

EXPERIMENTAL DYNAMIC EVALUATION OF THREE-PHASE REACTORS

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Abstract - The hydrodynamic parameters were determined for two different three-phase systems (fluidized-bed and trickle-bed) through the experimental evaluation of the dynamic residence time distribution in the gas or liquid phase. For different fluid phase flow rates the fluidized-bed system was operated with a tracer in the gas phase and in the trickle-bed with a tracer in the liquid phase. The analysis of the processes was achieved through the application of different models representing the dynamic behavior of tracers present in the gas and liquid phases with a simulation of experimental operations of the three-phase reactors. The transfer functions developed for the two systems made it possible to estimate the hydrodynamic parameters providing the gas holdup (2.22×10^{-2} to 8.42×10^{-2}) and the gas phase Peclet number (54.18 to 41.20) for the fluidized-bed and the liquid holdup (0.16 to 0.25), the liquid phase Peclet number (20.37 to 4.52) and wetting efficiency (0.34 to 0.56) for the trickle-bed.

Keywords: Dynamic evaluation; Fluidized-bed; Trickle-bed; Residence time distribution; Tracers.

INTRODUCTION

Three-phase reactions are processed in three main types of reactors, namely: slurry, fluidized-bed and trickle-bed reactors. Slurry reactors are used in non-continuous operations for relatively small-scale productions while fluidized-bed and cocurrent gas-liquid downflow trickle-bed reactors have found increasing use in industrial processes. These two reactor technologies are often used in petrochemical processes and petroleum industries (Korsten and Hoffman, 1996; Rajashekar et al., 1998; Pawelec et al., 2001). Three-phase fluidized (FBR) systems are finding industrial applications in reactive processing schemes of contacting gas, liquid and slurry (Fan, 1989; Deckwer, 1992). They find

widespread use in industry, for example in the Fischer-Tropsch synthesis of hydrocarbons from synthesis gas (Krishna et al., 1996). Trickle-bed reactors (TBR) are used extensively for hydrotreating and hydrodesulfurization applications in the refining industry, and hydrogenation, oxidation and hydrodenitrogenation applications in the chemical, biochemical and waste treatment industries (Al-Dahhan et al., 1997; Saroha and Nigan, 1996). Medium or small-scale treatments of raw materials with potential valorization can justify the application of both fluidized and trickle-bed reactors with continuous productions at smaller processing units. Hydrogenation and oxidation of soluble carbohydrates (Besson et al., 1995; Abbadi et al., 1997; Barbosa et al., 1999) are examples of

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catalytic reactions usually carried out in batch slurry reactors that may be transposed to continuous processes.

The performance of continuous three-phase reactors may be affected by many physical factors, such as interphase mass transfer, intraparticle diffusion, gas and liquid holdup, axial dispersion and incomplete catalyst wetting.

Results of Zahradnik and Fialová (1996) proved the essential effect of gas dispersion mode on the extent of gas and liquid axial mixing in bubble column reactors. The gas phase RTD curves justified the plug flow approximation for the homogeneous bubbling regime. The comparison of these data with the corresponding dependences of the gas holdup and volumetric gas-liquid mass transfer coefficient on the superficial gas velocity revealed the existence of direct links between the variation of RTD and the respective mass transfer characteristics induced by the changes of gas dispersion mode.

Iliuta et al. (1999) proposed a RTD model to describe the liquid trickle flow in a trickle-bed reactor packed with porous particles and operated under partially wetted conditions based on a simple representation of the liquid flow structure. Analysis of the dynamic tracer impulse-response data of the liquid revealed the significance of the mass transfer resistance between static liquid and adjacent wetted particles, intraparticle diffusion resistance, as well as partial wetting. In trickle flow, high liquid axial dispersion coefficients were obtained for low liquid velocities.

The effects related with the establishment of the fluidized pattern in the FBR are a function of the fluid phase flow, which is related to the minimum gas velocity to fluidize the catalysts in the liquid phase. In TBR, operations with high gas phase retention are associated with low liquid throughput and partially wetting conditions of the particles. The present work proposes a same methodology based on the importance on the determining effects of the fluid phases of the each system, to describe the physical behavior of the fluidized-bed and trickle-bed three-phase reactors, using three-phase process parameters obtained via dynamic analysis. The estimations resulting of the application were utilized for the development of wet oxidation and hydrogenation processes in the three phase fluidized bed (FBR) and trickle bed reactors (TBR) to produce fine chemicals continuously (Mathias and Rodrigues, 1994; Sales, 2001; Gallezot et al., 1998; Maranhão, 2001). The hydrodynamic results obtained in the fluidized-bed for the gas phase was compared with the data from Zahradnik and Fialová (1996),

determined at similar operation conditions and set-up. In the same way, for the trickle-bed system, Iliuta et al. (1999) served as a reference to compare their data with the results obtained in this study for the liquid phase.

DEVELOPMENT OF THE MATHEMATICAL MODELS

Modeling of the Gas Phase Dynamic Behavior in the Three-Phase Fluidized-Bed Reactor

To evaluate the behavior of the gas tracer in the three-phase fluidized-bed reactor, a two-zone model for the gas phase was proposed, based on the operational behavior of the system. The dynamic model assumed that the total gaseous volume V_G inside the reactor was distributed in two zones, represented by a tubular reactor of volume V_{G1} associated in series with a continuous stirred tank reactor of volume V_{G2} . The volumes of the two zones are related as $V_{G1}=aV_G$, $V_{G2}=(1-a)V_G$ and the gas holdup is expressed as $h_G=aV_G(V_L+aV_G)^{-1}$, where a is the volume fraction of gas inside the zone of high concentration of solid particles.

In zone 1 (V_{G1}) the gas flows in the liquid phase and a homogeneous axial dispersion model is proposed for the gaseous tracer, as follows:

$$\left(\frac{\partial C_{GT}}{\partial t} \right)_{z=L} = D_{ax,G} \frac{\partial^2 C_{GT}}{\partial z^2} - u_G \frac{\partial C_{GT}}{\partial z} \quad (1)$$

where C_{GT} , $D_{ax,G}$ and L are the gas tracer concentration through liquid phase (zone 1), the axial dispersion coefficient of the tracer and the length of the tubular reactor respectively, and u_G is the gas phase superficial velocity.

For zone 2 (V_{G2}) a CSTR behavior is considered, where the mass balance for the gas tracer is:

$$\frac{V_G(1-a)}{Q_G} \frac{dC_{GT_E}}{dt} = C_{GT_{z=L}} - C_{GT_E} \quad (2)$$

where C_{GT_E} is the zone 2 exit gas tracer concentration.

The initial and boundary conditions for the system of equations are (Danckwerts, 1953; Barber et al., 1998):

$$t = 0 \quad C_{GT} = C_G = 0 \quad (3)$$

$$u_G \left(C_{GT_I} - C_{GT_{z=0^+}} \right) + D_{ax,G} \left(\frac{\partial C_{GT}}{\partial z} \right)_{z=0^+} = 0 \quad (4)$$

$$\frac{V_G}{Q_G} (1-a) \frac{dC_{GT_E}}{dt} = - \frac{D_{ax,G}}{u_G} \left(\frac{\partial C_G}{\partial z} \right)_{z=L} \quad (5)$$

$$G_{FBR}(s) = \frac{2\beta \exp[M(1-\beta)]}{(1-a)s? \left[(1+\beta) - (1-\beta)e^{-2M\beta} \right] + \frac{1}{2} \left[(1+\beta)^2 - (1-\beta)^2 e^{-2M\beta} \right]} \quad (6)$$

where $\theta = Z/u_G$; $\beta = (1+2\theta a s M^{-1})^{1/2}$ and $M = (u_G Z / 2D_{ax,G})$.

Modeling of the Liquid Phase Dynamic Behavior in the Trickle-Bed Reactor

For the non-reactive trickle-bed operation a heterogeneous model with axial dispersion (HAD) was suggested. The mass balance for the tracer in the liquid phase flow is:

$$D_{ax,L} \frac{\partial^2 C_L}{\partial z^2} = u_L \frac{\partial C_L}{\partial z} + h_L \frac{\partial C_L}{\partial t} + \frac{3(1-e)}{R} D_{ef,L} \left. \frac{\partial C_{Lin}}{\partial r} \right|_R f_e \quad (7)$$

where C_L , C_{Lin} and $D_{ax,L}$ are the external and internal liquid phase concentrations and the axial dispersion coefficient of the tracer, respectively; u_L is the liquid phase superficial velocity, h_L the liquid holdup, $D_{ef,L}$ the internal effective diffusivity and f_e the wetting efficiency.

For the tracer in the intraparticle liquid phase, considering a spherical catalyst particle and linear kinetics of liquid-solid adsorption, the mass balances are written as:

$$\frac{\rho_b}{(1-e)} \frac{\partial q_L}{\partial t} + e_{in} \frac{\partial C_{Lin}}{\partial t} - D_{ef,L} \frac{1}{r^2} \frac{\partial}{\partial r} \left(r^2 \frac{\partial C_{Lin}}{\partial r} \right) = 0 \quad (8)$$

$$G_{TBR}(s) = \frac{1}{\left[1 - \left(\frac{u_L - \sqrt{u_L^2 + 4D_{ax,L}f(s)}}{2u_L} \right) \right]} \exp \left(\frac{u_L - \sqrt{u_L^2 + 4D_{ax,L}f(s)}}{2D_{ax,L}} z \right) \quad (13)$$

where C_{GT_I} is the gas tracer concentration inlet of tubular reactor.

The solution of the system of equations for the gas phase, in the Laplace domain, can be described as the transfer function $G_{FBR}(s)$ (Equation 6) of the dynamic process of non-reactive runs for the tracer in the fluidized-bed reactor.

$$\frac{\partial q_L}{\partial t} = k_{ad} \left(C_{Lin} - \frac{q_L}{K_{ad}} \right) \quad (9)$$

where q_L is the tracer concentration on the solid surface, ρ_b the bed density, ϵ the bed porosity, ϵ_{in} the catalyst internal porosity, k_{ad} the adsorption kinetic constant and K_{ad} the adsorption equilibrium constant.

The initial and boundary conditions are (Danckwerts, 1953; Barber et al., 1998):

$$t = 0; C_{Lin} = C_L = q_L = 0 \quad (10)$$

$$z = 0; C_{L0^-} = C_{L0^+} - \frac{D_{ax,L}}{u_L} \frac{\partial C_L}{\partial z} \quad (11)$$

$$z = Z; \frac{\partial C_L}{\partial z} = 0$$

$$r = 0; \frac{\partial C_{Lin}}{\partial r} = 0 \quad r = R; \quad (12)$$

$$D_{ef,L} \left. \frac{\partial C_{Lin}}{\partial r} \right|_R = k_{t,LS} (C_L - C_{Lin}|_R)$$

where $k_{t,LS}$ is the liquid-solid mass transfer coefficient of the tracer.

Based on the solution of the system of equations (Equations 7, 8 and 9) the transfer function $G_{TBR}(s)$ of the dynamic process of non-reactive runs for the liquid tracer in the trickle-bed reactor is expressed, as follows:

where,

$$f(s) = \frac{3(1-e)}{R} k_{t,LS} f_e \left[1 - \frac{1}{\frac{D_{ef,L}}{k_{t,LS}} \delta(s) \coth[\delta(s)R] + \left(1 - \frac{D_{ef,L}}{k_{t,LS}R}\right)} \right] \quad (14)$$

$$\delta(s) = \sqrt{\frac{s}{D_{ef,L}} \left(\frac{\beta_b}{1-e} K_{ad} + e_{in} \right)} \quad (15)$$

EXPERIMENTAL

For the fluidized-bed reactor a gas tracer was used while for the trickle-bed reactor an organic salt solution was used as a tracer for the liquid phase. At non-reactive conditions, RTD experiments in the fluidized and fixed-bed systems with gas and liquid tracers, respectively, were performed at three different flow rates. The non-reactive runs carried out for both flow systems were described by models of the dynamic behavior of the tracers present in the gas and liquid phases.

Fluidized-Bed Reactor

For the experiments to establish the dynamic RTD of the fluidized-bed the solid phase was γ - Al_2O_3 catalyst ($S_p=3.65 \times 10^5 \text{ m}^2/\text{kg}$, $\epsilon_{in}=0.59$, $dp=400\text{-}500 \mu\text{m}$, $\rho_{app}=7.84 \times 10^2 \text{ kg/m}^3$, $\rho_{abs}=2.41 \times 10^3 \text{ kg/m}^3$, Procatalyse). The tracer used in the gas phase (air) was methane (ONU 1971, 4.0, 99.95% purity, White Martins).

The fluidized bed reactor (FBR) was constructed in such a way that after the fluidizing zone (Zone 1) the liquid phase overflows the walls of the reactor into a chute, liberating the gas that flows upwards through a chamber (Zone 2) when it is discharged to the surroundings. The RTD experiments in the fluidized-bed reactor (1.40m in height and 0.30m in diameter) were performed under non-reactive conditions, the liquid flowed cocurrently upwards, with the air bubbling through it and keeping the γ -alumina catalyst suspended (Figure 1). Methane was fed as gas tracer (3.0% v.v.). Under isobaric (101.32kPa) and isothermal (300K) conditions the gas flow rate varied within a range from $0.50 \text{ m}^3 \text{ h}^{-1}$ to $1.50 \text{ m}^3 \text{ h}^{-1}$ while the liquid flow rate was fixed in 5.00 Lh^{-1} or 10.00 Lh^{-1} in each experiment in contact with 80.0g of particulate solid. A flame ionization detector (FID) connected to the reactor inlet or outlet analysed the methane concentration in a continuous way. The suppression of the methane in the air feed stream resulted in negative step concentration curves.

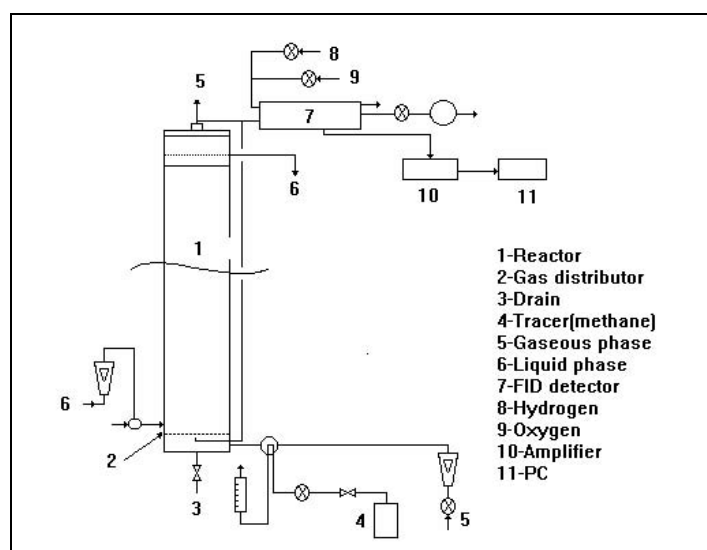


Figure 1: Experimental dynamic set-up of the three-phase fluidized-bed reactor

Trickle-Bed Reactor

In the dynamic RTD at trickle-bed experiments the solid phase utilized was the catalyst: activated carbon ($S_p=6.32 \times 10^5 \text{ m}^2/\text{kg}$, $\epsilon_{in}=0.33$, $d_p=4.00\text{--}6.00\text{ mm}$, $\rho_{app}=1.10 \times 10^3 \text{ kg/m}^3$, $\rho_{abs}=1.51 \times 10^3 \text{ kg/m}^3$, Carbomafra-119). Sodium gluconate (P.A., Merck) was utilized as tracer in water for the liquid phase.

In the micro-pilot scale trickle-bed reactor (0.80m high and diameter of 0.04m, Figure 2), operated under non-reactive conditions, the aqueous liquid phase was pumped under constant flow rate to the system feed section, mixed just before with the liquid tracer (0.50 gL^{-1} , sodium gluconate solution).

The liquid mixture and the gas stream flow down through a packed bed with 918.0g of activated carbon catalyst, under isobaric (202.64kPa) and isothermal (300K) conditions. Assuming that the reactor is irrigated with low liquid throughputs the solid particles were considered partially wetted. The liquid flow rate in the reactor varied within a range from 5.00 Lh^{-1} to 15.00 Lh^{-1} . Continuous samplings of the liquid mobile phase in contact with the fixed bed were analyzed through a conductivimeter with an AD/DA microcomputer interface. The suppression of the tracer solution in the liquid feed stream through the fixed bed results in negative step concentration curves at inlet and outlet of the bed.

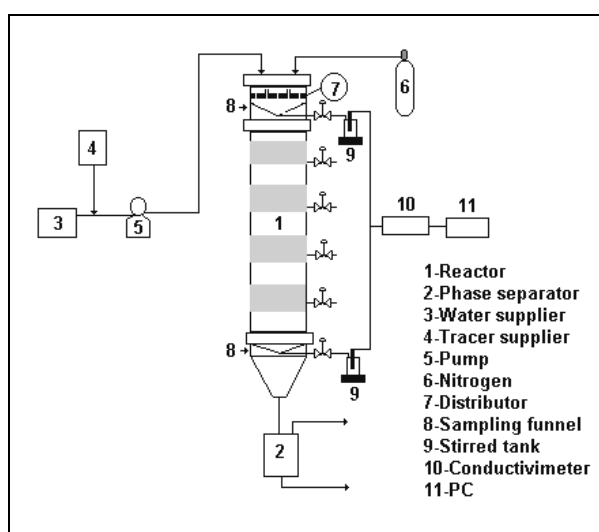


Figure 2: Experimental dynamic set-up of the trickle-bed reactor

RESULTS AND DISCUSSION

The transfer functions of the models of both the reactors (Equations 6 and 13), fluidized-bed and trickle-bed, are used to simulate the dynamic experiments. In the sequence to estimate model parameter values an optimization procedure (Box, 1965) was applied. Independent experimental values and estimations from correlations were used as initialization values for the numerical optimizations. Through the transfer functions, the numerical procedure searched to minimize the objective functions $f_o = \sum_{i=1}^n [C_{Ei}^{exp}(t) - C_{Ei}^{th}(t)]^2$, where C_{Ei}^{exp} and C_{Ei}^{th} are the respective tracer concentrations, obtained experimentally and calculated by the models at the exit of the three-phase reactors. In both cases the transfer functions (6) and (13) were numerically inverted by a fast inversion algorithm (FT^{-1}) in the Fourier domain performing the deconvolution expressed by $C_E^{th}(t) = FT^{-1}[(C_I^{exp}(t) * G(s))]$, where C_I^{exp} is the experimental tracer concentration at the inlet of the three-phase reactors. The values of the parameters that allowed the optimization were

considered as the final values indicating the order of magnitude of the hydrodynamic, mass transfer and fluid-solid interaction phenomena in the three-phase systems.

Parametric Evaluation and Simulation of the Experimental Results for the Three-Phase Fluidized-Bed Reactor

The application of the RTD methodology at gas phase in the fluidized-bed reactor using methane as a low solubility tracer in the liquid phase allowed to isolate, via dynamic model, the relative effects on the gas holdup and axial dispersion excluding the representative terms of the gas-liquid and liquid-solid mass transfer and adsorption steps. The three-phase system of fluidized-bed operation was identified under homogeneous regime operation from Ramachandran and Chaudhari (1983) classification, which shows that for a small distribution of gas bubbles, low interaction between the bubbles occurs. Furthermore, the operational conditions identify that this process occurs in the homogeneous region in the Deckwer diagram

(Deckwer, 1980). In addition to the experimental evaluation of the three-phase fluidized system experiments were performed without suspended solid at two-phase gas-liquid operations. Comparison of the results of the two-phase operation with the three-phase system may show the effects of the presence of solid particles on the behavior of the gas phase.

To quantify the hydrodynamic effects originated from the presence of the gas phase in the fluidized-bed reactor, an optimization procedure of the parameters gas holdup (h_G) and Peclet number (Pe_G) was achieved where experimental RTD's concentration curves obtained at different gas flows containing methane (3.0% v.v.), were compared with calculated values obtained by the fluidized-bed dynamic model.

The gas holdup values used for the initialization of the optimization procedure were obtained by the experimental technique consisting in shutting down simultaneously the gas and liquid inlets, with $h_G = 1 - H_{LS}/H$, where H and H_{LS} are the expanded and non-expanded slurry heights, respectively. The gaseous phase Peclet number was introduced from the experimental values of Zahradnik and Fialová (1996), determined at similar operation conditions in a three-phase fluidized reactor.

The optimized parameters are shown in Table 1 for the gas phase in the fluidized-bed reactor. Figure 3 shows the experimental and calculated profiles of tracer concentrations at the reactor outlet.

From the results presented in Table 1, an analysis of the influence of the gas phase superficial velocity and the presence of the particulate solid on the gas holdup and the Peclet number was achieved. It was observed that the values of both parameters do not vary significantly with the liquid flow rate at two-phase or three-phase operations. It was observed at the both operations in two-phase and three-phase systems, reductions of the values of the Peclet number with the increasing of the superficial velocity of the gas phase, maintained the superficial velocity of liquid phase constant. These effects are less evidenced in the three-phase system. This was attributed to a larger stability of the emulsion due to the presence of particulate solid. The values of the evaluated parameters are in comparable magnitude orders with those estimated by Zahradnik and Fialová (1996) under similar conditions.

The model elaborated for the FBR simulates its behavior at the conditions of the liquid flow of 10.00Lh^{-1} and the gas flow of $1,000.00\text{Lh}^{-1}$ (Figure 4).

Table 1: Optimized model parameters for the three-phase fluidized-bed reactor

Superficial velocities		Three-phase operations			Two-phase operations		
$u_L \times 10^2 (\text{cm s}^{-1})$	$u_G (\text{cm.s}^{-1})$	$h_G \times 10^2$	Pe_G	$a \times 10^2$	$h_G \times 10^2$	Pe_G	$a \times 10^2$
1.00	1.00	2.22	53.55	10.01	2.48	54.54	15.55
1.00	2.00	4.89	46.74	19.65	4.90	38.40	26.68
1.00	3.00	8.33	41.20	29.40	9.73	38.00	32.73
2.00	1.00	2.21	54.18	9.95	2.65	54.54	16.44
2.00	2.00	4.53	52.13	18.47	4.43	38.40	18.13
2.00	3.00	8.42	47.07	29.63	9.29	38.00	31.72

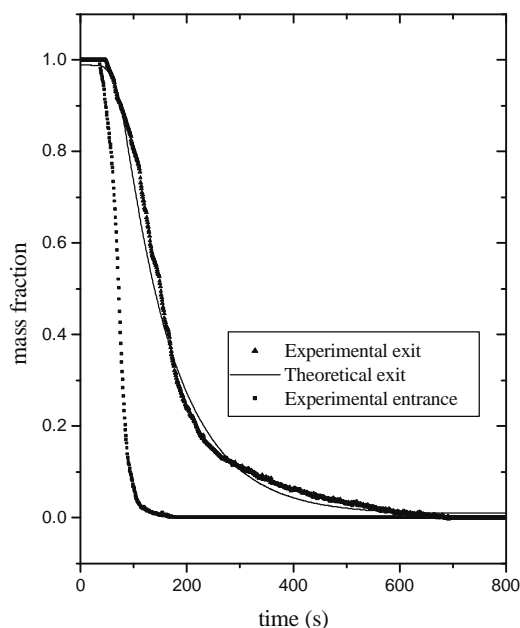


Figure 3: Dynamic evolution of concentration of gas phase in the three-phase fluidized-bed reactor.

System: $\text{CH}_4\text{-Air/Al}_2\text{O}_3$; $Q_G = 500.00\text{Lh}^{-1}$; $Q_L = 5.00\text{Lh}^{-1}$; $T = 300\text{K}$; $P = 101.32\text{kPa}$; $m_{\text{cat}} = 80.00\text{g}$.

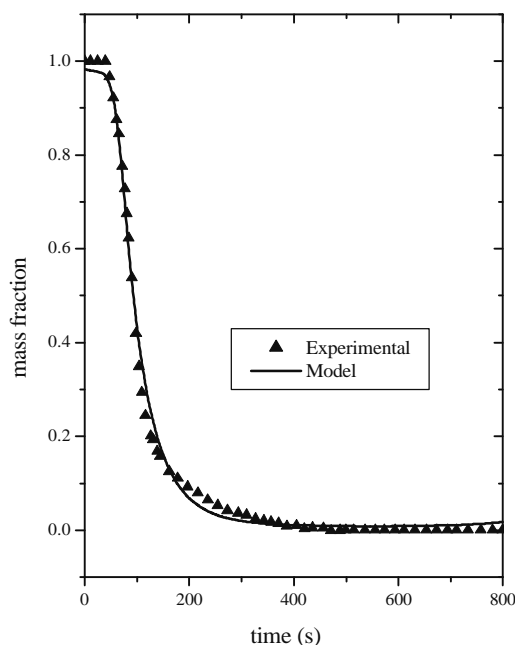


Figure 4: Simulation of the tracer dynamic behavior in the three-phase fluidized-bed reactor. System: CH₄-Air/Al₂O₃; Q_G=1,000.00Lh⁻¹; Q_L=10.00Lh⁻¹; T=300K; P=101.32kPa; m_{cat}=80.00g.

Parametric Evaluation and Simulation of Experimental Results for the Trickle-Bed Reactor

Moment analysis of the experimental RTD's curves obtained at different liquid flows, containing sodium gluconate (0.50gL⁻¹) used as a tracer in the fixed-bed reactor under trickle-bed regime, according to Charpentier and Favier diagram (Charpentier and Favier, 1975), allowed to evaluate the initialization values of model parameters associated to the first order moment of RTD and that of transfer function, as follow as:

- computation of experimental first order moments ($\mu_1^{exp} = t_{TBR}$) from the experimental dynamic responses ($F_{(t)}$) of the trickle-bed systems by

$$\mu_1^{exp} = \int_0^{\infty} tF(t)dt ;$$

- derivation of the moments of the transfer functions using the Van der Laan relation ($\mu_n = \lim_{s \rightarrow 0} [(-1)^n (d^n G/ds^n)]$), obtaining:

$$\mu_1 = \frac{V_R}{Q_L} \left[h_L + (1 - e) f_e \left(\frac{\gamma_b}{1 - e} K_{ad} + \epsilon_{in} \right) \right] \quad (16)$$

Equation 16 provides the estimation of the initialization values of the following parameters: liquid holdup (h_L) and wetting efficiency (f_e) for the liquid phase. Additionally, RTD's experiments with an auxiliary tracer (KCl), allow to use the Equation 16 twice (sodium gluconate, KCl) to form a system of equations having the values of h_L and f_e as solution. Equilibrium adsorption constants were evaluated in isolated batch adsorption experiments. Liquid-solid mass transfer coefficient ($k_{t,LS}$), effective intraparticle diffusivity ($D_{ef,L}$) and initialization values of the Peclet number (Pe_L) for the liquid tracer were estimated by the expressions written in Table 2, where their values are listed.

Table 3 shows the optimized parameters for the liquid phase in the trickle-bed reactor and Figure 5 shows the experimental and calculated dynamic profiles of tracer concentrations at the trickle-bed reactor outlet.

Table 2: Estimation of mass transfer parameters for the sodium gluconate in the trickle-bed reactor

Correlations	Parameters	References
$\frac{k_{t,LS}d_p}{D_{t,L}} f_e = 0.815 Re_L^{0.822} \left(\frac{\mu_L}{\gamma_L D_{t,L}} \right)^{0.333}$	$k_{t,LS} \times 10^5 \text{ms}^{-1}$ 6.11-11.81	Satterfield et al., 1978
$D_{ef,L} = 1.17 \times 10^{-16} \frac{(\gamma_{H_2O} M_{H_2O})^{1/2} T}{V_t^{0.6} \mu_{H_2O}}$	$D_{ef,L} \times 10^7 \text{m}^2 \text{s}^{-1}$ 1.21-2.52	Wilke and Chang, 1955
$Pe_L = c_1 Re_L^2 10^{c_3} Re_G$ $c_1=0.012 ; c_2=0.77 ; c_3=0.0011$	Pe_L 0,077 – 0,191	Iliuta et al., 1999

Table 3: Optimized model parameters for the trickle-bed reactor

$u_L \times 10^3 (\text{m s}^{-1})$	$u_G \times 10^2 (\text{m s}^{-1})$	f_e	h_L	Pe
1.00	2.13	0.38	0.17	4.84
2.00	2.13	0.48	0.21	11.44
3.00	2.13	0.56	0.25	21.12
1.00	6.39	0.34	0.16	4.52
2.00	6.39	0.36	0.20	9.57
3.00	6.39	0.45	0.21	20.37

Parameters h_L , f_e and $k_{t,LS}$ are of external nature related to the solid, being susceptible to liquid phase flow variations. For the parameters of intraparticle nature ($D_{ef,L}$; K_{ad}), the results obtained are satisfactory due to being at the same order of magnitude, with slight variations, to different liquid phase flows that means compatibility with the predicted constancy of effective diffusivity ($D_{ef,L} = 7.65 \times 10^{-10} \text{m}^2 \text{s}^{-1}$) and equilibrium adsorption constant ($K_{ad} = 8.20 \times 10^{-5} \text{m}^3/\text{kg}$).

The procedure applied to the results obtained from the trickle-bed operations led to the optimized values of the hydrodynamic parameters that were compared with the predictions from the literature correlations (Iliuta et al., 1999) obtained under similar experimental conditions. The authors estimated the parameters through the technique of

RTD, using a pulse of a tracer in the liquid phase. It was confirmed that the liquid phase superficial velocity has a significant influence on the hydrodynamic parameters such as liquid holdup, liquid phase axial dispersion and wetting efficiency. Only at higher liquid velocities the action of the gas flow rate can be considered as Iliuta et al. (1999) predicted. An increasing on the value of the liquid phase superficial velocity increases the liquid holdup, as well as the wetting efficiency, and the liquid phase Peclet number. Increasing the gas flow rate lowers the liquid holdup and wetting efficiency as well as the liquid phase Peclet number.

The model elaborated for the TBR simulates its behavior at the conditions of the liquid flow of 10.00Lh^{-1} and the gas flow of 300.00Lh^{-1} (Figure 6).

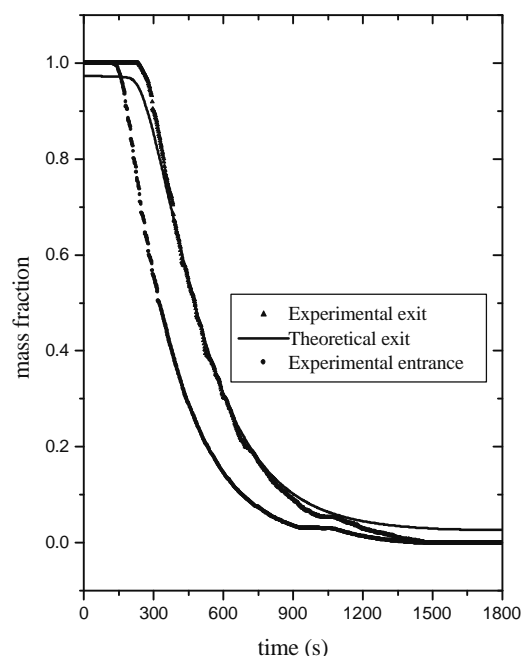


Figure 5: Dynamic evolution of concentration of liquid phase in the three-phase trickle-bed reactor. System: NaCl-H₂O/C; $Q_G=300.00 \text{Lh}^{-1}$; $Q_L=5.00 \text{Lh}^{-1}$, $T=300 \text{K}$; $P=202.64 \text{kPa}$; $m_{\text{cat}}=918.0 \text{g}$.

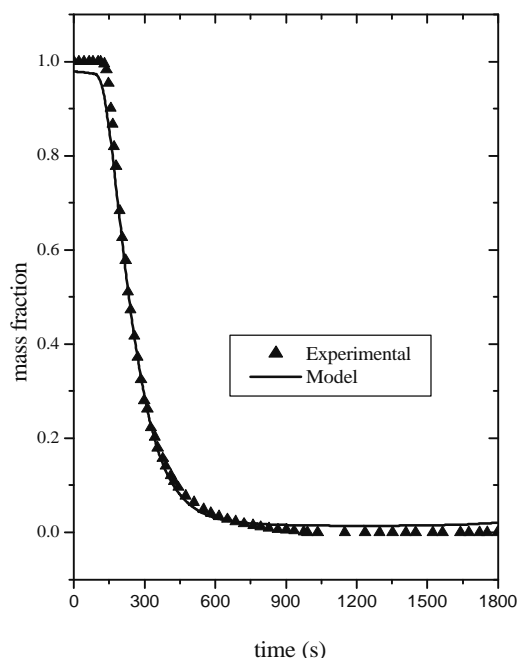


Figure 6: Simulation of the tracer dynamic behavior in the three-phase trickle-bed reactor. System: NaCl-H₂O/C; $Q_G=300.00 \text{Lh}^{-1}$; $Q_L=10.00 \text{Lh}^{-1}$, $T=300 \text{K}$; $P=202.64 \text{kPa}$; $m_{\text{cat}}=918.0 \text{g}$.

CONCLUSIONS

The results obtained by applying the experimental dynamic methodology associated with the phenomenological models in both three-phase systems operating in fluidized-bed and trickle-bed regimes, allowed to describe the dynamic behavior of the gas and liquid phases with methane and sodium gluconate tracers in gas and liquid phases, respectively.

On the basis of the estimated values of the parameters gas holdup and gas phase Peclet number for the three-phase fluidized-bed reactor, and of the parameters liquid holdup, wetting efficiency and liquid phase Peclet number for the trickle-bed reactor, an analysis was performed to determine the influence of the gas and liquid phase superficial velocities.

The three-phase fluidized-bed reactor operating under homogeneous regime at low gas velocity is submitted to influences of weak effects of the gas holdup and gas phase Peclet number. Increasing the gas velocity results in higher levels of Peclet number and gas holdup. A larger stability of the gas-liquid emulsion attributed to the presence of the solid particle catalyst justifies the reduction of the mixture effects.

For a trickle-bed system, an increase in the liquid phase superficial velocity leads to an increase of the liquid holdup, as well as the wetting efficiency and the liquid phase Peclet number. The gas flow rate increasing conducts the parameters liquid holdup and wetting efficiency to be diminished as well as the liquid phase Peclet number.

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NOMENCLATURE

C	Concentration,	mol m^{-3}
D_{ax}	Axial dispersion,	$\text{m}^2 \text{s}^{-1}$
D_{eff}	Effective diffusivity,	$\text{m}^2 \text{s}^{-1}$
d_p	Particle diameter,	m
f_e	Wetting efficiency	(-)
h_G	Gas holdup	(-)
h_L	Liquid holdup	(-)
K_{ad}	Adsorption equilibrium constant,	L mol^{-1}
k_{ad}	Adsorption constant	

$k_{t,LS}$	Liquid-solid mass transfer coefficient,	m s^{-1}
Pe_G	Peclet number of the gas phase in the three-phase fluidized-bed reactor $[Z \cdot u_G / (D_{ax} \cdot h_G)]$	(-)
Pe_L	Peclet number of the liquid phase in the three-phase trickle-bed reactor $[Z \cdot u_L / (D_{ax} \cdot \epsilon)]$	(-)
Q	Flow rate,	$\text{m}^3 \text{s}^{-1}$
q	Tracer concentration on the solid surface,	$\text{g (g of the support)}^{-1}$
Re	Reynolds number in the three-phase reactors $(u \cdot d_p \cdot \rho / \mu)$	
S_p	Superficial area (BET),	$\text{m}^2 \text{m}^{-3}$
u	Superficial velocity,	m s^{-1}
V	Volume,	m^3
Z	Reactor length,	m

Greek Letters

α	Volume fraction of gas in zone 1 (tubular reactor) of the model of the three-phase fluidized-bed reactor	(-)
ϵ	Bed porosity	(-)
ϵ_{in}	Internal particle porosity	(-)
ρ_{abs}	Absolute gravity,	kg m^{-3}
ρ_{app}	Apparent gravity,	kg m^{-3}
ρ_b	Bed gravity,	kg m^{-3}

Subscripts

G	Gas	(-)
L	Liquid	(-)
R	Reactor	(-)
S	Solid	(-)

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