Bubble Column Reactor Fluid-dynamic Evaluation at Pilot-plant for Residue and Extra-heavy Crude Oil Upgrading Technology

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Abstract: Bubble column reactors are multiphase contacting devices used in a wide variety of industrial application. Inrtevep S. A. is working on developing technologies to convert heavy and extra-heavy crude oil using this type of reactors. Volumetric gas hold up, flow pattern, average gas bubble size, average interfacial area, RTD (residence time distribution), dispersion coefficient, Péclet number are important design parameters for a proper scale up of them. Several cold model experiments have been proposed to determine the previously mentioned parameters at atmospheric conditions, using a plexiglas bubble column reactor at pilot plant scale unit (12 cm diameter). It was also evaluated our own design of internal trays (plates) in the reactor. Air-tap water and air-light oil systems have been used. A wide operating condition range was applied, superficial gas velocity between 0.5-10 cm/s, liquid flowrate between 15-65 l/h. Generally speaking, working without internal trays was found that gas hold up increase along the reactor and it was possible to identify heterogeneous bubble, transition and turbulent flow pattern areas for the air-light oil system. Average gas bubble size increase along the reactor at bubble regime from 2-5 mm but at turbulent regime, stay oscillating between 1-3 mm. Average interfacial area increases exponentially with superficial gas velocity at any reactor height, till 1,412 m²/m³ for the air-light oil system but, at bubble flow regime, the average interfacial area is lower than 100 m²/m³, which negatively impact the reactor performance. Internal trays in the reactor always increase gas hold up at any condition or system used. Residence time distributions curves, Péclet numbers and dispersion coefficients founded, show that this reactor with this kind of design internal trays still tends to be a complete mixing reactor under the operating conditions used.

Key words: Fluid dynamic, bubble column reactor, residue upgrading technology, trays.

1. Introduction

It is well known that bubble column reactor have a wide range of applications such as absorption, catalyst slurry reactor, bioreactions, coal liquefaction, etc. These reactors are preferred because the simplicity of its operation, easy construction and low operating costs. These main characteristics make bubble column reactors an excellent choice for extra-heavy oil upgrading technology as Aquaconversion[®].

Venezuela has extensive reserves of heavy and extra heavy crude oil in the eastern part of the country, including conventional production areas and the Orinoco Oil Belt. The Orinoco Oil Belt is divided in

four big areas named as Carabobo, Ayacucho, Junín and Boyacá, and has 297,7 MMMBLS of proved reserves of extra heavy crude oil 77 MKm² approx. The Orinoco Oil Belt crude has very high densities (5-9°API (American petroleum institute) gravity) which reduce his market value and a very high viscosity (10-40 Kcst at 50 °C) which impact the transportation cost. This crude also has high sulfur (2-5w%) and metal content (400-600 ppmw).

Production of these extra heavy crudes with gravities of 8-9° is currently done by dilution with light crude (24-30°) in the well to be lifted to the surface, where it goes through a process scheme including degasification, dehydration and desalting to be commercialized as a dilute crude of 13.5-16° API

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gravity and a viscosity less than 350 cst at 50 °C. The quantity of diluents could be around 38%-50%v of the total production, depending on the light crude oil used. One of the main constraints in the current production of extra heavy crude in eastern Venezuela is the availability of light or medium crudes to be used as diluents.

This situation will be even worse in the future due PDVSA aim to incorporate more heavy and extra heavy crude oil in its crude portfolio to reach production capacity of 6 million BPD (Barrel per day) by 2019. In order to overcome the constraints imposed by the potential shortage of light crudes required as diluent, PDVSA-Intevep has developed a new low-capital-cost technology for moderate conversion of residue and heavy or extra-heavy crude oil called Aquaconversion[®].

Aquaconversion[®] is a thermal catalytic steam cracking technology developed by PDVSA-Intevep to convert residue and heavy or extra heavy crude oil in the producing field into transportable upgraded crude, reducing the use of light/medium crude as diluents. The process is also utilized in petroleum refineries to increase conversion of vacuum residue without losing unconverted bottom-product stability [1].

In spite of the advantages, bubble column reactors used in a technology as Aquaconversion[®] are difficult to design and scale up, because of the flow regime complexity, and their unknown behavior under different sets of design parameters such a diameter, height, etc. Nonadjustable parameters like phase holdups, gas-liquid interfacial area, interfacial mass transfer coefficient, dispersion coefficient and heat transfer coefficient have a direct effect on the design of the bubble column reactor. Therefore, this study presents some experimental results to understand the fluid dynamic behavior of a bubble column at pilot plant scale, working under similar Aquaconversion[®] operating conditions of liquid and gas flow rates, but, using atmospheric conditions of pressure and temperature.

2. Experimental Sections

2.1 Experimental Set Up

A schematic diagram of the experimental apparatus is shown in Fig. 1. This unit was conformed by several sections. The liquid section had three tanks, one for recycling tests (gas hold up tests) another, for one through tests (dispersion coefficient test) and the last for disposal of liquid material. This section also contained the pumping system. The gas section capable of feeding air at required conditions, from the air pilot plant network into the unit has a pressure control valve, Brooks rotameter and an Endress Hauser mass flowrate meter used to accurately data measurement. In the reactor section was used a 12 cm diameter plexiglas column which is geometrically similar to the real Aquaconversion[®] reactor (see Fig. 2). The bubble column is a plexiglas tube with inner diameter of 12.3 cm, the cylindrical height of 160 cm, and a cone-shaped bottom also made of plexiglas without gas sparger. The special design cone-shape column top was made of metal and it was built as the real apparatus.

The column was equipped with a Rousemount 115 T DPT (differential pressure transmitter) which is connected to ABB Screen Master 200 videografic register to gather the column pressure drop. This unit had a pneumatic liquid tracer system which is used to inject both tracers used in this study, Thiophene and Potassium Dichromate. Those injections were used to determine the dispersion coefficient in the column. All tests were conducted at room conditions of temperature and pressure.

Two different systems were used in this study, air-tap water and air-light oil. The light oil was chosen because of some physical and chemical properties like, density, surface tension and viscosity, which has to be similar to the feedstock (vacuum residue) at operating conditions (180 psig of pressure and around 430 °C of temperature). These main liquid properties are shown in Table 1.

Fig. 1 Schematic diagram of the experimental apparatus.

Fig. 2 Plexiglas reactor.

The upgrading process use steam as gas phase, but air was used because of availability and economic aspects. Table 2 shows a comparison of the properties of the two gases. Akita [2] used water and four gases

of different densities, air, oxygen, carbon dioxide and helium in a column with 15×15 cm square cross section, showed that the effect of the gas density on the gas hold up could be neglected. Nevertheless, in very high pressure systems, in which the gas density can be affected in a higher proportion, the gas hold up, could increase significantly with the increase of the gas density [3].

A wide operating condition range was used, in order to reproduce the actual conditions of the pilot plan unit. Superficial gas phase velocities between 0.5-10 cm/s and liquid flowrate between 15-65 l/h (liquid phase velocity between 0.035-0.15 cm/s). Three different flowrates were used, 15, 35 and 65 l/h and for each one, a set of superficial gas velocity were tested from 1 cm/s to 10 cm/s with 1 cm/s steps. In some cases, it was necessary to operate at 0.5 cm/s of superficial gas velocity.

It was tested internal trays inside the column in order to define their impact on the gas hold up and liquid axial dispersion coefficient. The internal trays

Table 1 Liquid phase properties.

Fluid	Method	Viscosity (cst)	Density (Kg/m ³)	Surface tension (dyn/cm)
Vacuum residue	PROII	1.7	843	16.63
Light oil	Whihelmy plate	2-3.5 at 40 °C ASTM (D-445)	815 at 25 °C ASTM (D-1298)	27.36 at 25 °C (Lab. E&P)
Water	Perry, 1992	1.00	996	72.8

Table 2 Gas phase properties.

Fluid	Viscosity (cst)	Mol. weight	Density (Kg/m ³)	Surface tension (dyn/cm)
Steam	0.025	18	3.69	4.93
Air	0.019	29	1.18	1.18

consist on 10 perforated trays with 3 mm diameter holes. (Fig. 3). The number of holes change, at the bottom is around 232 holes and increases along the column till reach 264 holes at the top, in order to keep gas velocity constant in the column trough the holes because the gas formation inside the column when this internal tray is used inside the real reactor, following typical shell internal trays design used in visbreaking commercial reactor.

2.2 Methods

This part consists of the description of the procedure used to estimate gas hold-up, flow pattern, dispersion coefficient and average gas bubble size and average interfacial area.

2.2.1 Gas Hold Up

The unit was operating in recycling mode. Two different methodologies were used, the first one consist on determining the pressure drop in the column and the second methodology was based on phase trapping and measure both phases inside the column. The pressure drop is determined by using a DPT in the column. The differential pressure profile is deduced from a one-dimensional biphasic model, where friction losses and fluid acceleration are neglected. Then, the axial pressure profile in the column can be described by the Eq. (1). Finally, with Eqs. (1) and (2), it is possible to define a total gas and liquid hold ups in the section where the pressure drop is determined. This procedure was done in order to

Fig. 3 Internal trays tested.

determine the total and partial (top and bottom sections) hold up in the column. Both sections represent almost half column.

$$-\frac{dP}{dz} = (\varepsilon_L \rho_L + \varepsilon_G \rho_G) * g \quad (1)$$

where, dP is the pressure differential, dz is the height differential, ε_L and ε_G are the liquid and gas holds up, respectively and g is the gravitational constant.

$$\varepsilon_L + \varepsilon_G = 1 \quad (2)$$

The phases trapping methodology consists on stopping both liquid and gas flow rate sent into the column and closing the inlet and outlet reactor valves to measure the height of both phases in the cylindrical section of the column. Then, it is possible to determine total gas hold-up as a relation between the gas volume and the total column volume, as it is shown in Eq. (3):

$$\varepsilon_G = \frac{V_{c-top} + V_h}{V_T} \quad (3)$$

where, V_{c-top} , V_h , V_T are the volume of the cone at the top of the column, the obtained volume at specific height and the total column volume, respectively.

2.2.2 Flow Regime

A pneumatic system was used to inject into the reactor a pulse of a liquid tracer. Thiophene and Potassium Dichromate were used as tracer for air-light oil and air-tap water systems, respectively. When the pulse is injected, the systems stop working in recycle mode and the discharge is sent to the disposal liquid tank. Reactor outlet is sampled along the time in order to quantify tracer concentration; sulfur concentration (Thiophene) by WD-FRX using ASTM-D-2620-10 norm and Potassium concentration (Potassium Dichromate) by ICP (induce coupled plasma). Injections were used to determine the RTD (residence time distribution curve) for all the cases. Levenspiel [4] defined a correlation between Péclet number and the RTD curve, as follows:

$$\frac{\sigma^2}{\bar{t}^2} = \frac{2}{Pe} + \frac{8}{Pe^2} \quad (4)$$

where, Pe is the Péclet number, σ is the variance and t is the average time.

Average time can be determined using the next Eq. (5) [4]:

$$\bar{t} = \frac{\sum_{i=1}^n t_i C_i \Delta t_i}{\sum_{i=1}^n C_i \Delta t_i} \quad (5)$$

where, C_i is concentration of the component, Δt is the time different between samples, and t_i is time.

In the same way, variance can be determined by the following Eq. (6) [5]:

$$\sigma^2 = \frac{\sum_{i=1}^n (t_i - \bar{t})^2 C_i \Delta t_i}{\sum_{i=1}^n C_i \Delta t_i} \quad (6)$$

Finally, dispersion coefficient can be obtained by

the correlation commonly used in reactor design between the axial Péclet number based on the liquid superficial velocity and the dispersion coefficient [6]:

$$Pe = \frac{QL}{A_R D} = \frac{uL}{D} \quad (7)$$

where, u is the liquid linear velocity, L is the column length tangent-tangent, D is the dispersion coefficient.

The previous procedure was applied at different operating conditions: liquid flow rate of 35 l/h and three different superficial gas phase velocities 0.5, 3 and 6 cm/s.

2.2.3 Average Gas Bubble Size and Average Interfacial Area

It was used the DSD (dynamic gas disengagement) methodology to determine the average gas bubble size distribution [7]. This method is based on determining the average gas bubble size distribution when the gas bubble is disengaging of the gas-liquid bed inside the column. This measurement can be linked with the rising bubble velocity of different groups of bubble sizes. These velocities could be used to determine the average bubble size of each group of bubbles and then it is possible to estimate the Sauter diameter and the average interfacial area.

First, it is necessary to plot the bed height level as function of time when the gas inlet is closed by the following Eq. (8):

$$\frac{H_{(t)}}{H_0} = \frac{\varepsilon_G (V_{c-inf} + A H_s)}{H_0 (100 - \varepsilon_G) A} + \frac{H_s}{H_0} \quad (8)$$

where, H_0 is the bed level before closing the gas inlet, $H_{(t)}$ is the bed level any time, H_s is the bed level at the end, A is the columns transversal area, and V_{c-inf} is the inlet cone volume.

According to the shape of the curve obtained, the gas bubble sizes have to be defined as bi, trio, poly modal different sizes of groups of bubbles: small, medium and large group of bubble. For instance, each group of bubbles should be defined as a section of the curve obtained and represented by a straight line. The slope (S_i) and the y-axis cut point (b_i) should be

determined (Fig. 4).

Then, gas bubble hold up for each individual group of bubbles could be determined for as follows:

$$\varepsilon_{Gi} = \frac{H_s}{H_0} \left[\frac{1}{b_{i+1}} - \frac{1}{b_i} \right] \quad (9)$$

where, b_i is the y-axis cut point of each group of bubble.

And the following equations could be used to estimate the gas bubble rising velocities:

$$u_{bi} = \frac{H_0 [b_i S_{i+1} - b_{i+1} S_i]}{b_i - b_{i+1}} \quad (10)$$

where, S_i is the curve slope of each group of bubble.

For small bubble, which is normally used the terminal rising velocity (u_{bs}) based on a single bubble in an infinite medium determined by Marrucci equation [7]:

$$u_{bs} = \frac{u_{bs} \left[1 - \varepsilon_{GS}^{5/3} \right]}{(1 - \varepsilon_{GS})^2} \quad (11)$$

where, u_{bs} is the small bubble velocity.

Then, average gas bubble size diameter for each group of bubbles could be determined following Stokes's law, or Peebles and Garber, Clift or Abou-el-Hassan equations as it is shown in Table 3.

Finally, to estimate the Sauter diameter (d_s) of the bubble and the average interfacial area (a_s) can be obtained by the next two Eqs. (12) and (13) [7]:

$$d_s = \frac{\sum_{i=1}^n n_i d_b^3}{\sum_{i=1}^n n_i d_b^2} = \frac{\varepsilon_G}{\sum_{i=1}^N \varepsilon_{Gi} / d_{bi}} \quad (12)$$

$$a_s = \frac{6\varepsilon_G}{d_s} = 6 \sum_{i=1}^N \frac{\varepsilon_{Gi}}{d_{bi}} \quad (13)$$

where, ρ_L and ρ_G are liquid and gas density, μ_L is the liquid viscosity and the σ_L is the liquid surface tension.

3. Results and Discussion

To ensure system stability and quality of the results,

Fig. 4 Gas disengagement profile on time.

Table 3 Bubble size diameter equations.

Reference	Correlation
Stokes's law	$d_{bs} = \left[\frac{18\mu_L u_B}{g(\rho_L - \rho_G)} \right]^{0.5}$
Peebles and Garber	$d_{bs} = 4.76 \left[\frac{\mu_L}{\rho_L} \right]^{0.41} \frac{u_B^{0.78}}{g^{0.59}}$
Clift	$u_{bs} = \left[\frac{2.14\sigma_L}{\rho_L d_B} + 0.505 g d_B \right]^{0.5}$
Abou-el-Hassan	$V = 0.75 [\log(F)]^{0.5}$ $V = \frac{u_B (d_B \rho_L)^{2/3}}{(\mu_L \sigma_L)^{1/3}}$ $F = \frac{g d_B^{8/3} (\rho_L - \rho_G) \rho_L^{2/3}}{\mu_L^{4/3} \sigma_L^{1/3}}$

column pressure drop was monitored and registered for all sets of experiments done here. For all the experiments conducted, the transient period of the system was less than 5 min. It was reached operational stability after that time.

3.1 Gas Hold Up Profile and Flow Patterns

Fig. 5 is a plot of the experimental gas hold up and gas velocity in the bubble column reactor for the air-light oil system at 35 l/h of liquid flow rate at different superficial gas velocities (line with solid triangles). Gas hold up increases when superficial gas velocity increases, as expected, following a potential shape curve ($\varepsilon = aUg^N$ type) described by Shah [8].

Fig. 5 Gas hold up for air-light oil system at 35 l/h of liquid flow rate.

The equation found for the air-light oil system was $\epsilon = 0.0821U_g^{0.5493}$. According to Shah, if the exponential number is lower than 0.7, in this case, 0.5493, then the flow pattern tend to be churn turbulent.

Empty triangles line represents the calculated total gas hold up obtained using the pressure drop methodology based on both, top and bottom gas hold up by partial pressure drops, which was exactly the same value obtained when the total gas hold up was determined by the pressure drop methodology for the total column. All the results showed in this work are based on pressure drop methodology used. But, the difference between the phases trapping and the pressure drop methodologies was only of 2.9% on average [9].

Top and bottom sections gas hold up are shown at the Fig. 5 on square and circle lines, respectively. It can be seen that gas hold up at the top of the column is almost always higher than the gas hold up at the bottom. The differential between both curves increase when the superficial gas velocity is higher than 2 cm/s and it is even higher after 4 cm/s.

Only at superficial gas velocity of 0.5 cm/s, the gas hold up at the top and bottom are quite de same. This result confirm what was observed by Sardella [9] where at 0.5 cm/s of superficial gas velocity using the same condition and system it was defined the flow

pattern as heterogeneous bubble flow, where different size bubbles rise along the reactor almost vertically without changing significantly bubbles volume or with a bubble size increasing to the pressure drop on the column.

After 2 cm/s of superficial gas velocity, the gas hold up at the top of the column is clearly higher than the one at the bottom. As is mentioned by Sardella [9], after 2 cm/s of superficial gas velocity, the visual flow pattern can be clearly defined as churn turbulent because it can be seen different size bubbles which change the volume along the column, with random movements, back flow and even axial swirl, with the big “jet” type bubble at bottom of the column which disappear along the column breaking down in small one.

The results obtained in this paper support the previous explanation because, big bubbles rise faster than small one, and faster the bubble lower the resident time in the column decreasing the gas hold up. Billet [10] found similar results.

After 4 cm/s of superficial gas velocity the different between gas hold up at the top and bottom increase. This can be explained based on Sardella [9] where it was identify foam formation at the top of the column and due to the increasing of gas velocity, the big type

“jet” bubble at the bottom of the column increase.

These two phenomena could explain the results obtained in this paper. Pino and Guitian found similar results with foam formation [11, 12].

3.2 Average Gas Bubble Size and Average Interfacial Area

Fig. 6 shows the bed height level as function of time when the gas inlet valve to the column is closed. This result was obtained working at 6 cm/s of superficial gas velocity and with an initial liquid level of 155 cm in the column (total column). Similar behavior was obtained for the rest of the cases evaluated. As it was mentioned at the methodology section, it was necessary to split the curve in different lineal segments and categorize it as big, medium (if apply) and small bubbles.

During the test, it was observed that bigger the bubbles faster it disengages from the liquid. Finally the bed height level remains constant and equal to the beginning.

Fig. 7 shows the average gas bubble size using air-light oil system at three different heights of the column with 4 different superficial gas velocities: 0.5, 1.5, 3 and 6 cm/s.

First, it is important to point out that the average

gas bubble size decreases when the superficial gas velocity increases, at any column level. Similar results were obtained by Daly [7], using a 0.23 cm diameter column working with similar gas-liquid systems. These results confirm also that the bubble breaking phenomena is governing along the column when the gas flow rate is increasing, which is also more obvious because the “jet” type big bubble formation at bottom of the column. This is a typical phenomenon in a church turbulent flow regime.

Morsi [13] obtained an average bubble diameter of 1.5-3.5 mm, Shah [8] obtained 1-2 mm and Daly [7] obtained bubble diameter of 2-5 mm. Those values are similar to the ones obtained in this work.

As it was mentioned before, at 0.5 cm/s of superficial gas velocity, the average bubble size tend to increase along the column due to the pressure drop confirming the heterogeneous bubble flow pattern due to the coalescence phenomena. In spite of the average bubble diameter also increase along the reactor at 1.5 cm/s, typical of heterogeneous bubble flow pattern, according with Sardella [9] this condition was defined as transition zone which make very difficult to understand. After 3 cm/s, at the upper size of the column, it can be observed a breaking-coalescence bubble phenomena and the bubble diameter try to keep almost constants,

Fig. 6 Typical gas disengagement profile on time air-light oil system at 6 cm/s and 155 cm of height.

Fig. 7 Sauter diameter for air-light oil system at different superficial gas velocity.

increasing and decreasing just a little, which is a typical behavior of a churn turbulent flow pattern. At 155 cm of liquid height level, it can be seen also that the average bubblediameter tend to reach a constant value, after 3 cm/s, which also can explain the tendency to reach minimum constant value in the gas hold up at the top of the column shown in the Fig. 5, because it reached a dynamic equilibrium between breaking and coalescence phenomena. Those results also confirm the flow patterns defined by Sardella [9] and the gas hold up profile already shown in this paper.

As it is mentioned on Soong [14] Akita y Yoshida proposed an equation to estimate average bubble diameter d_{32} , to be used on column till 30 cm of diameter, and superficial gas velocity lower than 7 cm/s, based also on physical-chemical properties and column diameter as it is following shown:

$$d_{32} = 26D_c(Bo)^{-0.5}(Ga)^{-0.12}(Fr)^{-0.12} \quad (13)$$

The Fig. 8 shows that, in spite of the value are not exactly the same, the tendency of the average bubble diameter (d_{32}) as a function of the superficial gas velocity are very similar, and the Akita and Yoshida equation could be used to estimate the average bubble diameter in this column working with this systems.

The comparison was made using the average bubble diameter experimentally obtained at 155 cm of liquid height in the column, because it almost represents the whole column scenario. Fig. 8 also shows that the average bubble diameter tends to reach a minimum constant value at higher superficial gas velocities as it was commented previously.

The gas-liquid interfacial area is a very important variable which is impacted by the column geometry, the operating conditions and liquid properties. The gas-liquid interfacial area is very important in this kind of reactor because it could impact the mass transfer. Fig. 9 shows the gas-liquid interfacial area at different superficial gas velocities working a liquid flow rate of 35 l/h at several liquid levels in the column.

Due to the fact the gas-liquid interfacial area depend on the gas hold up and the average bubble diameter as it is shown in the Eqs. (12) and (13), higher the superficial gas velocity, higher the gas hold up and lower the average bubble diameter, impacting almost exponentially the gas-liquid interfacial area as it is shown in the Fig. 9 at any liquid level. Previous works have found similar gas-liquid interfacial area (900-1200 m²/m³) working with inorganic liquid and with superficial gas velocities between 3-14 cm/s [15].

Fig. 8 Bubble diameter comparison with Akita and Yoshida correlation.

Fig. 9 Interfacial area for air-light oil system.

Working at heterogeneous bubble flow pattern at 1.5 cm/s, the gas-liquid interfacial area obtained is very low (lower than $150 \text{ m}^2/\text{m}^3$), which could negatively impact the mass-transfer coefficient (Kla), with a potential decreasing of the vacuum residue conversion at the real reactor, dropping the reactor performance. Working at church turbulent flow pattern, at 3 cm/s or 6 cm/s, the situation change dramatically and the gas-liquid interfacial area increase till the expected value for this kind of moderate hidroconversion reactor. As reference,

according to Trambouze [16], the gas-liquid interfacial area for moderate hidrotreatment process should be around $200\text{-}1,000 \text{ m}^2/\text{m}^3$.

Fig. 10 shows the comparison between the air-tap water and the air-light oil systems, working at 155 of liquid level and at three different superficial gas velocities. As it can be seen, the average bubble diameter of the air-tap water system is always higher (at least 4 times) than the air-light oil system.

The tendency was very similar to the one found with light oil. The average bubble diameter is higher

Fig. 10 Air-tap water and air-light oil systems Sauter diameter comparison.

at bubble flow pattern than the one found at turbulent flow pattern. This result can be explained because as it is shown in Table 1, light oil has lower surface tension than the water. According to Lemoine [15] and Daly [7], when liquid surface tension decreases, the system become less coalescing and the gas hold up increase. This was observed by Sardella [9]. The lower the liquid surface tension, the lower the bubble size, increasing the interfacial area and making the system easier to transport. Smaller bubble produce larger gas holds up due to their low rise velocities. Kelkar and Daly found similar results [7, 17].

Daly [7] found average bubble diameter between 2-5 mm for tap water systems, but Lemoine [15] found values around 5-9 mm for the same system in a column of 7.8 cm diameter on church turbulent flow regime, which is more similar to the values obtained in this work.

3.3 Reactor Trays Evaluation

3.3.1 Reactor Trays Evaluation

Fig. 11 shows the gas hold up with tap water-air and light oil-air systems at a liquid flow rate of 35 l/h at different superficial gas velocities with and without trays inside the column. As it can be seen, trays inside the column increase clearly the gas hold up in both systems tested. Trays inside the column make the

system less coalescing, breaking more the bubbles, creating small bubbles. As it was mentioned before, smaller bubble produce larger gas holds up due to their low rise velocities. Kelkar and Daly found similar results [7, 17].

Alvaré [18] found that trays inside the column also reduce the liquid recirculation which then, reduces its hold up.

3.3.2 Flow Regime

Fig. 12 shows the obtained Residence Time Distribution curves for the air-light oil system working at 35 l/h of liquid flow rate and three different superficial gas velocities (0.5, 3 and 6 cm/s) with internal trays inside the column. In order to compare all the systems it was necessary to normalize the RTD curves, constructing which is normally called E curves. It can be seen that all curves follow the same shape. They have a peak at the beginning and then fall progressively. These curves are typical for well mixed flow regime systems. The higher the superficial gas velocities, the higher the tracer concentrations, because the liquid hold up is lower. The difference on time between all the peaks is very small, but, it is possible to identify that at higher gas velocity, the peak is found faster, as expected.

Table 4 shows dispersion coefficients and Péclet numbers found for air-light oil system at 35 l/h of

Fig. 11 Gas hold up with and without trays.

Fig. 12 E curves for air-light oil systems at 35 l/h of liquid flow rate with trays.

Table 4 Flow regime parameter for air-light oil system at 35 l/h of liquid flow rate.

Superficial gas velocity (cm/s)	Mean time (s)	Dispersion coefficient (cm ² /s)	Péclet number (adim)
0.5	1,970	12.6	6.42
3	1,476	111.6	4.30
6	1,130	304.8	3.15

liquid flow rate and three different superficial gas velocities. It can be seen that for all the cases, similar and small Péclet number values and similar and high dispersion coefficient values. Based on the Péclet numbers and the RTD curves, it is possible to define

that this reactor at 35 l/h of liquid flow rate and any gas velocities tends to be a well mixed flow reactor. Carberry [6]. Similar results were found for the air-tap water system.

Fig. 13 shows a comparison with and without

internal trays inside the column working at 35 l/h of liquid flow rate and 3 cm/s of superficial gas velocity, using Sardella [9] results. Both E curves have the same behavior, but due to the higher gas hold up on obtained with the internal trays, tracer concentration increases inside the column and the peak value in this system is higher than the value found for the column with internal. When it is used internal trays inside the column the peak of the tracer concentration appears later on time, almost the double of the time due to the increasing on the number of the mixing stages creating by the trays.

It was found a Péclet number of 4.30 is the same to the one obtained without internal trays Sardella [9], reason why this system also can be defined as a complete mixing reactor.

Another way to see the results is trough the F curves (Fig. 13), which represent the quantity of tracer removed as function of time. For the system with internal trays, the quantity of tracer removed out the reactor at the average residence time (2,119 s) is between 84%, around 10% more than the quantity of material obtained without trays, but in both cases the tracer remain inside the column until 8000 s (2 h and 13 min), almost four times the average residence time

defined for the liquid flow rate. This result indicate that the tray inside the column could improve products selectivity, because, less material remain in the reactor after at the desired residence time. At operating condition with the vacuum residue then, this situation could decrease coke formation.

The results obtained with the internal trays shows that dead volumes are obtained by mean of the internal trays, displacing the peak of the curve on time, without changing the flow regime, as it is mentioned on Fogler [19].

Alvaré [18] explain that the total free area of the holes should be lower than 10% of the total area of the tray in order to obtain an important change in the flow regime. Trays tested in this work have 14-16% of free area, which is higher the value mentioned at Alvaré [18], and it could be the reasons of the results obtained in this work. Dead zones are very common on fixed bed reactor, nevertheless, trays can also create dead zone just over it. Higher the superficial gas velocity, higher the backflow mixing between each pair of trays, reducing potential dead zone, making the RTD more similar to the one obtained without internal trays. Dead zone are volume with negligible velocities which could promote coke formation.

Fig. 13 E and F curves for air-light oil systems at 35 l/h of liquid flow rate with and without trays.

4. Conclusions

Without internal trays inside the column was found that gas hold up increase along the reactor. It was possible to identify heterogeneous bubble, transition and turbulent flow pattern areas at different superficial gas velocities for the air-light oil system.

It was also determine that average gas bubble size increase along the reactor at bubble regime from 2-5 mm but at turbulent regime, stay oscillating between 1-3 mm.

At the same reactor height level, average gas bubble size decreases with the superficial gas velocity.

Average interfacial area increases exponentially with gas superficial velocity at any reactor height, till $1,412 \text{ m}^2/\text{m}^3$ for the air-light oil system but, at bubble flow regime, the average interfacial area is lower than $100 \text{ m}^2/\text{m}^3$, that perhaps negatively impact the reactor performance.

Internal plates in the reactor always increase global gas hold up at any condition or system used. Residence time distributions curves, Péclet numbers and dispersion coefficients founded, show that this reactor with internal plates used here still tends to be a complete mixing reactor under the operating conditions used. The new RTD curve was displaced on time because the death space created by the internal inside the column

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