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Use of pressure to mitigate gas bypassing in fluidized beds of FCC catalyst particles

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ABSTRACT

Deep gas fluidized beds of low-fines fluid catalytic cracking (FCC) catalyst particles can have severe gas maldistribution due to gas bypassing. Tests were conducted in a 0.6-m-diameter unit using 3.2% and 4% fines less than 44 μm FCC catalyst particles to determine the influence of system pressure on gas bypassing in a fluidized bed of 3.66 m static bed height. The freeboard pressure was varied up to 207 kPag (30 psig). Differential pressure fluctuations were measured at four locations around the column, bubble void fraction was measured at two opposite locations close to the column wall, and radial bubble void fraction profiles were measured at axial elevations of 0.9 and 1.52 m. At no or low pressures, gas bypassing was present in the bed. With gas bypassing, differential pressure fluctuation intensities were significantly different around the column, significantly higher bubble void fractions were measured close to the inner wall on one side of the column than on the opposite side, and the radial bubble void fraction profiles were not symmetrical about the column axis. Increasing the system pressure weakened the intensity of gas bypassing. Gas bypassing disappeared at a freeboard pressure of about 100 to 140 kPag (15 to 20 psig).

1. Introduction

Bubbling fluidized beds have wide industrial applications because of their intimate gas–solids contacting. The traditional image of a bubbling fluidized bed is one in which gas voids (bubbles) rise through the bed, transporting solids in their wakes and accompanying drifts while simultaneously solids move downwards in bubble-free regions which leads to a vigorous mixing of the gas and solids. This rapid mixing also results in a nearly homogeneous temperature in the bed. This image of a well fluidized bed is, however, not always true. Fluidization tests with FCC catalyst particles by Knowlton [1] in a 0.3-m-diameter column and a static bed height of 1.83 m and by Wells [2] in a 2.44-m-diameter semicircular column and a static bed height of 4.88 m, with both units being transparent, showed significant bypassing of the fluidizing gas occurring in the beds at superficial gas velocities that are typical of the bubbling bed fluidization regime. Knowlton [1] observed that the fluidizing gas preferentially flowed through one side of the bed, whereas Wells [2] found that the streaming gas moved about the center near the faceplate, occasionally splitting and passing up the sides of the unit. In both studies gas–solids contacting was extremely poor. The remainder of the bed was a mass of nearly stagnant solids being either defluidized or poorly fluidized. Knowlton [1] observed that gas bypassing could be eliminated by increasing the fines content. Wells [2] found that neither fines content nor the type of gas distributor used had any effect on gas bypassing. The gas bypassing phenomenon is also referred to as streaming or jet streaming.

Apparently, the gas bypassing phenomenon is rarely reported in the literature. This is probably because most laboratory beds are generally not deep enough to cause jet streaming or because the steel construction of large commercial beds makes visual observation of gas bypassing impossible. It is also possible that the combinations of the type of solids particles, operating pressure and temperature, gas velocity and/or the presence of internals have prevented gas bypassing. Gas bypassing is detrimental if it occurs in industrial beds, such as FCC regenerators and strippers or in circulating fluidized bed (CFB) combustors. It can result in poor gas/solids contacting, afterburning in the freeboard, poorly fluidized entrances to standpipes and discharge regions of cyclone diplegs, and poor solids flows around CFB loop seals. Gas bypassing can also compromise the scale-up process.

Gas bypassing was explained (Knowlton [2], Karri et al. [3]) as being due to gas compression caused by the static head of fluidized solids in deep beds whereby the pressure at the bottom of a bed can be significantly greater than the pressure at the top. The compression of the gas reduces the volume of interstitial gas in the emulsion phase which causes the density of the emulsion phase to increase enough that it approaches the minimum fluidization density causing a partial defluidization of a vertical section of the bed. As a result, the solids become less permeable to gas flow. If the permeability is too low, it becomes easier for the gas to bypass the bed via a single or several bubble streams rather than fluidizing the bed uniformly. This phenomenon is regarded (Karri et al. [3]) to be similar to what happens in standpipes

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conveying Group A solids where aeration is added at intervals along the standpipe to compensate for the volume change caused by gas compression in order to maintain the standpipes in a fluidized state.

Studies (Knowlton [1], Karri et al. [3], Issangya et al. [4–6]) have shown that gas bypassing can be mitigated by lowering the bed height, increasing the fines content (material smaller than 44 μm), installing properly spaced horizontal baffles in the whole column and increasing the superficial gas velocity. Increasing the gas velocity was more effective in beds with high fines than beds of low fines materials. Different types of gas distributors, including sintered metal plates, perforated plates, bubble caps, pipe manifolds and ring spargers were used in those studies and no significant influence of the type of gas distributor on gas bypassing was observed. Bolthrunis et al. [7] observed uncharacteristically high amplitude vibrations in a 1.7 m diameter, 13 m tall fluidized bed pyridine reactor with a catalyst that was very similar in particle size and density to FCC catalyst particles. The reactor was operating at a superficial gas velocity at the gas distributor of 0.7 m/s, the bed depth was 5.4 m and the pressure at the grid was 1.54 bara. Judged from a video (Knowlton [1]) showing gas bypassing occurring in a large Plexiglas fluidized bed, it was suggested that gas bypassing could reasonably cause the forces that would explain the observed deflections and low frequency vibrations of the reactor. Installation of “subway grating” baffles (Issangya et al. [5]) in the reactor helped to eliminate the vibrations.

Karimipour and Pugsley [8] studied pressure fluctuations in a fluidized bed for different combinations of bed depth, gas velocity, particle size and distributor design and found that streaming flow emerged gradually as the bed depth was increased and remained the dominant phase in the bed at operating conditions which would otherwise give a bubbling fluidized bed regime. Karimipour and Pugsley [9] proposed a mathematical model that predicted the trends of the effect of bed height, gas velocity and particle size on the fluidization behavior in gas bypassing beds. The model, however, assumed an already streaming bed and as such, did not address the cause of streaming flow.

Some types of operational instabilities which could be due to gas bypassing have been found in fluidized bed strippers. Strippers are essentially flowing fluidized beds with baffles. Rivault et al. [10] conducted studies in a 0.5 m OD× 0.1 m ID semi-circular, annular stripper using 6.5% fines FCC catalyst. The bed height was 1.35 m and the solids circulation flux reached 108 kg/s.m^2. The unit developed flow instabilities, significantly high pressure fluctuations, and flooding when it was operated without baffles and at high solids fluxes. Senior et al. [11] discussed flooding, large-scale maldistribution, and “bridging” that can occur in fluidized bed strippers and suggested that the reason for the gas-flow misdistribution was that dense-phase gas compression in deep beds was sufficient enough to defluidize a section of the stripper. Rall and Pell [12] studied the performance of different types of stripping internals in a Plexiglas circulating fluidized bed unit with a stripping section of 0.66 m diameter and 1.98 m high using 8% fines (<44 μm) equilibrium FCC catalyst powder. The solids circulation flux in the stripping zone reached as high as 64 kg/s.m^2. It was observed in a test with no internals that tests could not be conducted with a full bed of catalyst because of excessive system vibration. The bed height had to be lowered from the planned 1.98 m to 1.22 m for the unit to be operable.

The application of pressure to fluidized beds has been found to significantly affect the fluidization behavior of gas–solids systems. For Group A solids, increasing the pressure was found to increase the bed voidage and bed expansion (Subzwari et al. [13]), decrease the emulsion-phase density (Weimer and Quarderer [14]), and decrease the bubble size (Barreto et al. [15], Chan et al. [16] and Rowe et al. [17]). Fluidization is also “smoother” when beds are operated at high pressure which has been attributed to the decrease in the bubble size with increasing pressure. Furthermore, it was pointed out (Knowlton [18]) that pressure affects the fluidization behavior through its influence on gas density and viscosity, but the pressure effect is also dependent upon particle size. The objective of this work was to determine if the application of pressure could solve the problem of gas bypassing in deep beds of FCC catalyst particles.

2. Experimental

Tests were conducted in a 0.6-m-diameter, 6.1 m tall fluidized bed unit shown in Fig. 1. Fluidizing air was supplied to the bed through a 76-mm-diameter line that contained an orifice plate for measuring the gas flow rate to the bed. The test unit had two gas distributors as shown in Fig. 2. For tests reported here, where the superficial gas velocities did not exceed 0.46 m/s, the air distributor was an octagon-shaped sparger (Fig. 3) with 15 cm sides constructed from a 38-mm-diameter PVC pipe. It had ninety ¼-in (6-mm)-diameter holes arranged in two rows facing downward 30° from the vertical. The sparger was located 38 cm above the bottom flange. The second gas distributor (Fig. 4), located below the octagonal distributor, was a pipe manifold that was used for tests conducted at superficial gas velocities greater than 0.46 m/s. Both air distributors were designed to give pressure drops in excess of 30% of the bed total pressure drop at all operating conditions. The first stage cyclone had a 25-cm-diameter dipleg that returned solids to the bottom of the column via an automatic L-valve. Air exiting the first-stage cyclone passed through two, parallel 15-cm-diameter second-stage cyclones and then through a pair of parallel air filters. A Y-fitting joined the second-stage cyclone diplegs to a 76-mm-diameter line that returned the solids to the primary cyclone dipleg via another automatic L-valve. The air leaving the two filters was combined into a 102-mm-diameter line that had a butterfly valve. A 25-mm-diameter bypass line with a pneumatically operated pressure control needle valve was installed around the

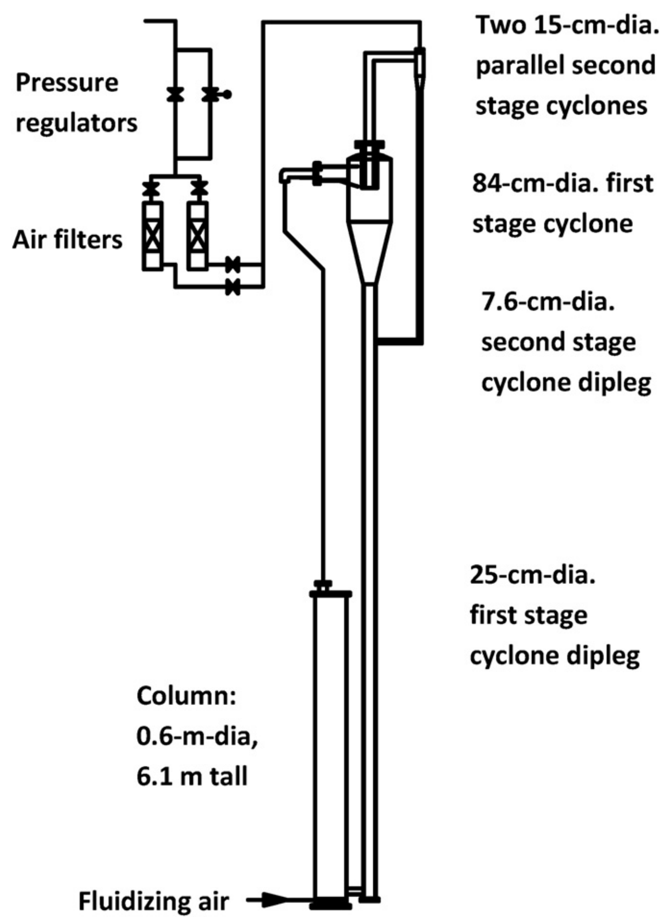

Fig. 1. Schematic drawing of the 0.6-m-diameter test unit.