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INDUSTRIAL CHALLENGES OF HDPE FLUID BED DRYING IN DIFFERENT GRADES

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RESUMEN: Saber de los retos en el secado de HDPE es muy esencial para mejorar el proceso. En este trabajo, el polímero en términos de producción con densidades de 940 kg / m3 o superior se produce mediante la técnica de polimerización en suspensión utilizando reactores CSTR. Las materias primas incluyen etileno, hidrógeno, hexano, buteno y catalizador Ziegler-Natta. El producto en suspensión se separa mediante un decantador (20-30% de hexano residual) y un secador de lecho fluidizado de 2 etapas en el que el polvo húmedo se seca adicionalmente hasta un contenido de hexano residual de 0,1%. El proceso de secado aceptable requiere fluidización adecuada. El estudio de este comportamiento hidrodinámico es un trabajo complejo, ya que su comprensión y modelización matemática sigue siendo un reto hasta la fecha y depende en gran medida de la recolección de datos empíricos significativos sobre plantas de procesamiento. El secado de partículas se realiza comúnmente usando secadores de lecho fluidizado en plantas HDPE donde su evaluación de rendimiento depende principalmente del comportamiento de las partículas y del patrón de flujo. Es importante comprender los factores que afectan al régimen de flujo de partículas de gas para alcanzar el proceso con buena eficiencia. Se han empleado comúnmente ciclones dobles para mejorar la separación y evitar que el polvo se transfiera a ubicaciones de plantas vulnerables a lo largo de la trayectoria de secado, incluyendo la placa distribuidora, las bandejas de torre de lavado y los impulsores de sopladores. Esto ayuda a describir los patrones de flujo y reduce aún más la aglomeración de partículas que conduce a un menor consumo de energía en los soplantes. Se investigan diferentes aspectos de los factores que afectan al proceso de secado en la planta capaz de manejar una alimentación húmeda de 56000 kg / h (40000 kg / h de polímero seco de diversos grados incluyendo PE100, BL3 y EX5 y 16000 kg / h de hexano).
Palabras clave: HDPE, secador de lecho fluidizado, desafíos industriales, fluidización.

ABSTRACT: Knowing of challenges in drying of HDPE is very essential to improve the process. In this work, the polymer in terms of production with densities of 940 kg/m^3 or higher is produced by slurry polymerization technique using CSTR reactors. Raw materials include ethylene, hydrogen, hexane, butene and Ziegler-Natta catalyst. The slurry product is separated by a decanter (20-30% residual hexane) and a 2-stage fluidized bed dryer where wet powder is dried further down to 0.1% residual hexane content. Acceptable drying process needs proper fluidization wheras, study of this hydrodynamic behaviour is a complex work as such it's understanding and mathematical modelling remains a challenge to date and it relies heavily on collection of meaningful empirical processing plant data. Drying of particles is commonly carried out using fluidized bed dryers in HDPE plants where their performance evaluation depends mainly on particles behavior and flow pattern. Understanding factors affecting gas-particle flow regime is important for reaching the process with good efficiency. Double-cyclones have commonly been employed to improve separation and prevent powder being transferred to vulnerable plant locations along the drying path including distributer plate, scrubbing tower trays and blowers impellers. This

assists to describe flow patterns and further reduces particle agglomeration leading to less energy consumption in blowers. Different aspects of factors affecting the drying process in the plant capable of handling a wet feed of 56000 kg/hr (40000 kg/hr dry polymer of various grades including PE100, BL3 and EX5 and 16000 kg/hr hexane) are investigated.

Keywords: HDPE, fluidized bed dryer, industrial challenges, fluidization.

1. INTRODUCTION

In petrochemical industry, grade changing is the everlasting demand of polyethylene processes. The increased competition, the narrow margin of the profit and the demand from the customers call for the high quality and end-use specified products. Slurry processes are still widely used, while slurry reactors have a very efficient heat removal and wide co-monomer range [1-4]. Based on fig 1, the process consists of two continuous stirred-tank reactors (CSTR) that can be operated in parallel or in series, depending on which grade of polymer is required. This process is designed to produce either unimodal (broad or narrow molecular weight distribution polymer) or bimodal polymer by using Ziegler-Natta catalysts. For bimodal polymer production, a high concentration of hydrogen in the first reactor and a low concentration of hydrogen plus a small amount of co-monomer in the second reactor are used. In the process as a copolymerization of HDPE, Ziegler-Natta catalyst has an important role to dictate particle size distribution on powder product in slurry media [5-7]. The size of catalyst particles is 5-20 micron, also heat and mass transfer properties of the growing particles are functions of the particle size [8,9]. Wax is a byproduct in this technology, which will be able to settle in powder particles affecting the adhesion of dry powder particles. Drying is usually the final stage of operations and its product will be final packaging. Liquids may be removed from solids mechanically by presses or centrifuges or thermally by vaporization. All industrial experiences in the majority of plants, which have been studied show it is better to reduce liquid content mechanically, and then the materials feed to thermal drying [10].

In HDPE fluid bed dryer, fluidization is the phenomenon of imparting the properties of a fluid to a bed of particulate solids by passing a fluid (liquid or gas) through the material. Fluidized beds are an important asset in many industrial processes because they present several advantages that include a high rate of heat and mass transfer, low-pressure drops, and uniform temperature distribution. Fluidized bed hydrodynamic behavior is very complex and must be understood to improve fluidized bed operations. Several parameters are used to know the behavior of a material the moment it is fluidized. In fluid bed drying entrainment is used to describe the ejection of particles from the surface of a bubbling bed and their removal from the vessel in the fluidizing gas.

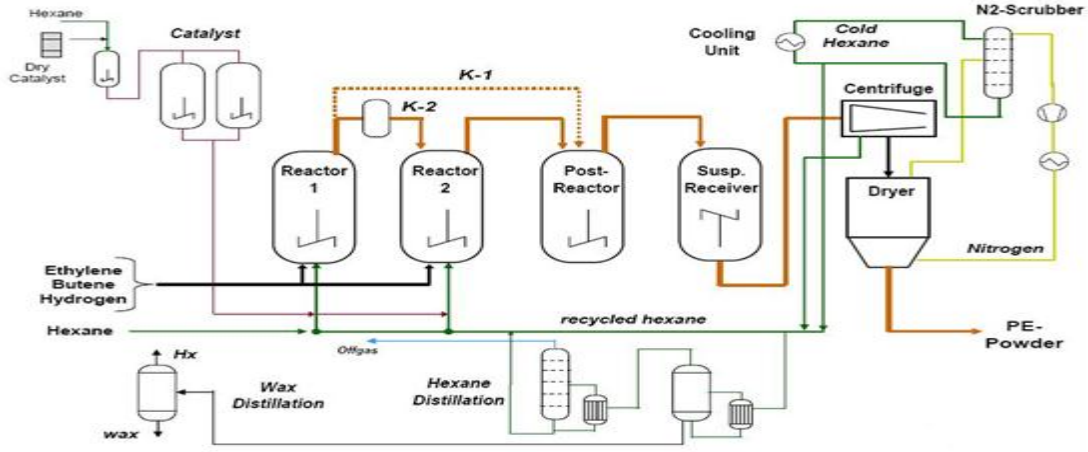


Fig. 1. Schematic diagram of slurry HDPE process

Other terms such as ‘carryover’ and ‘elutriation’ are often used to describe the same process. The factors affecting the rate of entrainment of solids from a fluidized bed are particle size distribution (PSD), terminal velocity, superficial gas velocity, particle density, gas properties and gas flow regime. In a moving gas stream with a number of particles with a range of particle size some particles may fall and some may rise depending on their size and their position. Thus, the entrainment of particles in an upward-flowing gas stream is a complex process.

The minimum fluidization velocity also constitutes a reference for evaluating fluidization intensity when the bed is operated at higher gas velocities [11]. In general, U_{mf} is a function of particle properties/geometry, fluid properties, and bed geometry. Gas holdup is another very important parameter that characterizes the fluidization quality, mixing, and process efficiency in a fluidization system, and is defined as the volume fraction of gas present within the bed.

When a fluid is passed upwards through a bed of particles the pressure loss in the fluid due to frictional resistance increases with increasing fluid flow. A point is reached when the upward drag force exerted by the fluid on the particles is equal to the apparent weight of particles in the bed. At this point, hot gas lifts the particles and the bed becomes fluidized. The force balance across the fluidized bed dictates that the fluid pressure loss across the bed of particles is equal to the apparent weight of the particles per unit area of the bed. Thus:

$$Pressure\ drop = \frac{Weight\ of\ particles - upthrust\ on\ particles}{Bed\ cross - sectional\ area}$$

For a bed of particles of density ρ_p, fluidized by a fluid of density ρ_g to form a bed of depth H and voidage ε in a vessel of cross-sectional area A:

$$\Delta P = \frac{HA(1-\varepsilon)(\rho_p - \rho_g)g}{A} \quad (1)$$

$$\Delta P = H(1-\varepsilon)(\rho_p - \rho_g)g \quad (2)$$

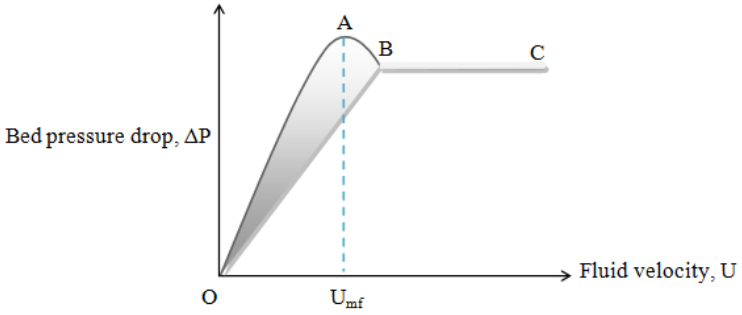
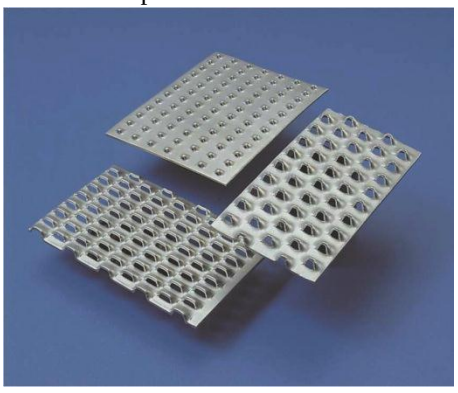


Fig.2. Bed pressure drop versus fluid velocity

Based on Fig.2, there is a plot of fluid pressure loss across the bed versus superficial fluid velocity through the bed. The straight-line region OA is the packed bed region.

The region BC is the fluidized bed region where Eq. (1) applies. At point A, it will be noticed that the pressure loss rises above the value predicted by Eq. (1). This rise is more marked in small vessels and in powders, which have been compacted to some extent before the test and is associated with the extra force required to overcome wall friction and adhesive forces between bed and the distributor. The superficial fluid velocity at which the packed bed becomes a fluidized bed is known as the minimum fluidization velocity, U_{mf}. This is also sometimes referred to as the velocity at incipient fluidization.

Gas distributor (fig 3) is a device designed to ensure that the fluidizing gas is always evenly distributed across the cross-section of the bed. It is a critical part affecting the fluidization regimes. Good design is based on achieving a pressure drop, which is a sufficient fraction of the bed pressure drop. Many operating problems can be traced back to poor distributor design [12-14]. Loss of fluidizing gas as an important challene will lead to collapse of the fluidized bed.



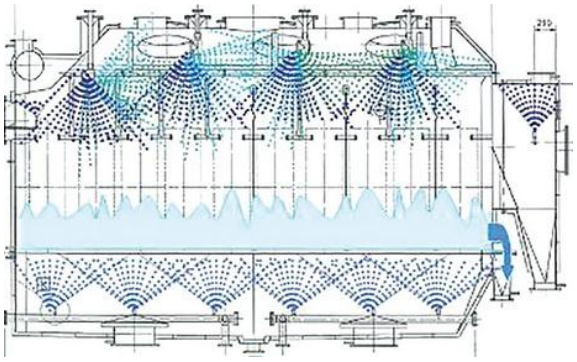
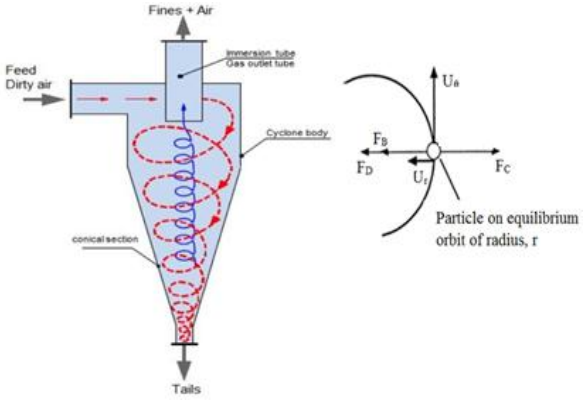


Fig.3. fluid bed dryer and gas distributor

Cyclone separators are often used in fluidized beds for separating entrained solids from the gas stream. Cyclones installed within the fluidized bed vessel would be fitted with a dip-leg and seal in order to prevent gas entering the solids exit. It consists of a vertical cylinder with a conical bottom, a tangential inlet near the top, and an outlet for dust at the bottom of the cone. The outlet pipe is extended into the cylinder to prevent short-circuiting of gas from inlet to outlet. Particles in the gas are subjected to centrifugal forces which move them radially outwards, against the inward flow of gas and towards the inside surface of the cyclone on which the solids separate (Fig.4)[15]. The performance of equipment is concluded using its static pressure drop between input and output [16].

Based on the Navier-Stokes equations, tangential velocity does not depend on the "Z" coordinate; radial velocity is constant throughout the flow and flow is turbulent [17]. The main part of the pressure drop about 80%, is considered to be pressure losses inside the cyclone due to the energy dissipation by the viscous



stress of the turbulent rotational flow [18].

Fig.4. Cyclone work mechanism

The remaining 20% of the pressure drop are caused by the contraction of the fluid flow at the outlet, expansion at the inlet and by fluid friction on the cyclone wall surface.Tangential inlets are preferred for the separation of solid particles from gases [15,19]. The relationship between pressure and 3D velocity can be simplified by neglecting the axial effects. The centrifugal force F_C at radius r is equal to mU_θ^2/rg_c, where m is the mass of the particle and U_θ is its tangential term. A large-diameter cyclone has a much lower separation factor at the same velocity. The general performance of the equipment is concluded using its static pressure drop between input and output and following efficiency expression [11].

$$\Delta P_C = \xi_c \frac{\rho_g V_i^2}{2}$$

The cyclone pressure drop (ΔP_C) with the square of the velocity (V_i^2), resistance coefficient (ξ_c) and gas density (ρ_g) is directly related and Euler number (E_u) for the cyclone will be defined according to Eq. (4):

$$E_u = \frac{\Delta P_C}{\frac{1}{2}\rho_g V_i^2}$$

Solids feeders as various devices are available for feeding solids into the fluidized bed. The choice of device depends largely on the nature of the solids feed. Screw conveyors, spray feeders and pneumatic conveying are in common use [20]. Particles in a real powder will seldom be of uniform size but will be characterized by a PSD, which will affect their rheological and fluidization behavior. For example, a wide size distribution has been observed to lead to smoother fluidization and better gas-solid contacting [21], whereas a narrow distribution may enhance bed stability (e.g., reduce segregation)[22]. In addition, the presence of fines (particles of 45μm diameter or less) may improve the performance of fluidized-bed mediums [23,24]; however, excessive fines may cause the powder to be too cohesive for proper fluidization [25]. Particle size and PSD will influence the magnitude of the interparticle force. As the size distribution increases, the minimum fluidization velocity decreases, while the minimum bubbling velocities increases.

2. THEORETICAL

Fluidized bed process is ideal for heat-sensitive and non-sensitive range of products and uniform process conditions because of gas passes through the layer of materials based on the controlled speed will be fined, and then this situation create the fluidized state [26]. In fluid bed drying, heat is provided by flow of hot gas with or without additional heat transfer provided by immersed heat exchangers in the bed. Fluid bed drying offers significant advantages compared with other methods of particulate drying [27]. In reality, this operation offers fluidizing of particles, easy transfer of material and high rate of change of energy with required thermal efficiency along preventing extreme heating of particles individually. Conventional fluidized bed is formed by passing a gas stream from the bottom of a medium of particulate solids. At low gas velocities, the bed is static (packed). The bed of particles rests on a gas distributor plate. The fluidizing gas passes through the distributor

and it is uniformly distributed across the bed. According to Fig.5, there are various regimes of the particulate bed from packed to bubbling bed when the gas velocity is increased. A fluidized bed is operated at superficial gas velocities higher than the minimum fluidization velocity, U_{mf} normally at 2–4 U_{mf}. The minimum fluidization velocity is typically obtained from experiments. There are several ways to determine the minimum fluidization velocity experimentally. It can also be estimated using various correlations. A list of minimum fluidization velocities can be obtained from Gupta and Sathiyamoorthy[28].

It should be noted that these correlations have limitations such as particle size, column dimensions, and operating parameters. Thus, they are valid only for a certain range of criteria and under certain operating conditions. The effect of wetness of the particles is, however, not included. Particles with high initial moisture content require a higher minimum fluidization velocity than a similar bed of dry particles.

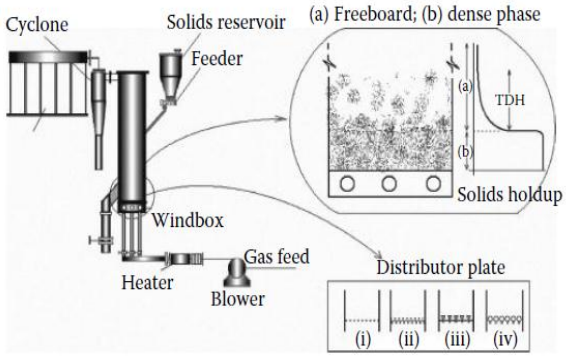


Fig.5. Typical fluid bed drying, zones, types of distributer plates, and various regims at different gas velocites

For the case of dry (or partially dry, no surface moisture) particles, if the fluidizing gas is further increased, the bed of particles goes through different types of fluidization regimes depending on the type of particles with reference to the Geldart classification of powders [29,30].

3. INDUSTRIALL EXPERIMENT

3.1 Experimental material

Hexane as the moisture content of the polymer; nitrogen (dry gas carrier); water steam for heating of jacket within the drying bed and the different grades of high density polyethylene such as PE100, BL3 and EX5 as bimodal polymer are materials used in our work that their details are as follows:

Table1. Physical specifications of experimental materials

Descriptions	Values
Nitrogen (N2)	
ρ_g *(kg* m^{-3}*)*	1.251
μ_g*(kg* $m^{-1}s^{-1}$*)*	2.07×10^{-5}
Normal hexane (HX)	
ρ*(kg* m^{-3}*)*	659
μ *(kg* $m^{-1}s^{-1}$*)*	3×10^{-4}
HDPE-PE100 (based on ZN -1)	
ρ_p*(kg* m^{-3}*)*	948±0.002
ρ_{Bp}*(kg* m^{-3}*)*	450
HDPE-BL3 (based on ZN -2)	
ρ_p*(kg* m^{-3}*)*	953±0.002
ρ_{Bp}*(kg* m^{-3}*)*	392
HDPE-EX5 (based on ZN -2)	
ρ_p*(kg* m^{-3}*)*	948±0.002
ρ_{Bp}*(kg* m^{-3}*)*	396

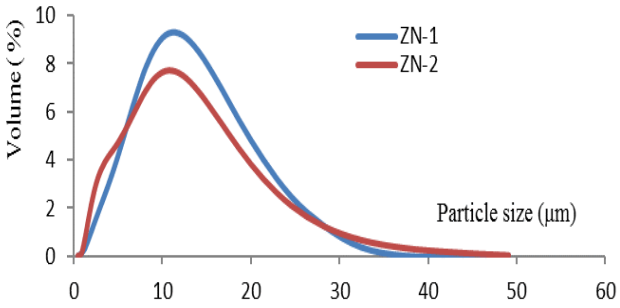


Fig.6 Particle size distribution of ZN- catalysts

Table 2. Sieve size for PE100, BL3 and EX5 respectively as HDPE grades.

Size (µm)	volume (%)	volume (%)	volume (%)
450	2.92	0.68	0.28
357.5	7.8	1.44	1.32
282.5	6.56	0.04	0.76
225	30.04	8.8	11.36
180	24.36	28	25.64
142.5	16.32	34.12	27.76
94	5.32	24.92	29.28
63	3.28	1.84	3.12

3.2 Experimental equipment

Based upon Fig.7; equipment include fluidized bed dryer (with dimensions as shown in table3), inner coil bed, nitrogen heater, gas distribution plate, gas blowers (BL-1,2), scrubbing tower, double cyclone (CY-1,2), rotary valves (BW-1,2), differential pressure gauges (PDI-1,2,3,4,5,6,7), temperature gauges (TI-1,2,3,4,5,6,7,8,9,10,11,12,13), pressure gauges (PI-1,2)

and flow meter (FI-1,2) forming the investigated fluidized bed drying system.

Table 3. Key dimensions of HDPE fluid bed dryer used in study.

Descriptions	Values
Length(m)	10.2
Width(m)	4.18
Height (m)	6.26

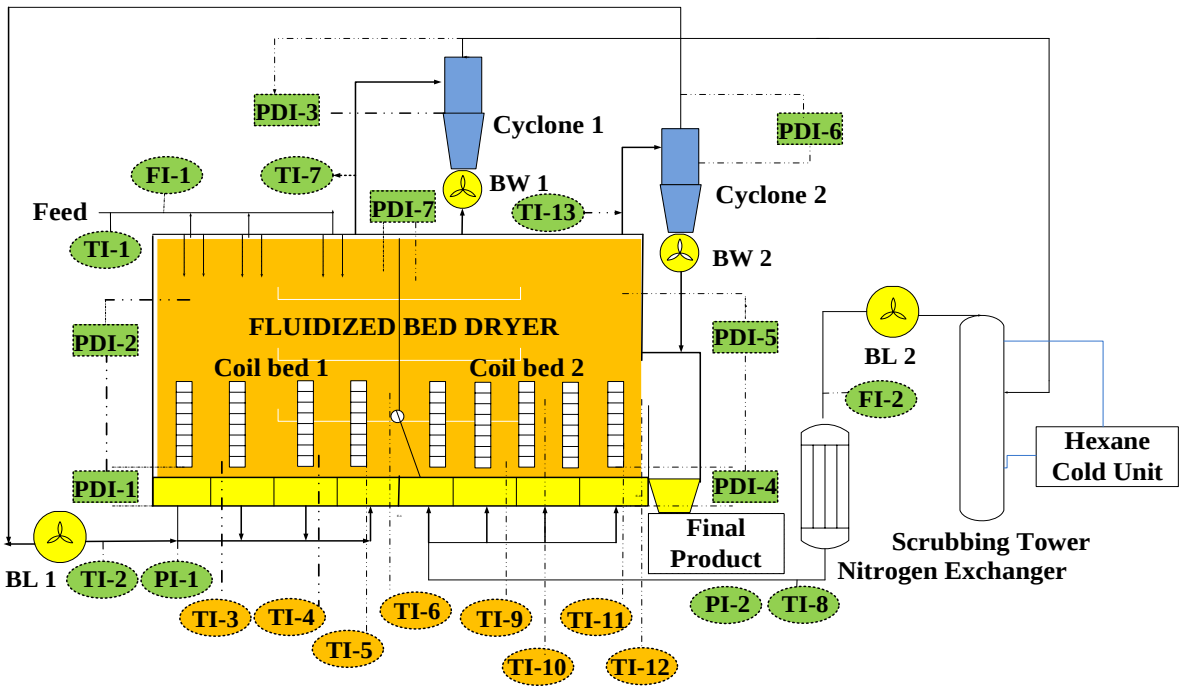


Fig.7. Schematic diagram of the HDPE fluidized bed drying unit

Double-cyclones include the spiral hood, cylinder, cone and triangular dust hopper that secondary gas streams from the pressurized dryer back to the dust hopper are prevented by means of the rotary valves BW-1/2 which do also control / even out the dust flow rate from the hoppers back into the dryer.(Fig.8)

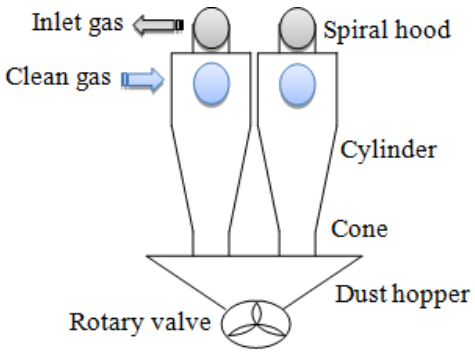


Fig.8. Schematic diagram of the HDPE double cyclone unit

3.3 Experimental method

To identify challenges in the field of HDPE drying, we use an industrial fluidized bed dryer as experimental equipment (Fig.7). In this work, the pre-dried product reaches the second stage where final drying to the required hexane content of 0.1%. The hot exhaust gas of drying stage 2 of approx. 78-80°C is de-dusted in CY-2 before being recompressed in BL-1 to be used to flash-dry the wet HDPE coming in the first drying stage to about 1% hexane content (dry basis). Because of the high dust content of the drying gas, the gas distributor has a special slotted plate.(Fig.9)

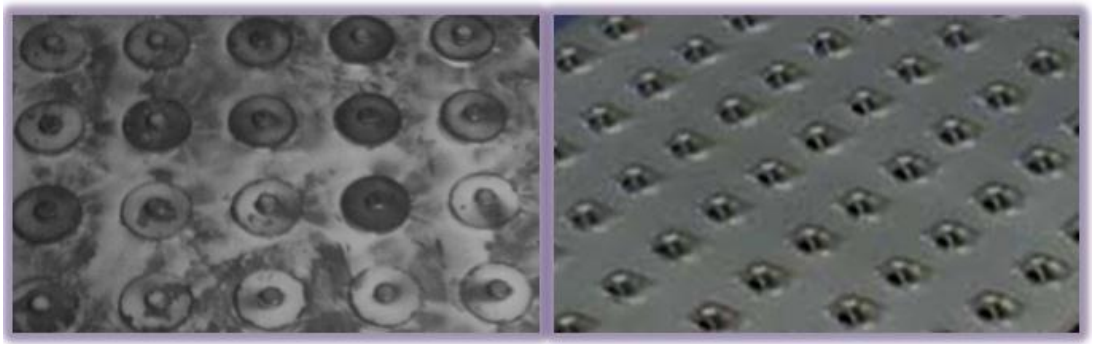


Fig.9. Gas distributor plate schematic in our test

In our test, distributor plates have a distinct advantage due to their unique Jet-shaped conical hole, which enable a sharp nitrogen flow of intense heat. One most important reason for an increasing number of bulk materials being thermally treated in fluidized bed units, is the resulting high economic efficiency achieved from the intensive heat and material exchange between solid material and fluid medium. This is largely due to type of gas distribution plate or distribution deck, which is an essential component of any fluid bed dryer installation [13,14].

The nitrogen gas which is re-circulated from stage 2 is simultaneously loaded with additional hexane which corresponds to about 35% relative humidity.The rich gas will be leaving the first drying stage at approx. 60°C to be de-dusted in CY-1 then, the gas must be regenerated by condensing the hexane that has been evaporated in the dryer by using scrubber tower. Therefore, nitrogen will be able to absorb n- hexane as the inner moisture based on the high driving force of mass transfer. Due tohigh dust loadings can be handled in our system, cyclones certainly should be a high performance equipment in separation the dried particles from gas leaving the fluid bed dryer. The capacity of the system with three separate centrifuges is 56,000 kg/hr of wet feed and 40,000 kg per hour of dry feed. The wet cake includes initial moisture content of 35% and inlet temperature of about 32°C. At the beginning of industrial test, by using of a fluid bed dryer, which has been cleaned before our work, we determined operating parameters shown in Table 4.

In this work as an idustrial drying operation, the change of parameters such as pressure drop and temperature versus the wet cake flow rate, polymeric grade, nitrogen volumetric flow rate will be resulted in the following table5 and 6.

Table 4. Operating parameters with cleaned HDPE fluidized bed dryer.

Descriptions	Values	Descriptions	Values
FI-1(kg hr $^{-1}$)	0	*TI-1(˚C)*	32
FI-2(m^{3}hr $^{-1}$)	10000	*TI-2(˚C)*	93
PI-1(kpa)	14.5908	*TI-3(˚C)*	81.1
PDI-1(kpa)	0	*TI-4(˚C)*	81.3
PDI-2(kpa)	0	*TI-5(˚C)*	87.8
PDI-3(kpa)	0.303975	*TI-6(˚C)*	88.9
PI-2(kpa)	14.5908	*TI-7(˚C)*	86
PDI-4(kpa)	0	*TI-8(˚C)*	93
PDI-5(kpa)	0	*TI-9(˚C)*	85.9
PDI-6(kpa)	0.506625-	*10(˚C)*	87.8
PDI-7(kpa)	2.22915	*TI-11(˚C)*	89.2
		-12(˚C)	89.9
		TI-13(˚C)	86

Table 5. Pressure drop of bed versus nitrogen volumetric flow in PE100, BL3 and EX5 respectively with 38000 kg hr $^{-1}$.

FI-2(m^{3}hr $^{-1}$)	PDI-2(kpa)	PDI-2(kpa)	PDI-2(kpa)
10000	1.11457	1.114575	1.114575
11000	1.519875	1.41855	1.41855
12000	2.0265	1.925175	1.925175
13000	2.533125	2.22915	2.0265
13500	3.03975	2.8371	2.735775
14000	3.44505	3.03975	3.141075
14011	3.749025	3.141075	3.141075

14020	3.749025	3.141075	3.141075
14040	3.85035	3.03975	3.03975
14040	3.85035	3.03975	3.03975
14040	3.85035	3.03975	3.03975
14050	3.749025	2.938425	2.938425
14050	3.749025	2.938425	2.938425
14100	3.6477	2.8371	2.938425
14100	3.6477	2.8371	2.8371
14100	3.6477	2.8371	2.8371

According to Table 5, PDI-2 is steady after achieving FI-2 of 14040, 14011 and 14000 for PE100, BL3 and EX5 respectively.

Table 6. Operating parameters in FI-2 of 13981for BL3 grade.

Descriptions	Values	Descriptions	Values
FI-1(kg hr $^{-1}$)	38000	*TI-1(˚C)*	32
FI-2(m^{3}hr $^{-1}$)	13981	*TI-2(˚C)*	89
PI-1(kpa)	19.65705	*TI-3(˚C)*	61.8
PDI-1(kpa)	1.51987	*TI-4(˚C)*	64
PDI-2(kpa)	3.2424	*TI-5(˚C)*	64.1
PDI-3(kpa)	2.0265	*TI-6(˚C)*	65.2
PI-2(kpa)	19.65705	*TI-7(˚C)*	62
PDI-4(kpa)	2.330475	*TI-8(˚C)*	98
PDI-5(kpa)	3.2424	*TI-9(˚C)*	88.7
PDI-6(kpa)	1.2159	*TI-10(˚C)*	86.8
PDI-7(kpa)	-0.20265	*TI-11(˚C)*	82.9
Volatile(%)	0.07	*TI-12(˚C)*	87.9
		TI-13(˚C)	82

Table 7. Operating parameters in FI-2 of 14060 for BL3 grade.

Descriptions	Values	Descriptions	Values
FI-1(kg hr $^{-1}$)	38000	*TI-1(˚C)*	32
FI-2(m^{3}hr $^{-1}$)	14060	*TI-2(˚C)*	87
PI-1(kpa)	19.65705	*TI-3(˚C)*	59
PDI-1(kpa)	1.51987	*TI-4(˚C)*	62.2
PDI-2(kpa)	3.03975	*TI-5(˚C)*	62
PDI-3(kpa)	2.0265	*TI-6(˚C)*	63.2
PI-2(kpa)	19.65705	*TI-7(˚C)*	60
PDI-4(kpa)	2.22915	*TI-8(˚C)*	98
PDI-5(kpa)	3.2424	*TI-9(˚C)*	86.8
PDI-6(kpa)	1.2159	*TI-10(˚C)*	84.6
PDI-7(kpa)	0.101325	*TI-11(˚C)*	87
Volatile(%)	0.04	*TI-12(˚C)*	86.3
		TI-13(˚C)	80

Furthermore, this table will presente minimum fluidization flow rate. After gathering the recent results for minimum fluidization, we test fluid bed drying on BL3 in two different nitrogen flow rates. (Table6,7). Based on Fig.5,10 and Table 3, by the use of check in sight glass instaled in fluidized bed we will see a fluidized height of 1.5-1.7 m that result in 2.3-2.5 m for the transport disengagement height (TDH).

4. RESULT AND DISCUSSIONS

According to the named theoretical issues and the present experiments, it is obvious that the quality of fluidization in fluid bed dryer will be the root of challenges in HDPE drying. Particle HDPE will reach a terminal velocity when the forces of gravity, buoyancy and drag are balanced. Empirical approach defines coarse particles as particles whose terminal velocities are greater than the superficial gas velocity ($U_T > U$) and fine particles as those for which $U_T < U$. One of the most important parameters to characterize fluidized bed conditions is the minimum fluidization velocity (U_{mf}), which quantifies the drag force needed to attain solid in the gas phase. U_{mf} increases with particle size and particle density and is affected by fluid properties. According to table 5, it is obvious that the minimum fluidization flow rate (FI-2) varies with particle size and particle density especially with bulk density of grades [31]. Pressure drop across the bed increases as the fluidizing gas velocity is increased. At a certain gas velocity, the bed is fluidized when the gas stream totally supports the weight of the whole bed. This state is known as minimum fluidization condition and the corresponding gas velocity is called minimum fluidization velocity, U_{mf}. Pressure drop across the bed remains nearly the same as the pressure drop at minimum fluidization even if the gas velocity is increased further. (Table5)

Particle size distribution of two types of catalysts (ZN-1,2) given in Fig.6. Based on Fig.6 and Table2, in the copolymerization process of HDPE, Ziegler-Natta catalyst dictate particle size distribution on powder product [5-7]. Therefore the type of catalyst affect the quality of fluidization in the experiment.

The bubbling fluidized bed (Fig.5) is divided vertically into two zones, namely, a dense phase and a freeboard region (lean phase or dispersed phase). The dense phase is located at the bottom; above the dense phase is the freeboard in which the solids holdup and density decrease with height. Fluidizing gas, after passing through the bed of particles, enters the freeboard region and carries with it fine particles with terminal velocities smaller than the operating gas velocity. This phenomenon is known as elutriation. Solids holdup in the freeboard region decreases as the freeboard height is increased until a height beyond which the solids holdup remains unchanged. This point is known as the transport disengagement height (TDH). TDH can be estimated from several empirical correlations; these correlations are expressed in terms of one or two operating parameters. Thus, the predictions are generally poor. However, there is no universally accepted equation for calculating TDH. As a practical result in our study, it is

best to determine the transport disengaging height experimentally. After determining TDH of 2.3-2.5 m, it is essential to control it for decreasing entrainment of particles.

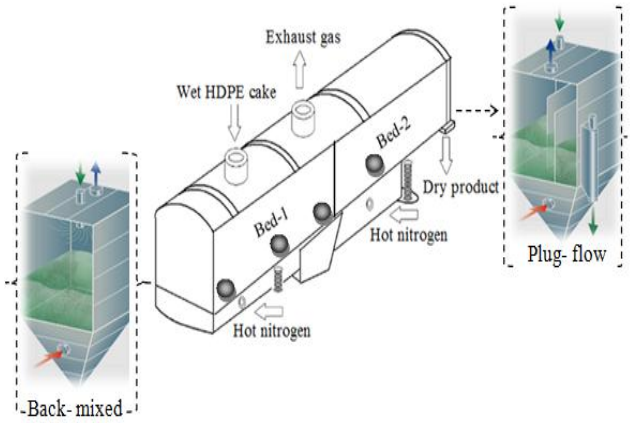


Fig.10. Flow pattern in HDPE fluid bed dryer

The drying system is designed in two stages (Fig.10). based on the shape of distributor plate in fig.9, back mixture mechanism and plug flow regime created at the first stage and the second stage respectively. In this situation, hot nitrogen and heating coils embedded in the dry layer of powder will play the drying role. As the bed of particles is perfectly mixed, the bed temperature is uniform and is equal to the product and exhaust gas temperatures (Table7). However, particle residence time distribution is necessarily wide, thus resulting in a wide range of product moisture content. On the other hand, as the feed material is continuously charged into the fluidized bed of relatively dry particles, this gives the added advantage of enhanced fluidizability and better fluidization quality. In the system, a well-mixed continuous pattern has been incorporated with plug-flow regime to give better drying performance. In plug-flow regime, vertical baffles are inserted to create a narrow particle flow path, thus giving relatively narrow particle residence time distribution. Particles flow continuously as a plug from the inlet toward the outlet through the path. This ensures nearly equal residence time for all particles irrespective of their size and ensures uniform product moisture content. Various paths can be designed such as straight or spiral paths. Operational problems might occur at the feed inlet because wet feedstock must be fluidized directly rather than when mixed with drier material as in the case of a well-mixed unit. To overcome the problem of fluidizability at the feed inlet, the inlet region may be agitated with an agitator, or by applying back mixing of solids, or by using a flash dryer to remove the surface moisture prior to plug-flow fluidized bed drying. In our experiment, back mixing way used for decreasing the surface moisture. The present volaties of 0.07 and 0.04% in table 6,7 show that, the first part of the moisture is separated in the first bed and another part is dried in the second bed of dryer. Reports on recent development and research findings of well-mixed fluidized bed and plug-flow FBDs can be obtained from several recent publications [31–34]. Based on fluidization quality, powders can be classified into four groups: group A (aeratable particles, easy to fluidize when dry), group B (sand-like particles, easy to fluidize when dry), group C (fine and ultrafine particles, difficult to fluidize due to dominated cohesive forces between particles), and group D (large and dense particles, poor fluidization quality due to formation of large bubbles in the bed) [29]. Fluidized bed dryers are normally operated in the regimes of smooth and bubbling fluidization. After passing through the fluidized bed, the gas stream is introduced into gas-cleaning systems to separate fine particles (dusts) from the exit gas stream before discharging it to the atmosphere. A typical fluidized bed drying system consists of a gas blower, gas-cleaning systems such as cyclone, bag filters, precipitator, and scrubber. To save energy, sometimes the exit gas is partially recycled. In the process, double cyclones in series form installed as a gas-cleaning system to separate fine particles from gas. (Fig.11) [35-39].

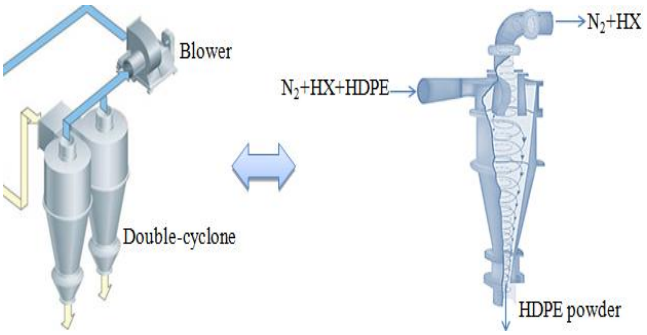


Fig.11. double cyclones as agas-cleaning system

HDPE powder is polydispersed with wide particle size distribution (Table2). It is important to take note about the occurrence of entrainment of fine particles. Solids in the fluidized bed can be classified into fine and coarse products based on the catalyst type. Particles that are elutriated by the fluidized gas stream are known as fine products, whereas particles retained in the bed are known as coarse products. For processes that require a certain degree of dedusting (removal of undesirable fine particles) or classification, operating gas velocity and location of gas exit should be chosen carefully in order to achieve the appropriate product cut size. Cut size refers to the critical size that separates the fine (elutriated) and coarse (remain in bed) particles. According to the present flow regime and particle size distribution, performance of the double cyclone will show the entrainment rate for fine particles. Based on table 6, PDI-3 and PDI-6 present the cyclone performance and the rate of carryover of particles. Drying with FI-2 of 14060 m^3hr^{-1} is proper to produce

the minimum elutriation of powder with optimal pressure drop across the two double-cyclones (CY-1, 2). Therefore, if we use improper nitrogen flow rate (FI-2), the quality of fluidization will be reduced and then the rate of entrainment of solids from a fluidized bed can be unsuitable and this case affect the electrical energy consumption of gas blowers (BL-1,2). By the use of presser drop resulted in table 6, poor fluidization is obvious and also this condition affect heat and mass transfer in fluidized bed drying because we discuss a volatile of 0.07% compared with volatile of 0.04% for final dry product in table 7. Pressure drops between two beds of dryer (PDI-7) represent the pneumatic transport of powder. Unfortunately, we can see PDI-7 of -0.20265 less than 0.101325 kpa, which show the negative effect of poor fluidization on powder transport in drying beds. If the process involves the emission of heat then this heat will not be dissipated as well from the packed bed as it was from the fluidized bed. All parts of the fluidized bed unit are subject to erosion by the solid particles. Heat transfer tubes within the bed or freeboard are particularly at risk and erosion here may lead to tube failure. Erosion of the distributor may lead to poor fluidization. Loss of fine solids from the bed reduces the quality of fluidization and reduces the area of contact between the solids and the gas in the process[13].

Gas distribution plates which are used in our test prove very satisfactory when handling a variety of HDPE grades with different humidities, physical and chemical properties. These plates are precisely selected for a required pressure drop, to ensure the fluid medium is distributed evenly to the solid material, but also to prevent the material from falling through the plate or clogging in the event of zero gas pressure beneath the plate[14,36].

The even distribution of the fluid medium is achieved by using a minimum of pressure under the gas distribution plate, but also dependent on the product characteristics and the dumping height in the installation. Specific material selection and finish also play an important part in avoiding adhesion. We detected that another important factor of the gas distribution plates is the gas velocity over the distribution plate. This has to be smaller than the sinking speed of the smallest particles in the fluidized product bed, otherwise too much of the product will be carried along by the gas flow, requiring recovery in a filter or cyclone.

5. CONCLUSION

It is obvious that improving drying process needs proper fluidization. Moreover, based upon theoretical and experimental results, on the way to reach this fluidization with good quality there are several industrial challenges. Theses challenges are physical specifications of different applied polymeric grades (PE100, BL3 and EX5) named PSD, density, operating parameters like wet cake rate, moisture content, gas temperature and medium bed temperature, the type of ZN- catalyst and hydrodynamic characteristics such as pressure drop, superficial gas flow rate, minimum fluidization velocity, bed height, gas holdup, TDH and gas distribution plate type. The whole of these issues create the hydrodynamic behavior and fluidization regimes affecting HDPE dry product with proper quality.

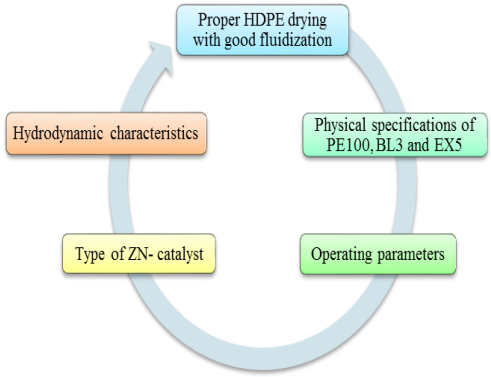


Fig.12. Schematic diagram of industrial challenges

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