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Particle-liquid mass transfer in solid–liquid fluidized beds

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HIGHLIGHTS

- Solid–liquid circulating multistage fluidized bed (SLCMFB) has been investigated.
- Values of mass transfer coefficient (k_{SL}) have been measured for SLFB and SLMFB.
- Effect of the distributor and bed height on k_{SL} has been measured experimentally.
- Correlations have been developed for the estimation of k_{SL}.

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ABSTRACT

Solid–liquid mass transfer coefficient (k_{SL}) was measured in both conventional solid–liquid fluidized bed (SLFB) and solid–liquid multistage fluidized bed (SLMFB) by using the system of dissolution of benzoic acid in water. The particle size was varied in the range of 165–890 μm and the voidage in the range of 0.4–0.9. The dependence of solid–liquid mass transfer coefficient on important variables associated with the distributor design has been studied for SLFB. It was observed that the orifice density (number of orifices per unit area), per cent free area and the pressure drop across the distributor play important role on the values of k_{SL}. The effect of the inerts (glass beads) on the mass transfer coefficient was investigated. For the voidages below 0.6, intensification in k_{SL} was observed due to the presence of inerts in the system. As the voidage increases, the presence of inerts was found to have increasing effect on the mass transfer coefficient. For SLMFB, an increment of up to 15% in mass transfer rate was observed in comparison to that in conventional SLFB under a wide range of operating conditions like superficial liquid velocity and particle diameter. All the past correlations have been critically analyzed and suitable recommendations have been made. A new generalized correlation has been proposed for the estimation of mass transfer coefficient for both SLFB and SLMFB based on the experimental data. These correlations have been shown to be valid for all the available data in the published literature.

1. Introduction

Separation processes using packed bed and fluidized bed of ion exchange resins (and other stationary phases) are widely used in industry to recover the dissolved solute from a variety of mixtures owing to the good regeneration ability of adsorbents. It is difficult to use the conventional physical separation methods such as fractional distillation, liquid–liquid extraction, solvent aided separation, crystallization, membrane separation, etc. to reduce the impurity to the new standards. In view of this, the chromatographic separation offers an edge over the conventional methods owing to its ability to achieve higher levels of purity. The chromatographic separation process occurring in classical fluidized beds with large production rates, generally possesses three major problems. Firstly, the overall separation is performed in a batch-wise manner due to the plurality of the processes involved such as adsorption (or ion exchange), washing and desorption (or regeneration). Secondly, from hydrodynamic point of view the mobile phase has to overcome a high pressure drop offered by the stationary phase, resulting in extra energy investment. The third major disadvantage of the batch operation is that the active zone is small fraction of the total bed. As, the utilization of the bed is restricted to the active zone and the rest of the adsorbent is either spent or is waiting for utilization. Thus in the above case, the capital expenditure is several times higher as compared to the case

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Nomenclature

$\underline{a}$	effective interfacial area, m^{-1}
$\underline{a}_P$	effective solid–liquid interfacial area, m^{-1}
A	average surface area of single benzoic particle, m^2
C_0	initial concentration of Benzoic acid, kmol m^{-3}
CC	correlation coefficient, (–)
C_B	bulk concentration, kmol m^{-3}
C_D	drag coefficient, (–)
C_{OUT}	benzoic acid concentration at the outlet, kmol m^{-3}
C_O	orifice coefficient, (–)
C_t	concentration of benzoic acid at time t, kmol m^{-3}
C_S	surface or saturation concentration, kmol m^{-3}
ΔC	difference between surface and bulk concentration, kmol m^{-3}
D	column diameter, m
D_A	diffusivity, $m^2\ s^{-1}$
d_O	orifice diameter, m
d_P	particle diameter, m
FA	free area, (m^2)
g	acceleration due to gravity, ms^{-2}
G	mass flow rate, kg $m^{-2}\ s^{-1}$
H	bed height, (m)
H_S	static bed height, (m)
H_E	expanded bed height, (m)
h	length of the cylindrical or width of the parallelepiped particle, m
j_D	Chilton-Colburn mass transfer factor, (–)
k_{SL}	solid–liquid mass transfer coefficient, ms^{-1}
L	column height, m
l	other dimension of the parallelepiped particle, m
n	Richardson–Zaki exponent (Eq. (11)), (–)
N_A	specific rate of mass transfer, kmol $m^{-2}\ s^{-1}$
P	average perimeter of single benzoic acid particle projection, m
ΔP	pressure drop across the bed, kg $m^{-1}\ s^{-2}$
ΔP_d	pressure drop across the distributor, kg $m^{-1}\ s^{-2}$
q	exponent in Eq. (14)
r	rate of mass transfer, kmol $m^{-3}\ s^{-1}$
S	average projected area of a single benzoic acid particle, m^2
t	time, s
U_m	optimal liquid mixing velocity for binary fluidized bed, ms^{-1}
U_{mf}	minimum fluidization velocity, ms^{-1}
V	average volume of single benzoic acid particle, m^3
V_O	orifice velocity, ms^{-1}
V_L	superficial liquid velocity, ms^{-1}
$V_{S\infty}$	terminal settling velocity of a particle, ms^{-1}

Dimensionless numbers

$Ar = \frac{\rho_L(\rho_L-\rho_S)gd_P}{\mu_L^2}$	archimedes number, (–)
$B = Ga \times Mv \times \left(\frac{\mu_L}{\rho}\right)$	constant in Eq. (4), (–)
$f' = \frac{gd_P(1-\epsilon_L)\rho_L(\rho_S-\rho_L)}{2G^2}$	modified friction factor
$Fr = \frac{V_L^2}{gd_P}$	Froude number, (–)
$Ga = \frac{d_P^3\rho_L(\rho_S-\rho_L)g}{\mu_L^2}$	Galileo number, (–)
$j_D = \frac{k_{SL}}{V_L}Sc^{\frac{2}{3}}$	dimensionless mass transfer factor, (–)
$Mv = \frac{(\rho_S-\rho_L)}{\rho_L}$	density number, (–)
$\mathrm{Re} = \frac{D\rho_L V_L}{\mu_L}$	Reynolds number, (–)
$\mathrm{Re}_P = \frac{d_P\rho_L V_L}{\mu_L}$	particle Reynolds number based on superficial liquid velocity, (–)
$\mathrm{Re}_{P\infty} = \frac{d_P\rho_L V_{s\infty}}{\mu_L}$	particle Reynolds number based on terminal settling velocity of particle, (–)
$\mathrm{Re}'_P = \frac{d_L V_L\rho_L}{\mu_L\epsilon_L}$	modified Reynolds number including voidage, (–)
$\mathrm{Re}''_P = \frac{d_P V_L\rho_L}{\mu_L(1-\epsilon_L)}$	modified Reynolds number including voidage, (–)
$\mathrm{Re}_P = \frac{6V_L\rho_L}{\underline{a}\mu_L}$	modified Reynolds number based on interfacial area, (–)
$\mathrm{Re}_m = \frac{\rho_L U_m d_P}{\mu_L}$	mixing Reynolds number, (–)
$Sc = \frac{\mu_L}{\rho_L D_A}$	Schmidt number, (–)
$Sh = \frac{d_P k_{SL}}{D_A}$	Sherwood number, (–)

Greek letters

$\Delta\rho$	density difference between solid and liquid phases, kgschmidtm^{-3}
ϵ_L	liquid holdup, (–)
ϵ_S	solid hold-up, (–)
ϵ_{SS}	solid hold-up at fixed bed conditions, (–)
λ	boundary layer thickness, m
μ	liquid viscosity, kg $m^{-1}\ s^{-1}$
ρ	density, kg m^{-3}
ϕ_S	sphericity, (–)
Φ	shape factor, (–)

Subscripts

GB	glass beads
L	liquid phase
P	particle
mf	at minimum fluidization
∞	at terminal settling conditions
S	external surface/solid phase

where only active zone is present in the entire column all the time. In addition to this, the fixed bed also suffers from bed clogging.

Although the problem of clogging can be addressed by use of expanded beds, the three major disadvantages mentioned above need to be addressed in such a way so as to increase the fraction of active zone in the column and controlling the pressure drop up to a certain useful level. The use of circulating fluidized bed not only addresses the above mentioned disadvantages but also makes the operation continuous. The bed can be divided into a number of stages by sectionalizing the column, and the differential contact of solid and liquid phase gets improved. The equipment having both the facilities of adsorption and regeneration and an additional facility of keeping only active zone (in the adsorption section) is called solid–liquid circulating multistage fluidized bed (SLCMFB). A schematic diagram of the SLCMFB can be found elsewhere [1]. Despite the fact that SLCMFBs are very attractive for industrial applications like catalytic reactions and chromatographic separations, substantial amount of additional information is still needed for the rational design of these equipment.

The present research work addresses the following key issues related to the design of SLCMFB: (1) To estimate the solid–liquid mass transfer coefficient (k_{SL}) over a wide range of the particle size and voidage in solid–liquid multistage fluidized bed (SLMFB) since, the operation of SLMFB is similar to that of the down comer of SLCMFB and in solid–liquid fluidized bed (SLFB) whose operation is similar to the riser column of SLCMFB. (2) As the downer of the SLCMFB is sectionalized into multiple stages, the k_{SL} strongly depends on the design of the perforated sieve plate of every stage. Therefore, the design of distributor plays a crucial role. In spite of the importance of distributor design, there is practically no