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Overall Efficiency Evaluation of Commercial Distillation Columns with Valve and Dualflow Trays

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The main objective of this work is to establish appropriated ways for estimating the overall efficiencies of industrial distillation columns with valve trays with downcomer and dualflow trays. The knowledge of efficiencies has fundamental importance in the design and performance evaluation of distillation columns. Searching in the literature, a tree of alternatives was identified to compose the tray efficiency model, depending on the mass transfer models, the liquid distribution and vapor flow models on the tray, the liquid entrainment model, the multicomponent mixture equilibrium model, the physical properties models, the height of froth on the tray model and the efficiency definition. In this work, different methods to predict the overall efficiency of distillation columns with valve and dualflow trays were composed and compared with data from three commercial distillation columns under different operating conditions. The models were inserted in the Aspen Plus 12.1 simulator, in Fortran language, together with tray geometrical data, fluid properties and operating data of the distillation columns. For each column, the best thermodynamic package was chosen by checking the temperature profile and overhead and bottom compositions obtained via simulation against the corresponding actual data of industrial columns. A modification in the fraction of holes evaluation that is jetting parameter of the Garcia's hydraulic model of dispersion above the tray was proposed. This modification produced better results than the original model to predict the fraction of holes that are jetting and in the efficiency of dualflow trays and similar results to Garcia model in the efficiency evaluation of valve trays. © 2010 American Institute of Chemical Engineers *AIChE J,* 56: 2323–2330, 2010

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Introduction

In the petrochemical industry, trayed column is one of the most used equipment in the separation of liquid mixtures

and vapor–gas mixtures.[1] The single-pass and the dualflow tray are among the possible flow-path arrangements used in these columns, and according to their liquid–vapor flow in the tray, it can be, respectively, classified as crossflow and countercurrent flow.

In crossflow trays, the liquid flows between trays through connecting downcomers and the vapor flows through holes, valves, or bubble caps in the bubbling area. In the less-known dualflow trays, there is no downcomer, and the descending liquid and the ascending vapor pass through the same tray openings; in special forms, they may have other names such as turbogrid trays. The dualflow trays have a higher capacity and present a lower pressure drop than those trays with downcomers because the flow occurs through the whole cross-sectional area of the column.[1] Such trays also provide a reasonable mass transfer with a low capital investment in its range of application, where a high turndown ratio is not essential. They are used for special services especially when openings of a crossflow tray might fouled.[2] Despite the more frequent use of dualflow trays in petrochemical and oil industries, publications related to the efficiency estimation of such trays are rare. As mentioned in Garcia and Fair,[2] more and more crossflow trays are being replaced by dualflow trays to prevent problems of severe fouling.

The first proposals for estimating global efficiency of distillation columns were based solely on experimental data analysis, resulting in empirical models. In 1946, O'Connell,[3] based on industrial plant data, proposed an empirical correlation for global efficiency as function of liquid viscosity and relative volatility evaluated at average temperature between top and bottom column temperature and feed composition. This correlation, based on the previous work of Drickamer and Bradford,[4] is largely used in industry because of its simplicity and reasonable agreement with plant data, being slightly conservative. MacFarland et al.[5] proposed an empirical correlation based on dimensionless groups for estimating the Murphree tray efficiency with absolute mean deviation from experimental data around 10%.

In 1958, the first theoretical method was proposed to evaluate the point efficiency based on the two-film mass transfer theory known as AIChE[6] method and being the origin of most methods available nowadays. Chan and Fair[7] improved the AIChE method using a large separation data bank of binary mixtures with the price of including an additional flooding factor parameter. Zuiderweg,[8] based on data from FRI (Fractionation Research Institute), proposed a model for point efficiency using semiempirical correlation for the mass transfer coefficients. Chen and Chuang[9] proposed a mechanistic model for point efficiency that needs the adjustment of two parameters, which was further improved by Klemola and Ilme.[10]

Garcia[11]/Garcia and Fair[12] presented a fundamental theoretical model devised for estimating the efficiency of tray with downcomer. Such model, designed to be applied to organic systems, was an improvement of the model suggested by Prado and Fair[13] for air–water systems, which divided the tray into sections and applied to each one the two-layer mass transfer theory and the type of hydraulic regime. According to Prado and Fair,[13] the dispersion above a sieve tray can be divided, vertically, into three sections: the one closest to the holes, where the gas may be either at jetting or bubbling condition; the bulk froth zone, at the middle section, which is composed of gas bubbles dispersed in the liquid; and the spray zone, at the top section, which is a continuous gas. Until then, the correlation established by Chan and Fair[7] was the only one developed using more experimental data and the most recommended in literature for using with valve trays.[14]

Garcia and Fair[2] extended the application of the zones model devised by Garcia[11]/Garcia and Fair[12] to dualflow trays. Until then, the most recent attempt of modeling the efficiency of this type of tray had been made by Xu et al.[1]

More recently, Syeda et al.[15] proposed a fundamental model to predict point efficiency based on the hydrodynamics of a small experimental sieve tray column. This model was tested against the FRI efficiency data of hydrocarbon systems within $\pm 10\%$ and, according to these authors, is able to predict the trend of tray efficiency from weeping to the flooding point more closely than any other model.

The main goal of this work is to establish appropriated ways for estimating the overall efficiencies of industrial distillation columns with valve trays, downcomer, and dualflow trays. The equation sequence and correlations of the methods described by Garcia[11]/Garcia and Fair,[12] Xu et al.,[1] and Chan and Fair[7] were implemented in the Aspen Plus 12.1 simulator for the prediction of overall efficiency of columns with dualflow trays and valve trays with downcomer, making these methods available for industrial use and also to compare the overall efficiency predicted by such models with the overall efficiency evaluated from plant data of industrial columns. In addition, a modification in the evaluation of the FJ parameter (fraction of active holes that are jetting) by Garcia[11] was proposed. Finally, a comparison was made to define the best method to be used according to the tray type.

Model Development

This section presents the theoretical model proposed by Prado and Fair[13] and improved by Garcia[11]/Garcia and Fair.[12] A modification in the calculation of the jetting volume fraction (FJ) proposed in the model of Garcia[11]/Garcia and Fair[12] is also presented. The proposed methodology is based on these works because they take into account geometric parameters of the tray, tray hydrodynamics, physical properties, and interaction among the components, not considered in most available methods, which is desirable for design purposes. Besides, the theoretical basis of the method allows a better understanding of the phenomena and complex behaviors occurring on the tray.

The predicted values of efficiency using this model are compared, in the next section, to the model of Chan and Fair[5] for valve trays with downcomer and to the model of Xu et al.[1] for dualflow trays.

Model structure

The modeling suggested by Prado and Fair[13] for water–air systems and improved by Garcia[11]/Garcia and Fair[12] for organic systems divides the dispersion above the sieve tray into six zones, as shown in Figure 1. At any given condition,

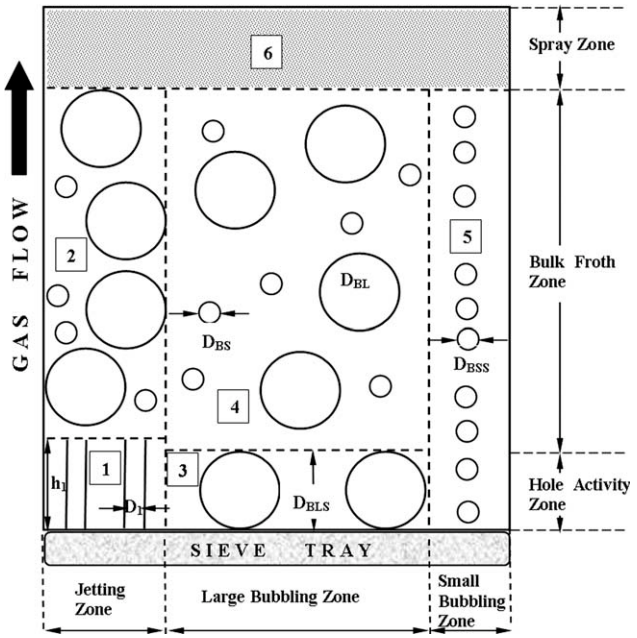

Figure 1. Hydraulic model of the dispersion above a sieve tray (Prado and Fair[13]).

zones are characterized by one of the following types of vapor flow dispersion: jetting, large bubbling, or small bubbling. The size of each of these zones varies according to the liquid and vapor volumetric flow rate.

Each zone is modeled separately in terms of the number of mass transfer units in the gas and liquid phases (N_G e N'_L). Each zone makes a contribution to the prediction of point efficiency (E_{OG}) that is determined by Eq. 1, where N_{GFJ} and N_{GFLB} are obtained using Eqs. 2 and 3, respectively:

$$E_{OG} = \mathrm{FJ}(1 - e^{-N_{GFJ}}) + \mathrm{FLB}(1 - e^{-N_{GFLB}}) + \mathrm{FSB}(1 - e^{-N_{G5}}) \quad (1)$$

$$N_{GFJ} = N_{G1} - \ln\{1 - [\mathrm{AJ}(1 - e^{-N_{G2S}}) + (1 - \mathrm{AJ})(1 - e^{-N_{G2L}})]\} \quad (2)$$

$$N_{GFLB} = N_{G3} - \ln\{1 - [\mathrm{AJ}(1 - e^{-N_{G4S}}) + (1 - \mathrm{AJ})(1 - e^{-N_{G4L}})]\} \quad (3)$$

The correlation for estimating FSB is given by Eq. 4, FJ is obtained as $\mathrm{FJ}_{\mathrm{modified}}$ through Eqs. 5 and 6, and FLB is given by Eq. 7.

$$\mathrm{FSB} = 165.65 d_H^{1.32} \phi^{1.33} \quad (4)$$

$$\mathrm{FJ}_{\mathrm{modified}} = \frac{u_A 0.60}{u_{AT}} \quad (5)$$

$$u_{AT} = 0.04302 \rho_G^{-0.5} \rho_L^{0.692} \sigma^{0.06} \phi^{0.25} \left(\frac{Q_L}{L_W}\right)^{0.05} d_H^{-0.1} \quad (6)$$

$$\mathrm{FLB} = 1 - \mathrm{FJ} - \mathrm{FSB} \quad (7)$$

The $\mathrm{FJ}_{\mathrm{modified}}$ calculation given in Eq. 5 is suggested in this work to be used in place of the original FJ calculation of the model of Garcia[11]/Garcia and Fair.[12] It is in accordance with the statement made by Johnson (apud Prado et al.[16]) that 60% of the tray holes are under jetting condition, when the superficial gas velocity based on active area is at the transition point from froth to spray condition (u_{AT}). The correlation for the u_{AT} calculation was presented by Johnson (apud Prado et al.[16]) and is given in Eq. 6.

The original correlation suggested by Prado and Fair[13] for the calculation of FJ, which was used by Garcia[11]/Garcia and Fair,[12] is shown in Eq. 8.

$$\mathrm{FJ} = \frac{u_A - u_{A,0}}{u_{A,100} - u_{A,0}} \quad 0 < \mathrm{FJ} < 1 \quad (8)$$

where $u_{A,0}$ is the superficial gas velocity based on active area at 0% jetting, and $u_{A,100}$ is the superficial gas velocity based on active area at 100% jetting. According to Prado et al.,[16] the superficial gas velocity (u_A) based on active area at minimal and maximum jetting condition can be calculated using the empirical correlations given in Eqs. 9 and 10:

$$u_{A,0} = 0.1 \rho_G^{-0.5} \rho_L^{0.692} h_W^{0.132} d_H^{-0.26} \phi^{0.992} \left(\frac{Q_L}{L_W}\right)^{0.27} \quad (9)$$

$$u_{A,100} = 1.1 \rho_G^{-0.5} \rho_L^{0.692} h_W^{0.132} d_H^{-0.26} \phi^{0.992} \left(\frac{Q_L}{L_W}\right)^{0.27} \quad (10)$$

The relationships used to calculate N_G and N'_L in each zone, which are based on the two-film mass transfer model, are: $N_G = k'_G a' t_G$ and $N'_L = k'_L \overline{a} t_L$.

Residence times are calculated using the jet and bubble ascendance velocities and the froth height on the tray (h_F). For the h_F calculation of dualflow trays, the correlations given in Garcia and Fair[2] were used; for valve trays, the Dhulesia (apud Lockett[17]) and Todd and VanWinkle (apud Lockett[17]) correlations were used.

Mixture models

Once the point efficiency (E_{OG}) is obtained, it is necessary to relate it to the Murphree tray efficiency (E_{MV}). The simulator uses the Murphree tray efficiency in the gas phase in the calculation of the column. In dualflow trays (counter-current flow), the point efficiency is the same as the Murphree tray efficiency, that is, Peclet number is zero. For the valve trays, as recommended by Lockett,[17] the mixture model developed by Lewis was used. It is given by Eq. 11, if $\mathrm{Pe} \geq 20$ (plug flow), or by Eq. 12, if $\mathrm{Pe} < 20$ (partial mixture).

$$E_{MV} = \frac{(e^{\lambda E_{OG}} - 1)}{\lambda} \quad (11)$$

$$\frac{E_{MV}}{E_{OG}} = \frac{1 - [e^{-(\eta + \mathrm{Pe})}]}{(\eta + \mathrm{Pe})\left[1 + \frac{(\eta + \mathrm{Pe})}{\eta}\right]} + \frac{e^{\eta} - 1}{\eta\left[1 + \frac{\eta}{(\eta + \mathrm{Pe})}\right]} \quad (12)$$

Although the two component efficiencies in a binary mixture are identical, in a multicomponent mixture, the

Table 1. Columns Specifications

	(A) C6 Column	(B) Butene-1 Column	(C) Propylene Column
Number of trays	60	138	224
Feed tray	30	100	140
Trays spacing, TS (m)	0.5	0.4	0.4
Light key component	Benzene	butene-1	propylene
Heavy key component	Toluene	*n*-Butane	propane
Equation of state/Activity coefficient model	NRTL	SRK	RK-SOAVE
Data base to binary interaction parameters	Aspen – VLE LIT e GMEHLING et al.	Aspen – Ethylene	Aspen – Pure 12
Feed flow rate (kg/s)	6.86	4.40	6.65
Vapor product flow rate (kg/s)		0.056	
Distillate flow rate (kg/s)	3.96	1.46	6.34
Bottom product flow rate (kg/s)	2.91	2.88	0.31
Reflux rate (kg/s)	4.22	32.60	116.57
Reboiler heat duty (GJ/s)	0.208	0.700	2.443
Top pressure (MPa)	0.029	0.664	1.147
Bottom pressure (MPa)	0.062	0.785	1.255
Feed temperature (°C)	127.0	35.0	33.1
Compounds	Nitrogen, cycle-pentane, *n*-hexane, methyl-cyclepentane, benzene, *n*-heptane, methyl-cycle-hexane, toluene, *n*-Propyl-cycle-pentane, *o*-xylene, *m*-xylene *p*-xyleno, ethyl-benzene, ethyl-heptane, *m*-ethyltoluene, (1,3)-dimethyl-4-ethylbenzene, *n*-undecane, water	Hydrogen, methanol, propane, *n*-butane, 1-butene, *cis*-2-butene, *Trans*-2-butene, (1,3)-butadiene, isobutane, isobutene, cyclebutane, isopentane, MTBE, DME, water	Ethane, ethylene, propane, propylene, cycle-propane, propadiene, methylacetylene, isobutane, Green-oil = hexane+C12
Composition (%)	0.0125, 0.0500, 2.0000, 6.0200, 48.6467, 0.6200, 2.2700, 22.7000, 0.8500, 1.5300, 3.1000, 1.0500, 9.0000, 0.4700, 1.0100, 0.3500, 0.3100, 0.0033	0.0700, 0.0030, 0.0300, 11.1000, 41.4963, 10.8700, 33.5000, 0.0017, 2.3500, 0.1100, 0.2900, 0.0510, 0.0080, 0.0850, 0.0350	0.0018, 0.00001, 4.6410, 95.2754, 0.0570, 0.0150, 0.0065, 0.00035, 0.0029
Composition base	wt	mol	mol

component efficiencies are usually all different. The treatment of such a mixture as a pseudo-binary mixture based on two key components is by far the most common procedure used in practice, according to Lockett,[17] who describes the pseudo-binary approach used here to calculate the slope of equilibrium curve (m), where the compositions of the light and heavy key components in the liquid and vapor phases were taken into consideration. The slope of equilibrium curve is used to calculate λ.

Liquid entrainment and weeping

The liquid entrainment in vapor reduces the tray efficiency because it represents an internal recirculation of the liquid. For dualflow trays, the effect of entrainment on efficiency was considered using the relationship suggested by Garcia and Fair[2] as function of the flooding factor (FF) and tray capacity (C_{SB}), whereas the Colburn's equation (apud Lockett[17]) together with the Zuiderweg[8] correlation were used for valve trays.

In relation to the effect of weeping, the relationship suggested by Garcia and Fair[2] was used for dualflow trays, whereas for valve trays the effect of weeping on the tray efficiency was not considered because "[…] there is not an equation that is conveniently simple and analogous to the one suggested by Colburn (1936), designed to determine the apparent efficiency under weeping conditions" (Lockett,[17] p. 175). In this work, only the weeping flow rate was determined, which is expected to be low, or even nonexistent, in the case of valve trays.

Methodology

The following three commercial distillation columns were used as reference in this study and have the specifications given in Table 1:

(A) C6 Fractionator: its function is to separate a C5–C9 hydrocarbon mixture, in a C6 cut at the top (rich in benzene) and a C7 + cut at the bottom.

(B) Butene-1 Fractionator: its function is to separate a C4 hydrocarbon mixture, producing butene-1 at the top with a minimal amount of *n*-butane.

(C) Propylene Fractionator: its function is to separate high purity propylene at the top from propane.

In the case of the C6 Fractionator, several activity coefficient models with different binary parameters were tested to compare the actual temperature profile with those obtained in the simulator and the expected Murphree efficiency. It was chosen the NRTL model, which describes the equilibrium liquid–vapor and liquid–liquid solutions of strongly nonideal. This model requires binary parameters that were obtained from literature (Gmehling et al.[18]) and from linear regression of experimental data, already included in the database of the simulator Aspen Plus 12.1 (called VLE LIT).

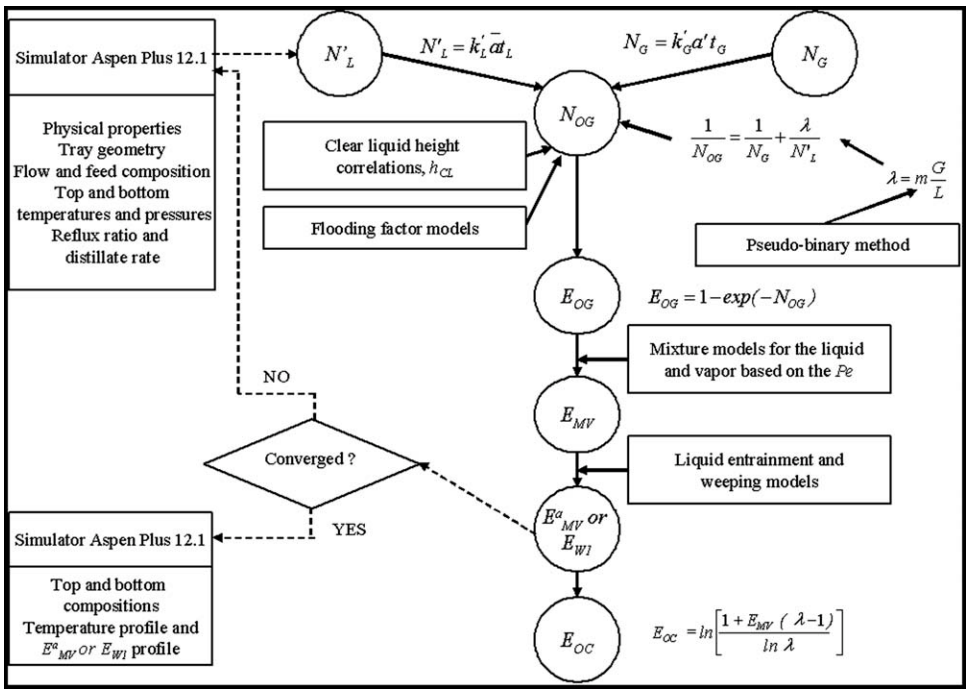

Figure 2. Algorithm implemented in the Aspen Plus 12.1 simulator to calculate the apparent Murphree vapor phase tray efficiency (E^a_{MV} or E_{W1}).

Regarding the Butene-1 Fractionator, the best thermodynamic package was chosen by checking the temperature profile and overhead and bottom compositions obtained via simulation against the corresponding actual data of commercial scale columns, checking the Murphree efficiency expected for this type of tray (dualflow), comparing the data of liquid–vapor equilibrium obtained from the simulation with the work of Carmichael et al.,[19] and following the recommendations of the technical support of the Aspen simulator 12.1. It was chosen the equation of state of Soave-Redlich-Kwong (SRK option in the simulator Aspen 12.1, with the database Ethylene for the parameters of binary interaction).

In the case of the Propylene Fractionator, the choice of the equation of state was based on the profiles of temperature and Murphree efficiency expected for this type of tray (dualflow). The results obtained with the equation of state of Soave-Redlich-Kwong option RK-SOAVE - Pure 12 were those that best agreed with the aforementioned plant data. In the Table 1, it is shown a summary of thermodynamic choices.

The Aspen Plus 12.1 simulator was used for carrying out the simulations. The implementation of the efficiency models, correlations for froth height on the tray, correlations for liquid entrainment and weeping, and mixture models of liquid on the tray was done in Fortran 77 programing language in the calculator block, and the algorithm used is summarized in Figure 2. More details and the full program can be found in Domingues.[20]

Results and Discussion

The results of overall column efficiency prediction as function of the flooding factor for the C6 (valve tray), Butene-1 (dualflow tray) and Propylene fractionators (dualflow tray), and the overall efficiency of the corresponding industrial columns, obtained by adjustment to plant data are given in Figures 3–5, respectively. For the valve tray column, Figure 3, the prediction of overall column efficiency was done using the Garcia[11] model, the modified Garcia model (using the $FJ_{modified}$ as proposed in this work), and the Chan and Fair[7] model developed for crossflow tray. For the dualflow columns of Figures 4 and 5, the prediction of overall column efficiency was done using the same models (but with the appropriated corrections for liquid entrainment and weeping), except the last one, which was replaced by the Xu et al.[1] model, developed for counter-current flow tray.

In all these figures, the effect of the proposed modification in the calculation of the jetting volume fraction (FJ) can be

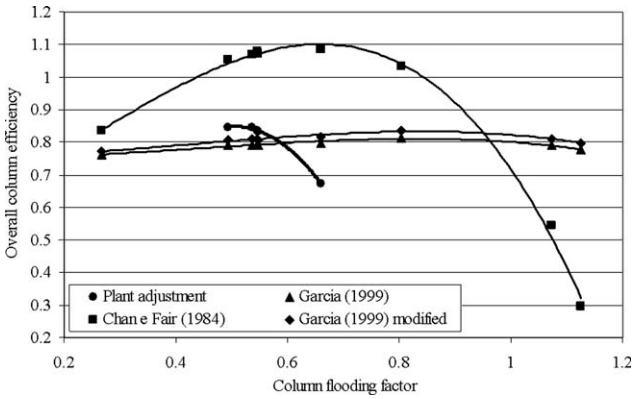

Figure 3. Comparison of adjusted and predicted overall efficiency of C6 column (valve tray).

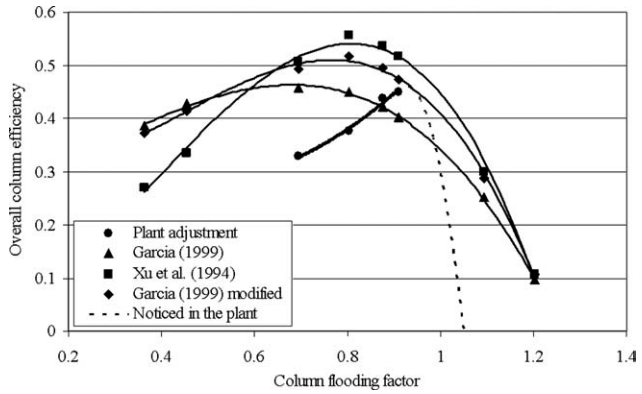

Figure 4. Comparison of adjusted and predicted overall efficiency of the butene-1 column (dualflow tray).

checked against the original model of Garcia,[11] using the overall efficiency prediction results. According to several industrial column data, the proposed modification in the Garcia[11] model was a significant improvement in the original model for the dualflow trays (Figures 4 and 5).

In relation to the C6 column (valve tray, Figure 3), there was a slight modification in the prediction of the column overall efficiency.

For the Butene-1 column (dualflow tray, Figure 4), the models did not represent appropriately the expected sudden loss of efficiency above the 1.0 flooding factor, according to the actual behavior of the Butene-1 column represented in dotted line. When a comparison is made between the estimated overall efficiency by the application of the original Garcia[11] model and the modified as suggested in this work, both have produced similar results, although the modified Garcia[11] model was more accurate at the efficiency peak.

For the Propylene column (dualflow tray, Figure 5), the $FJ_{modified}$ improved the overall efficiency estimation (E_{OC}) when compared with the Garcia[11] model.

Finally, parity plots of overall efficiency observed in the plant versus the predicted from each model, given in Figures

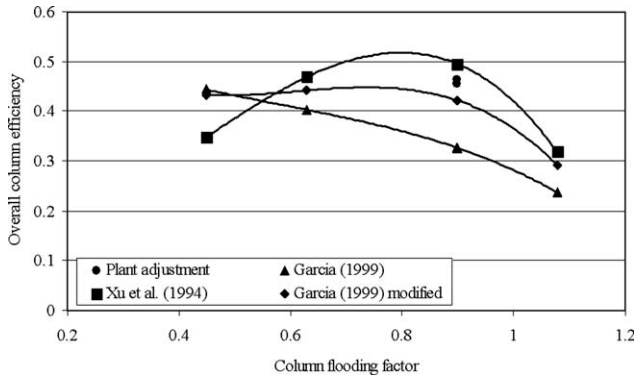

Figure 5. Comparison of adjusted and predicted overall efficiency of the propylene column (dualflow tray).

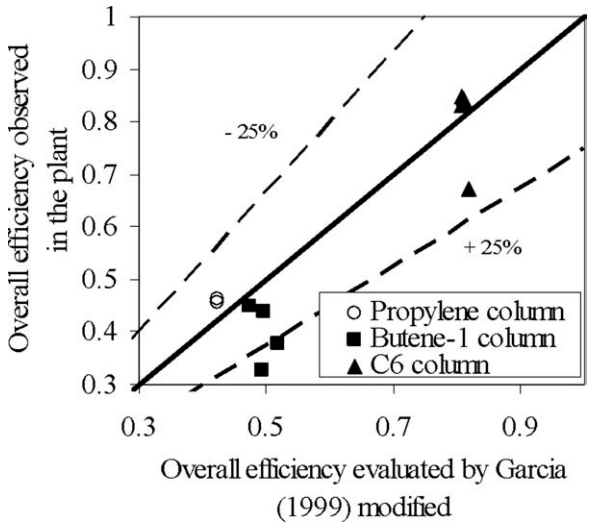

Figure 6. Parity plot, observed in plant versus evaluated overall efficiency calculated using modified Garcia model.

6–8, show the compiled full data from the three columns. Limits of ±25% are shown in these plots, which are considered reasonable limits for the adjustment of such diverse plant data, corresponding to different geometries, operating conditions, analytical procedures, nonideal thermodynamic behavior, deviations in material balances, and other small deviations. It can be observed that with the Garcia modified model, only 1 among 11 points is located out of limits. With the Garcia[11] model, without the modification in the FJ parameter, 3 points are out of limits. Finally, for the more used models of Chan e Fair[7] for valve trays, and Xu et al.[1] for dualflow trays, 3 points are out of ±25% limits.

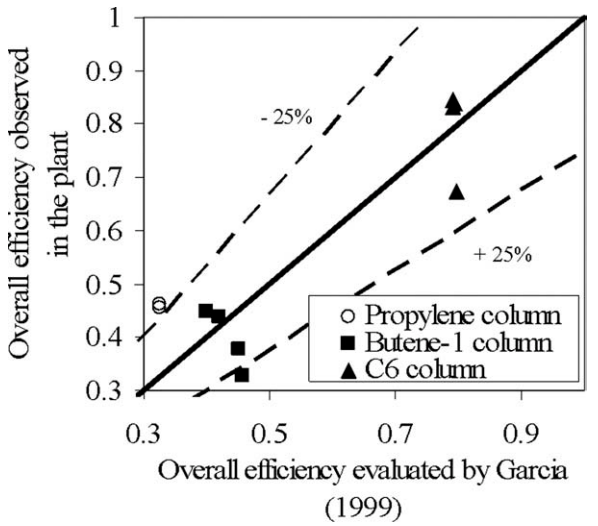

Figure 7. Parity plot, observed in plant versus evaluated overall efficiency calculated using Garcia[11] model.

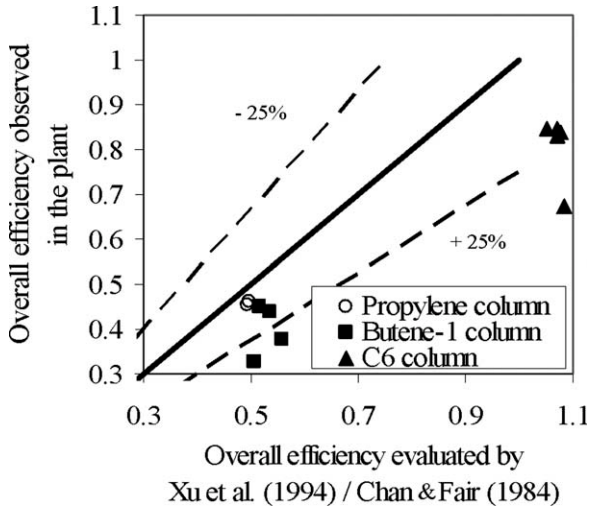

Figure 8. Parity plot, observed in plant versus evaluated overall efficiency calculated using Chan and Fair[7] and Xu et al.[1] models.

Conclusions

The modified Garcia model for estimating the efficiency by contribution zones of mass transfer proposed in this work coupled with the modifications suggested by Garcia e Fair[2] for dualflow trays have successfully represented the dualflow tray column performance (counter-current flow). This is one of the first attempts to apply the Garcia[11]/Garcia and Fair[12] model with the modifications suggested by Garcia and Fair[2] to dualflow tray columns. The modification on the FJ parameter calculation seemed to be coherent with the jetting fraction expected for such columns. Furthermore, this is also one of the first attempts for using the correlations suggested by Garcia and Fair[2] for estimating the tray capacity, liquid entrainment, and weeping in dualflow tray columns.

The performance of butene-1 and propylene distillation columns were better represented by the modified model suggested in this work than by the previous models given by Garcia[11]/Garcia and Fair[12] and Xu et al.[1]

It should also be mentioned that for dualflow tray columns, the modified Garcia model represented the efficiency peak closer to the overall column efficiency obtained from plant data adjustment than using the original FJ suggested by Prado and Fair[13] and used by Garcia.[11] The peak efficiency is in the flooding factor range from 0.9 to 1.0 in which the studied dualflow columns runs during almost 99% of time, so it is the most important operating range for the prediction of the overall column efficiency.

The use of the modified Garcia[11] model for the C6 column (valve tray) still deserves a certain reserve, as it overestimate the FJ and the original Garcia[11] model underestimate this factor. Both models were better than Chan and Fair[7] model. The satisfactory results in efficiency prediction of the new model developed for crossflow tray by Syeda et al.[15] suggest its inclusion in further studies regarding the C6 column.

Another important contribution of this work was the implementation, in the Aspen Plus 12.1 simulator, of the algorithm to calculate the apparent Murphree vapor phase tray efficiency (E_{MV}) that can be used to estimate efficiencies in new similar distillation column designs in industry.

Notation

A_A = active or tray bubbling area ($A_T - 2A_D$), m^2
A_H = hole area of the tray, m^2
AJ = fraction of small bubbles present in the bulk froth zone
a', $\overline{a}$ = Geometrical interfacial area per volume of vapor and liquid, $m^2\ m^{-3}$
C_{SB} = tray capacity, $m\ s^{-1}$
De = Eddy diffusivity in the liquid phase, $m^2\ s^{-1}$
d_1 = jet diameter, m
d_{BL}, d_{BS} = arithmetic mean bubble diameter of large and small bubbles in the zones 2 e 4, m
d_{BLS} = Sauter mean bubble diameter of large bubbles in the zone 3, m
d_{BSS} = Sauter mean bubble diameter of small bubbles in the zone 5, m
d_C = column diameter, m
d_H = hole diameter, m
E_{OC} = overall column efficiency
E_{OG} = point efficiency
E_{MV} = Murphree vapor phase tray efficiency
E^a_{MV} = apparent Murphree vapor phase tray efficiency, accounting the effects of entrainment and weeping of liquid in the valve trays
E_{W1} = apparent Murphree vapor phase tray efficiency, accounting the effects of entrainment and weeping of liquid in the dualflow trays
FF = flooding factor
FJ = fraction of active holes that are jetting
FLB = fraction of active holes that are issuing large bubbles
F_S = superficial F factor based on A_A ($u_A\rho_G^{0.5}$), $kg^{0.5}\ m^{-0.5}\ s^{-1}$
FSB = fraction of active holes that are issuing small bubbles
h_1 = jet height, m
h_{CL} = clear liquid height, m
h_F = froth height, m
h_W = weir height or wave height in dualflow trays, m
k'_G, k'_L = vapour and liquid phase mass-transfer coefficient, $m\ s^{-1}$
L_W = weir length, m
M_G, M_L = vapor and liquid mass flow rate, $kg\ s^{-1}$
MW_G, MW_L = molecular weight of vapor and liquid mixture, $kg\ kmol^{-1}$
m = slope of equilibrium curve, dy/dx
N_G = number of vapor phase mass-transfer units: N_{G1} (Zone 1); N_{G2L} (Zone 2, large bubbles); N_{G2S} (Zone 2, small bubbles); N_{G3} (Zone 3); N_{G4L} (Zone 4, large bubbles); N_{G4S} (Zone 4, small bubbles); N_{G5} (Zone 5); N_{G6} (Zone 6); N_{GFJ} (Zones 1, 2 e 6); N_{GFLB} (Zones 3, 4 e 6); N_{GFSB} (Zones 5 e 6)
N'_L = number of liquid phase mass-transfer units
Q_G, Q_L = vapor and liquid volumetric flow rate, $m^3\ s^{-1}$
Ts = tray spacing, m
t_G = mean residence time of gas in dispersion, s
t_L = mean residence time in the liquid phase, s
u_A = superficial gas velocity based on A_A or A_b, $m\ s^{-1}$
u_{AT} = superficial gas velocity based on active area at transition point from froth-to-spray, $m\ s^{-1}$
$u_{A,0}$, $u_{A,100}$ = superficial gas velocity, based on active area at 0% and 100% jetting, $m\ s^{-1}$
Pe = Peclet number, dimensionless: $\mathrm{Pe} = \left(\frac{M_L}{MW_L}\right) d_C \Big/ L_W h_{CL} \left(\frac{\rho_L}{MW_L}\right) \mathrm{De}$

Greek letters

ε, α = vapor and liquid holdup fraction
λ = ratio of slope of equilibrium curve to slope of operating line: $\lambda = m\left[\left(\frac{M_G}{MW_G}\right) \Big/ \left(\frac{M_L}{MW_L}\right)\right]$
μ_G, μ_L = Vapor and liquid viscosity, Pa s = $kg\ m^{-1}\ s^{-1}$ = N s m^{-2}

ρ_G, ρ_L = vapor and liquid density, kg m^{-3}

σ = surface tension, N m^{-1}

ϕ = fractional free area (A_H/A_A)

η = constant of Eq. 12: $\eta = \frac{Pe}{2}\left[\sqrt{1+\frac{4\lambda E_{OG}}{Pe}} - 1\right]$

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