of the presented correlation can be increased by updating this correlation with more quality data points having a wide range of system-pressure combinations.

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APPENDIX A

DRICKAMER AND BRADFORD CORRELATION DATA

Table A.1: Data used for Drickamer and Bradford correlation development (Adopted from: Drickamer, H. G., and Bradford, J. R. (1943). Overall plate efficiency of commercial hydrocarbon fractionating columns as a function of viscosity. *Transaction of American Institute of Chemical Engineers, 39*, 319-360.)

	Tower Type	Avg. Temp., F	Avg. Press., psia	Molal average viscosity, cP	% Tray Eo
1	Cracking Unit Stabilizer	312	357	0.066	83.0
2	Cracking Unit Stabilizer	315	347	0.072	84.0
3	Cracking Unit Stabilizer	313	366	0.091	84.2
4	Cracking Unit Stabilizer	314	350	0.098	80.0
5	Cracking Unit Stabilizer	319	363	0.105	81.0
6	Cracking Unit Stabilizer	309	363	0.098	77.0
7	Debutanizer*	275	117	0.165	59.0
8	Deisopentanizer*	280	116	0.160	64.0
9	Deisopentanizer*	270	116	0.172	62.0
10	Deisopentanizer	260	117	0.152	73.0
11	Deisopentanizer	262	116	0.158	67.7
12	Deisopentanizer	264	115	0.162	69.0
13	Cracking Unit Debutanizer	245	115	0.190	60.0
14	Cracking Unit Debutanizer	252	130	0.213	61.5
15	Cracking Unit Debutanizer	248	125	0.210	64.0
16	Butane Depropanizer	162	235	0.148	68.0
17	Butane Depropanizer	161	235	0.170	64.0
18	Poly Plant Stabilizer	228	365	0.220	58.0
19	Poly Plant Stabilizer	198	362	0.222	55.0
20	Poly Plant Debutanizer	238	145	0.310	46.0
21	Poly Plant Debutanizer	242	147	0.322	44.0
22	Poly Plant Absorber	126	255	0.405	42.0

Continued on next page

(Table A.1 continued)

23	Poly Plant Absorber	127	260	0.410	39.0
24	Poly Plant Absorber	124	265	0.500	38.0
25	Poly Plant Absorber	120	260	0.480	36.0
26	Hydroformer Stabilizer	285	350	0.355	44.0
27	Kerosene Stripper	448	68	0.205	57.0
28	Kerosene Stripper	435	68	0.190	64.0
29	Virgin Gas Oil Stripper	507	60	0.250	49.0
30	Virgin Gas Oil Stripper	481	58	0.270	48.0
31	Virgin Gas Oil Stripper	492	60	0.280	52.0
32	Absorber	138	267	0.220	56.4
33	Absorber	132	254	0.310	50.0
34	Absorber	117	94	1.410	10.4

*Bottom 11 trays perforated in tower

APPENDIX B

TRAY PICTURES

This appendix provides the pictures of the trays included in the FRI valve tray datasets.

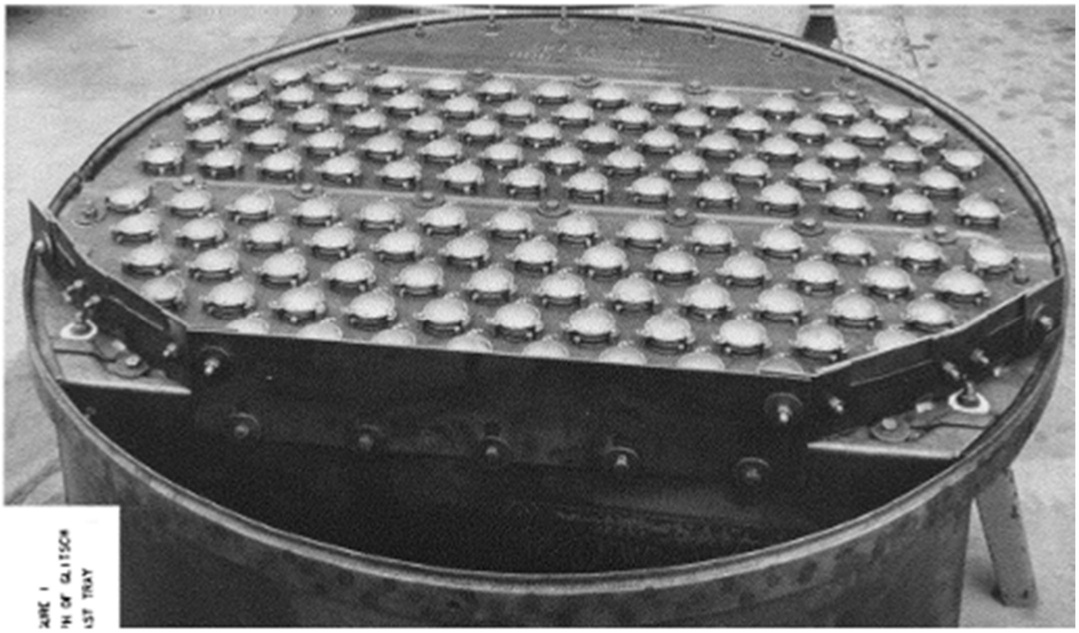

Figure B.1: Glitsch Ballast tray (124 Moving valves)

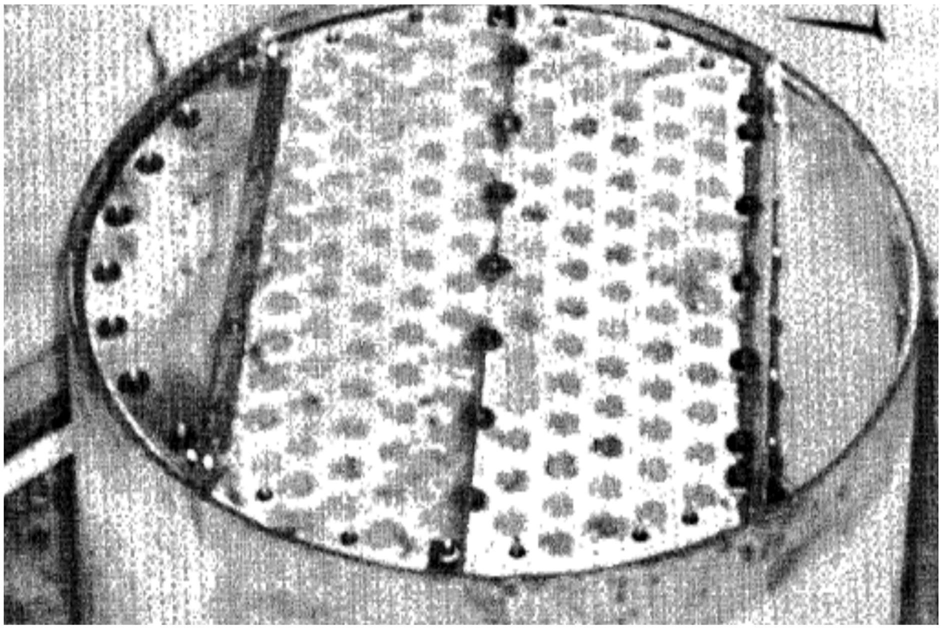

Figure B.2: Glitsch Ballast Tray V1 (140 Moving valves)

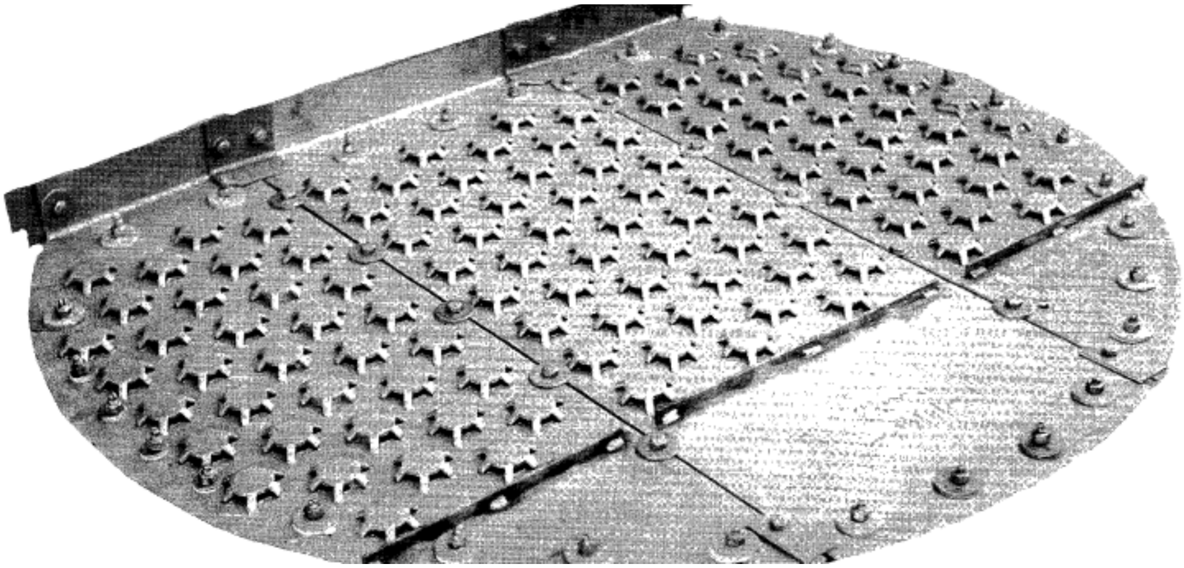

Figure B.3: Glitsch V-O Ballast tray (114 Fixed valves)

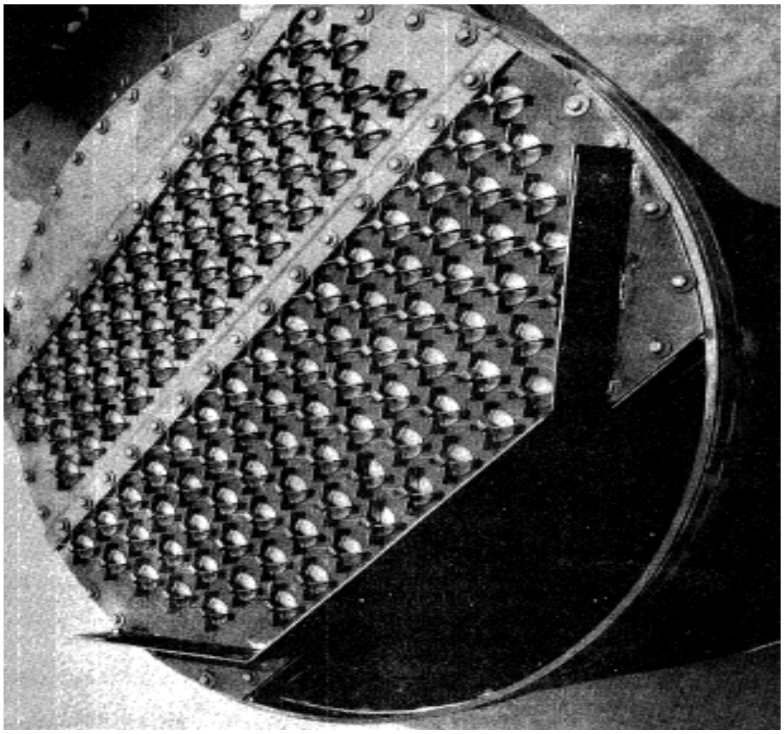

Figure B.4: Koch Flexitray (122 Moving valves)

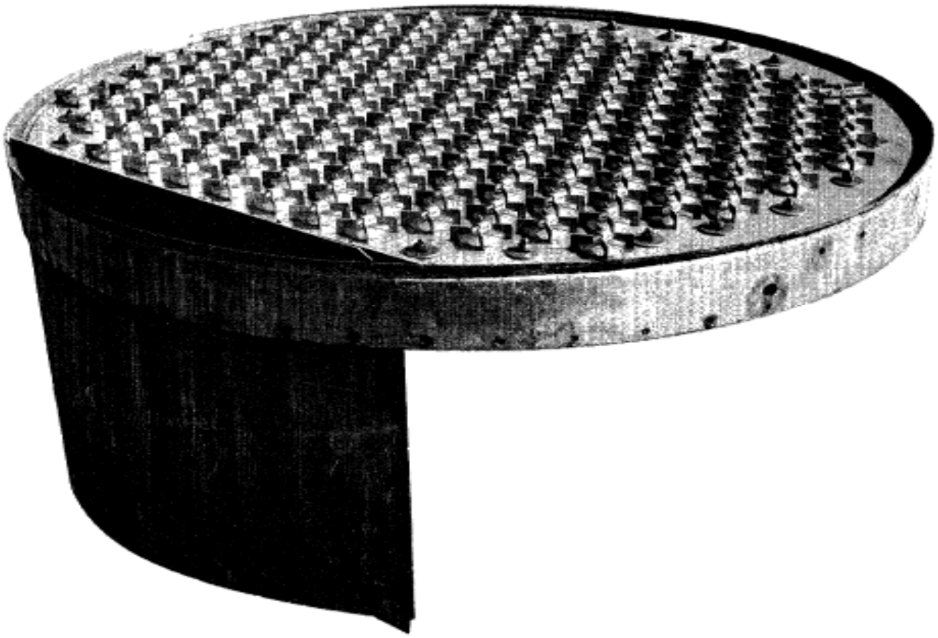

Figure B.5: Koch Type T Flexitray (167 Moving valves)
[Two different valve weights were used on alternate rows]

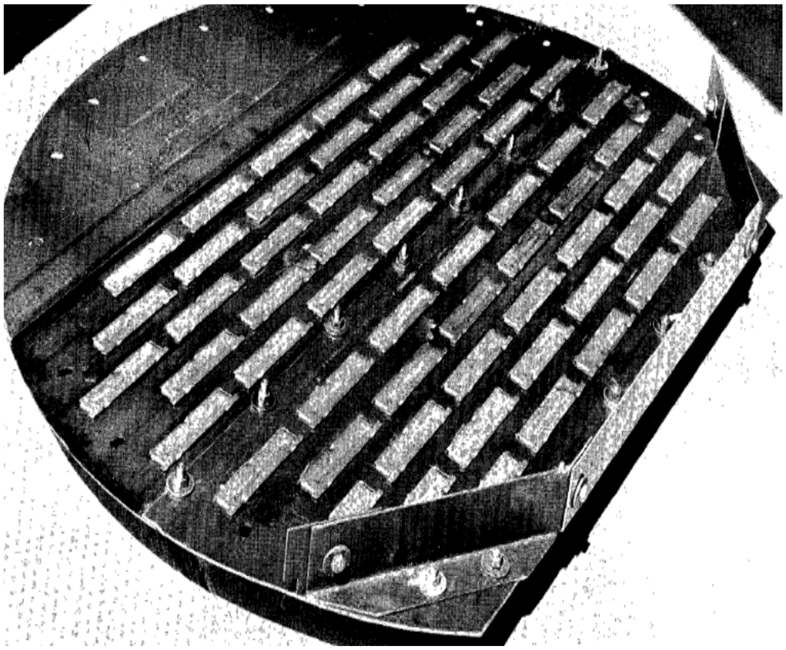

Figure B.6: Nutter Float Valve P-437 and Nutter Float Valve D-437 (62 Moving valves)

[The difference between Nutter Float Valve P437 and Nutter Float Valve D437 is only the slot area. The slot area per tray for D437 and P437 is 1.44 and 1.34 ft^2, respectively.]

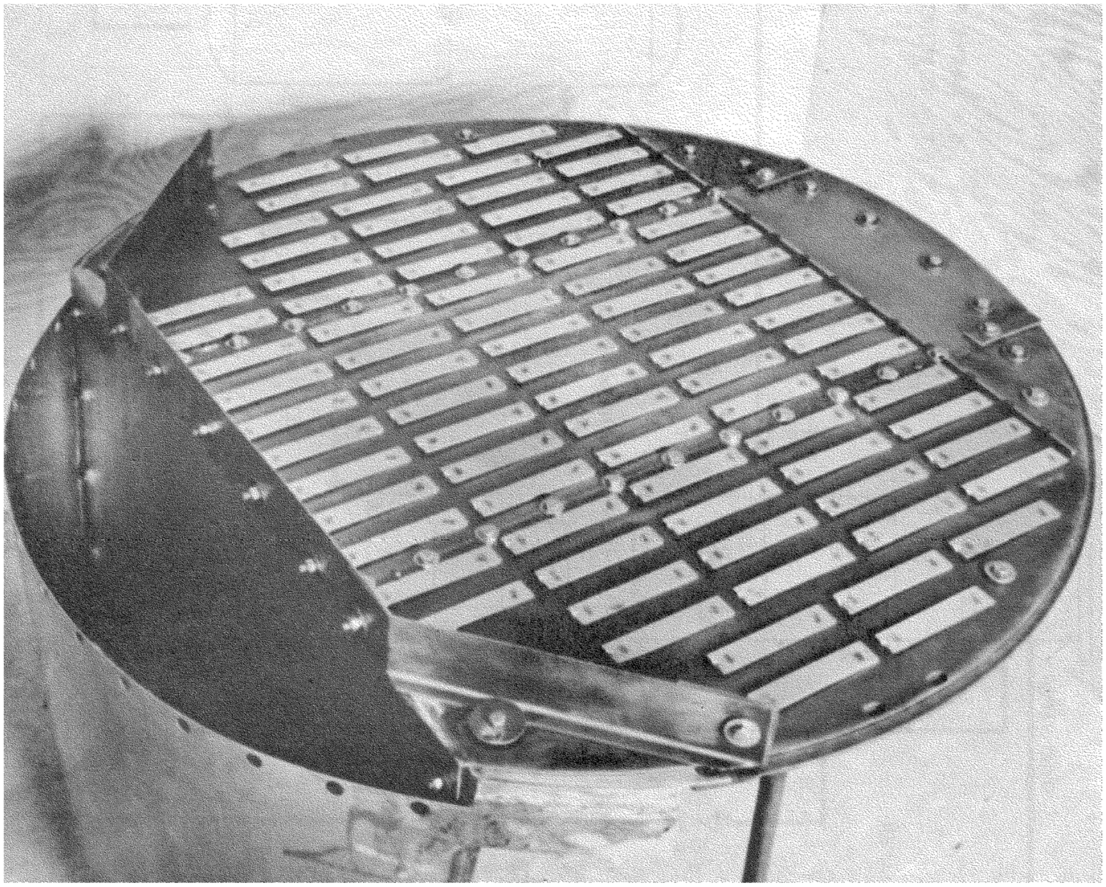

Figure B.7: Nutter Type B Float Valve and Nutter Type B Float Valve with larger slot area (81 Moving valve)

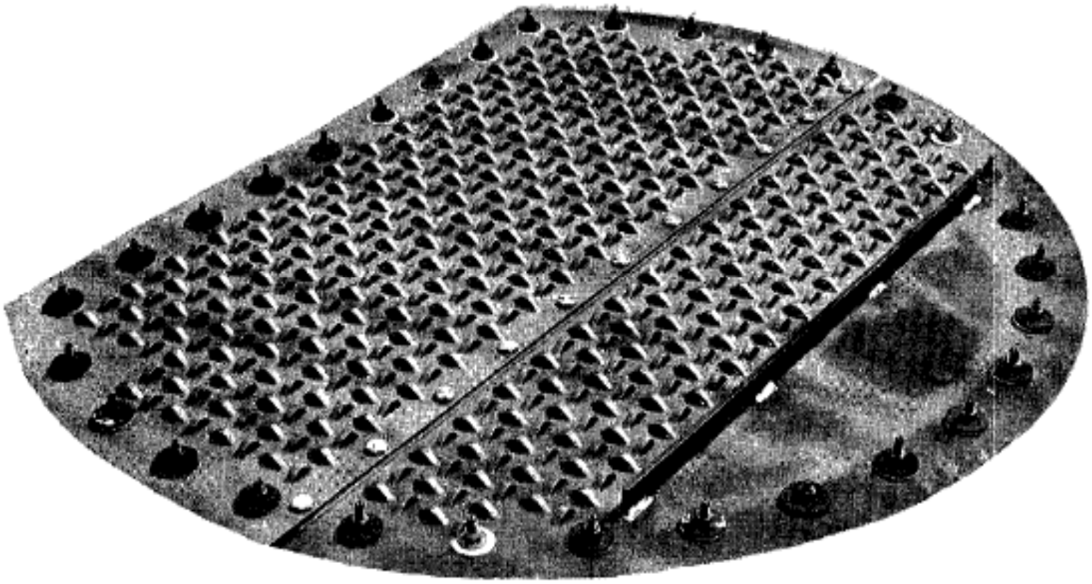

Figure B.8: Zink Bi-Mix Tray No. 1 and Zink Bi-Mix Tray No. 2 (526 Fixed valves)

[The difference between Zink Bi-Mix Tray No. 1 and Zink Bi-Mix, Tray No. 2 is height of inlet weir. The height of inlet weir for Zink Bi-Mix Tray No. 1 and Zink Bi-Mix, Tray No. 2 is 0.5 and 1.0 inch, respectively.]

APPENDIX C

DIFFERENCES IN TRAY GEOMETRIES

Table C.1: Difference between Nutter Float Valve D-437 and Nutter Float Valve P-437 trays.

Tray type	Nutter Float Valve D-437	Nutter Float Valve P-437
Slot area, ft^3	1.44	1.34

Table C.2: Differences between Glitsch Ballast, Glitsch V-O Ballast, and Glitsch Ballast V-1 trays.

Tray type	Glitsch Ballast tray	Glitsch V-O Ballast tray	Glitsch Ballast tray V-1
Moving/Fixed	Moving	Fixed	Moving
Weir height, inch	3	2.5	2
No. of valves	124	114	136
Bubbling area, ft^2	9.43		9.3
Weir length, inch	52	37	36
Inlet weir	No	Yes	Yes
Number of ballast plates	2		1

Table C.3: Differences between Koch Flexitray and Koch type T Flexitray.

Tray type	Koch Flexitray	Koch type T Flexitray
Valve weights	Same weight valves	Different weight valves
Downcomer	sloped	straight
Weir height, inch	3	0.75
Weir length, inch	52	30
No. of valves	122	167

Table C.4: Differences between Nutter Float Valve D-437, Nutter Float Valve P-437, Nutter Type B Float Valve, and Nutter Type B Float Valve (with larger slot area) trays.

Tray type	Nutter Float Valve D-437	Nutter Float Valve P-437	Nutter Type B Float Valve	Nutter Type B Float Valve (with larger slot area)
No. of valves	62	62	81	81
Weir height, inch	2.85	2.85	3	3
Weir length, inch	50	50	54.3	54.3
Valve orientation	Perpendicular to liq. flow	Perpendicular to liq. flow	Parallel to liq. flow	Parallel to liq. flow

Table C.5: Difference between Zink Bi-Mix Tray No. 1 and Zink Bi-Mix Tray No. 2 trays.

Tray type	Zink Bi-Mix Tray No. 1	Zink Bi-Mix Tray No. 2
Inlet weir height, inch	0.5	1.0

APPENDIX D

TRAY GEOMETRY INFORMATION

This appendix provides the available detailed information for the trays considered in this study.

Table D.1: Tray details of Glitsch Ballast tray (Adapter from: FRI (1958). Fractionation Research, Inc. Topical Report No. 15, Obtainable from the Oklahoma State University Library)

Glitsch Ballast tray

Tray spacing, inches	24
Number of trays	9
Inside column diameter, inches	47.75
Column cross-sectional area, sq. ft.	12.44
Bubbling area, sq. ft.	9.43
Area of downcomer at bottom, sq. ft.	1.04
Area of downcomer at top, sq. ft.	1.38
Area under downcomer, sq. ft.	0.53
Clearance under downcomer, inches	2.25
Length of downcomer seal, inches	32
Length of outlet weir, inches	52
Height of outlet weir, inches	3
Number of Ballast units per tray	124
Ballast assembly, 410 SS:	
Orifice seat diameter, inches	1.53
Orifice cover diameter, inches	1.875
Ballast plate, light weight	16 ga
Ballast plate, heavy weight	14 ga
Slot areas per tray:	
Orifice covers open, sq. ft.	0.49
Ballast plates open, sq. ft.	1.31

Table D.2: Tray details of Glitsch Ballast tray V1 (Adapter from: FRI (1959). Fractionation Research, Inc.Topical Report No. 19, Obtainable from the Oklahoma State University Library)

Glitsch Ballast tray V-1

Tray spacing, inches	24
Number of trays	9
Inside column diameter, inches	47.75
Column cross-sectional area, sq. ft.	12.44
Bubbling area, sq. ft.	9.3
Area of downcomer at bottom, sq. ft.	1.04
Area of downcomer at top, sq. ft.	1.38
Area under downcomer, sq. ft.	0.47
Clearance under downcomer, inches	2.0
Length of downcomer seal, inches	33.125
Length of outlet weir, inches	36
Height of outlet weir, inches	2
Length of inlet weir, inches	38
Height of inlet weir, inches	1.0
Number of Ballast units, per tray	136

Glitsch Ballast tray V-1 Ballast unit consists of a single part.

Table D.3: Tray details of Glitsch V-O Ballast tray (Adapter from: FRI (1967b). Fractionation Research, Inc. Topical Report No. 41, Obtainable from the Oklahoma State University Library)

Glitsch V-O Ballast tray

Tray spacing, inches	24
Number of trays	10
Inside column diameter, inches	48.0
Column cross-sectional area, sq. ft.	12.56
Area of downcomer at bottom, sq. ft.	1.5
Area of downcomer at top, sq. ft.	1.5
Clearance under downcomer, inches	2.25
Length of outlet weir, inches	37
Height of outlet weir, inches	2.5
Height of inlet weir, inches	0.5
Tray deck thickness	10 ga
Number of V-O units per tray	114

Table D.4: Tray details of Koch Flexitray (FRI (1958). Fractionation Research, Inc. Topical Report No. 10, Obtainable from the Oklahoma State University Library)

Koch Flexitray

Tray spacing, inches	24
Number of trays	10
Area of downcomer at bottom, sq. ft.	1.05
Area of downcomer at top, sq. ft.	1.35
Area under downcomer, sq. ft.	0.65
Length of outlet weir	52
Height of outlet weir	3

Table D.5: Tray details of Koch type "T" Flexitray (Adapter from: FRI (1968). Fractionation Research, Inc.Topical Report No. 44, Obtainable from the Oklahoma State University Library)

Koch type "T" Flexitray

Tray spacing, inches	24
Number of trays	6
Inside column diameter, inches	47.75
Column cross-sectional area, sq. ft.	12.44
Bubbling area, sq. ft.	11.2
Area of downcomer at bottom, sq. ft.	0.53
Area of downcomer at top, sq. ft.	0.53
Clearance under downcomer, inches	0.75
Length of outlet weir, inches	30.0
Height of outlet weir, inches	0.75
Number of valve units, per tray	167

Table D.6: Tray details of Nutter Float valve D-437 and P-437 (Adapter from: FRI (1962). Fractionation Research, Inc.Topical Report No. 27, Obtainable from the Oklahoma State University Library)

Nutter Float valve trays

	D-437	**P-437**
Tray spacing, inches	24	24
Number of trays	10	10
Inside column diameter, inches	47.75	47.75
Column cross-sectional area, sq. ft.	12.44	12.44
Bubbling area, sq. ft.	7.79	7.79
Tower free area, sq. ft.	10.84	10.84
Slot area per tray, sq. ft.	1.44	1.34
Area of downcomer at top, sq. ft.	1.50	1.50
Clearance under downcomer, inches	2.375	2.375
Downcomer escape area at bottom, sq. ft.	0.67	0.67
Length of outlet weir, inches	50	50
Height of outlet weir, inches	2.875	2.875
Length of inlet weir, inches	40.75	40.75
Height of inlet weir, inches	0.75	0.75
Number valve units, per tray	62	62

Table D.7: Tray details of Nutter Type B Float valve (Adapter from: FRI (1964). Fractionation Research, Inc.Topical Report No. 31, Obtainable from the Oklahoma State University Library)

Nutter Type B Float valve

Tray spacing, inches	24
Number of trays	10
Column diameter, inches	48.0
Column cross-sectional area, sq. ft.	12.56
Area of downcomer at top, sq. ft	1.44
Area of downcomer at bottom, sq. ft	0.95
Downcomer escape area at bottom, sq. ft.	0.58
Length of outlet weir, inches	54.3
Height of outlet weir, inches	3.0
Number valve units, per tray	81

Nutter Type B Float valves are of two type. In Nutter Type B Float valves (same as B with larger slot area), only slot area is larger. However, slot areas are not known.

Table D.8: Tray details of Zink Bi-Mix trays (Adapter from: FRI (1967a). Fractionation Research, Inc.Topical Report No. 40, Obtainable from the Oklahoma State University Library.)

Zink Bi-Mix trays

	Tray No. 1	**Tray No. 2**
Tray spacing, inches	24	24
Number of trays	7	7
Inside column diameter, inches	48.0	48.0
Column cross-sectional area, sq. ft.	12.56	12.56
Bubbling area, sq. ft.	9.25	9.25
Area of downcomer at top, sq. ft.	1.50	1.50
Area of downcomer at bottom, sq. ft.	1.50	1.50
Clearance under downcomer, inches	1.75	1.75
Length of outlet weir, inches	37.0	37.0
Height of outlet weir, inches	2.0	2.0
Length of inlet weir, inches	30.9	30.9
Height of inlet weir, inches	0.5	1.0

VITA

Dadasaheb B. Salunke

Candidate for the Degree of

Master of Science

Thesis: THE OSU-FRI VALVE TRAY CORRELATION FOR PREDICTION OF OVERALL EFFICIENCY OF VALVE TRAY COLUMNS

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Institution: Oklahoma State University Location: Stillwater, Oklahoma

Title of Study: AN O'CONNELL TYPE CORRELATION FOR PREDICTION OF OVERALL EFFICIENCY OF VALVE TRAY COLUMNS

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Scope and Method of Study: The prediction of overall column efficiency directly affects the design and cost of a distillation column. Therefore, there is an economic incentive to predict column efficiency as accurately as possible. The O'Connell correlation is widely recommended for efficiency predictions and was developed using bubble-cap and sieve tray data. In this work, a new O'Connell type correlation for prediction of overall efficiency of valve tray columns is developed using valve tray data collected at Fractionation Research, Inc. (FRI) that have been made public. The correlation was developed using data for ortho-xylene / para-xylene, isobutene / normal-butane, cyclohexane / n-heptane, and n-octanol / n-decanol systems at pressures ranging from 0.2 to 164.7 psia. The data were collected on the 4 feet diameter FRI column between 1956 and1968. All of the FRI data corresponds to total reflux operation.

Findings and Conclusions: An O'Connell type correlation (The OSU-FRI valve tray correlation) for prediction of overall efficiency of valve tray columns has been developed. The form of the correlation is similar to that of the O'Connell correlation. The mean absolute relative error obtained for the new valve tray correlation is 5.6 %. This model is expected to predict the efficiency of valve tray column in reasonable agreement, thereby help design engineers economically design valve tray distillation columns. The new OSU-FRI valve tray correlation is expected to predict efficiency over a wide range of pressures, as it was developed using data that were available at a wide range of pressures from vacuum to super-atmospheric. In addition, the correlation is directly applicable to commercial column designs without any scale-up factors, as data used for the development of the correlation were taken from a commercial column.

ADVISER'S APPROVAL: Dr. James R. Whiteley