Mechanistic Model for Structured-Packing-Containing Columns: Irrigated Pressure Drop, Liquid Holdup, and Packing Fractional Wetted Area

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An implicit one-dimensional two-zone two-fluid mechanistic model was developed for the prediction of the irrigated two-phase pressure drop, the total liquid holdup, and the packing fractional wetted area in gas−liquid countercurrent columns containing structured packings and operated in the preloading zone. The model was an offshoot of the well-known “single-slit” mechanistic approach to cocurrent down-flow trickle-bed reactors. It mimicked the actual bed void by means of two hypothetical, recurrent, and geometrically similar inclined slits consisting of a dry slit and a wet slit. This mechanistic model required *no single* adjustable parameter and proved powerful in the prediction of the column hydraulics under various operational conditions such as atmospheric scrubbing or high-pressure/temperature distillation conditions. In this context, a collection of data relative to the irrigated pressure drop, liquid holdup, and packing fractional wetted area obtained under low/high pressure/temperature has been compiled from the literature for columns equipped with structured packings and operated below the loading point and under partial wetting. This databank provided pertinent information for successful validation of the model. The satisfactory results obtained highlight the breadth of applicability of the proposed approach, especially for new designs or for optimal rating of existing equipment.

Introduction

In recent years, environmental concerns such as the greenhouse effect, acid rain and malodorous emissions from the chemical industries have triggered criticisms from the public urging governmental and international observatory institutions to impose increasingly strict regulations on pollution discharge. As the “tracked” pollutants are more or less water-soluble, packed towers, because of their cheap operation, are very appealing devices for accomplishing pollution abatement tasks. Packed towers with gas−liquid countercurrent flows are indeed omnipresent in such conventional areas as sour gas scrubbing and stripping, but also in distillation and rectification.[1] Furthermore, such devices are increasingly evoked as promising configurations in catalytic distillation and petroleum refining operations.[2,3] The permanent market demand in the pursuit of cost-effective units is continually imposing the need for well-designed and highly efficient packings. This trend is created by the need to respond to industrial economic requirements (e.g., high mass-transfer capacities, low pressure drops) and, at the same time, to comply with the stringent discharge tolerances.

As opposed to their classical random packing homologues, structured packings permit the attainment of high mass-transfer efficiency at lower power consumption, thus entraining an economy in column scale and in operating costs. Althoughstructured packings have been known for almost 30 years, only a limited amount of experimental data has been released in the open literature, especially regarding their hydraulic and mass-transfer capacity behavior, and current models for predicting their performance have not been validated adequately.[4] This state of affairs is due, on one hand, to the reluctance of the packing suppliers to disclose their proprietary knowledge. On the other hand, the relative novelty of these packings makes the available body of information far less abundant than that inherited for random packings.

To reduce operating costs, the industry requires packings having high specific surface areas, high mass-transfer efficiencies, and low-pressure drops. The growing problem is to suggest reliable conceptual design models for the evaluation of the hydrodynamic and the mass-transfer parameters of structured packings. A model often used for predicting mass-transfer and hydrodynamic performance of structured packings was developed several years ago at the Separations Research Program (SRP) of The University of Texas at Austin.[5−8] The basis for the model is a modified wetted-wall flow arrangement through the channels of the packing. An alternate model has been developed more recently at Delft University of Technology[9−11] and is also founded on liquid film flow down inclined corrugated plates. Several packing macro-geometry characteristics, having an effect on performance, have been explicitly incorporated in the model. Both models are valid in the preloading zone where the gas−liquid interaction is minimal. The comprehensive SRP model led to satisfactory prediction results, but because of the large number of adjustable parameters, its application in ranges

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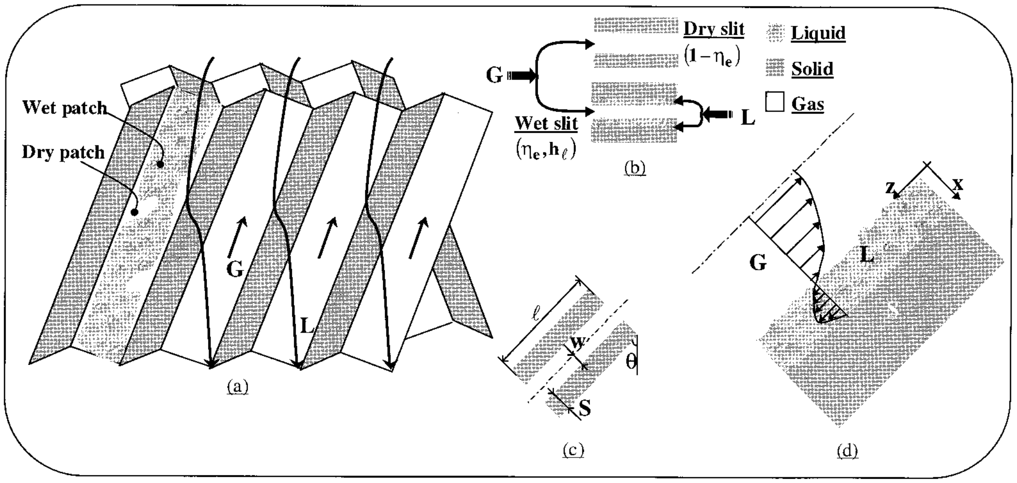

Figure 1. (a) Actual and (b) conceptual gas–liquid flow pattern in partially wetted structured packing, (c) slit geometrical configuration, and (d) details of the gas–liquid flow structure in the wet slit.

outside the region in which the parameters were tuned can lead to sizable errors.[12] The Delft model assumes *complete wetting* of the packing surface, and thus, the liquid holdup is expressed as a function of an average liquid film thickness and packing surface area. As an alternative to these two approaches, Brunazzi and Paglianti[12] developed a mechanistic model based on solving mass- and momentum-conservation equations in an idealized wetted-wall arrangement approximating the actual geometry of the structured packing. The model views the channels within the packing as a bundle of identical columns inclined with respect to the horizontal axis by an angle equal to the corrugation angle. The model relied on several assumptions including (i) the packing fractional wetted area is unaffected by the gas load and gas–liquid interfacial shear stress, (ii) the interfacial shear stress is related to the effective liquid velocity and not to the interfacial velocity, and (iii) dynamic liquid holdup is known a priori.

Some common features shared by these approaches are that they treated distinctly and separately the issue of prediction of liquid holdup and irrigated pressure drop[13–16] from the issue of prediction of gas–liquid interfacial area.[17–19] They are semiempirical in nature and based on the assumption of liquid flow over an inclined wall. Provided that no ripples develop at the gas–liquid interface, as is the case in the preloading zone, the gas–liquid interface is admittedly assumed to be smooth, so that most mass-transfer correlations agree in identifying the area available to the mass transfer as the packing fractional wetted area of the geometrical surface.

The objective of this work is to develop a two-fluid mechanistic model for gas–liquid countercurrent structured packings wherein the interrelationship between the irrigated pressure drop, the total liquid holdup and the packing fractional wetted area is, for the first time, *quantitatively* highlighted. It further allows for the *simultaneous* prediction of these three flow variables in the preloading zone. The approach is designed to approximate the actual two-phase flow topography in structured packings using two inclined and interconnected slits consisting of a dry slit solely fed by gas and a gas–liquid slit fed by liquid and remaining gas. A comprehensive model validation has been performed most satisfactorily using published literature data relative to the fractional wetted area, pressure drop, and liquid holdup obtained under various operational contexts relevant to scrubbing and distillation hydraulics in low/high pressure/temperature conditions.

Model Development

Conceptual Gas–Liquid Topography. Typically, the countercurrent gas–liquid flow structure over partially wetted corrugated sheets (or gauzes) of a structured packing, especially at low irrigation liquid flow rates, can be decomposed into two elementary contacting patterns, dry and wet patches on the solid support (Figure 1a). This configuration is typically encountered in the preloading zone and consists of (i) a dry region having a fractional area of $1 - \eta_e$ with gas–solid contact and (ii) a wet region composed of the liquid film (holdup, h_l) with a fractional area equal to η_e.

This simple dichotomy of the flow structure highlights some key hydrodynamic features, the prediction of which plays a key role in descriptions of the hydrodynamics of a structured packed column. According to Figure 1b, there are three variables that are interlinked to each other: the packing fractional wetted area, η_e; the total liquid hold-up, h_l; and the overall irrigated pressure drop, $-\Delta P/H$.

Structured Packing Geometrical Approximate Model. A two-zone mechanistic model is proposed herein with the attempt to forecast simultaneously the three aforementioned hydrodynamic variables. The basic idea behind this approach consists of conceptualizing the gas–liquid structure depicted in Figure 1a using the double-slit architecture sketched in Figure 1b. This approach is borrowed from the hydrodynamic modeling of cocurrent down-flow randomly packed trickle-bed reactors.[20,21] It postulates that the bed-scale macroscopic hydrodynamics can be approximated and mapped from knowledge of the gas–liquid flow distribution and the hydrodynamics occurring on a representative element of the structured packing. An intuitive repetitive element to be taken as a reference can be the interstitial space confined between the viewing faces of two immediate corrugated sheet neighbors and the flow developing between them (Figure 1a).

Thus, the double-slit architecture is represented by a recurrent assemblage of two parallel inclined and interconnected slits: a gas slit representing a dry region fed by a fraction of the gas flow rate, region $a_s(1 - \eta_e)$, and a gas−liquid slit representing region $a_s\eta_e$ moistened by the totality of the liquid flow rate (holdup, h_l) in the form of a constant-thickness (δ) liquid film, where a second remaining fraction of the gas flows countercurrently in the central core.

The mapping between the slits and the structured packing column is based on the following assumptions: (i) The slits described above have identical half-wall thicknesses S, half-void thicknesses w, and stream-wise slit lengths l (see Figure 1c). The widths of the dry and wet slits are proportional to $1 - \eta_e$ and η_e, respectively. (ii) The solid surface area per unit solid volume is assumed to be uniform across the two slits and equal to that of the bed

$$S = \frac{1-\epsilon}{a_s} \tag{1}$$

(iii) The void-to-solid and liquid-to-solid volume fractions within the slits and the whole packed bed are also considered to be identical

$$\frac{w}{S} = \frac{\epsilon}{1-\epsilon} \quad \text{for both slits} \tag{2}$$

$$\frac{\delta\eta_e}{S} = \frac{h_l}{1-\epsilon} \quad \text{for the wet slit} \tag{3}$$

(iv) The slits are inclined by an angle θ corresponding to the corrugation angle of the structured packing (with respect to the vertical axis, see Figure 1c). Thus, the quantity $1/(\cos\theta)$ is a direct measure of the bed tortuosity, or the ratio between the actual flow path and vertical bed height.

Irrigated Pressure Drop, Liquid Holdup, Wetted Fraction Model. Further assumptions on the nature of the gas−liquid flow are also made as follows:

(i) The flow regime in the bed corresponds to the *preloading zone*. Therefore, no ripples (or waves) occur at the gas−liquid interface, thus leading to a smooth and stable liquid film. Note that these ripples and interfacial instabilities are more likely to occur in the *loading zone*, where the packing wetting phenomenon is no longer an issue.

(ii) The dominant liquid texture components are contributed by films and rivulets which are referred to as "liquid film" throughout this paper. This structure far outweighs the droplet contribution so that the packing fractional wetted area (η_e) is tantamount to the effective gas−liquid interfacial area (a_e)

$$\eta_e = \frac{a_e}{a_s} \tag{4}$$

Therefore

$$h_l = a_s\eta_e\delta = a_e\delta \tag{5}$$

(iii) The (Newtonian) liquid film and gas flows remain acceleration-free and at steady state.

(iv) The average phase interstitial velocities in the bed and the slits are assumed to be identical (the index z stands for the stream-wise flow in idealized slit architecture, Figure 1c)

$$\text{Dry slit} \quad \langle u_g\rangle = \frac{v_{sg}}{\epsilon\cos\theta} \tag{6}$$

$$\text{Wet slit} \quad \langle u_g\rangle = \frac{v_{sg}}{(\epsilon - h_l/\eta_e)\cos\theta} \tag{7}$$

$$\text{Wet slit} \quad \langle u_l\rangle = \frac{v_{sl}}{h_l\cos\theta} \tag{8}$$

(v) The friction factor (f_i) at the gas−liquid interface and the wall friction factor (f_w) at the fluid−solid interface are equal

$$f_i = f_w \tag{9}$$

[20]

(vi) There are no discontinuities in the velocity and shear stress profiles at the gas−liquid interface

$$\text{Velocity continuity} \quad u_{ig} = u_{il} \quad \text{at } x = 0 \tag{10}$$

$$\text{Shear stress continuity} \quad \tau_{ig} = \tau_{il} \quad \text{at } x = 0 \tag{11}$$

(vii) The total pressure gradient is the same across the bed and the idealized slit network

$$-\frac{\Delta P}{H} - \rho_g g = \frac{1}{\cos\theta}\left(-\frac{dP}{dz} - \rho_g g\right) = \rho_g g\Psi_g \tag{12}$$

$$-\frac{\Delta P}{H} + \rho_l g = \frac{1}{\cos\theta}\left(-\frac{dP}{dz} + \rho_l g\right) = \rho_g g\Psi_l \tag{13}$$

$$1 + \Psi_g = \frac{\rho_l}{\rho_g}(\Psi_l - 1) \tag{14}$$

Using the above assumptions, the stream-wise projection of the gas momentum balance equation over the dry and wet slits become, respectively

$$\text{Dry slit} \quad (1-\eta_e)\tau_{wg} = \left(-\frac{dP}{dz} - \rho_g g\cos\theta\right)w(1-\eta_e) \tag{15}$$

$$\text{Wet slit} \quad \eta_e\tau_{ig} = \left(-\frac{dP}{dz} - \rho_g g\cos\theta\right)(w-\delta)\eta_e \tag{16}$$

Therefore, by combining eqs 15 and 16, the slit pressure drop expression can be written as

$$-\frac{dP}{dz} + \rho_g g\cos\theta = \frac{\eta_e\tau_{ig} + (1-\eta_e)\tau_{wg}}{(w-\delta)\eta_e + w(1-\eta_e)} \tag{17}$$

Similarly, the stream-wise projection of the momentum balance equation for the liquid is

$$\text{Wet slit} \quad -\frac{dP}{dz} + \rho_l g\cos\theta = \frac{\tau_{wl} - \tau_{il}}{\delta} \tag{18}$$

In eqs 17 and 18, $\tau_{i\alpha}$ and $\tau_{w\alpha}$ designate, respectively, the shear stress on the α-phase side of the gas−liquid interface in the wet slit, and the wall shear stress exerted by the slit wall on the α phase (wet and dry slits). Moreover, the gas−liquid interfacial stress and the wall−liquid and wall−gas shear stresses are assumed to have laminar and turbulent contributions derived from integration of the universal velocity profile within each fluid

$$\tau_{ig} = 2\frac{(u_g + u_{ig})\mu_g}{w - \delta} + f_i\rho_g(u_g + u_{ig})^2 \tag{19}$$

$$\tau_{wl} = 2\frac{u_l\mu_l}{\delta} + f_w\rho_l u_l^2 \tag{20}$$

$$\tau_{wg} = 2\frac{u_g\mu_g}{w} + f_w\rho_g u_g^2 \tag{21}$$

By means of the bed-to-slit mapping relations in eqs 1−3, 12, and 13, aided by the integration of eqs 19−21, it is possible to express eqs 17 and 18 using the structure-packing-bed framework. The resulting equivalent equations, derived after a lengthy development and finally conveniently expressed in dimensionless form, are, respectively

$$\Psi_g = \frac{32}{\cos^2\theta}\frac{1-\eta_e}{\epsilon - h_l}\left(1 + \frac{f_w Re_g}{8\epsilon\cos\theta}\right)\frac{Re_g}{Ga_g} + \frac{32}{\cos^2\theta}\frac{\epsilon^2\eta_e^2}{(\epsilon - h_l)(\eta_e\epsilon - h_l)^2}\left\{1 + \frac{f_w[\eta_e Re_g + (\eta_e\epsilon - h_l)Re_i]}{8\eta_e\epsilon\cos\theta}\right\}\frac{\eta_e Re_g + (\eta_e\epsilon - h_l)Re_i}{Ga_g} \tag{22}$$

$$\Psi_l = \frac{32\eta_e^2}{\cos^2\theta}\frac{\epsilon^2}{h_l^3}\left(1 + \frac{f_w Re_l}{8\eta_e\epsilon\cos\theta}\right)\frac{Re_l}{Ga_l} - \Psi_g\frac{\rho_g}{\rho_l}\left(\frac{\epsilon\eta_e}{h_l} - 1\right) \tag{23}$$

These two equations establish an implicit two-fluid one-dimensional mechanistic model relating the irrigated pressure drop to the liquid holdup and wetting efficiency in the preloading zone. Solutions of the above equations require a number of auxiliary relationships describing the wall friction factor, f_w; the interfacial Reynolds number, Re_i (or the interfacial velocity u_{il}); and the packing fractional wetted area, η_e.

The channel wall friction factor, f_w, in structured packings can be correlated for both laminar and turbulent conditions by taking advantage of the dry pressure drop measurements. As usually recommended, f_w can be correlated by means of an Ergun-type equation.[12,13]

Performing a z-momentum balance in the wet slit over a differential liquid element of cross-stream thickness Δx and between planes z and $z + \Delta z$ gives

$$\frac{d\tau_{l,xz}}{dx} = \rho_l g\cos\tau - \frac{dP}{dx} \tag{24}$$

subject to the boundary condition

$$\tau_{l,xz} = \left(-\frac{dP}{dz} - \rho_g\cos\theta\right)(w - \delta) \quad \text{at } x = 0 \tag{25}$$

A first integration of eq 24 using boundary condition (eq 25) yields

$$\frac{du_{l,z}}{dx} = -\frac{x}{\mu_l}\left(\rho_l g\cos\theta - \frac{dP}{dz}\right) + \frac{w-\delta}{\mu_l}\left(\rho_g g\cos\theta + \frac{dP}{dz}\right) \tag{26}$$

In addition to the boundary conditions in eqs 10 and 11 above, eq 26 also satisfies the boundary condition

$$u_{l,z} = 0 \quad \text{at } x = \delta \tag{27}$$

Integration of eq 26 leads to the velocity distribution of the liquid in the wet slit. By replacing $x = 0$, one obtains the expression for the liquid (or gas) velocity at the gas−liquid interface

$$u_{il} = \frac{h_l g\cos\theta}{\eta_e^2 a_s^2}\left(\frac{h_l\Psi_l}{2\nu_l} + \frac{(\eta_e\epsilon - h_l)\Psi_g}{\nu_l}\frac{\rho_g}{\rho_l}\right) \tag{28}$$

Double integration of eq 26 leads to the liquid-film average velocity

$$\langle u_{l,z}\rangle = \frac{1}{\delta}\int_0^\delta u_{l,z}\,dz \tag{29}$$

Using various manipulations on a combination of eqs 8 and 29, one finally obtains the last relationship for the packing fractional wetted area required for closure of the mechanistic model. A suitable derivation of the dimensionless equation expressing η_e as a function of the bed variables is

$$\eta_e = \frac{a_e}{a_s} = \frac{h_l^2\cos^2\theta}{64\epsilon}\frac{Ga_l}{Re_l}\left(\Psi_l - 1 - \frac{\rho_g}{\rho_l}\right) + \frac{h_l^2\cos\theta}{8\epsilon}\left(\frac{Ga_l}{Re_l}\right)^{1/2}\sqrt{\frac{4}{3h_l}\left[\Psi_l - \frac{3}{2}\left(\Psi_l - 1 - \frac{\rho_g}{\rho_l}\right)\right] + \frac{\cos^2\theta}{64}\frac{Ga_l}{Re_l}\left(\Psi_l - 1 - \frac{\rho_g}{\rho_l}\right)^2} \tag{30}$$

In the preloading zone, the irrigated pressure drop is small enough that, in eq 30, the approximation $\Psi_l \approx 1$ can be assumed, thus leading to the simplified wetting relationship

$$\eta_e = \frac{a_e}{a_s} = -\frac{h_l^2\cos^2\theta}{64\epsilon}\frac{\rho_g}{\rho_l}\frac{Ga_l}{Re_l} + \frac{h_l^2\cos\theta}{8\epsilon}\left(\frac{Ga_l}{Re_l}\right)^{1/2}\sqrt{\frac{4}{3h_l}\left(1 + \frac{3}{2}\frac{\rho_g}{\rho_l}\right) + \frac{\cos^2\theta}{64}\left(\frac{\rho_g}{\rho_l}\right)^2\frac{Ga_l}{Re_l}} \tag{31}$$

Method of Solution. Equations 22, 23, and 31 establish the nonlinear implicit mechanistic model as a set of three coupled algebraic equations and three unknowns, i.e., $-\Delta P/H$, h_l, and η_e. The input variables to be fed to this model are the fluid superficial velocities, the fluid physical properties, the bed and packing characteristics (void fraction, specific surface area, corrugation angle, and dry pressure drop data[12,13]), and the interfacial velocity estimated by means of eq 28. If all of these variables are known a priori, then the model requires *no adjustable* parameters. The nonlinear model is solved using an iterative Newton−Raphson algorithm.

Experimental Database

To validate the proposed model, data relative to the irrigated pressure drop, the liquid holdup, and the packing fractional wetted area in the preloading zone have been collected from the literature for columns containing the following structured packing: Flexipac, Gempak, Mellapak, Montz-Pak, and coiled screen packing[7,9,12,22−25,28−31] (see also Table 1). The experimental conditions included low and high pressures, ambient and high temperatures, superficial liquid ve-

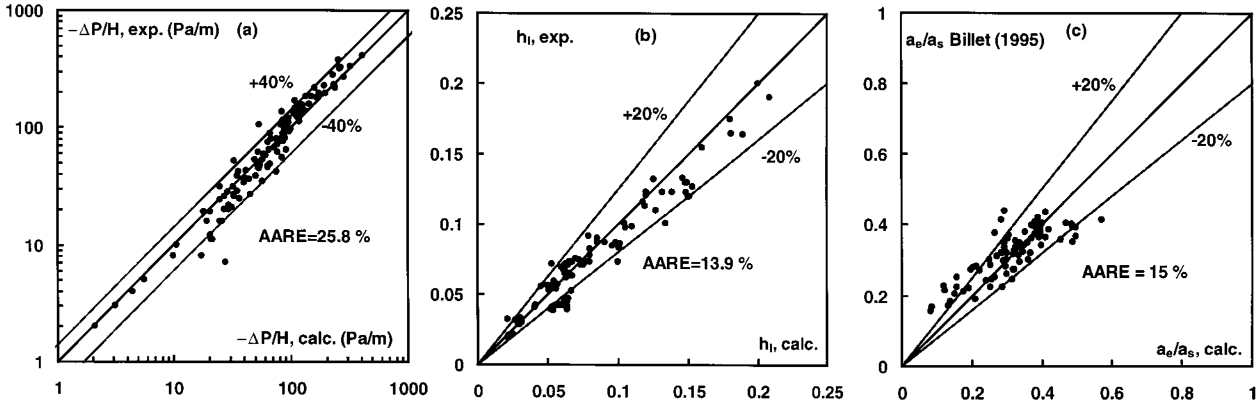

Figure 2. Predicted versus experimental (a) irrigated two-phase pressure drop, (b) total liquid holdup, (c) packing fractional wetted area (predicted versus simulated data from Fitz et al.[28] using the Billet model[27]). All data refer to the preloading zone.

Table 1. Packings and Column Characteristics

packing type	a_s (m^{-1})	ϵ (%)	angle/vert.	column Ø (m)
Flexipac 1Y	443	91.0	45	0.914
Flexipac 2Y	223	95.0	45	0.914
Flexipac 3Y	223	96.0	45	0.914
Gempak 1A	115	96.0	45	0.914
Gempak 2A	223	95.0	45	0.914
Gempak 4A	453	91.0	45	1.0
Mellapak 250X	250	98.0	60	1.0
Mellapak 250Y	250	95.0	45	1.2, 0.295
Montz-Pak B1-250	244	98.5	60	0.8
Montz-Pak B1-400	394	96.0	60	0.43
coiled screen packing	628	83.6	61.7	0.105

locities from 0.6 to 50 mm/s, and superficial gas velocities changing between 0.1 and 2.6 m/s. The fractional wetting values simulated by Fitz et al.[28] (Figure 13 in their paper) using the Billet model[27] for the distillation at total reflux of the system *i*-butane/*n*-butane on a Mellapak 250Y structured packing have been compared to the predictions of our mechanistic model.

Discussion

An examination of the parity plot diagrams, shown in Figure 2a–c, provides the ability to compare the model prediction results with the corresponding experimental data obtained in the preloading zone. As a result, it should be noted that the irrigated pressure drop and the liquid holdup are predicted with mean absolute relative errors (MAREs) of 25.8% and 13.8%, respectively. The model computation of the packing fractional wetted area for the distillation conditions reported by Fitz et al.[28] at high temperature in the pressure range $P = 0.002-2.8$ MPa in a 1200-mm-i.d. column is also very encouraging. Using the simulated data of Fitz et al.[28] from the Billet model[27] after surface tension effects have been handled properly,[28] this hydrodynamic variable is predicted with an estimated mean absolute relative error of 15%.

Because of its mechanistic nature, the model can be helpful in exploring variations in the irrigated pressure drop, liquid holdup, and effective interfacial area as a function of changes in the operation of the structured-packing-containing columns. Computations were undertaken to determine whether likely variations are expected from changes in the packing geometry, the fluid throughputs, and the temperature/pressure.

In accordance with the experimental observations (Figure 3), the mechanistic model predicts very well the increase in the liquid holdup with an increase in gas capacity factor and superficial velocity and in liquid superficial velocity for the air–water system studied in a Flexipac 2Y column by Fair and collaborators.[7,26] In the preloading zone, the liquid holdup increases only slowly with gas capacity, as expected, whereas the trend is well-predicted over the whole range swept by the gas load. This weak sensitivity to the gas capacity factor can be explained by the slight influence on the liquid film thickness[32] and on the liquid film spreading with increasing gas load.

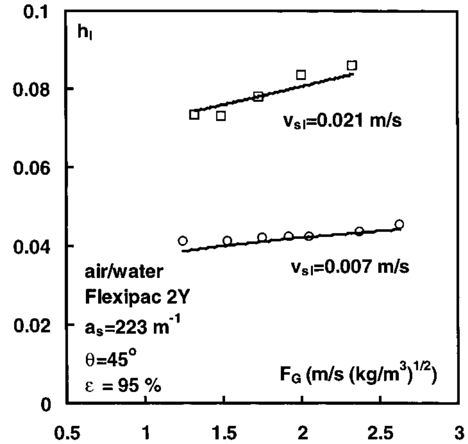

Figure 3. Effect of gas capacity factor and liquid superficial velocity on the total liquid holdup. Lines show simulated behavior using eqs 22, 23, and 31 (Flexipac 2Y stainless steel, column diameter 914 mm, air–water system, data of Fair et al.[7,26]).

The model is also capable of predicting the impact of fluid throughputs on the irrigated two-phase pressure drop. Figure 4 illustrates the increase in the irrigated pressure as a function of gas capacity factor for the air–water system across Montz B1 and Flexpic 2Y packings using the data of Fair and colleagues.[7,26]

Consistent with the prediction of the effective interfacial area using the Billet model,[27] Figure 5 illustrates the effect of liquid superficial velocity on the total reflux flow of an *i*-butane/*n*-butane mixture over Mellapak 250Y at 125 °C and 2.6 MPa. The model yields effective interfacial areas close to those simulated using the Billet model. Furthermore, both models predict the correct variation trend between the liquid velocity and the interfacial area. As the liquid irrigation velocity increases, both the liquid holdup and the irrigated pressure drop increase, which improves the liquid spreading on the packing.

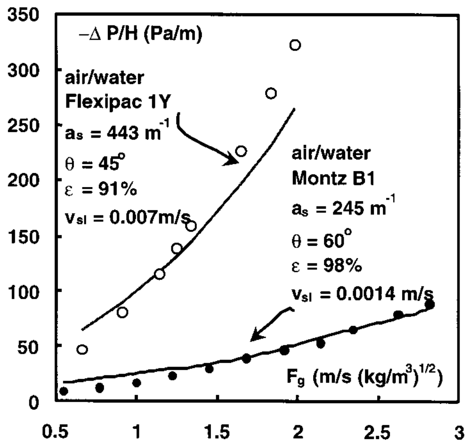

Figure 4. Effect of gas capacity factor and liquid superficial velocity on the irrigated pressure drop. Lines show simulated behavior using eqs 22, 23, and 31 (Montz B1 stainless steel, column diameter 450 mm, air−water system, data of Olujic;[9] Flexipac 2Y stainless steel, column diameter 914 mm, air−water system, data of Fair et al.[7,26]).

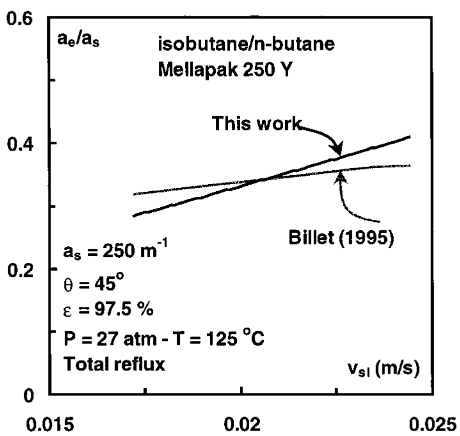

Figure 5. Effect of liquid superficial velocity on the packing fractional wetted area. Lines show simulated behavior using eqs 22, 23, and 31 (this work) and Billet model[27] for the distillation at total reflux of *i*-butane/*n*-butane on Mellapak 250Y metal, column diameter 1200 mm (operating conditions of Fitz et al.[28]).

Summary

A mechanistic hydraulic model was developed for the prediction of the irrigated pressure drop, the liquid holdup, and the effective interfacial area (or packing fractional wetted area) in the preloading zone of structured-packing-containing columns operated in countercurrent flow of gas and liquid under partial wetting conditions. The model mimics the two-phase flow by a double-slit network consisting of a dry slit and a wet slit. The slits were subject to Couette−Poiseuille gas and liquid flows where the fluid−wall and the fluid−fluid interfacial shear stresses and the (stream-wise) pressure gradient were incorporated. The model was successful in predicting the irrigated pressure drops, liquid holdups, and effective interfacial areas under various conditions simulating the scrubbing or the distillation industrial conditions.

Nomenclature

a_e = gas−liquid interfacial area (m^{-1})
a_s = packing specific surface area (surface packing area/column volume) (m^{-1})
d_e = hydraulic equivalent diameter, $d_e = \dfrac{4\epsilon}{a_s}$ (m)
f_i = interfacial friction factor
f_w = wall friction factor
F_g = gas capacity factor, $F_g = v_{sg}\rho_g^{0.5}$
g = gravity acceleration (m/s^2)
Ga_α = α-phase Galileo number, $Ga_\alpha = d_e^3 g \rho_\alpha^2/\mu_\alpha^2$
h_l = liquid holdup
l = slit length (m)
MARE = mean absolute relative error,

$$\mathrm{MARE} = \frac{1}{N}\sum_{i=1}^{N} \left|\frac{y_{\mathrm{calc},i} - y_{\mathrm{exp},i}}{y_{\mathrm{exp},i}}\right|$$

P = pressure (MPa)
Re_i = interfacial Reynolds number, $Re_i = \rho_g d_e u_{il} \cos\theta/\mu_g$
Re_α = α-phase Reynolds number, $Re_\alpha = v_{s\alpha} d_e \rho_\alpha/\mu_\alpha$
S = slit half-wall thickness (m)
T = temperature (K)
u_α = α-phase local velocity in slit (m/s)
$u_{i\alpha}$ = α-phase interfacial velocity (m/s)
$v_{s\alpha}$ = α-phase superficial velocity (m/s)
w = slit half-void thickness (m)
x = slit cross-wise coordinate (m)
y = hydrodynamic parameter ($y = h_l$, $-\Delta P/H$, or η_e)
z = slit stream-wise coordinate (m)

Greek Letters

δ = liquid film thickness (m)
$-\Delta P/H$ = irrigated pressure drop (Pa/m)
ϵ = bed void fraction
η_e = packing fractional wetted area
μ_α = α-phase dynamic viscosity (Pa s)
ν_α = α-phase kinematic viscosity (m^2 s)
θ = slit inclination or packing corrugation angle (°)
ρ_α = α-phase density (kg/m^3)
$\tau_{i\alpha}$ = α-phase interfacial shear stress (Pa)
$\tau_{w\alpha}$ = wall−α phase shear stress (Pa)
Ψ_α = α-phase dimensionless body force

Subscripts

calc = calculated
exp = measured
g = gas
i = gas−liquid interface
l = liquid
w = wall
x = in slit cross-wise direction
z = in slit stream-wise direction
$\langle\,\rangle$ = averaging operator

Literature Cited

(1) Strigle, R. F. Pollution control. In *Packed Tower Design and Applications: Random and Structured Packings*; Gulf Publishing Company: Houston, TX, 1994; pp 113−149

(2) Dudukovic, M. P.; Larachi, F.; Mills, P. L. Multiphase reactors−Revisited. *Chem. Eng. Sci.* **1999**, *54*, 1975.

(3) Malone, M. F.; Doherty, M. F. Reactive distillation. *Ind. Eng. Chem. Res.* **2000**, *39*, 3953.

(4) Fair, J. R.; Seibert, A. F.; Behrens, M.; Olujic, Z. Structured packing performance. Experimental evaluation of two predictive models. *Ind. Eng. Chem. Res.* **2000**, *39*, 1788.

(5) Bravo, J. L.; Rocha, J. A.; Fair, J. R. Mass transfer in gauze packings. *Hydrocarbon Proc.* **1985**, *64*, 91.

(6) Bravo, J. L.; Rocha, J. A.; Fair, J. R. A comprehensive model for the performance of columns Containing structured packings. *Inst. Chem. Eng. Symp. Ser.* **1992**, *128*, A489.

(7) Rocha, J. A.; Bravo, J. L.; Fair, J. R. Distillation columns containing structured packings: A comprehensive model for their performance. 1. Hydraulic models. *Ind. Eng. Chem. Res.* **1993**, *32*, 641.

(8) Rocha, J. A.; Bravo, J. L.; Fair, J. R. Distillation columns containing structured packings: A comprehensive model for their performance. 2. Mass transfer model. *Ind. Eng. Chem. Res.* **1996**, *35*, 1660.

(9) Olujic, Z. Development of a complete simulation model for predicting the hydraulic and separation performance of distillation columns equipped with structured packings. *Chem. Biochem. Eng. Q.* **1997**, *11*, 31.

(10) Olujic, Z. Effect of column diameter on pressure drop of a corrugated sheet structured packing. *Trans. Inst. Chem. Eng.* **1999**, *77*, 505.

(11) Olujic, Z.; Kamerbeek, A. B.; de Graauw, J. A. A corrugation geometry based model for efficiency of structured distillation packing. *Chem. Eng. Process.* **1999**, *38*, 683.

(12) Brunazzi, E.; Paglianti, A. Mechanistic pressure drop model for columns containing structured packings. *AIChE J.* **1997**, *43*, 317.

(13) Bravo, J. L.; Rocha, J. A.; Fair, J. R. Pressure drop in structured packing. *Hydrocarbon Proc.* **1986**, *56*, 45.

(14) Stichlmair, J.; Bravo, J. L.; Fair, J. R. General model for prediction of pressure drop and capacity of countercurrent gas/liquid packed columns. *Gas Sep. Purif.* **1989**, *3*, 19.

(15) Robins, L. A. Improve pressure drop prediction with a new correlation. *Chem. Eng. Prog.* **1991**, *87*, 87.

(16) Spiegel, L.; Meier, W. A. Generalized pressure drop model for structured packings. *Inst. Chem. Eng. Synp. Ser.* **1992**, *128*, B85.

(17) Spiegel, L.; Meier, W. A. Correlations of the performance characteristics of the various Mellapak types. *Inst. Chem. Eng. Symp. Ser.* **1987**, *104*, A203.

(18) Brunazzi, E.; Nardini, G.; Paglianti, A.; Petarca, L. Interfacial area of Mellapak packing: Absorption of 1,1,1-trichloroethane by Genosorb 300. *Chem. Eng. Technol.* **1995**, *18*, 248.

(19) Nardini, G.; Paglianti, A.; Petarca, L.; Viviani, E. Sulzer BX gauze: Fluid dynamics and absorption of acid gases. *Chem. Eng. Technol.* **1996**, *19*, 20.

(20) Holub, R. A.; Dudukovic, M. P.; Ramachandran, P. A. A phenomenological model for pressure drop, liquid holdup and flow regimes transition in gas−liquid trickle flow. *Chem. Eng. Sci.* **1992**, *47*, 2343.

(21) Iliuta, I.; Larachi, F.; Al-Dahhan, M. H. Double-slit model for partially wetted trickle flow hydrodynamics. *AIChE J.* **2000**, *46*, 597.

(22) McNulty, K.; Hsieh, C. L. Hydraulic performance and efficiency of Koch Flexipac structured packing. AIChE National Meeting, Los Angeles, CA, Nov 1982.

(23) Chen, G. K.; Kitterman, L.; Shieh, J. H. Development of a new generation of high efficiency packing for mass transfer operations. *Chem. Eng. Prog.* **1982**, *79*, 48.

(24) Chen, G. K.; Kitterman, L.; Shieh, J. H. *Glitsch High-Efficiency Packings. Preliminary Design Information*; Addendum to Glitsch Publication No. 40283; Glitsch, Inc.: Dallas, TX, 1983.

(25) Süss, P.; Spiegel, L. Holdup of Mellapak structured packings. *Chem. Eng. Process.* **1992**, *31*, 119.

(26) Fair, J. R.; Bravo, J. L. Distillation columns containing structured packings. *Chem. Eng. Prog.* **1990**, *Jan*, 19−29.

(27) Billet, R. *Packed Towers in Processing and Environmental Technology*; VCH Verlagsgesellschaft: Weinheim, Germany, 1995.

(28) Fitz, C. W., Jr.; Kunesh, J. G.; Shariat, A. Performance of structured packing in a commercial scale column at pressures of 0.02−27.6 bar. *Ind. Eng. Chem. Res.* **1999**, *38*, 512.

(29) Meier, W.; Hunkeler, R.; Stocker, W. D. Sulzer Mellapak−Eine Neue, Geordnete Packung für Stoffaustausch-Apparate. *Chem. Ing. Tech.* **1979**, *51*, 119.

(30) Verschoof, H.-J.; Olujic, Z.; Fair, J. R. A general correlation for predicting the loading point of corrugated sheet structured packings. *Ind. Eng. Chem. Res.* **1999**, *38*, 3663.

(31) Svinukhov, A. G.; Klykov, M. V. Hydraulic resistance and dynamic holdup of liquid in beds of screen packing. Chem. *Technol. Fuels Oils* **1978**, *12*, 900.

(32) Woerlee, G. F. Hydrodynamics and mass transfer in packed columns and their applications for supercritical separations. Ph.D. Thesis, Technical University of Delft, Delft, The Netherlands, 1997.

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