20 Liquid and liquid–gas spouting of solids

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The spouted bed technique permits agitation of particles too coarse for fluidizing with a gas when excellent heat and mass transfer characteristics and intimate fluid–particle contacting are important. Liquid spouting has attracted much less interest, as coarse particles can be easily fluidized in this medium. However, recent advances in biotechnology and renewed interest in wastewater treatment have sparked new applications of liquid-spouted beds, particularly those incorporating a gas phase.

20.1 Liquid spouting

As discussed elsewhere in this book, a spouted bed can form when a fluid jet blows vertically upward along the center line of a vertical column, forming a spout in which fast-moving fluid and entrained particle mixing occur, surrounded by an annular region densely packed with particles moving slowly downward and inward. The spout is topped by a spillover fountain. Fluid percolates through the annulus from the spout. In a spout-fluid bed (see Chapter 6), additional fluid is introduced at the bottom of the annulus. Several fluid–particle patterns are possible, depending on the magnitude of the external annular fluid introduced to the annulus bottom:

(1) Spouting with irrigation: beds in which the external annular fluid velocity, \( U_{a0} \), is restricted to a velocity that keeps the annular velocity \( U_{aH} \leq U_{mf} \).
(2) Spout-fluidization: beds in which the annulus is partly or completely fluidized; the level at which fluid velocity reaches \( U_{mf} \) depends on the external annular flowrate. If \( U_{a0} = U_{mf} \), the annulus is completely fluidized.
(3) Jet-fluidized beds: beds in which the bed depth is deeper than the maximum spoutable bed height. When the fluid is a liquid, such beds exhibit two zones: a lower spouted bed region and an upper fluidized bed zone.

A typical regime map for a water spout-fluid bed system is shown in Figure 20.1. The conventional spouted bed \( (V_{a0} = 0) \) falls between points E and D along the abscissa, whereas the fluidized bed is depicted along the ordinate between points G and A. Below the minimum spout-fluid line, GHE, the bed is fixed; the outer boundary curve, ABCD,

represents the locus of annulus and spout flowrates, whose sum equals $V_T$. One could also surmise from Figure 20.1 that because the experimental value of $V_{ms}$ is approximately equal to $V_{mf}$, the actual bed height is close to $H_m$. Actual bed heights can be substantially less than $H_m$, however, even when $(U_{ms}/U_{mf})$ is close to unity.

A spout-fluid bed of height $H \leq H_{mSF}$ looks much the same as a spouted bed with $H < H_m$. As $(V_{a0})_{max}$ represents the inlet annulus flowrate required to fluidize the top of the annulus in a spout-fluid bed of height $H$, the annulus flowrate at point H is $(V_{a0})_{max}$. In the region EHCD, Regime I (item 1 in the preceding list), the total flowrate $> V_{mSF}$, with $V_{a0} \leq (V_{a0})_{max}$. The initial bed height is equal to or less than $H_{mSF}$ and the bed is very similar in appearance to that of a conventional spouted bed. As $V_{a0}$ is increased at constant $V_{n0}$ along line a–a, the top of the annulus becomes incipiently fluidized at point M, as $V_{a0}$ equals $(V_{a0})_{max}$. An increase in the external annulus flowrate, $V_{a0}$ (along line a–a), results in the fluidization of the annulus, initiated at lower and lower levels. Finally, when $V_{a0}$ reaches $V_{mf}$ (point N on line a–a), the bottom of the annulus becomes just fluidized. These observations reflect the fact that $H_{mSF}$ decreases as $V_{a0}$ increases at constant $V_{n0}$. Along line HC, despite the fact that the spout diameter increases as $V_{n0}$ is increased, the annulus height actually decreases, owing to the increased solids holdup in the spout and the fountain. Note that the data points lie above HC when $V_{n0} > 9(V_{n0})_{mSF}$. The fountain height and spout diameter are large, and the annulus height is well below the initial bed height. In region HGBC, Regime 2 (item 2 in the preceding list), the bed forms two zones, in which a fluidized bed sits on top of a spout-fluid bed with a
well-defined annulus and spout. This bed is very similar in appearance to a spouted bed with \( H > H_m \) (i.e., to item 3 on the list). Along GB, \( U_{a0} = U_{mf} \), thus fluidizing the entire annulus, and the total volumetric flow rate is above \( V_{mf} \) except at point G. Regime 3 is bounded by region AGB, in which the fluid velocity > \( U_{mf} \) everywhere in the annulus, and a liquid jet penetrates into the bed.

### 20.1.1 Pressure drop versus flow rate

The plot of pressure drop versus flow rate for liquid spouting is generally similar to that of a gas spouting system (see Chapter 2). With an increase in liquid superficial velocity, above \( U_{ms} \), a constant pressure drop is established in the system. The difference between the peak pressure drop (\( \Delta P_m \)) and the pressure drop at minimum spouting (\( \Delta P_{ms} \)) decreases with increasing bed height. For gas phase systems at the minimum spouting velocity, a small reduction in the gas flowrate causes the spout and fountain to collapse, whereas in liquid spouting at \( V_{ms} \) the fountain is unexpanded with respect to the surface.

### 20.1.2 Minimum spouting velocity

In liquid-spouted beds the minimum spouting velocity at the maximum spoutable bed height is approximately equal to \( U_{mf} \). Based on the assumption that \( (U_s)_{Hm} = (U_a)_{Hm} = U_{mf} \) and that \( A_s \ll A \), Grbavčić et al.\(^5\) derived the following scaled equation for \( U_{ms} \):

\[
\frac{U_{ms}}{U_{mf}} = 1 \left( 1 - \frac{H}{H_m} \right)^3.
\] (20.1)

Eq. (20.1) is based on the Mamuro-Hattori\(^6\) equation for \( U_{aH} \), assuming that at minimum spouting, \( U_{aH} = U_{ms} \). The original Mathur-Gishler\(^7\) equation, when scaled in this fashion, produces a different dependency:

\[
\frac{U_{ms}}{U_{mf}} = \left( \frac{H}{H_m} \right)^{0.5}.
\] (20.2)

This equation predicts lower values of \( U_{ms}/U_{mf} \) than Eq. (20.1) when \( H/H_m \gtrsim 0.15 \), and the converse when \( H/H_m < 0.15 \).

Littman and Morgan\(^8\) developed a general correlation for \( U_{ms}/U_{mf} \) as a function of \( (f_1/f_2 U_{mf}) \), \( H/D \) and \( \theta_f \). That correlation was based on the following equation, derived from the annular pressure gradient at the top of the bed under minimum spouting
conditions:

\[
\frac{U_{ms}}{U_{mf}} = -\frac{f_1}{2f_2U_{mf}} + \left[\left(\frac{f_1}{2f_2U_{mf}}\right)^2 + \left(1 + \frac{f_1}{f_2U_{mf}}\right)C_p\right]^{1/2};
\]

\[
\frac{H}{D} \geq \frac{H^*}{D} \text{ and } A_f \cdot g(\phi_s) > 0.02,
\]

where

\[
C_p = 1 - Y - \frac{H}{D}X^2 \left[\frac{2Y + (X - 2) + (X - 0.2) - (3.24/\theta_f)}{2Y + 2(X - 0.2) - 1.8 + (3.24/\theta_f)}\right],
\]

\[
\theta_f = 7.18 \left[A_f \cdot g(\phi_s) - D_i/D\right] + 1.07,
\]

\[
X = 1/(1 + H/D),
\]

\[
Y = 1 - \Delta P_{ms}/\Delta P_{mf},
\]

and

\[
A_f = \frac{\rho}{\rho_p - \rho} \frac{U_{mf}U_T}{gD_i}.
\]

The dimensionless parameter, \(C_p\), has an important effect on the \(U_{ms}/U_{mf}\) ratio and provides a connection to the viscous or inertial term in the Ergun equation and bed geometry through the parameter \(\theta_f\). This model covers a wide range of experimental conditions, including water-spouted systems of spherical and nonspherical particles. For spherical particles, the \(g(\phi_s)\) parameter of Eq. (20.5) approaches unity asymptotically.

Littman et al.\(^{10,11}\) showed that the minimum spout-fluid flowrate in a liquid phase spout-fluid bed follows the simple linear relationship:

\[
(V_{a0})_{mSF} = V_{mf} - \frac{V_{mf}}{V_{ms}}(V_{a0})_{mSF}.
\]

As depicted in Figure 20.2 for such systems, the total minimum spout-fluid flowrate for \(H/H_m \leq 1\) is given by

\[
V_{mSF} = (V_{a0})_{mSF} + (V_{a0})_{mSF}.
\]

For liquid spouting in a conical vessel, Legros et al.\(^{12}\) proposed a model for the minimum spouting velocity based on the momentum transfer between the liquid jet and the bed. Beds consisting of different particle size distributions were used to substantiate this model. Good agreement was obtained, even when a large fraction of fine particles was present in the initial particle mixture.

### 20.1.3 Pressure drop at minimum spouting

\(\Delta P_{ms}\) is commonly calculated by integrating the one-dimensional longitudinal pressure gradient in the annulus over the height of the bed using one of the annulus models. This
Figure 20.2. Relationship of spout and annular flows at the minimum spout-fluid flowrate, for water and glass beads in a 117 × 9.2-mm rectangular column with $D_i = 9.3$ mm, after Littman et al.\textsuperscript{11}

procedure leads to quite reasonable results, although it ignores radial pressure gradients across the annulus and assumes that $\Delta P_S = \Delta P_{ms} = (\Delta P_a)_{ms}$. Equations for predicting $\Delta P_{ms}/\Delta P_{mf}$ have been proposed by Mamuro and Hattori,\textsuperscript{6} Lefroy and Davidson\textsuperscript{13} (for $H = H_m$), Epstein and Levine,\textsuperscript{14} Grbavčić et al.,\textsuperscript{5} and Morgan and Littman\textsuperscript{15} (for $H \leq H_m$). The correlation of Morgan and Littman is

$$\Delta P_{ms}/\Delta P_{mf} = 1 - Y,$$

(20.11)

where $Y$ is obtained from:

$$Y^2 + [2(X - 0.2) - 1.8 + (3.24/\theta_f)]Y + [(X - 2)(X - 0.2) - (3.24X/\theta_f)] = 0.$$

(20.12)

The ratio $\Delta P_{ms}/\Delta P_{mf}$ is a function of $A_f$, $g(\phi_s)$ and $D_i/D$, and was found to fit a wide range of experimental data involving both air- and water-spouted systems.

Based on their $U_{ms}/U_{mf}$ model for spouted beds, Littman et al.\textsuperscript{11} developed a relationship for the minimum spout-fluid bed pressure drop ratio, $\Delta P_{msf}/\Delta P_{mf}$. That model predicted a linear relationship between the spout-fluid pressure drop and the auxiliary velocity, $U_a0$.

### 20.1.4 Maximum spoutable bed depth

Littman et al.\textsuperscript{16} also developed a model for predicting the maximum spoutable bed height ($H_m$) for coarse particles when the controlling mechanism was fluidization of
the annulus. That model was based on a detailed axisymmetric analysis of the flow in the annulus, of the spouted bed employing interfacial pressure and velocity distributions, a fluid continuity equation, a vectorial Ergun equation\(^9\) for the local pressure in the annulus, and appropriate boundary conditions. The following relationship was established between the maximum spoutable height, \(H_m\), and the average spout diameter, \(D_s\), for spherical particles:

\[
\frac{H_m D_s}{D^2 - D_s^2} = 0.345 \left( \frac{D_s}{D} \right)^{-0.384}.
\]  

(20.13)

When evaluated with experimental values for \(D_s\), predictions of \(H_m\) using Eq. (20.13) differed on average by 8.5 percent from experimental data for both air- and water-spouted beds. Assuming that \(H_m\) was proportional to the inlet fluid momentum, Littman et al.\(^{17}\) subsequently developed an equation for predicting \(H_m\) for coarse spherical particles as a function of the inlet tube diameter, the column diameter, and the properties of the fluid–particle system:

\[
\frac{H_m D_i}{D^2} = 0.218 + \frac{0.0050}{A_f}.
\]  

(20.14)

Eqs. (20.13) and (20.14) were combined to obtain a relationship for the \(D_s/D_i\) ratio as a function of \(D/D_i\) and parameter \(A_f\).

Grbavčić et al.\(^5\) presented the following simpler empirical correlation for predicting \(H_m\) in water-spouted beds:

\[
\frac{H_m}{D} = 0.347 \left( \frac{D}{D_i} \right)^{0.41} \left( \frac{D}{d_p} \right)^{0.31}.
\]  

(20.15)

In spout-fluid bed systems, the maximum spoutable bed height decreases with increasing external annular flow. Grbavčić et al.\(^{18}\) assumed that incipient fluidization of the solids at the top of the annulus is the limiting/controlling mechanism and proposed that \(H_{mSF}/H_m = f(U_{a0}/U_{mf}, C)\), where the parameter \(C\) is the ratio of the fluid velocity at the top of the annulus at \(H_m\) or \(H_{mSF}\) to the minimum fluidization velocity. Their combined spout-fluid bed data for both air and water as the fluid gave a best-fit value of \(C = 0.935\).

### 20.1.5 Fine particle systems

The spouting of fine particles (\(d_p < 1\) mm) differs in two fundamental ways from large particle behavior:

1. Fine particles have a much smaller range of acceptable inlet diameter that result in stable spouting. Under operating conditions in which the inlet diameter exceeds the maximum stable inlet diameter, instabilities originate in the bed at either the spout inlet or outlet. Littman and Kim\(^{19}\) characterized such particles at the minimum spouting conditions and examined the effect of both the inlet orifice diameter and the pressure distribution along the spout–annulus interface on the basic spouting parameters (\(U_{ms}\) and \(\Delta P_{ms}\)).
2. Fine particles can cause a change in the spout inlet boundary condition. With gasspouted systems the spout becomes unstable, bubble generation commences within the spout, and solid circulation is impaired. However, with liquid-phase spouting of fine particles, this instability manifests itself as a change in the inlet voidage. No longer is the inlet voidage equal to 1, but rather is a value less than 1. Dense transport is established throughout the spout, and $\Delta P_{ms}/\Delta P_{mf}$ exceeds 0.785. Morgan et al.\(^\text{20}\) characterized such fine particle systems as low $A_f$ systems when the values of this dimensionless parameter are less than 0.002 and thus indicative of a low inlet orifice momentum. In this regime, they found that a modified $A_f$ parameter correlated the system dynamics better. Kim and Ha\(^\text{21}\) studied the annulus flow of fine-particle waterspouted systems and found that the axisymmetric model of flow, which assumes Darcy flow in the annulus and uses the experimental spout–annulus interfacial condition, predicts the annulus fluid streamlines well.

### 20.1.6 Bed expansion

With an increase in liquid flow rate, a bed’s spout and fountain expand, thereby increasing the average bed voidage even though the annular voidage increases minimally. Kmiec\(^\text{22}\) investigated overall bed expansion to the top of the fountain using glass spheres, silica gel particles, ion-exchange resins, and sand spouted with water in an 88-mm-diameter conical-cylindrical column with included cone angles of 30°, 60°, and 180°. On a logarithmic plot, the voidage versus liquid superficial velocity relationship exhibited a change in slope at a voidage $\approx 0.85$. Similar behavior has been reported by several investigators of liquid fluidized systems (Di Felice\(^\text{23}\)). The following correlations were proposed by Kmiec\(^\text{22}\) for low and high-voidage regions:

For $\varepsilon < 0.85$,

$$\varepsilon = 0.838 \cdot Re^{0.182} Ar^{-0.0866} (H_0/D)^{-0.243} \theta^{-0.067},$$  \hspace{1cm} (20.16)

where $Re = d_p U \rho / \mu$ and $Ar = d_p^3 \rho (\rho_p - \rho) g / \mu$. For $\varepsilon > 0.85$,

$$\varepsilon = 1.102 U^{0.0774} (H_0/D)^{-0.105} \theta^{-0.047},$$  \hspace{1cm} (20.17)

with $U$ in m/s. Eqs. (20.16) and (20.17) are valid for $3.5 < Re < 350$, $2.1 \cdot 10^3 < Ar < 2.6 \cdot 10^6$, $0.19 < H_0/D < 1.55$, $\pi/6 < \theta < \pi$ radians, and $0.023 < U < 0.095$ m/s.

Ishikura et al.\(^\text{24}\) investigated the maximum bed expansion in liquid-spouted and fluidized beds undergoing upward piston movement under conditions in which $U$ exceeds the single particle terminal velocity.

### 20.1.7 Mass transfer

Hadžismajlović et al.\(^\text{25}\) and Kim et al.\(^\text{26}\) studied the liquid–solid mass transfer coefficients in liquid spouted and spout-fluid beds of ion exchange resins. They found that the traditional spouted bed is less effective than an equivalent fluidized bed. Such liquid spout-fluid beds can be successfully used to treat liquids contaminated with suspended
solid particles, but the bed height should be significantly greater than $H_m$ to obtain efficiency comparable with fluidized beds.

20.1.8 Spouting of binary mixtures and elutriation

Relatively little research has been done on liquid spouting of poly-disperse particle systems. Ishikura and Nagashima\textsuperscript{27} investigated the spouting of binary particle mixtures with glass spheres of $d_p = 1.124$ mm as the coarse particles and glass spheres of diameters 233 and 474 μm as the finer component. The spouting fluids were water and aqueous sucrose solution. The addition of a small amount of the finer particles to the coarse ones produced a much lower minimum spouting velocity for the mixture than that for the coarse particles alone. An increase in the liquid viscosity led to a decrease of the minimum spouting velocity and hence the bed pressure drop. Ishikura et al.\textsuperscript{28,29} also investigated the elutriation of the fine particles from a liquid spouted bed of binary solids in conical-based cylindrical columns of 66 and 100 mm diameter. These results were compared to similarly sized fluidized beds employing a perforated plate instead of an inlet nozzle. The initial fines entrainment velocity, $U_{be}$, was found to be smaller than the free-settling terminal velocity of the fines, but larger than the velocity required to entrain the same species from a liquid-fluidized bed. It was also found that $U_{be}$ decreased with a reduction in freeboard height, making the particles more subject to elutriation. An empirical correlation for the elutriation coefficient was strongly dependent on a modified Froude number. Other variables – the column diameter, the initial total particle holdup, initial fines mass fractions, and nozzle diameter – were found to have little effect on this elutriation coefficient.

20.2 Liquid–gas spouting

Functionally, liquid–gas spouting systems are analogous to three-phase fluidization, and thus many expressions used to predict the dynamics of the latter hold for the former. Basically, two types of contactors have been investigated: contactors in which the gas is a dispersed phase, including liquid–gas spouted beds with draft tube, and contactors in which the liquid is a dispersed phase. The principal applications of these systems are similar to those for three-phase fluidized beds.

20.2.1 Gas as a dispersed phase

20.2.1.1 Liquid–gas spouted and spout-fluid beds

Nishikawa et al.\textsuperscript{30} studied gas absorption in a liquid–gas spouted bed by monitoring oxygen dissolution into a copper-catalyzed sodium sulfite solution in a conical-based cylindrical column of diameter 150 mm and length 1400 mm. Air and water were introduced at the cone base through a nozzle whose diameter varied from 10 to 45 mm. Glass spheres of diameters 1.01, 2.59, 3.10, and 4.87 mm were the solid particles. Bed mass varied between 2 and 10 kg. For the largest solids mass loading, the estimated
$H/D$ ratio was ∼3. Experiments conducted at gas and liquid superficial velocities of 4.7 and 3.8 cm/s, respectively, showed that particles smaller than 3 mm were not effective because of channeling in the central portion of the bed. With particles larger than 3 mm, the gas–liquid volumetric mass transfer coefficient was more than 1.75 times larger than that of a column without particles because of efficient bubble breakage in both the spout region and in the fluidized bed region above the spout.

Wang et al. proposed the use of a three-phase spouted bed with large particles and a downward annular liquid flow for gas–liquid reaction systems in which deposition and adhesion on the nozzle or the reactor wall limited effectiveness. Large particles were found to reduce fouling because of their high impact momentum. Glass beads of 10.5, 12.0, and 16.0 mm diameter were used as the solid phase. The column was 230 mm in diameter and was equipped with a slotted conical base. Water in the column leaked into a reservoir through the slots and was pumped back to the column for spouting the glass particles through a 16-mm-diameter nozzle. Air was introduced through a ring-type distributor equipped with six 8-mm-diameter gas injection tubes. Wang et al. identified three characteristic flow regimes: a packed bed regime, a transition regime, and a spouting regime. The experimental minimum liquid spouting velocity was found to increase with increasing static bed height and to decrease slightly with increasing gas velocity. The spout was effective for bubble breakup, whereas frequent collisions of the particles with the reactor wall and the nozzles prevented fouling and promoted self-cleaning. Gas holdup increases accompanied corresponding increases of the superficial gas and liquid circulation velocities. Gas holdup varied between 0.1 and 0.5, depending on bed operating conditions.

Anabtawi et al. investigated liquid-phase mass transfer from the dispersed gas in a three-phase spout-fluid bed at conditions above and below the minimum spout-fluid values. They followed oxygen absorption in carboxymethylcellulose solutions under varying rheological conditions. The gas-absorbing solution was introduced into the surrounding annular section and the spouting air through a central nozzle. Spherical glass particles of diameter 1.75 mm were the solids. These authors stated that the main advantage of this contacting system was the enhanced mixing associated with a vigorous circulatory motion caused by rising bubbles.

### 20.2.1.2 Liquid–gas spouted and spout-fluid beds with a draft tube

The gas–liquid–solid fluidized bed has emerged in recent years as one of the most promising devices for three-phase operations. A new design with a spouting orientation incorporates a draft tube that is coaxially located inside the bed with a gas and liquid injector under the draft tube. The draft tube promotes bulk circulation of gas, liquid, and solids between the draft tube and annulus, thereby achieving intimate contacting between phases (Figure 20.3). If there are no particles flowing in the draft tube, the system is, in essence, an internal loop gas lift reactor. Fan et al. conducted the most detailed study of this type of contacting system. They identified flow regimes and studied an array of variables: pressure profiles and pressure drops, bubble penetration depth in the annulus, overall gas holdup, apparent liquid circulation rate, and bubble size distribution.
in both the draft tube and annulus. Three flow regimes were identified: a packed bed mode, a fluidized bed mode, and a circulated bed mode. The circulated bed mode is of particular interest, as this bed mode generates both high gas holdup and efficient contacting between phases. The transition points between the respective regimes are a complex function of particle size, particle density, and particle loading.

The overall gas holdup increases with an increase in either gas or liquid superficial velocities. In all cases the overall gas holdup of a draft tube three-phase spouted bed is significantly higher than that for the corresponding three-phase fluidized bed system.

Different hydrodynamic aspects of liquid–gas spouted and spout-fluid beds with a draft tube have also been investigated by Karamanev et al., Hwang and Fan, Vunjak-Novaković et al., Kundaković et al., Cecen Erbil, Nitta and Morgan, Olivieri et al., and Pironti et al.

20.2.2 Liquid as a dispersed phase

Fluidized bed contactors containing large, low-density spheres considerably improve gas–liquid contacting in scrubbers, absorbers, distillation columns, and direct cooling towers. They operate at much higher gas and liquid rates before flooding than conventional packed towers and have lower static bed heights. Lighter packing and higher fluid flowrates also make them efficient low-pressure-drop contactors. Movable packing
systems allow for the handling of gases and liquids containing particulate matter and precipitates, and, for all practical purposes, clogging is eliminated. The highly turbulent states induced in such contactors reveal transfer rates two orders of magnitude higher than those in packed beds.\textsuperscript{43} These fluidized bed contactors do possess certain shortcomings – they tend to channel, and liquid backmixing makes true countercurrent operation impossible. Vuković et al.\textsuperscript{44} developed a three-phase spouted bed contactor in which the solid particles are spouted in a countercurrent flow of gas and liquid. A schematic diagram of their system is shown in Figure 20.4. Hollow 10-mm-diameter polyethylene spheres ($\rho_p = 320$ kg/m$^3$) were spouted in a 194-mm-diameter column. A perforated 60$^\circ$ conical section, 86 mm high, served both as a bed support and water drain. Air was introduced axially through a converging nozzle 30 mm in diameter, and water was admitted through a distributor containing sixty 2-mm-diameter tubes.

Two general flow regimes were observed in this three-phase spouted bed contactor as the gas flowrate was raised. In the higher-flowrate regime, the pressure drop, total liquid holdup, and bed expansion all increased with gas flowrate. The “active” holdup, given by the ratio of the measured bed pressure drop/pressure drop owing to the total weight of the bed, rises rapidly in this flow rate regime and it is likely that interphase transfer processes then occur at the highest rate. A spouted bed contactor with gas and liquid mass flowrates of 2.18 and 1.88 kg/m$^2$s, respectively, had similar pressure drop per unit area of particle surface, total holdup per unit of operating bed volume, and “active” holdup as a comparable fluidized bed contactor.
20.2.3 Processing in liquid–gas spouted beds

Several possible applications of liquid–gas spouted beds have been reported. Kechagiopoulos et al. investigated a spouted bed reactor system for hydrogen production via reforming of ethylene glycol, which served as a representative compound for aqueous phase bio-oil. An olivine–nickel alloy was a suitable catalyst with high reforming activity, excellent anticoking characteristics, and exceptional mechanical strength. Coke formation, a major problem in most reforming processes, was drastically reduced because of rapid and effective mixing of the hot solid particles and the cold reactants. Wright and Raper investigated fluidized and spouted beds for the biological filtration of gases because of their high specific gas flowrate and vigorous mixing that facilitates enhanced gas–biomass contacting. Switching from a conventional fluidized bed to a spouted bed improved yields by 10 percent to 20 percent. Nieto et al. and Tailleur studied the butene–isobutane alkylation reaction on a super-acid solid catalyst in a three-phase spouted bed reactor. The main result was that the three-phase draft tube spouted bed reactor improved both gas–solid contact and liquid mixing because of internal and external recycling. Dabhade et al. investigated the degradation of phenol by immobilized Nocardia hydrocarbonoxydans (actinomycetes) on granular activated carbon in a gas–liquid–solid three-phase spouted bed contactor. The effect of an array of variable flow rates, influent concentrations, and solid loadings on phenol degradation was examined. It was found that adsorption was dominant during the initial reaction phase, whereas biodegradation dominated after adsorption equilibrium was reached. In general, higher liquid flow rate increases shear between the spout and the annulus and enhances turbulence; unfortunately, in biofilm reactors, this turbulence caused biomass detachment, resulting in poor degradation of phenol. Elmaleh et al. evaluated the oxygen transfer characteristics of a high compact multiphase reactor (HCMR) for wastewater treatment employing cells immobilized on spouting particles. The high turbulence levels caused significant particle–particle attrition, limited the biofilm thickness around the particles, and lowered the biomass holdup. Even though biomass holdup was relatively low in these HCMRs, they were effective low-energy and low-sludge-producing reactors.

Chapter-specific nomenclature

$A_f$ parameter defined by Eq. (20.8)
$C_p$ parameter defined by Eq. (20.4)
$f_1 = 150((1 - \epsilon)^2/\epsilon^3) \mu/d_p^2$
$f_2 = 1.75((1 - \epsilon)/\epsilon^3) \rho/d_p$
$g(\phi_s)$ particle sphericity function, $g(\phi_s) = 1$ for spherical particles
$H_{mSF}$ maximum spoutable bed height at the minimum spout-fluid condition, m
$H^*$ lowest bed height for a one-dimensional model based on Eq. (20.3), m
$U_{a0}$ superficial fluid velocity at the inlet to the annulus ($= V_{a0}/A_a$), m/s
$U_G$ superficial gas velocity in three-phase bed, m/s
$U_L$ superficial liquid velocity in three-phase bed, m/s
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$U_s$ superficial fluid velocity in the spout, m/s
$U_T$ terminal velocity of a single particle, m/s
$V$ total fluid flow rate ($= V_{a0} + V_{n0}$), m$^3$/s
$V_{a0}$ inlet annulus flowrate, m$^3$/s
$V_{mf}$ minimum fluidizing flow rate, m$^3$/s
$V_{ms}$ minimum spouting flow rate, m$^3$/s
$V_{mSF}$ total fluid flow rate at the minimum spout-fluid condition, m$^3$/s
$V_{n0}$ fluid flow rate in spout inlet tube, m$^3$/s

$V_T = U_T \cdot A$, m$^3$/s

$X = 1/(1 + H/D)$
$Y = 1 - \Delta P_{ms}/\Delta P_{mf}$

**Greek letters**

$\varepsilon_g$ gas holdup in three-phase bed
$\theta_f$ parameter defined by Eq. (20.5)
$\phi_s$ sphericity

**Subscripts**

$a$ in the annulus
$mf$ minimum fluidization
$ms$ minimum spouting
$mSF$ minimum spout-fluid condition

**References**


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