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Problems in Catalyst Recirculation in a Catalytic Cracking Plant

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Summary

In a fluidized bed catalytic cracking plant, smooth catalyst recirculation between a reactor and a regenerator is important. Spent catalyst flows at rates up to 120 tonne min^{-1} from the reactor down a standpipe to the regenerator. The regenerated catalyst flows through another standpipe to be transported back to the reactor via a riser. Operating problems such as insufficient pressure build-up in the pipes, pressure reversal, inadequate solid flow, flow instability or complete stoppage sometimes occur. These problems can often be traced to mal-operation in the standpipes. Remedial action by injecting steam (aeration) into the standpipes to counteract compressibility of the gas sometimes produces an effect counter to that intended. In this paper the causes of operating problems are discussed in the light of recent advances in the understanding of gas-solid flow in standpipes (Dries, 1980; Ginestra et al., 1980; Jones et al., 1980). In particular the effects of slide valve opening, aeration gas injection rate and catalyst size on flow instability will be discussed.

Nomenclature

A	cross-sectional area, m^2
C_d	discharge coefficient
D	diameter of pipe, m
d	particle diameter, m
f	friction factor
G	mass flux, kg m^{-2}s^{-1}
$\dot{M}$	mass flow rate, kg s^{-1}
P	pressure, N m^{-2} or Pa
U	superficial velocity, m s^{-1}
U_{mb}	minimum superficial velocity when bubbles appear, m s^{-1}
U_{mf}	minimum fluidization velocity, m s^{-1}
U_{sl}	slip velocity, m s^{-1}
V_w	continuity wave velocity, m s^{-1}
z	coordinate in vertical direction, positive upwards, m
Δp	pressure drop, N m^{-2}
ϵ	voidage
ϵ_c	vibrated bed voidage
ϵ_{mf}	bed voidage at minimum fluidization
μ	shear viscosity, kg m^{-1} s^{-1}
ϱ	density, kg m^{-3}
σ	normal stress, N m^{-2} or Pa
τ	shear stress, N m^{-2} or Pa
ϕ	shape factor

Subscripts

c	vibrated bed
g	gas
mb	minimum bubbling
mf	minimum fluidization
or	orifice
p	particle
sl	slip
w	wall

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Introduction

A catalytic cracking plant usually consists of reactor, stripper and regenerator interconnected through pipes known as standpipes or risers as depicted in Fig. 1.

Hydrostatic head difference in the sides of the circuit enables the catalyst to circulate at enormous rates of up to 120 tonne min^{-1} (Matsen, 1976). On the regenerator side, solids descend down the standpipe by gravity in a dense form, through the slide valve; then are conveyed pneumatically up the riser into the reactor in a very dilute fashion.

The main function of a standpipe is often to transport solid (and gas) from a region of low pressure to a region of higher pressure. During normal operation pressure increases in the downwards direction as a result of gravity head and there is a positive pressure difference across the slide valve (i.e., $P_2 > P_3$ in Fig. 1). The slide valve controls solid circulation rate and prevents oil vapour from getting into the regenerator which otherwise may form an explosive mixture in the regenerator. Such occurrence is prevented in practice by the automatic shut off of the slide valve when (P_2 minus P_3) falls below a preset value. Such valve shut off can lead to costly shutdown of the entire plant. Thus a stable and adequate pressure build-up in the standpipe is important to ensure smooth catalyst circulation. In this paper, the causes of the common problems will be discussed in the light of recent advances in understanding of standpipe flow.

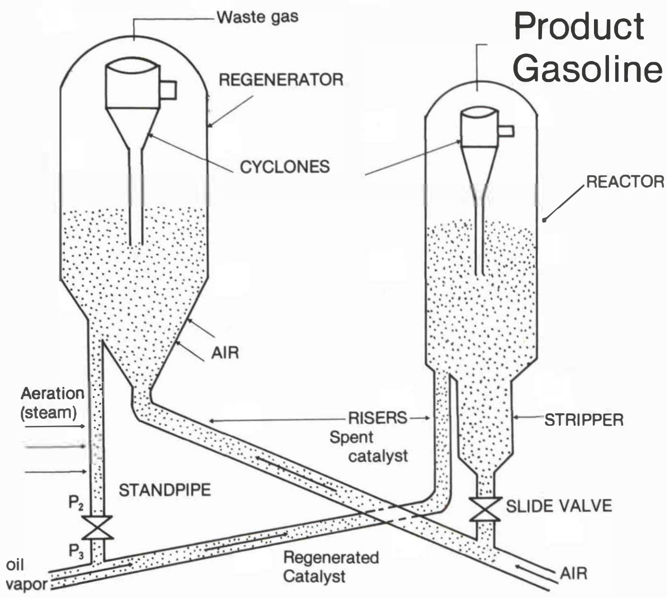

Fig. 1: Hydrocarbon catalytic cracking process

2. Reported Industrial Problems

Matsen (1976) reported the normal and unusual operating experience of some 65 catalytic crackers and fluid cokers, with standpipe diameters up to 1.5 m and heights 40 m, circulation rates to 120 tonne min^{-1}, and pressure build-up of almost 3 bar. He explained that poor pressure build-up and flow interruptions with subsequent loss of solid circulation were due to the presence of large bubbles in the pipes.

Dries (1980) recently reported some erratic flow and insufficient pressure build-up in a 21 m long, 0.86 m diameter standpipe in a catalytic cracking plant equipped with a number of aeration points. Problems encountered include low circulation rate, interruption of recirculation and loss of the pressure drop across the slide valve. The problems lead to costly shut down and the low recirculation rate restricts plant throughput. Dries pointed out the importance of adding fines to the catalyst inventory to promote an increase in solid recirculation rate. He stressed the importance of setting the correct aeration rate into the standpipe to counteract the effect of gas compression. Insufficient aeration rate would result in low solid recirculation while excessive aeration rate would cause arching of solids in the pipe at the aeration point.

3. Flow Regimes

Classification of the four flow modes is made possible as a result of the works of Kojabashian (1958), Leung and Jones (1978 a,b) and Staub (1980) which also indirectly involved the ideas of Lapidus and Elgin (1957), Slis et al. (1959), Wallis (1969) and Matsen (1973). All these works are summarized by Leung (1980) who has now adopted the criteria to demarcate the four flow modes as the following:

(i) Type I fluidized flow

$U_{sl} > (U_{mf}/\epsilon_{mf})$

$(\partial U_g/\partial\epsilon)_{U_p} < 0$

or $V_w < 0$ [continuity wave downwards]

(ii) Type II fluidized flow

$U_{sl} > (U_{mf}/\epsilon_{mf})$

$(\partial U_g/\partial\epsilon)_{U_p} > 0$

or $V_w > 0$ [continuity wave upwards]

(iii) Transition packed bed flow

$0 < U_{sl} < (U_{mf}/\epsilon_{mf})$

$\epsilon_c < \epsilon < \epsilon_{mf}$

$\epsilon = \epsilon_c + (\epsilon_{mf} - \epsilon_c)\, U_{sl}/(U_{mf}/\epsilon_{mf})$

(iv) Packed bed (moving bed) flow

$U_{sl} < 0$

$\epsilon = \epsilon_c$

Note that all velocities are defined positive upwards. The slip velocity, U_{sl} is defined by:

$$U_{sl} = U_g/\epsilon - U_p/(1-\epsilon) \tag{1}$$

$(\partial U_g/\partial\epsilon)_{Up} = 0$ is the flooding point which defines the minimum gas velocity possible for a constant solid velocity. Conversely, at a constant gas velocity, flooding point defines the maximum possible downward solid velocity. The flooding point is related to that where the continuity wave V_w is zero, defined by:

$$V_w \equiv [\partial U_g/\partial\epsilon]_{(U_p + U_g)}$$
$$= (1-\epsilon)(\partial U_g/\partial\epsilon)_{U_p} \tag{2}$$

These four flow modes theoretically can occur in a standpipe, however, the flow behaviour can be complicated by the coexistence of flow regimes and also by the supplementary injection of aeration gas.

4. Coexistence of Flow Regimes

Kojabashian (1958), Kunii and Levenspiel (1969) have long recognised the coexistence of flow pattern in standpipes. Adopting Wallis' drift flux plot and Staub's continuity wave approaches and supported by experimental observations, Jones (1980) concluded that the three common types of coexistence are:

(i) Type I fluidized flow (with relatively high voidage) on top of non-fluidized flow (see Fig. 2).

(ii) Type I fluidized flow (with relatively low voidage) on top of non-fluidized flow as in Fig. 3.

(iii) Type I fluidized flow (with relatively high voidage) on top of type II fluidized flow (with relatively low voidage) as in Fig. 4.

The above three classes of coexistence can occur without the injection of aeration gas into the tube (see Judd and Rowe, 1978; Judd and Dixon, 1976) and with no gas compression effects as a result of change in pressure in the standpipe (Jones, 1980). Staub has suggested that coexistence of the two flow patterns is possible when the continuity waves on either side of the interface propagate towards the interface.

Jones et al. (1980) have shown that this condition is satisfied in the above classes of flow coexistence.

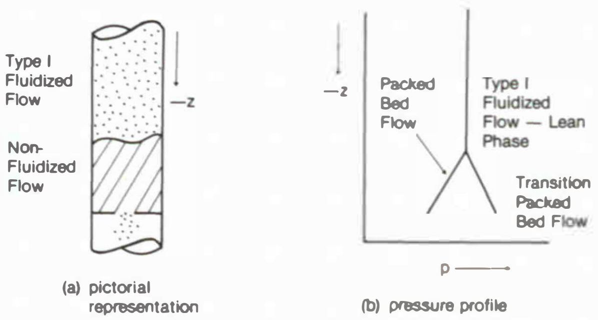

Fig. 2: Coexistence of Type I fluidised flow (lean phase) with non-fluidised flow

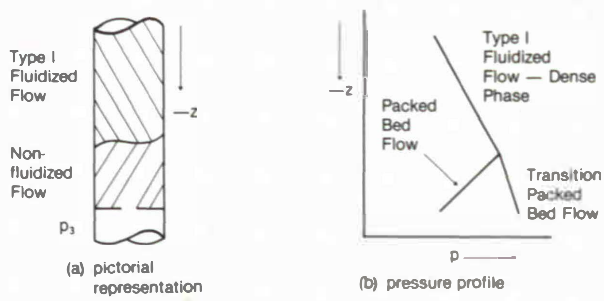

Fig. 3: Coexistence of Type I fluidised flow (dense phase) with non-fluidised flow

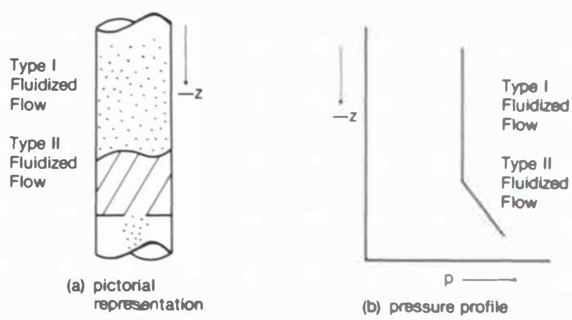

Fig. 4: Coexistence of Type I und Type II fluidized flow.

Dries (1980) and Ginestra et al. (1980) in their analyses have pointed out that addition of aeration gas can create other forms of flow regime coexistence.

5. Equations for Standpipe Flow

5.1 Non-Fluidized Flow

Two equations are available for describing gas pressure gradient and normal stress acting on the solid in non-fluidized flow. Yoon and Kunii (1970) showed that the pressure gradient can be written in terms of slip velocity using a modified Ergun (1952) equation, giving

$$-\frac{dp}{dz} = K_1 U_{sl} + K_2 U_{sl} |U_{sl}| \tag{3}$$

where

$$K_1 = [150\mu (1-\epsilon)^2] / (\phi d\epsilon)^2$$
$$K_2 = 1.75\, \varrho_g (1-\epsilon) / (\phi d \epsilon)$$

The mean normal stress $\bar{\sigma}$ acting on a horizontal plane is related to the pressure gradient for fully developed flow by (Grossman, 1975; Spink and Nedderman, 1979)

$$\frac{d\bar{\sigma}}{dz} + \frac{dp}{dz} + \frac{4\tau_w}{D} + (\varrho_p - \varrho_g)(1-\epsilon) g = 0 \tag{4}$$

τ_w, the wall shear stress is related to the mean stress, the internal angle of friction and the wall angle of friction by different equations of varying sophistication (Janssen, 1895; Brown and Richards, 1960; Walker, 1966; Walters, 1973).

5.2 Fluidized Flow

In fluidized flow, $\bar{\sigma} = 0$. Neglecting the gas momentum and the gas wall friction term, the pressure difference between two points in a standpipe can be calculated from (Hinze, 1962)

$$[p]_0^z = \Delta p = \int_0^z + \varrho_p (1-\epsilon) g\, dz + \left[\frac{G_p^2}{\pi_r (1-\epsilon)} \right]_0^z$$

$$+ \int_0^z \left[2 f_p G_p^2 / [\varrho_p (1-\epsilon) D] \right] dz \tag{5}$$

For fully developed fluidized flow, voidage can be obtained from an appropriate fluidized bed expansion equation such as the Matsen equation (1973), the Richardson-Zaki (1954) equation or an experimentally determined correlation.

5.3 Fluidized Flow Through an Orifice

Solid flow. Jones and Davidson (1965) treated the flowing mixture as an inviscid liquid and by applying Bernoulli's theorem, the solid mass flux can be predicted as

$$|G_p| = C_d (A_{or}/A) \sqrt{\varrho_p (1-\epsilon_{mf}) \Delta p_{or}} \tag{6}$$

C_d = 0.5 to 0.65 agrees with extensive results obtained by Massimilla et al. (1961), Stemerding et al. (1963), Burkett et al. (1971), de Jong and Hoelen (1975) and Do (1976). Despite this agreement, Jones and Davidson cautioned the analogy of liquid and gas-solid systems because the exit from the orifice of the gas-solid mixture did not show any "vena contracta"!

Gas flow. For flow of fluidized gas-solid mixture through an orifice, Stockel (1962) was the first to consider the effect of voidage variation. He described ΔP_{or} as a function of slip velocity. This principle is further applied by many workers.

De Jong and Hoelen (1975) adapted the Ergun equation to relate the solid and gas flowrates as:

$$-\Delta p_{or} = K_3 U_{sl} + K_4 U_{sl} \, |U_{sl}| \tag{7}$$

where

$$K_3 = K_1 D_{or} (A/A_{or})/4$$
$$K_4 = K_2 D_{or} \, (A/A_{or})/24$$

This equation was derived by assuming that the streamlines of solid and gas are straight lines converging towards the orifice, with isobar planes forming the surfaces of hemispheres concentric with the orifice, and of constant voidage. In spite of these unrealistic assumptions, equation (7) does agree with experimental observations and predictions from the more sophisticated analyses of Burkett et al. (1971) and Do (1976).

Coupled with the solid flow equation (6), equation (7) permits the calculation of gas flow through an orifice if Δp_{or} is known.

5.4 Non-Fluidized Flow Through an Orifice

Brown and Richards (1960, 1965), Beverloo et al. (1961), Zenz (1976), Davidson and Nedderman (1973) and Williams (1977) have presented equations for solid flow. Beverloo et al. (1961) presented the equation,

$$\dot{M}_p = 1.84 \times 10^{-8} \varrho_p (1 - \epsilon_{mf}) g^{1/2} (D_{or} - Kd)^{2.5} \tag{8}$$

K varies for different types of solid.

Equation (8) shows good agreement with six types of solids. Crewdson et al. (1977) modified equation (8) to account for the pressure gradient while Harmens (1963) went a step further to account for the internal angle of friction and half angle of hopper. Further work to model moving bed flow through an orifice at the bottom of a standpipe is needed.

Little work has been done on effects of pressure gradient and aeration on the flow of non-fluidized solids through an orifice. Ginestra et al. (1980) and Jones (1980) have presented theoretical equations for relating gas and solid flow-rates through the moving bed and the orifice. Much different from fluidized flow, this is now complicated by $\bar{\sigma}$ in addition to Δp. Careful experiments will be required to test their equations.

6. Stability

At least four aspects of stability are relevant in standpipes:

(i) the hydrodynamic stability of downwards uniform suspension flow;
(ii) "flooding" instability as represented by $(\partial U_g/\partial \epsilon)_{U_p} = 0$.
(iii) multiple steady state instability
(iv) pseudobridge due to aeration.

6.1 Hydrodynamic Stability of Downwards Uniform Suspension Flow

Jackson (1963) and Pigford and Baron (1965) had shown that the state of uniform fluidization is inherently unstable. Instability is affected by ϱ_p/ϱ_g and particle size, supported by the experimental evidence of Wilhelm and Kwauk (1948), Geldart (1973), El-Kaissy and Homsy (1976). Grace and Tuot (1979) extended the Jackson analysis to cocurrent gas-solid upflow. Jones (1980) presented a similar analysis for solid downflow. Like Jackson, they show that the state of uniform suspension flow is inherently unstable even at voidage remote from ϵ_{mf}. The rate of growth of voidage waves varies greatly from system to system as it does in a stationary fluidized bed. They conclude that all gas-solid flows tend to form "clusters" or "streamers" or "bubbles" in gas-solid downflow. Such a conclusion corresponds to the experimental observations of Judd and Dixon (1976), Matsen (1976) and Judd (1977).

6.2 Flooding Instability

The flooding phenomenon in two phase flow refers to a limiting flow-rate condition at which no steady-stage operation is possible. Lapidus and Elgin (1957), Quinn et al. (1961), Kwauk (1963, 1974, 1980), and Wallis (1969) all have discussed the phenomenon of flooding.

In a standpipe, flooding occurs when $(\partial U_g/\partial \epsilon)_{U_p} = 0$. Matsen (1973) suggested that instability sets in when the solid downflow velocity is equal to the velocity of rise of the single bubbles in the system. This corresponds to the definition of $(\partial U_g/\partial \epsilon)_{U_p} = 0$. Hence the industrial problem whereby a bubble is held stationary in the pipe, corresponds to the flooding point for that system. Also, Matsen observes that at this point the stationary bubble grows downwards until it occupies the whole pipe. If we now consider $(\partial \epsilon/\partial U_g)_{U_p} \rightarrow \infty$ whence $(\partial U_g/\partial \epsilon)_{U_p} = 0$, for a small variation in U_g, the voidage has a run away response. Perturbation of U_g and ϵ can be the consequence of pressure fluctuation or aeration rate or change of valve opening. So Matson's observation that a bubble can grow leaving a great portion of the pipe empty seems to have some theoretical backing.

6.3 Stability of Multiple Steady States

Under certain operating conditions for a particular system, the system equations may exhibit multiple roots. The prediction of the stable steady states can be obtained using a supply-demand analysis in the style of Ledinegg (1938) as discussed by Jones et al. (1980). The existence of multiple steady states can lead to hysterisis effects. Increase of aeration rate beyond a critical value can result in a change of flow pattern from high solid circulation to low solid circulation with a subsequent loss of pressure in standpipe. This is known to occur in some industrial standpipes. Subsequent reduction of aeration rate to below the critical value may not revert the system back to high solid flow.

The analysis of Jones et al. (1980) concluded that for one flow mode to occur throughout the standpipe, fluidized flow is stable within a very narrow range of voidage around ϵ_{mf} and a much wider range of voidage near 1. This conclusion is supported by the experimental observations of Judd and Rowe (1978). This conclusion may throw light on some industrial problems whereby it is difficult to maintain and operate standpipes at their optimum design capacity. Often, to achieve a high pressure build-up and to have a high solid rate, the standpipe must be designed to operate in single flow mode, that is, a dense fluidized flow at $\epsilon \approx \epsilon_{mf}$. However, the range of stability in the vicinity of ϵ_{mf} is limited, any small fluctuation in operating conditions may swing its operation away from the ϵ_{mf} zone. As a result of such swings, significant pressure loss may occur, thereby resulting in flow interruption or significant reduction in solid circulation rate.

6.4 Pseudobridge

Industrially, Dries (1980) has observed the occurrence of an arch in which solids are supported by aeration gas and wall friction to form a stable arch above the aeration point. Laboratory tests on the catalyst rule out the possibility of the powder having such enormous cohesive force to be able to span the width of the pipe forming a stable arch. Dries described the arch as a "pseudobridge", the existence of which is also predicted by a one dimensional analysis of Ginestra et al. (1980). The formation of a pseudobridge will cause interruption of solid circulation and could lead to costly plant shutdown.

7. Flow Regime Diagram

A quantitative flow regime diagram mapping out the regions of stability and instability will go a long way in assisting in the design and operation of standpipes. Unfortunately, such a stage has not been reached yet. Nevertheless, the flow regime diagrams introduced by Kojabashian (1958), Zenz (1960), Leung and Jones (1978), Ginestra (1980) and Dries (1980) have all contributed to the understanding of flow behaviour in solid-gas circulatory systems. Kwauk (1963, 1980) has unified the theories in the particulate fluidization of liquid-solid systems and presented a number of nomographs from which some commercial plants could be designed. He has also opened up the idea that his unified theory of general fluidization may be extended to aggregative fluidization of gas-solid systems.

To define a flow pattern, specification of solid velocity alone is not adequate, a knowledge of the voidage is also necessary. Direct measurement of voidage without disturbing the flow requires expensive radiation techniques such as x-ray or γ-ray attenuation. A knowledge of the pressure gradient alone will be insufficient in assessing flow pattern in a pipe.

8. Fluidity of Particles

Recently Tsutsui and Miyauchi (1980 have studied the effect of fines and size distribution on the "fluidity" of particles with mean diameter ranging from 43 to 169 μm and below 44 μm fines ranging from 0.3 to 32 weight %. A system is deemed to exhibit good "fluidity" if it has a large U_{mb}/U_{mf} and the bubbles present are small. They concluded that for systems with the same surface-volume mean diameter, good fluidity is exhibited by systems with wide size ranges and with high percentage of fines. This conclusion is in support of the empirical observation that the presence of fines may improve operation in a stand-pipe. It is likely that by increasing the range of bubble free operation velocity (i.e. between U_{mf} and U_{mb}), the range of voidage for stable standpipe operation also increases. Further experimental work will be useful to establish this effect on a quantitative basis.

9. Conclusions

Although there has been recent progress in the understanding of standpipe flow, much remains unknown. Major problems such as the effect of aeration pattern, maldistribution of gas at an aeration point, the effect of particle size distribution and mean particle diameter in flow instability, the effect of standpipe inlet and outlet designs, effect of angle of inclination on flow in standpipes, and flow of a dilute mixture through a slide valve have yet to be resolved. Industrial problems on gas-solid circulation and the success or failure of any remedial actions taken should be discussed in the open literature. This will provide valuable information to check the theoretical analyses reported in the literature. Systematic experimental work on large scale standpipes has to be carried out to verify these theoretical analyses.

References

[1] Beverloo, W.A., Leniger, H.A., & van de Velde, J., "The Flow of Granular Solids Through Orifices", Chem. Eng. Sci., **15**, 260 (1961).

[2] Brown, R.L. and Richards, J.C., "Profile of Flow of Granules Through Apertures", Trans. Inst. Chem. Engrs., **38**, 243 (1960).

[3] Brown, R.L. and Richards, J.C., Rheologica Acta, **4**, 153 (1965).

[4] Burkett, R.J., Chalmers-Dixon, P., Morris, P.J., Pyle, D.L., "On the Flow of Fluidized Solids Through Orifices", Chem. Eng. Sci., **26**, 405 (1971).

[5] Crewdson, B.J., Ormond, A.L. and Nedderman, R.M., "Air-Impeded Discharge of Fine Particles from a Hopper", Powder Technology, **16**, 197 (1977).

[6] Davidson, J.F. and Nedderman, R.M., "The Hour Glass Theory of Hopper Flow", Trans. Inst. Chem. Engrs., **51**, 29—35 (1973).

[7] Do, D.D., "Flow of a Fluidized Solid and Gas Through an Orifice and the Design of a Standpipe in the Fluidized Flow Regime", B.E. Thesis, University of Qld., Australia (1976).

[8] Dries, H.W.A., "Cocurrent Gas/Solids Downflow in Vertical Cat. Cracker Standpipes. Effects of Gas Compression and Solids Compaction", Proceedings of Engineering Foundation Conference, Henniker, N.H. (1980).

[9] El-Kaissy, M.M. and Homsy, G.M., "Instability Waves and the Origin of Bubbles in Fluidized Beds", Int. Jnl. Multiphase Flow **2**, 379—395 (1976).

[10] Ergun, S., "Fluid Flow Through Packed Columns", Chem. Eng. Prog., **48** (2), 89—94 (1952).

[10a] Geldart, D., "Types of Gas Fluidization", Powder Technology, **1**, 285 (1973).

[11] Ginestra, J.C., Rangachari, S. and Jackson, R., "Flow Regimes in a One-Dimensional Model of a Standpipe", Proceedings of Eng. Foundation Conf., Henniker, N.H, (1980).

[12] Grace, J.R. and Tuot, J., "A Theory for Cluster Formation in Vertically Conveyed Suspension of Intermediate Density", Trans. Inst. Chem. Engrs., **57**, 49—54 (1979).

[13] Grossman, G., "Stresses and Friction Forces in Moving Packed Beds", AIChE J., **21** (4), 720—730 (1975).

[14] Harmens, A., "Flow of Granular Material Through Horizontal Apertures", Chem. Eng. Sci., **18**, 297—306 (1963).

[15] Hinze, J.O., "Momentum and Mechanical Energy Balance Equations for a Flowing Homogeneous Suspension with Slip Between the Two Phases", App. Sci. Res. **A**, **11**, 33 (1962).

[16] Jackson, R., "The Mechanics of Fluidized Beds: Part I: The Stability of the State of Uniform Fluidization", Trans. Inst. Chem. Engrs., **41**, T13—21 (1963).

[17] Janssen, H.A., "Versuche über Getreidedruck in Silozellen", Z. Verein Deutscher Ingenieure, **39**, 1045 (1895).

[18] Jones, D.R.M. and Davidson, J.F., "The Flow of Particles from a Fluidized Bed Through an Orifice", Rheologica Acta **4**, (3), 180 (1965),

[19] Jones, P.J., Teo, C.S. and Leung, J.S., "The Stability of Vertical Gas-Solid Downflow in Bottom-Restrained Standpipes", Proceedings Third Engineering Foundation Conference on Fluidization, Henniker (1980).

[20] Jones, P.J., Ph. D. Thesis, University of Qld. (1980).

[21] de Jong, J.A.H., "Aerated Solids Flow Through a Vertical Standpipe Below a Pneumatically Discharged Bunker", Powder Technology, **12**, 197—200 (1975).

[22] Judd, M.R. and Dixon, P.D., "The Flow of Fine Dense Solids Down a Vertical Standpipe", Paper presented at AIChE Annual Conference, Chicago (1976).

[23] Judd, M.R., Proceedings Particle Technology, Nuremberg, Ed. Brauer, H. and Molerus, O., **3**, Di18 (1977).

[24] Judd, M.R. and Rowe, D.N., "Dense Phase Flow of a Powder Down a Standpipe", Proceedings International Fluidization Conference, Ed. Davidson, J.F., Cambridge University Press, 110—115 (1978).

[25] Kojabashian, C., "Properties of Dense-Phase Fluidized Solids in Vertical Downflow", D.Sc. Thesis, Massachusetts Institute of Technology, Cambridge, Massachusetts (1958).

[26] Kunii, D. and Levenspiel, O., *Fluidization Engineering,* John Wiley and Sons, New York (1969).

[27] Kwauk, M., "Generalized Fluidization I: Steady State Motion", Scientia Sinica, **12** (4), 587—612 (1963).

[28] Kwauk, M., "Particulate Fluidization in Chemical Metallurgy", Proceedings of the First Iranian Congress of Chemical Engineering, (ed.) Davalloo, P. et al., Vol. 2, 539—558, Elsevier (1974).

[29] Kwauk, M., "Towards a Unified Hypothesis for Fluidized Systems", Proceedings 8th Australian Chemical Engineering Conference, Melbourne (1980).

[30] Lapidus, L. and Elgin, J.C., "Mechanics of Vertical-Moving Fluidized Systems", AIChE J., **3**, 63 (1957).

[31] Ledinegg, M., "Unstabilität der Strömung bei natürlichem und Zwangumlauf", Die Wärme, **61** (48), 891—898 (1938).

[32] Leung, L.S. and Jones, P.J., "Coexistence of fluidized solids flow and packed flow in standpipes", Proc. International Fluidization Conference, Ed. Davidson, J.F., Cambridge University Press, 116 (1978a).

[33] Leung, L.S. and Jones, P.J., "Flow of Gas-Solid Mixtures in Standpipes — A Review", Powder Technology, **20**, 145—160 (1978b).

[34] Leung, L.S., "The Ups and Downs of Gas-Solid Flow — A Review", Proc. Third Engineering Foundation Conference on Fluidization, Henniker (1980).

[35] Massimilla, L., Betta, V. and Della Rocca, C., "A Study of Streams of Solids Flowing from Solid-Gas Fluidized Beds", AIChE J., **7**, 502 (1961).

[36] Matsen, J.M., "Flow of Fluidized Solids and Bubbles in Standpipes and Risers", Powder Technology, **7**, 93—96 (1973).

[37] Matsen, J.M., "Some Characteristics of Large Fluid-Solids Circulation Systems", Proceedings in Fluidization Technology, eds. Keairns, D. et al., Vol. II, 135—153, Hemisphere Publishing Corporation, Washington D.C. (1976).

[38] Pigford, R.L. and Baron, T., "Hydrodynamic Stability of a Fluidized Bed", Ind. Engng. Chem. Fundam., **4**, 81 (1965).

[39] Quinn, J.A., Lapidus, L. and Elgin, J.C., "The Mechanics of Moving Vertical Fluidized Systems: V. Concurrent Cogravity Flow" AIChE J., **7** (2), 260—263 (1961).

[40] Richardson, J.F. and Zaki, W.N., "Sedimentation and Fluidization: Part I.", Trans. Inst. Chem. Engrs., **32**, 35—53 (1954).

[41] Slis, P.L., Willemsee, Th.W. and Kramers, H., "The Response of the Level of a Liquid Fluidized Bed to a Sudden Change in the Fluidizing Velocity", App. Sci. Res. **A**, **8**, 209 (1959).

[42] Spink, C.D. and Nedderman, R.M., "Gravity Discharge Rate of Fine Particles from Hoppers", Powder Technology, **21**, 245—261 (1978).

[43] Staub, F.W., "Steady-State and Transient Gas-Solids Flow Characteristics in Vertical Transport Lines", Powder Technology, **26**, 147—159 (1980).

[44] Stemerding, S., de Groot, J.H. and Kuypers, G.M.J., Proceedings of the Fluidization Symposium Society of Chemical Industries, London (1963).

[45] Stockel, I.H., "High Speed Flow of Fluidized Solids in Changing Area", Chem. Eng. Prog. Symp. Series, **58** (38), 106—120 (1962).

[46] Tsutsui, T., Miyauchi, T., "Fluidity of a Fluidized Catalyst Bed and Its Effect on the Behaviour of the Bubbles", Int. Chem. Eng., **20** (3), 386—393 (1980).

[47] Wallis, G.B., *One-Dimensional Two-Phase Flow*, McGraw-Hill, New York (1969).

[48] Walker, D.M., "An Approximate Theory for Pressures and Arching in Hoppers", Chem. Eng. Sci., **21**, 975—997 (1966).

[49] Walters, J.K., "A Theoretical Analysis of Stresses in Silos with Vertical Walls", Chem. Eng. Sci., **28**, 13—21 (1973).

[50] Wilhelm, R.H. and Kwauk, M., "Fluidization of Solid Particles", Chem. Engng. Prog., **44** (3), 201 (1948).

[51] Williams, J.C., "The Rate of Discharge of Coarse Granular Materials from Conical Mass Flow Hoppers", Chem. Eng. Sci., **32**, 247—255 (1977).

[52] Yoon, S.M. and Kunii, D., "Gas Flow and Pressure Drop Through Moving Beds", Ind. Engng. Chem. Proc. Design Dev., **9**, 559 (1970).

[53] Zenz, F.A., "Bulk Solids Efflux Capacity in Flooded and Streaming Gravity Flow" in Fluidization Technology, Vol. II, ed. Keairns, D.L. et al., 239, Hemisphere Publishing Corporation, Washington (1976).