Study of Electrostatic Charging in Gas-Solid Fluidized Bed Polyethylene Reactors

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***Abstract*—Electrostatic charge generation due to contact charging gives rise to significant problems including reactor wall fouling in commercial-scale gas-solid fluidized bed polyethylene reactors. To gain a better understanding of the occurrence of electrostatic charging in such reactors, a high-pressure pilot-scale gas-solid fluidized bed, housing two online Faraday cups, was used to fluidize polyethylene particles (received directly from industrial reactors), while the particles charging behaviour was investigated. The fluidization system was operated at atmospheric and 3 barg, which provided some insight as whether increasing the reactor pressure affects the magnitude of reactor fouling. In addition, the effect of fluidization flow regimes within the bed was studied by varying the fluidization gas velocity. The fluidizing particles charge distribution was measured in three different regions of the fluidized bed including the bulk of the bed consisting of relatively large particles, and near the reactor wall where mid-sized range particles were adhered to the wall. The results showed that the degree of wall fouling increased by elevating the system pressure as well as operation in turbulent flow regime. At atmospheric condition, the wall fouling composed of mainly positively charged particles. However, at 3 barg the fouling consisted of two layers, the outer layer and the inner layer that were positively and negatively charged, respectively.**

I. INTRODUCTION

Electrostatic charge generation due to contact charging gives rise to significant problems including reactor wall fouling and particle agglomeration in some commercial scale gas-solid fluidized bed reactors particularly those found in polyolefin industry. In polyethylene production process, due to the electrostatic charge generation during fluidization, polyethylene particles adhere to the reactor wall and form a sheet of fused particles [1-4]. If fouling continues, the sheet gets thicker, dislodges and drops onto the distributor plate, resulting in reactor shutdown for cleanup. Although numerous works reported in literature have aimed at understanding the underlying charging mechanisms in polyethylene reactors, the problem still persists. Thus further research is essential to gain a better understanding of factors that affect the degree of reactor charging and wall fouling. Commercial polyethylene gas-solid fluidized bed reactors operate at gas velocities within the turbulent flow regime and at high pressures close to 30 bar. Therefore, it is beneficial to conduct any studies in relation to polyethylene reactor electrifications under industrially-related operating conditions. A pilot-scale gas-solid fluidization infrastructure housing an online charge measurement technique was used in this study to investigate the effect of operating pressure and fluidizing gas velocity on the extent of reactor wall fouling. Contact charging in gas-solid fluidized beds is anticipated to be influenced by the bed hydrodynamics. It has been known that increasing the fluidization operating pressure makes fluidization smoother, reduces the gas bubbles size and in turn increases their rise velocity [5-10]. Such changes consequently affect the contacts between particles, and particles and the column wall; and thus, the magnitude of the electrostatic charge generation within the fluidized bed. As the fluidizing gas velocity increases, beyond minimum fluidization velocity, various flow regimes are achieved including bubbling, slugging and turbulent flow [11-13]. Each flow regime will results in different particles movement and mixing patterns therefore affecting the contact charging behavior of the particles. Majority of works found in open literature in relation to the effect of fluidizing gas velocity are conducted in bubbling or slugging flows. However, industrial polyethylene reactors are operated in turbulent flow regime. Thus, in this work the bed electrification and reactor fouling was also investigated for two fluidizing gas velocities, representing bubbling and turbulent flow regimes.

II. EXPERIMENTAL SETUP AND PROCEDURE

Fig. 1 illustrates the high-pressure pilot plant fluidization system. The three-dimensional gas-solid fluidization column was 0.15 m in diameter and 4.5 m in total height, with a 2.5 m fluidization section. The two expanded sections at the top and bottom of the column were 0.34 m in diameter. The entire system was made of stainless steel and designed to be operated at pressures up to 25 bar.

The same measurement method as Sowinski el al. [14-15] was applied in this system. Two cupper cups each with the diameter of 0.254 m were placed into the top and bottom expanded sections where both were electrically isolated from the column which was grounded. The expanded sections enclosing the cupper cups were then acted as Faraday cups and were connected to Keithley digital electrometers (Model 6514). A filter bag was housed in the top Faraday cup in order to capture the entrained particles during fluidization process. The distributor plate was designed to be opened in order to allow the bulk and wall particles to drop into the bottom Faraday cup for their charge measurement after fluidization completed. Two manways were installed at the top and bottom of the fluidization column allowing the easy access and removal of the Faraday cups upon the completion of the fluidization process. In each run, the cumulative charge of the entrained particles throughout the fluidization period, and the net charge of the particles adhered to the column wall and those in the bulk of the bed upon the completion of a fluidization run were measured.

The high pressure pilot system can be operated in two modes: once through and closed loop circulation enabling operations at pressures up to 3 barg and 25 barg, respectively. Compressed building air or nitrogen from a cylinder can be used as fluidizing gas source in both modes. The system was operated in once through mode in this work by using compressed building air with temperature of 23 °C and relative humidity of 3%. Since the source pressure was at a maximum of 6 barg, the fluidization operating pressure was limited to 3 barg. This pressure was achieved by first filling up the system with air and controlling the pressure using a back pressure regulator located at the exit of the column. Fluidization gas flow rate was measured and controlled by a mass flow controller. LabVIEW software was used for all data collection and control.

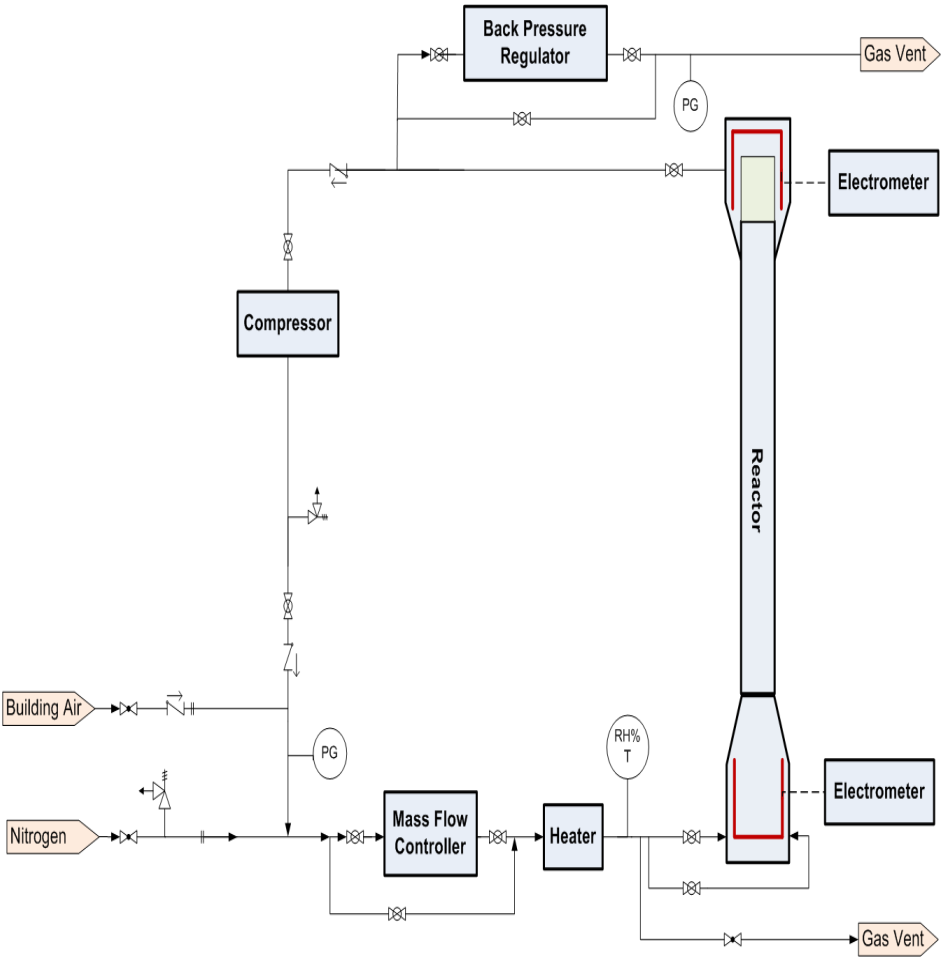

Fig. 1. Schematic diagram of the high-pressure fluidization system.

Effect of operating pressure was examined up to 3 barg by conducting the experiments in the bubbling flow regime. Due to some operational limitations, the effect of gas velocity was only examined at atmospheric condition. Polyethylene resin produced using metallocene catalyst in an industrial gas-solid fluidized bed reactor was used as bed material. The resin had a particle density of 918 kg/m^3 and a wide size distribution of 20-2000 micron with an average particle size of approximately 700 micron. For experiments in bubbling regime in both atmospheric and 3 barg, fluidizing gas velocity was set at 1.25 times the minimum fluidization velocity (U_{mf}). For experiments in turbulent regime, gas velocity was 6.5 times of U_{mf}. For all trials, fluidization was conducted for one hour.

An external Faraday cup was used to measure the charge of particles before they were placed inside the fluidization column. The filter bag was then mounted at the top of the column and the top manway was closed. The top Faraday cup was connected to the electrometer. The charge of entrained particles (fines) captured by the filter bag was cumulatively measured by electrometer during the fluidization process. After fluidizing for 60 minutes, the fluidizing gas was stopped and the column was depressurized (for the case of 3 barg) before the top manway was opened. Top Faraday cup and filter bag were removed, and the mass of fines were obtained. The distributor plate was opened, allowing the bulk particles drop into the bottom Faraday cup in order to measure their charge. After opening the bottom manway, the bottom Faraday cup was removed and the mass of bulk particles were measured. In the case that any particle fouling was present on the inner column wall, images were taken. The bottom Faraday cup was replaced and pressurized air was then used to remove any particles adhered to the column wall. These particles were divided into two groups of tightly bound (TB) and loosely bound (LB), according to the lower and higher gas velocity of the air used to remove them, respectively. Samples obtained from the initial, fines, bulk, and wall region were analyzed for their particles size distribution. Results were normalized with respect to the mean particle diameter of the initial particles.

III. Results and Discussion

Experiments were conducted at atmospheric condition in bubbling and turbulent fluidization flow regimes and at 3 barg in bubbling flow regime in order to investigate the effect of fluidization gas velocity and pressure on fluidized bed electrification. All trials began with initial particles having a small net charge-to-mass ratio (-0.02 $\pm$ 0.075 μC/kg).

Fig. 2 shows the effect of fluidizing gas velocity on the magnitude of the reactor wall fouling and fines collected. The mass of the collected particles in turbulent flow regime was significantly higher than that in bubbling regime. This was expected due to the higher fluidizing gas velocity. In atmospheric condition, the mass of wall particles in the experiments operated in turbulent regime was larger than that in bubbling regime, while the mass of tightly bound particles was comparable for both conditions. Results also indicate that the higher operating pressure led to a significantly higher amount of tightly and loosely bound particles. Almost no fine particles were detected in experiments in bubbling regime at 3 barg and therefore their results will not be further discussed.

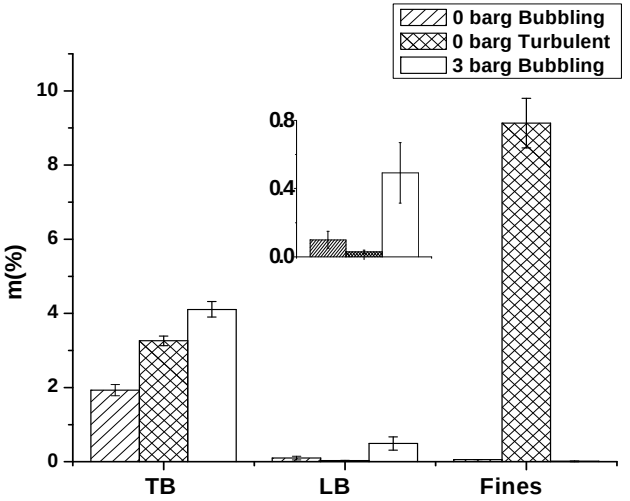

Fig. 2. Results of the mass percentage of particles collected off of the column wall and entrained fines.

Samples taken from various regions of the fluidized bed were analyzed for their particle size distribution (PSD). As can be seen in Fig. 3, for all of the three operating conditions, particles adhered to the column wall were smaller than those in the bulk of the bed. This indicated that the particle size decreased in the radial direction from the center of the bed to the column wall.

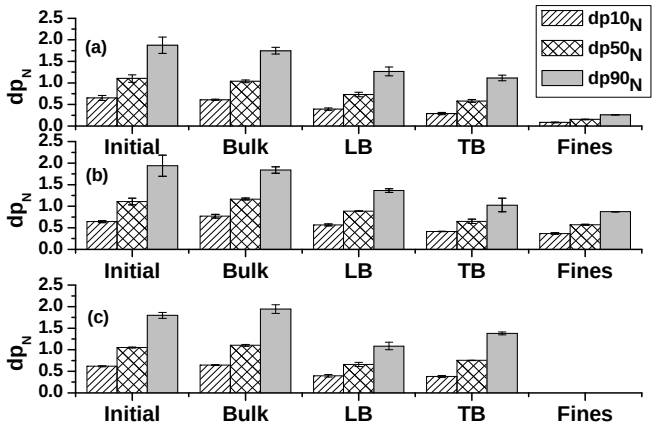

Fig.3. Normalized particle size ($dp10_N$, $dp50_N$, $dp90_N$) in different regions of the fluidized bed. (a) Bubbling flow regime at atmospheric; (b) Turbulent flow regime at atmospheric; (c) Bubbling flow regime at 3 barg.

Fig. 4 illustrates the effect of fluidization flow regime on the particle size distribution of loosely bound, tightly bound, and fine particles. Fig. 4a and Fig. 4b show the normalized PSD results for particles that adhered to the column wall for all three scenarios. It is clear that under atmospheric conditions in the bubbling flow regime, the wall layer consisted of some smaller particles that were not found in the turbulent flow regime. The loosely bound particles had a wider particle size distribution at atmospheric condition in comparison to that at higher pressure (Fig. 4a). As can be seen in Fig. 4c, fine particles collected in turbulent regime also consisted of some significantly larger particles than those in the bubbling. The higher gas velocity in turbulent flow generated a larger drag force that enabled the entrainment of larger particles from the bed. This resulted in the elutriation of smaller particles that were previously found in the wall region during the bubbling experiments. As a result, there were less smaller particles adhered to the wall in turbulent experiments than that in bubbling experiments.

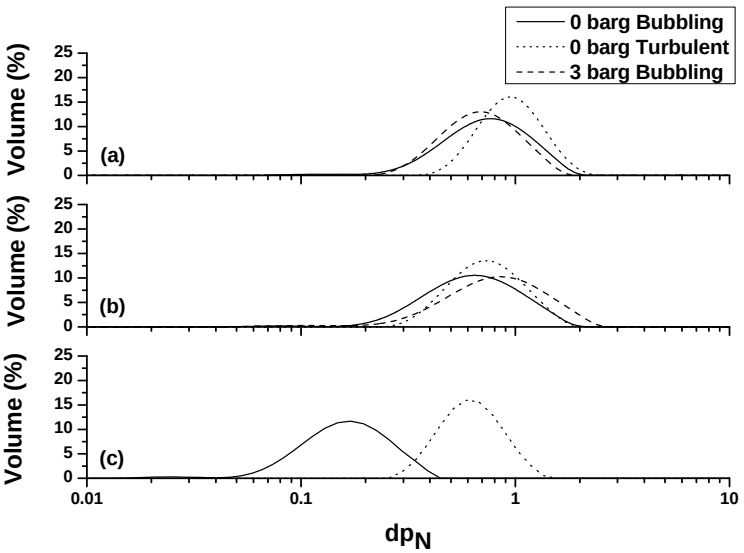

Fig. 4. Normalized particle size distribution. (a) Loosely bound; (b) Tightly bound; (c) Fine particles.

The net q/m of particles collected from various regions of the fluidization column is presented in Fig. 5. The bulk particles were predominantly positively charged with the magnitude of the charge close to zero, similar to that of the initial particles. In the turbulent regime, the particles adhered to the wall were predominantly positively charged, while in bubbling regime loosely and tightly bound particles were charged positively and negatively,

respectively. In addition, the net specific charge of the wall particles in bubbling regime was on average slightly higher than that in turbulent regime (55.91 μC/kg and 46.48 μC/kg, respectively). In bubbling regimes at both atmospheric and 3 barg, tightly bound particles were highly charged with an absolute value 1.5 and 2.5 times higher than that of loosely bound particles, respectively. Fine particles were negatively charged under both operating conditions for atmospheric condition. The specific net charge of fine particles in turbulent flow was very low, while quite high in bubbling flow.

In the turbulent flow regime, in comparison to that of the bubbling, the fluidizing gas velocity was much higher resulting in a greater degree of particles mixing; and thus, enhancing the degree of particles contact with each other and those with the fluidization column wall. Therefore, more charges were generated in turbulent than in bubbling flow regime. Similarly, as the operating pressure is increased, the bed hydrodynamics will change such that gas bubble size reduces, resulting in higher degree of mixing of particles and their contact charging. Overall, both effects would result electrostatic forces dominating those of gravity and drag force and more fluidizing particles adhering to the column wall, which agrees well with the results obtained in this study where the mass of the wall particles in the turbulent flow regime and at 3 barg was higher than what was found in the bubbling flow under atmospheric conditions.

Since the loosely bound particles, especially those in the bubbling flow regime at both pressures consisted of some highly negatively charged particles, it was critical to further investigate the charge distribution of these particles at various conditions. This was achieved by using a charged particle separator apparatus detailed elsewhere [16]. As can be seen in Fig. 6, the loosely bound particles in both bubbling and turbulent experiments were predominantly positively charged; however, the amount of positively charged particles in turbulent experiments was slightly higher than that in bubbling, while those in bubbling experiments at 3 barg were higher than both bubbling and turbulent regimes under atmospheric conditions. As previously mentioned, since the fluidizing gas velocity was lower in the bubbling flow regime it would have not enabled the entrainment of the smaller and negatively charged particles (Fig. 5). Thus, allowing these particles to contact the fluidized bed column wall, and adhere to it. In turbulent experiments, most of the smallest particles, which were negatively charged, were blown out of the column at the beginning of the fluidization. Hence, most of the particles left in the bulk of the bed were positively charged and thus formed a predominantly positively charged layer on the wall.

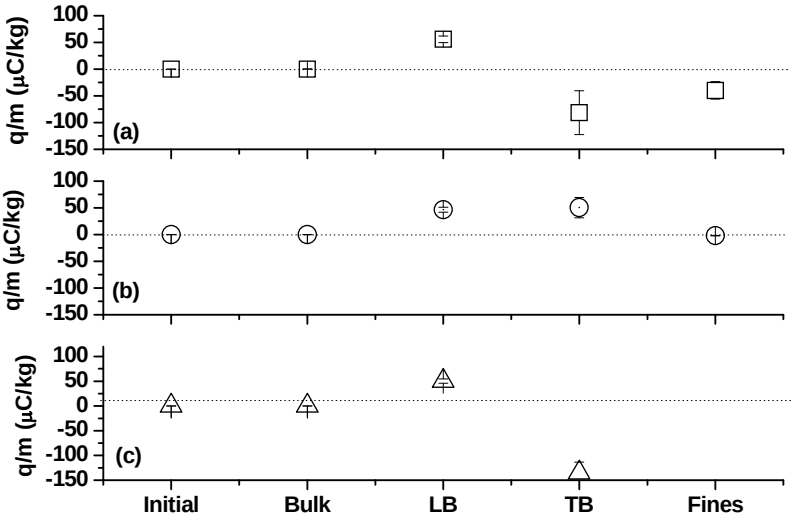

Fig. 5. Results of the net charge-to-mass ratio of the initial, bulk, loosely (LB) and tightly bound (TB), and fine particles. (a) Bubbling flow regime at atmospheric; (b) Turbulent flow regime at atmospheric; (c) Bubbling flow regime at 3 barg.

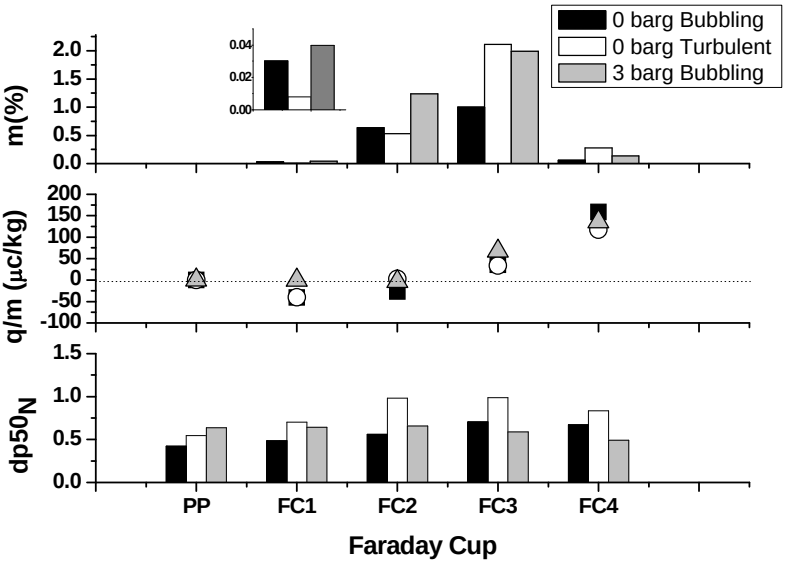

Fig. 6. Results of the charged particle separator of loosely bound particles. PP stands for positive plate.

In this work the degree of column wall fouling was also examined visually by taking photos of particle wall coating from the bottom of the fluidized column after the removal of

the bulk particles. As can be seen in Fig. 7, the wall layer in bubbling experiments extended farther into the column than that in turbulent flow. In bubbling flow, there was also a second layer of particles formed a few centimeters above the main wall layer, while this did not occur in turbulent experiments. This could be due to the bubbles bursting at the surface of the expanded bed height, projecting the charged particles towards the column wall and enhancing their chance to attach to the wall.

(a) (b) (c)

Fig.7. Images of wall fouling. (a) Bubbling flow regime in atmospheric; (b) Turbulent flow regime in atmospheric; (c) Bubbling flow regime at 3 barg.

IV. Conclusion

Increasing the gas velocity (i.e., transition from bubbling to turbulent flow regime) as well as increasing the operating pressure from atmospheric to 3 barg resulted in a higher degree of wall fouling. In addition, this indicated that since the net specific charge of the loosely bound particles varied slightly, then particles charge must have also increased for these operating conditions. This agrees well with the fact that at high pressures and gas velocities more mixing and contacts occur between particles and the column wall, resulting in a higher degree of bed electrification. In turbulent flow regime, particles on the column wall were predominantly positively charged, while loosely and tightly bound particles in the bubbling flow regime at both pressures were charged positively and negatively, respectively. The reason for the particles bipolar charging in bubbling flow regime was associated to the removal of the smaller negatively charged particles in turbulent flow regime due to their entrainment. This finding in turn indicated that bipolar charging occurred in the system studied where smaller polyethylene particles were oppositely charged (i.e. negatively) to those of the larger particles.

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