Recent Progress in Efficient Gas–Solid Cyclone Separators with a High Solids Loading for Large-scale Fluidized Beds

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Abstract
Circulating fluidized beds (CFBs) are important technical equipment to treat gas–solid systems for fluid catalytic cracking, combustion, gasification, and high-temperature heat receiving because their mass and heat transfer rates are large. Cyclones are important devices to control the performance of CFBs and ensure their stable operation; heat-carrying and/or solid catalyst particles being circulated in a CFB should be efficiently separated from gas at a reduced pressure loss during separation. In commercial CFBs, a large amount of solids (> 1 kg-solid (m\(^3\)-gas\(^{-1}\)) or > 1 kg-solid (kg-gas\(^{-1}\)) is circulated and should be treated. Thus, gas–solid cyclones with a high solids loading should be developed. A large number of reports have been published on gas–solid separators, including cyclones. In addition, computational fluid dynamics (CFD) technology has rapidly developed in the past decade. Based on these observations, in this review, we summarize the recent progress in experimental and CFD studies on gas–solid cyclones. The modified pressure drop model, scale-up methodology, and criteria for a single large cyclone vs. multiple cyclones are explained. Future research perspectives are also discussed.

Keywords: gas–solid cyclone, circulating fluidized bed, high solids loading, design

1. Introduction
Circulating fluidized beds (CFBs) are commercially used for fluid catalytic cracking (FCC), and combustion and gasification of coal, biomass, and waste material because their heat and mass transfer rates are very large (Basu P., 2015; Grace J.R. et al., 1997; Gräbner M., 2015; Lettieri P. and Macri D., 2016; Li C.-Y. et al., 2017; Pecate S. et al., 2019; Scala F., 2013; Stolten D. and Scherer V., 2011). Recently, novel CFB solar receivers have been developed for concentrated solar power generation (Ansart R. et al., 2017; García-Triñanes P. et al., 2016, 2018). In a CFB, because heat-carrying and/or catalyst particles are circulated to transfer heat between each reactor, fast and efficient gas–solid separation is extremely important for a stable operation and reducing particle loss due to entrainment. Cocco R et al. (2017) investigated particle entrainment and clustering in a fluidized bed. Cyclones are widely used as primary gas–solid separators owing to their simple configuration and ease of operation. A stable operation of cyclones is important to reduce load in secondary gas–solid separators, makeup of the heat-carrying and/or catalyst particles, and pressure loss. Thus, much attention has been paid to the research and development of cyclones over the past several decades (Hoffmann A.C. and Stein L.E., 2010).

In commercial CFB boilers where carbonaceous solids are combusted to generate heat, a high solids-loading gas \( C_T > 10 \text{ kg-solid (kg-gas)}^{-1} \) as defined in Eq (1) should be treated in the primary gas–solid separator in order to efficiently separate solids from the gas (Van de Velden M. et al., 2007; Dewil R. et al., 2008).

\[
C_T = \frac{\text{total mass flow rate of solid to cyclones (kg-s⁻¹)}}{\text{total mass flow rate of gas to cyclones (kg-s⁻¹)}}
\]

Cortés C. and Gil A. (2007) extensively reviewed the models developed for the flow behavior, velocity profiles, pressure drop and collection efficiency in inverse-flow cyclones under the conditions of \( C_T < 0.23 \text{ kg-solid (kg-gas)}^{-1} \). Huard M. et al. (2010) comprehensively reviewed gas–solid separators including cyclones, impact separators, and other separators with a riser top and downer
bottom configuration. They summarized the solids-separation mechanism and separation efficiency of each separator and reported that recent research on reverse-flow cyclones is directed toward the influence of high solids loadings on cyclone performance and computational fluid dynamics (CFD) simulations. However, in the past, CFD calculations were not highly reliable for large solids-loading cyclones and fast separators due to limitations on computational power.

On the basis of such reviews and reports, in the current review, we summarized progress in the past decade in terms of experimental and numerical studies on high solids-loading gas–solid separators (mainly cyclones) in fast solids-circulation systems in CFBs.

2. High solids-loading cyclones

2.1 Pressure-drop model

Chen J. and Shi M. (2007) analyzed pressure drop in a reverse-flow cyclone with a tangential inlet by dividing the total pressure loss into the four components, namely (1) expansion loss at the cyclone inlet, (2) contraction loss at the entrance of the outlet tube, (3) swirling loss due to gas flow and cyclone walls, and (4) dissipation loss of gas dynamic energy in the outlet tube. They reported that component (2) has limited effect on the total pressure loss and can be omitted. Based on this result, they proposed the Chen-Shi (C-S) model to predict pressure drop in cyclones operating with (i.e., an inlet solid concentration \( C_s \) of \( 0.002 \) to \( 2 \) kg-solid (m\(^3\)-gas)\(^{-1}\)) or without dust gas at ambient and high temperatures. Li S. et al. (2011) investigated cyclone pressure drop at a very high \( C_s \) (\( 0.125–4.42 \) kg-solid (m\(^3\)-gas)\(^{-1}\)) and inlet gas velocities \( (v_i) \) of \( 13.2, 17.5, \) and \( 21 \) m s\(^{-1}\). They improved the C-S model to predict cyclone pressure drop more accurately and reported that component (4) in the C-S model should not be included; further, they suggested that the swirl exponent may be zero for high solids-loading conditions.

The improved C-S model can be described as follows to estimate pressure loss in a cyclone (Li S. et al., 2011, Chen J. and Shi M., 2007).

\[
\Delta p_x = \frac{\rho_g}{2} \left( \frac{v_i^2 + v_{ze}^2}{2} \right) = \left[ \frac{\bar{v}_e^2}{(\bar{v}_e D_e)^2} + \frac{1}{K_A \left( D_e^2 - \bar{v}_e^2 \right)^2} \right] \frac{\rho_g v_i^2}{2}
\]  \hspace{1cm} (4)

\[
\Delta p_l = \left( 1 + \frac{C_s}{\rho_g} \right) \left( \frac{1}{1 + 2(1 - \bar{v}_e) b - D_e} \right) \frac{\rho_g v_i^2}{2}
\]  \hspace{1cm} (5)

\[
\Delta p_b = \frac{4 \varepsilon_A \bar{v}_b (v_i w_i C_s)^{1.5}}{2 \times 0.9 Q} \times 1.11/K_A \bar{v}_b D_e^{-1.5} \frac{\rho_g v_i^2}{2}
\]  \hspace{1cm} (6)

\[
f = f_{uu}(1 + 3\sqrt{C_s/\rho_g})
\]  \hspace{1cm} (7)

\[
K_A = \frac{\pi D_e^2}{4ab}
\]  \hspace{1cm} (8)

\[
\lambda_s = \frac{4 \varepsilon_A}{\pi D_e^2} = (1 - \bar{D}_e^2) + 4\bar{D}_e \bar{f} + 4\bar{H}_b + \left( 1 + \bar{B} \right) \frac{4(\bar{H} - \bar{H}_b)^2}{(1 - \bar{B}^2) + 4\bar{D}_e \bar{H}_h}
\]  \hspace{1cm} (9)

\[
v_{bw} = \frac{1.11K_A^{-0.21}\bar{D}_e^{0.16} Re^{0.06}}{1 + f_{uu} \lambda_s \sqrt{K_A \bar{D}_e}}
\]  \hspace{1cm} (10)

\[
Re = \frac{v_i \rho_g D_e}{\mu}
\]  \hspace{1cm} (11)

\[
\bar{v}_b = \frac{\bar{v}_{bw}}{1 + 0.35(C_s/\rho_g)^{0.27}}
\]  \hspace{1cm} (12)

\[
n = 0
\]  \hspace{1cm} (13)

\[
\bar{r}_e = 0.38 \bar{D}_e + 0.5 \bar{D}_e^2
\]  \hspace{1cm} (14)

Fig. 1 Comparison between model-predicted and experimental data \( (v_i = 17.5 \) m s\(^{-1}\)). Reprinted with permission from Ref. (Li S. et al., 2011). Copyright: (2011) WILEY-VCH Verlag GmbH & Co. KGaA, Weinheim.
In the improved C-S model, geometric and velocity variables were normalized with respect to cyclone diameter. As shown in Fig. 1, the performance predicted by the improved C-S model agrees well with experimental data when \( C_s = 1–5 \) kg-solid (m\(^3\)-gas\(^{-1}\)), thus effectively amending the overprediction of the original C-S model.

### 2.2 Multi-cyclone and non-uniform distribution of particles

In large-scale industrial reactors with high gas–solid flow rates, small parallel cyclones are often preferred to achieve a high separation efficiency when the distribution of gas–solid flow at each cyclone inlet is uniform. However, it is difficult to place a refractory in such small cyclones (Nowak W. and Mirek P., 2013) and there is strong evidence that gas–solid flow in parallel cyclones is non-uniform, which reduces the total separation efficiency (Zhang C. et al., 2016).

Masnadi M.S. et al. (2010) examined gas–solid flow distribution across two parallel and identical cyclones based on an analytical model and compared flow distribution through parallel pipes. They confirmed the consistency of their analytical model by comparing the experimental data of two identical cyclones (barrel diameter \( D = 101.6 \) mm). They reported that a non-uniform (or maldistribution) gas–solid flow is unavoidable for a high solids loading (upstream solids hold up > 0.01 %) and that fouling can significantly affect maldistribution of gas–solid flow through the identical cyclones. Zhang C. et al. (2010) calculated a three-dimensional full-loop CFB boiler model with two parallel cyclones using an Eulerian granular multiphase model. In their study, they found minor differences in the average solids mass flux (\( G_s \)) in the two cyclones (5.74 and 6.05 kg m\(^{-2}\) s\(^{-1}\)) and pointed out that the maximum \( G_s \) alternates between these two cyclones. Zhou X. et al. (2012) investigated gas–solid flow through six parallel cyclones located asymmetrically on the left and right walls of the riser (i.e., three cyclones on the left and three cyclones on the right) in a CFB cost test apparatus. They observed that the distribution of gas–solid flow was non-uniform across three cyclones on one side and that the middle cyclone on each side exhibited higher particle velocities while their \( G_s \) was lower than that of other cyclones. Jiang Y. et al. (2014) conducted numerical calculations on gas–solid flow hydrodynamics at a CFB boiler test facility with six parallel cyclones using an Eulerian-Lagrangian model and computational particle fluid dynamics (CPFD). They validated simulation data using experimental data obtained by electrical capacitance tomography (ECT) in cold model tests (at ambient temperature and atmospheric pressure). The geometry of the six cyclones was either axis-symmetric or point-symmetric, as shown in Fig. 2.

They also observed that the solids concentrations of the four cyclones located at the corners of the chamber were greater than those of others and stated that an “axis-symmetric” arrangement (case A) for cyclones is better than a “point-symmetric” arrangement (case B) from the viewpoint of uniform distribution of solids (Fig. 3).

Wang S. et al. (2017) investigated the hydrodynamics of gases and solids in six parallel cyclones with central symmetry and axial symmetry arrangements combined with a full-loop CFB by CFD and a discrete elemental method (CFD-DEM, Fig. 4). They reported that for a uniform distribution of solid mass flux in parallel cyclones, axial symmetry is better than central symmetry. The middle
cyclones on both sides have higher solid velocity and solid holdup than other corner cyclones (Fig. 5).

Shuai D. et al. (2017) also investigated the flow behavior of gas and solids in six cyclones in parallel in an annular furnace. They observed a non-uniform distribution of $G_s$ and cyclone pressure drop ($\Delta p_c$) in the six cyclones. However, no regularity could be observed. The relative deviation of $G_s$ in the six cyclones was 8.0 % under typical operating conditions.

2.3 Design principles of multi-cyclones

The design principles of small multi-cyclones and a large single cyclone were analyzed. Mo X. et al. (2015) investigated the influence of wall friction and solid acceleration on the non-uniform distribution of gas–solid flow in two parallel cyclones. They reported that pressure drop in cyclones has an inflection point with respect to the mass flow rate ratio between solids and the gas ($C_{T,inf}$, Eq. (1)), which is around 0.5–3.33 kg-solid (kg-gas)$^{-1}$, as calculated from the reported values. They also suggested that the inflection point increases with an increase in the wall friction but decreases with an increase in solid acceleration. The inflection point has a large effect on non-uniform gas–solid distribution in the two parallel cyclones. As summarized in Table 1, the solid flow distribution is non-uniform when $C_T$ approaches $C_{T,inf}$.

Zhang C. et al. (2016) analyzed multi-phase interactions and investigated instabilities in uniformity in two parallel cyclones installed after a fluidized bed reactor. They provided a novel design principle to avoid non-uniform distribution of solids using $C_T$ and dimensionless vortex finder diameter ($d_r$) defined in Eq. (15). Fig. 6 shows a phase diagram of uniformity of the two parallel cyclones as a function of fraction of solid flow to cyclone 1 (i.e. path 1) and $C_T$. They reported that under the condition that inlet velocity of air is 15 m s$^{-1}$, Reynolds number is $3.08 \times 10^5$ and glass beads are used at 20 °C and 101.3 kPa, a low solids loading ($C_T < 1.35$), the uniform distribution is stable but at $C_T = 1.35$, there occurs a turning point from uniform to non-uniform distribution. When $C_T > 1.35$, the maldistribution of solids is stable.

![Fig. 4 Schematic of 6 cyclones in CFD calculation](a) Front view of simulation geometry (b) Central symmetry (c) Axial symmetry arrangements (units: mm). Reprinted with permission from ref. (Wang S. et al., 2017). Copyright (2017) Elsevier B.V.

![Fig. 5 Time-averaged solid flux distribution in cyclones](Reprinted with permission from ref. (Wang S. et al. 2017). Copyright (2017) Elsevier B.V.)

| Table 1 Summary of the three phases in two parallel cyclones. Reprinted with permission from ref. (Mo. X. et al. 2015). Copyright (2015) Elsevier B.V. |
|--------------------------------------------------|-----------------|-----------------|-----------------|
| $C_{T,inf}$ | $C_T << C_{T,inf}$ | $C_T \sim C_{T,inf}$ | $C_T > C_{T,inf}$ |
| Main contributor | Wall friction | Both greatly influence the pressure drop | Solid acceleration |
| Gas fraction in cyclone 1 ($\psi$) and solids fraction in cyclone 1 ($\gamma$) | $\psi > 0.5$ and $\gamma > 0.5$ or $\psi < 0.5$ and $\gamma < 0.5$ | $\psi > 0.5$ and highly uneven solid distribution at certain flow perturbation | $\psi < 0.5$ and $\gamma > 0.5$ |
They found that $d_r$ should be less than 0.32 and that $d_r$ strengthens the swirl in a cyclone for enhancing the stability of uniformity and cyclone efficiency under this condition.

Fig. 7 shows a phase diagram of stability based on the data in literature (Zhang C. et al., 2016). For parallel cyclones that treat high solids loading, $d_r$ should be small enough to ensure the stable operation. These criteria are very useful in the selection of single large cyclones or multi-cyclones for uniform and steady operation.

$$d_r = \frac{vortex\ finder\ diameter\ [m]}{cyclone\ barrel\ diameter\ [m]}$$

2.4 Cyclones in series

Wu X. et al. (2011) experimentally measured and calculated the pressure drop of two scroll-type cyclones connected in series for a wide range of solids loading ($C_T = 0–30\ kg\-solids\ (kg\-gas)^{-1}$) and cyclone inlet velocity ($v_i = 16–24\ m\ s^{-1}$). The experimental data showed that the cyclone pressure drop decreased dramatically as $C_T$ increased to 7.5 kg-solids (kg-gas)$^{-1}$ after which it remained almost constant.

Fushimi C. and Guan G. et al. developed a triple bed combined CFB for coal gasifiers (Fushimi C. et al., 2011, 2014; Guan G. et al., 2010, 2011) and experimentally investigated the hydrodynamics of silica sand (average particle diameter of 126 μm) at a high flux ($G_s$ in the range of 200–546 kg m$^{-2}$ s$^{-1}$) in cold model tests. In these experiments, three cyclones were connected in series to separate solids from air. Separation efficiency at the outlet of the third cyclone was 99.998% when $G_s$ was...
490–510 kg m\(^{-2}\) s\(^{-1}\). Meanwhile, Hoffmann A.C. and Stein L.E. (2010) recommended that the underflow pipes of series-connected cyclones should be independently sealed.

### 2.5 Combination of an inertial separator and a cyclone for a high-flux CFB

Wang X. et al. (2015) developed a high-flux CFB reactor, comprising of a fuel reactor, an air reactor, a J-valve, a downcomer, and solid separation systems, for in situ gasification chemical looping combustion (iG-CLC). They optimized the iG-CLC system by developing an inertial separator for primary gas-solids separation and a cyclone for secondary gas-char and remaining solids separation to improve operation stability and solid separation efficiency. They reported that the selective separation efficiency of coal particles in the first-stage inertial separator was 77.7\% when the mass flow rate of the solid was 1021 kg h\(^{-1}\) and the overall separation efficiency after the second-stage cyclone reached 99.5\% when the mass flow rate to cyclone was 35 kg h\(^{-1}\) in their cold model tests.

In their subsequent work (Shao Y. et al., 2017), they carried out three-dimensional full loop CFD calculation using an Eulerian-Eulerian two-fluid model combined with the standard \(k\)-\(\varepsilon\) turbulence model for the gas phase and the kinetic theory of granular flow for the solid phase to investigate the hydrodynamics of coal and iron ore particles on the basis of the experimental results of cold model tests. They reported continuous and stable solid circulation in the iG-CLC model when \(G_s\) was around 400 kg m\(^{-2}\) s\(^{-1}\).

### 2.6 Design principles of the cyclone inlet and body for high solids-loading multi-cyclones

Hoffmann A.C. and Stein L.E. (2010) recommended that 1) bends should be located at a distance of ten pipe diameters or its equivalent before cyclone inlets to avoid non-uniform distribution of gases and solids in the cyclone inlet piping and 2) a scroll or wrap-around type of inlet should be used. At high solids or liquid loadings (>\(\sim\)10 vol%), care must be taken not to restrict the discharge opening or the annular space around a vortex stabilizer.

### 2.7 Flow behavior of gas and solids in a dipleg below cyclone

Wei Z.G. et al. (2016) examined the pressure of a catalyst powder (Sauter mean particle diameter of 63.6 \(\mu\)m) flow in a dipleg (150 mm inner diameter and 9 m height) when \(G_s\) was in the range of 50.0 to 385.0 kg m\(^{-2}\) s\(^{-1}\).
Fig. 9 shows the experimental apparatus consisting of a riser of a fluidized bed (600 mm i.d. and 8 m height), a riser (200 mm i.d. and 12.5 m height), and a dipleg (150 mm i.d. and 9 m height) connected with a cyclone (400 mm i.d.). The bottom of the dipleg was immersed into the fluidized bed. Fig. 10 shows the observed fluidization pattern in the dipleg. At small vortex (cf. Fig. 3a, b), a dilute-dense coexisting falling flow. The swirl flow (just below the cyclone), the dilute particle falling flow (in the middle) and the dense falling flow. The same researcher summarized the scaling relationship of cyclones by setting up the following model (Eqs. (16)–(21)). In this model, the total separation efficiency of a cyclone \( \eta_{\text{tot}} \) is the sum of the efficiency at the wall \( \eta_{\text{wall}} \) and in the inner vortex \( \eta_{\text{vtx}} \) (Eq. (16)). A portion of the incoming solids not collected by the wall is collected by the inner vortex, transformed. Further increase in particle concentration reached a maximum in the fluidization pattern being no interface between the dilute particle falling flow and the dense flow. When the solids loading is higher than the limit loading (\( \mu_e > \mu_{\text{lim}} \)), dense particles cannot flow through the rotating/swirling stream because of their high inertia; in this case, these particles hit the outside wall of the cyclone and descend to the bottom.

Mirek P. also described the particle size distribution of the carryover, \( R_i(d) \), and it can be determined using Eq. (19).

\[
R_i(d) = \frac{\mu_{\text{lim}}}{\mu_e} \left( 1 - \eta_e(d) \right) \frac{\Delta R_e(d)}{\Delta R_i(d)}
\]  

The total separation efficiency is determined by Eq. (20).

\[
\eta_e(d) = 1 - \left( 1 - \eta_{\text{tot}} \right) \frac{\Delta R_e(d)}{\Delta R_i(d)}
\]

The same researcher summarized the scaling relationship using five dimensionless parameters sets shown in Table 2. They compared estimation curves based on the

Table 2  Scaling relationships and parameters used for the analysis of the cyclone performance in cold model tests (cf. Fig. 11). Reprinted with permission from ref. (Mirek P., 2018). Copyright (2018) Elsevier B.V.

| Scaling relationships | Parameters chosen independently | Notation |
|-----------------------|-------------------------------|----------|
| \( \frac{U_0}{u_T} \cdot \frac{G_s}{\rho_s U_0} \cdot \text{PSD} \) | \( d_{50}, \rho_p, \rho_g, D_e, \mu_e, g \) | Set (1) |
| \( \frac{R_{p_0}}{u_T} \cdot \frac{U_0}{\text{Ar, PSD}} \) | \( \rho_p, \rho_g, D_e, \mu_e, g \) | Set (2) |
| \( \frac{U_0}{u_T} \cdot \frac{G_s}{\rho_s U_0} \cdot \text{Fr, PSD} \) | \( \rho_p, \rho_g, D_e, \mu_e, g \) | Set (3) |
| \( \frac{U_0}{u_T} \cdot \frac{G_s}{\rho_s U_0} \cdot \text{Fr, PSD} \) | \( \rho_g, D_e, \mu_e, g \) | Set (4) |
| \( \frac{F_r, R_{p_0}}{\rho_g} \cdot \frac{G_s}{\rho_s U_0} \cdot \frac{d}{D_r} \cdot \text{PSD} \) | \( \rho_g, \mu_e, g \) | Set (5) |
Fig. 11 Influence of a set of scaling relationships on the total grade efficiency curve based on the parameter sets shown in Table 2 and the Stokes equation. Reprinted with permission from ref. (Mirek P., 2018). Copyright (2018) Elsevier B.V.

five dimensionless parameter sets and the Stokes equation (Eq. (21)).

\[ \text{Stk}_{\text{in}, i} = \frac{(\rho_p - \rho_g) d_{50} v_i}{18 \mu_g D_b} \]  

(21)

Fig. 11 shows the result. All the proposed scaling relationships resulted in a very high separation efficiency (> 99.4%). However, the results strongly depended on the selected scaling relationships set (Mirek P., 2018).

Wang J. et al. (2019) recently conducted experiments extensively and introduced sophisticated models to improve the prediction of grade efficiency of cyclone separators. They constructed generalized linear mixed-effects (GLME) models that are functions of the parameters \((\text{St}, d_{50}, \sigma_\xi, d_{50}, \text{Re}, Fr, H)\) to understand and control grade efficiency \((\eta_F(d))\) variation. Note \(\sigma_\xi\) is the size deviation in particle diameter at the inlet.

In their GLME models, the observed grade efficiency vector \(\eta\) is considered to be a summation of the expected grade efficiency vector \(\mu\) and the noise vector \(\epsilon\), which is assumed to follow a normal distribution with mean 0 (Eq. (22)).

\[ \eta = \mu + \epsilon \]  

(22)

The expected grade efficiency \(\mu\) is represented by the Gaussian regression model shown in Eq. (23),

\[ g(\mu) = \beta_0 + (\beta_1 + \phi_1) \ln(\text{St}) + (\beta_2 + \phi_2) \ln\left(\frac{d}{d_{50}}\right) + \beta_3 \ln(\sigma_\xi) + \beta_4 \ln\left(\frac{d_{50}}{D_b}\right) + (\beta_5 + \phi_5) \ln(\text{Re}) + \beta_6 \ln(Fr) \]  

(23)

where \(\phi_i\) is a random effect corresponding to the \(i^{th}\) component and has a normal distribution with a variance \(\sigma_i^2\). The \(g(\mu)\) and the expected grade efficiency \(\mu\) are linked in the following equations ((24), (25)).

\[ g(\mu) = -\ln(-\ln(\mu)) \]  

(24)

\[ \mu = \exp[-\exp(-g(\mu))] \]  

(25)

Table 3 lists the values estimated for the parameters \((\beta_0, \beta_1, \sigma_1, \sigma_2, \sigma_6)\) in Eq. (23), including a 95% confidence interval for random effects. Fig. 12 shows the parity plot of experimental results and fitted values predicted by using the GLME model. It can be seen that the GLME modeling predicts the experimental grade efficiency accurately. It is considered that the GLME model can provide setting conditions for cyclone separators (Wang J. et al., 2019).

| Parameter | Estimate | Std. Error | P-value |
|-----------|----------|------------|---------|
| \(\beta_0\) | 64.952   | 4.7019     | 1.7896e–38 |
| \(\beta_1\) | –6.8608  | 0.64062    | 7.4357e–25 |
| \(\beta_2\) | 14.721   | 1.2805     | 4.303e–28  |
| \(\beta_3\) | 23.673   | 3.2352     | 7.0839e–13 |
| \(\beta_4\) | 14.794   | 0.98611    | 3.2965e–44 |
| \(\beta_5\) | 1.6486   | 0.40073    | 4.3607e–05 |
| \(\beta_6\) | –2.2869  | 0.36419    | 6.0287e–10 |
| \(\sigma_1\) | 0.0011382| 0.00035134 | 0.0036874  |
| \(\sigma_2\) | 0.16095  | 0.13411    | 0.19315    |
| \(\sigma_6\) | 0.038589 | 0.028213   | 0.052781   |

Table 3 GLME model estimation parameters. Reprinted with permission from ref. (Wang J. et al., 2019). Copyright (2019) Elsevier B.V.

Fig. 12 The parity plot of fitted values based on the GLME model and experimental results of the grade efficiency. Reprinted with permission from ref. (Wang J. et al., 2019). Copyright (2019) Elsevier B.V.
Table 4: Summary of recent research on gas-solid cyclones.

| Method | References | Barrel diameter, $D$ [m] | Inlet gas velocity, $v_i$ [m s$^{-1}$] | Particle size, $d_p$ [μm] | Solid loading, $C_s$ [kg-solid (kg-gas)$^{-1}$] | Solid concentration, $C_s$ [kg-solid (m$^3$-gas)$^{-1}$] | Particle recovery ratio [%] | Inlet type | CFD method Above: Fluid Below: Solid |
|--------|-------------|--------------------------|-------------------------------|-----------------------------|---------------------------------------------|---------------------------------------------|-----------------------------|-----------|----------------------------------|
| exp.   | Mirek P. (2018) | 0.6 | 1.87, 7.87 | Tangential | — |
| exp.   | Demir S. (2014) | 0.29 | 10–18 | Tangential | — |
| exp.   | Lee H. and Yook S.J. (2014) | 0.133 (Single) 4.4–12.5 3.1–8.9 | 0.094 (Dual) 100–500 | 95 | Tangential | — |
| exp.   | Haig C.W. et al. (2014) | 0.04 | 15–45 0.2–5.5 | 0.00009–0.0003 | Tangential | — |
| exp.   | Oh J. et al. (2014) | 0.15 | 5.2–13.0 | (<1.5 vol%) 99.5 | Tangential | — |
| exp.   | Xiong Z. et al. (2014) | 0.168 | 19–39 9.26 | 0.0001–0.0005 0.0002 | 99 | Axial | — |
| exp.   | Avcı A. et al. (2013) | 0.04 | 2.4–37.2 | 100–500 | Tangential | — |
| exp.   | Yoshida H. et al. (2010) | 0.072 | 2.7–19.1 1.8 | 86.02–95.72 | Tangential | — |
| exp.   | Ahuja S.M. (2010) | 0.3 | 6.63–15.11 35 (Fly ash) 7 (Talc) | 86.02–95.72 | Tangential | — |
| exp. & sim. | Wang W. et al. (2010) | 0.148–0.428 | 12–16 10.2 | 0.0015 82–96 | Tangential | — |
| exp. & sim. | Zhang P. et al. (2019) | 0.24 | 17–36.5 2.4 | 0.002–0.010 97 | Tangential | — |
| exp. & sim. | Karagoz I. et al. (2013) | 0.19 | 10–25 14 | 80–90 | Tangential | — |
| exp. & sim. | Hsiao T.C. et al. (2015) | 0.025 | 2–6 | Tangential | — |
| exp. & sim. | Mazyan W.I. et al. (2017) | 0.19 | 14 4–150 0.008 | 90–95 | Tangential | — |
| exp. & sim. | Wu X. et al. (2011) | 0.15 | 16–24 55, 128 0–30 | Tangential | — |
| exp. & sim. | Li S. et al. (2011) | 0.282 | 13.2–21 263 | 0.125–4.42 | — |
| exp. & sim. | Wei Z.G. et al. (2016) | 63.6 | (G$:50.0–385$ kg m$^{-2}$ s$^{-1}$) | — |
| exp. & sim. | Prabhansu et al. (2017) | 0.100 | 7–30 100–1000 | 86–98 | Tangential | — |
| exp. & sim. | Song J. et al. (2017) | 0.5 | 16–29 2–120 | 0.04 > 90 | Tangential | Euler Lagrange |
| exp. & sim. | Huang A.N. et al. (2018b) | 0.0723 | 15 2.1 | 80.7 |
| exp. & sim. | Huang A.N. et al. (2017a) | 0.072 | 11–21 2.1 | 0.00149–0.00078 0.00179–0.000936 | Tangential | — |
| exp. & sim. | Huang A.N. et al. (2017b) | 0.547 | 13.6–18.8 16.9 | 0.00025–0.00059 | Tangential with slits | — |
| exp. & sim. | Gao Z. et al. (2019) | 0.14 | 12–18 | RSM— |
| exp. & sim. | Wei J. et al. (2017) | 0.3 | 4–12 2000 0.72–8.64 | Tangential | — |
| exp. & sim. | Yu X. et al. (2018) | 1.33–4.64 300, 600 | Tangential | — |
| Method | References | Barrel diameter, \( D \) [m] | Inlet gas velocity, \( v_i \) [m s\(^{-1}\)] | Particle size, \( d_p \) [μm] | Solid loading, \( C_T \) [kg-solid (kg-gas)\(^{-1}\)] | Solid concentration, \( C_s \) [kg-solid (m\(^3\)-gas)\(^{-1}\)] | Particle recovery ratio [%] | Inlet type | CFD method |
|--------|------------|-----------------|----------------|----------------|----------------|----------------|----------------|------|----------------|
| Sim.   | Su Y. et al. (2011) | \( 0.200 \times 0.200 \) (square) | 17.98, 20.21 | | | \( 0.0084, 0.0088 \) | | | |
|        | Surmen A. et al. (2011) | 2–20 (L/D) | | | | | | | |
|        | Qiu Y. et al. (2012) | 0.33 | 8–22 | 12.8 | \( < 1.5 \text{vol}\% \) | 99.5 | | |
|        | Winfield D. et al. (2013) | 0.0066 | 16.70, 16.74 | | | | > 92 | |
|        | Shi B. et al. (2013) | 0.04 | 30 | | | | 99 | |
|        | Bogodage S.G. and Leung A.Y.T. (2015) | 0.3302 | 10 | 0–18 | | \( 0.001 \) | | |
|        | Hsiao T.C. et al. (2015) | 0.025 | \( 0.0025–0.01 \) | 2–6 | | | | |
|        | Misiulia D. et al. (2015) | 4.21–5.07 (H/D) | 3.5 | 0.2–30 | | \( 0.003 \) | | |
|        | De Souza F.J. et al. (2015a,b) | 0.031 | 10.67 | 0.5–6.0 | | | | |
|        | Yang J. et al. (2015) | 0.3, 1.2, 0.9, 0.6, 0.186 | | | | | | |
|        | Xu W. et al. (2016) | 0.25 | 23 | 10 | | \( 88 \) | | |
|        | Schneiderbauer S. et al. (2016) | 2.5 | 13.4 | 0.6–400 | \( 0.224 \) | 93 | | Lagrange-Euler hybrid |
|        | Kozolub P. et al. (2017) | 0.2 | 5.55–9.49 | 40–80 | \( 1.0–2.7 \) | | | |
|        | Mazyan W.I. et al. (2017) | 0.19 | 14 | | | | | |
|        | Siadaty M. et al. (2017) | 0.3 | 16.04 | 2–6 | \( (3 \text{vol}\%) \) | > 91.11 | | |
|        | Sun X. et al. (2017) | 0.29 | 16.34 | 1–5 | | | | |
|        | Wang S. et al. (2017) | 0.07 | 1600 | | \( < 0.03 \text{vol}\% \) | | | |
|        | Wasilewski M. and Brar L.S. (2017) | 3.35 | 8.84 | | | | 80.4–82.4 | |
|        | Huang A.N. et al. (2018a) | 0.0723 | 15, 18 | 0.339–11.565 | \( 0.0016-0.153 \) | > 99 % | | |
|        | Safikhani H. and Mehrabian P. (2016) | 1, 1.7, 2.3 (L/D) | 17 | 20, 21, 23, 19, 13 | | | | Axial |
|        | Safikhani H. et al. (2018) | 1.7 (L/D) | 8.75, 17.5, 35 | 20, 21, 23, 19, 13 | | | 60–62 | Axial |
|        | Zhang N. et al. (2010) | 8.08 | 200 | | \( (G_c: 4.0–8.5 \text{kg m}^{-2}\text{s}^{-1}) \) | | | Euler |
|        | Chu K.W. et al. (2011) | 0.2 | 20 | 2000 | \( 0–2.5 \) | | | |
|        | Chu K.W. et al. (2017) | 1,000 | 3.9 | 200–600000 | 5–11 Medium to Coal | | | Tangential |
|        | Jang K. et al. (2018) | 10–30 | 0.4–10 | | | 57–75 | | Tangential |
Table 4 summarizes the experimental and numerical studies on gas–solid cyclones conducted in the last decade. Recently, a number of numerical studies using CFD simulations have been conducted to investigate the velocities and volume fractions of gases and solids and pressure drop in cyclones by varying their configurations. Most of these experimental and numerical studies were conducted under dilute conditions ($C_s$ or $C_T$ < 0.001). Huang A.N. et al. (2018a) investigated the effect of particle mass loading on the hydrodynamics of solids and separation efficiency using the Eulerian-Lagrangian approach with a two-way coupling method for CFD and compared with the experimental results of laboratory-scale cyclones at $C_s$ of 0.0016–0.153 kg-solid (m$^3$-gas)$^{-1}$ and $v_i = 15$ and 18 m s$^{-1}$. They reported that the overall separation efficiency increased from 74.5% for a conventional cyclone to 80.7% for a cyclone with a lower cone slit. More detailed experimental and numerical studies under high solids loading conditions on gas–solid cyclones are expected in the future.

5. Conclusions and future prospects

In this review, recent progress in high solids-loading gas–solid cyclones has been summarized. The improved C-S model proposed by Li S. et al. (2011) is highly useful for predicting pressure drop in cyclones. For commercial-scale CFBs, 2 to 6 multi-cyclones are often used in parallel. However, the non-uniform distribution of solids in the parallel section is a problem. Mo et al. investigated the effect of wall friction and solid acceleration on the non-uniform distribution of gas–solids flow and found that the inflection point has a significant influence on the maldistribution of gas–solids in two cyclones in parallel. Zhang C. et al. (2016) found that the phase graph of the stability of uniform distribution/maldistribution of solids in two parallel cyclones as functions of $C_T$ (kg-solid (kg-gas)$^{-1}$) and $d_c$.

To improve the scale-up methodology, Mirek P. (2018) summarized and modified the Muschelknautz model by applying five sets of scaling relationships and found that all the employed sets of scaling relationships resulted in a very high separation efficiency (> 99.7%). The GLME model developed by Wang J. et al. (2019) can predict grade efficiency with a large prediction variability.

It is expected that CFB boilers will be used for biomass co-combustion to reduce CO$_2$ emissions due to power generation. In waste-treatment plants, CFB incinerators may be used in the future because of their high thermal efficiency and operability. In CFB boilers used for power generation, non-steady-state operations (rapid changes in load) are required. In these operations, a high solids separation efficiency and control over the cut-off diameter are required. We expect for future research, flow behaviours of gas and solids, and separation efficiency of solids in cyclones under non-steady state or transient operation conditions should be investigated.

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Nomenclature

$A_c$: Total area of the contact surfaces between gas flow and the cyclone wall, $A_c = \frac{\pi D_c^2}{4}$ [m$^2$]

$A_s$: Sedimentation area of the cyclone wall [m$^2$]

$a$: Cyclone inlet height [m]

$b$: Cyclone inlet width, $b = \frac{2}{3} D_b$ [m]

$B$: Diameter of the cyclone dust exit, $B = \frac{2}{3} D_b$ [m]

$c$: Width of the inlet cutting into the cyclone body, $c = \frac{2}{3} b$ [m]

$C_i$: Inlet solid concentration [kg-solid (kg-gas)$^{-1}$]

$C_T$: Mass flow rate ratio between solids and gas [kg-solid (kg-gas)$^{-1}$]

$C_{T,inf}$: Solids loading at the inflection point [kg-solid (kg-gas)$^{-1}$]

$d$: Feed particle diameter [m]

$d_c$: Cut size for separation at wall [m]

$d_{50,i}$: Dimensionless vortex finder diameter [–]

$d_{50,s}$: Sauter mean particle diameter [m]

$d_{50,A}$: Median particle diameter in feed [m]

$D_c$: Cyclone barrel diameter [m]

$D_s$: Cyclone hopper diameter, $D_s = \frac{4}{3} D_h$ [m]

$D_c$: Diameter of core flow, $D_c = 2r_c$ [m]

$D_e$: Cyclone outlet diameter, $D_e = \frac{4}{3} D_b$ [m]

$D_r$: Riser hydraulic mean diameter, $4A$ (cross-sectional area of the flow)/P (wetted perimeter of the cross-section) [m]

$e$: Noise vector [–]

$f$: Friction coefficient [–]

$f_w$: Friction coefficient for dust-free gas flow [–]

$F_r$: Froude number based on $D_s$, $F_r = \frac{v_t}{\sqrt{gH}}$ [–]

$F_{r,s}$: Froude number based on $H$, $F_{r,s} = \frac{v_t}{\sqrt{gH}}$ [–]

$g$: Gravitational acceleration [m s$^{-2}$]

$G_c$: Solids mass flux [kg m$^{-2}$ s$^{-1}$]

$H_s$: Cyclone barrel height, $H_s = \frac{4}{3} D_h$ [m]

$H$: Cyclone body height, $H = \frac{4}{3} D_b$ [m]
Greek alphabets

\( \mu \): Expected grade-efficiency vector [-]
\( \mu_i \): Initial load [kg kg\(^{-1}\)]
\( \mu_G \): Gas viscosity [Pa s]
\( \mu_{lim} \): Limited load ratio [kg kg\(^{-1}\)]
\( \rho_G \): Gas density [kg m\(^{-3}\)]
\( \rho_p \): Particle density [kg m\(^{-3}\)]
\( \eta \): Observed grade-efficiency vector [-]
\( \eta_i(d) \): Grade efficiency [-]
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