

*Research Paper*

# **INTERACTION BETWEEN HYDRODYNAMICS AND CHEMICAL REACTION RATE IN VERTICAL GAS- SLURRY COCURRENT TURBULENT TUBE REACTOR.**

## **II. APPLICATIONS**

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***Running title:*** Tube Reactor in Turbulent Flow

***Key words:*** Turbulent Tube Reactor, Upflow-Downflow, Performance Indicators,  
Continuous Hydrogenation

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## **Abstract**

The behavior of an isothermal, vertical tube reactor in a dispersed bubble flow environment in either upflow, downflow or upflow-downflow configurations is examined, using the performance indicators defined in Part I. The hydrogenation of a coalescent organic system, like the one found in the mild hydrogenation process of soybean oil, is taken as a reference system. The likelihood of the axial changes of the local gas-slurry mass transfer efficiency and, then, of the controlling regimes of the process rate is discussed. Further effects of the configuration on the conversion of the liquid reactant, the unconverted mass of the gas phase, and the pressure drop are also analyzed.

## **Introduction**

This paper is devoted to the analysis of what we regard as the most important features of the complex interaction between hydrodynamics and chemical reaction rate in an isothermal turbulent tube reactor (TTR) in cocurrent flow. To this end, a local effectiveness factor associated with the efficiency of gas-slurry mass transfer, the conversion of the liquid reactant, the local unconverted mass fraction of the gas phase, and the pressure drop along the tube are taken as performance indicators. These performance indicators were developed in the theoretical Part I of this work (Cantero et. al, 1999). Their mathematical expressions are summarized in **Table 1**, and are applied here to a specific case.

TTRs are used in commercial processes for the manufacture of commodities, such as hydrogenated 2-ethyl-antraquinone, hydrogen peroxide (Käbisch, 1969), and hydrogenated edible oils to meet specifications of stability to oxidation while keeping their dietary properties (Coombes et al., 1974; Hastert, 1981; Cantero, 1993). In this part we examine the isothermal continuous hydrogenation of soybean oil (a coalescent organic system) for dispersed bubble flow pattern (DBP) in upflow (U), downflow (D) and alternating upflow-downflow (A) configurations.

### **1. Process characteristics and physical properties**

Results are presented for a set of operating conditions typically found in the mild catalytic hydrogenation of highly unsaturated vegetable oils (e.g., soybean oil; iodine value: 135-137), which are usually hydrogenated using supported Ni catalysts (0.01-0.10 % w/w) at moderate temperatures (410-430 K) and pressures (0.2-0.6 MPa). More specifically, an isothermal analysis is done at  $T = 423$  K,  $P = 0.6$  MPa. In order to span a wide range of process conditions having either chemical reaction rate- or gas-slurry mass transfer rate-controlled regimes as asymptotic behaviors, several catalyst loadings and/or catalytic activities at several rates of energy dissipation per unit mass of slurry are considered.

The friction factor ( $f_L$ ) and the Lockhart-Martinelli correction factor ( $\phi_L$ ) were corrected for dispersed bubble flow (DBP) pattern (Wallis, 1969; Cantero et al., 1999). The values of the Lockhart-Martinelli correction factor are in the 0.25-1.00 range depending on the local gas holdup ( $\varepsilon_G$ ), which cannot exceed 0.5 to fulfill the conditions needed to ensure the DBP (Cantero *et al.*, 1996). The gas-slurry mass transfer coefficient ( $k_L$ ) was estimated from a well-known correlation (Calderbank and Moo Young, 1961; Nagel, 1981). The gas-slurry interfacial area ( $a$ ) was estimated from the Sauter mean diameter of the gas bubbles (Calderbank, 1967; Deckwer and Schumpe, 1987). The gas holdup was calculated from values of unconverted mass fraction of gas reactant. The remaining physical parameters, constants and/or properties can be found elsewhere. (Cantero et al., 1996). The steady-state cocurrent flow patterns map for the hydrogen-soybean oil system is shown in Part I of this contribution (Cantero *et al.*, 1999). For the simulations, the following initial conditions were used:  $T = 423$  K;  $P(\xi=0) = P^o = 0.6$  Mpa;  $w_A(\xi=0) = w_A^o = 0.05$ ;  $w_H(\xi=0) = w_H^o = 0.0$ . Stoichiometric amounts of gas and liquid reactant were used, i.e.:  $G_G/G_L(\xi=0) = M_H w_A^o/M_A$ , where  $M_H = 2$  kg mol<sup>-1</sup> and  $M_A = 294$  kg mol<sup>-1</sup>.

## 2. Downflow and Upflow Configurations

### 2.1 Local values of the $E_H/E_{Hmin}$ ratio

**Figure 1** shows the profiles of the local energy dissipation rate per unit mass of slurry ( $E_H$ ) relative to the minimum one ( $E_{Hmin}$ ) which is needed to sustain the DBP throughout the TTR for the D- and U-configurations. The values of the ratio  $E_H / E_{Hmin}$  were calculated using the gas holdup profiles obtained solving the system of integral equations, Eqs. (1)-(4), for three levels of mean superficial velocity of slurry phase ( $Fr = 10, 20$  and  $30$ ) at extreme values of the chemical reaction rate ( $Da_H = 10^{-5}$  and  $10^0$ ). The DBP can be sustained neither in the region under the dashed lines labelled  $\aleph$ , nor on the right-hand side of the dashed lines labelled  $\beth$  and  $\Re$  which represent the operational limit beyond which  $\varepsilon_G > 0.5$ , for  $Da_H = 10^{-5}$  and  $Da_H = 10^0$ , respectively.

In the D-configuration (see Fig. 1a), the regular increase of local pressure leads to a moderate decrease of gas holdup, even for extremely low values of chemical reaction rate per unit volume of TTR (fine lines,  $Da_H = 10^{-5}$ ). A moderate decrease of the  $E_H / E_{Hmin}$  ratio along the TTR results, meaning that only if the DBP can be established at the entrance of the reactor it will be safely maintained along it (for  $Fr > 15$  in this example). Conversely, the gaseous reactant is rapidly depleted at high chemical reaction rate (full lines,  $Da = 10^0$ ). In this case, the smallest values of the  $E_H / E_{Hmin}$  ratio are found closer to the reactor entrance. Further downstream, a fast recovery follows, and the entrance conditions for sustaining the DBP are again reached and surpassed, due to the pronounced decrease of  $E_{Hmin}$  when the gas depletion is approached.

For the U-configuration (see Fig. 1b), the gas holdup can notably increase at low values of the chemical reaction rate per unit volume of TTR ( $Da_H = 10^{-5}$ , fine lines) as the local pressure steadily decreases in the axial direction. However, this growth in the gas holdup can be attenuated by consumption of the gaseous reactant in the other extreme case ( $Da_H = 10^0$ , full lines). In the first case, as a consequence of the increased gas holdup along the tube,  $E_H$  grows faster than  $E_{Hmin}$ , and therefore the  $E_H/E_{Hmin}$  ratio improves, until eventually the DBP becomes unstable above  $\varepsilon_G > 0.5$ . At high levels of chemical reaction rate, the gas holdup decreases upwards owing to the fast gas consumption and, consequently, the  $E_H/E_{Hmin}$  ratio may decrease at first. Further upwards, depressurization due to a lower hydrostatic pressure prevails, and the gas holdup also increases demanding, then, a large energy dissipation per unit mass of slurry to sustain the DBP.

A minimum level of slurry mean superficial velocity is required to generate the DBP for both U- and D-configurations ( $Fr > 15$  in this exemplification). Also, in upflow, DBP can be sustained farther along the TTR as the reaction rate increases.

## 2.2 Gas-slurry mass transfer effectiveness factor

The local effectiveness factor profiles predicted by Eq. (1) with Eq. (5), for D- and U-configurations and a wide range of values of the  $Da_H/St$  ratio, are shown in **Fig. 2**. Since pressure drops are nearly linear along the tube reactor, these plots also represent the changes of the local effectiveness factor vs. pressure along the TTR.

For D-configuration (see Fig. 2a), the highest values of the local effectiveness factor are reached, almost linearly, near the reactor entrance. The lower the  $Da_H/St$  ratio, the larger  $\eta(\xi)_{\max}$ . Further changes of  $\eta$  are minimal under both the chemical reaction rate-controlled ( $\eta(\xi) \geq 0.91$ ) and the gas-slurry mass transfer rate-controlled regimes ( $\eta(\xi) \leq 0.09$ ), but they turn quite significant for intermediate values of the  $Da_H/St$  ratio. Equations (1) and (5) fully account for this behavior. Near the reactor entrance, the Green function tends to unity and thus, on the RHS of Eq. (1), a linear-source term  $[\varepsilon_G(\xi) St]$  prevails, leading to gas saturation of the slurry phase in linear form, too. Farther along the tube, the exponential source term of Green's function counterbalances the former effect, the more so the higher the  $Da_H/St$  ratio is.

Remarkably, close to the reactor entrance the process rate is controlled by the gas-slurry mass transfer rate, in every case. Downstream, the local effectiveness factor is practically uniform throughout most of the TTR for both extreme controlling regimes, because neither the gas holdup nor the liquid reactant composition change substantially under these extreme conditions. However, intermediate values of the  $Da_H/St$  ratio allow gas holdups and/or reactant concentrations which may enable transitions either from a regime controlled by the chemical reaction rate to an intermediate one ( $0.1 < \eta(\xi) < 0.9$ ) or from the latter regime to one controlled by the gas-slurry mass transfer rate.

For the U-configuration (see Fig. 2b) these transitions reverse because the gas holdup increases as a consequence of the prevailing effect of decompression caused by the upward decrease of the hydrostatic pressure. A secondary effect adds-up, though, because increased gas desorption accelerates the flow which, in turn, raises the pressure drop and feedbacks the increased

desorption of the dissolved gas. Then, the local concentration of the dissolved gas can be even higher than its saturation solubility, due to the local decompression of the gas phase.

Summarizing, for the U-configuration the following transitions are possible either from the gas-slurry mass transfer rate-controlled regime to an intermediate one with an eventual further transition to chemical reaction rate-controlled regime or from an intermediate regime to chemical reaction rate-controlled regime.

### 2.3 Conversion of the liquid reactant vs. unconverted mass fraction of the gas phase

**Figure 3** shows the profiles of the local conversion of the liquid reactant (full lines) and the local unconverted mass fraction of the gas phase (fine lines) for D- and U-configurations, respectively. Even for high values of the chemical reaction rate ( $10^{-2} < Da_H$ ), only the D-configuration enables the achievement of total conversion since in the U-configuration the total admissible length of the tube is bounded because the pressure drop leads to an accelerated expansion and desorption of the gaseous reactant, with the consequent departure from the DBP at  $\varepsilon_G > 0.5$ .

Another remarkable feature comes from the opposing effects that the Froude number has on the conversion. Low Froude number values are most desirable to yield high conversions, as well as to minimize pumping costs because the energy dissipation per unit mass of slurry phase increases with a 1.5-order dependency of the Froude number [Cantero et al., 1999; Eq. (4)]. Nevertheless, as high Froude number values are needed to establish and to keep the DBP, savings in both capital and operating costs are limited by the lower boundary for which the operation of the TTR is no longer feasible. The detrimental impact of a large Froude number on the conversion can be compensated with a higher geometrical ratio ( $Ge = L/D$ ), but at the expense of impractical vertical arrangements. For instance, even under extremely high intrinsic chemical reaction rates, a TTR with a 0.05 m tube diameter would require a height of about 60 m (Cantero, 1993). Thus, an A-configuration becomes unavoidable for usual layouts.

### 3. Alternating Configuration

An A-configuration can be set by assembling  $N$  consecutive sections of pairs of upflow and downflow stretches, each of them with the same diameter and length. To illustrate matters, the following analysis is made considering the following geometry: tube diameter,  $D = 0.05$  m; stretch length,  $L = 10$  m.

#### 3.1 Gas-slurry mass transfer effectiveness factor

**Figure 4** shows the dimensionless pressure drop and the local effectiveness factor profiles along the  $N$  sections of the TTR, for an intermediate Froude number value ( $Fr = 20$ ) and extreme  $Da_H/St$  ratios. The discrete abscissa,  $N$ , indicates the exit from each of the upflow-downflow sections of the tube reactor. The overall appearance of the  $\eta(\xi)$  vs.  $N$  plots resemble those of the U-configuration of the TTR, since frictional losses lead to net pressure drops in both cases.

At the highest  $Da_H/St$  ratio it is observable that, after the initial region under the gas-slurry mass transfer rate-controlled regime is surpassed, the reactor operates inside an intermediate regime, which is a direct consequence of having moderate gas holdup levels (not shown). Conversely, very low  $Da_H/St$  ratios lead to operating conditions under which the fast decrease in local pressure due to high frictional losses produces an excessive increase of the gas holdup which, unchecked by an adequate consumption of the gaseous reactant, may rapidly lead to the critical flow limit and the consequent unstabilization of the DBP (from the fourth section downwards in the figure).

#### 3.2 Conversion of the liquid reactant vs. unconverted mass fraction of the gas phase

**Figure 5.a** shows the progress of the local conversion of the liquid reactant and the unconverted mass fraction of the gas phase at the exit from each  $N$  section of the TTR in the A-configuration, for a wide range of values of the reaction rate. Continuous lines were drawn between consecutive sections in the figure, to ease the visual analysis. Under gas-slurry mass transfer-rate



controlled regime, the overall performance of the A-configuration is quite similar to that of the U-configuration because the frictional losses lead to net pressure drops and gas expansion in both cases (See Fig. 3.b). However, unlike the U-configuration, and like the downflow one, it is now possible to attain complete conversion of the liquid reactant given enough catalytic activity or catalyst loading.

**Figure 5.b** compares the performances of the U-, D- and A-configurations for the extreme values of the Damköhler number used herein. It shows that conversion of the liquid reactant in the D- or A- configurations evolves similarly, regardless of the degree of chemical reaction rate. However, the unconverted mass fraction of the gas phase evolves similarly in the U- and A-configurations under the chemical reaction rate-controlled regime, indicating that desorption of the dissolved gaseous reactant takes place with the pressure drop.

We have already mentioned the conflicting needs of having small Froude number values to maximize conversion, on the one hand, and of requiring moderate values to sustain the DBP, on the other. **Figure 6.a** shows that the effect of the Froude number on conversion is larger when the Damköhler number is lower, because the higher the Froude number, the shorter the residence time and the lower the conversion, under chemical reaction rate controlling conditions. However, under the gas-slurry mass transfer-controlled regime there is a modest improvement in the ratio of the gas-slurry mass transfer rate to the convective transport rate at higher Froude number values ( $St \propto Fr^{0.56}$ ), which largely compensates for any decreases in the residence time ( $t_R \propto Fr^{-0.5}$ ).

**Figure 6.b** depicts values of the local concentration of the dissolved gas reactant. They differ by about three orders of magnitude when going from low ( $Da_H = 10^{-5}$ ) to high ( $Da_H = 10^0$ ) chemical reaction rates, which certainly favors a high conversion of the liquid reactant in the latter case. Here, for a gas-slurry mass transfer-rate controlled regime, the local concentration of the gaseous reactant remains essentially constant downstream of the initial section as long as enough liquid reactant is still made available. Also, it can be seen straightforwardly that the better the ratio of

convective transport rate to chemical reaction rate becomes, the higher the availability of the gaseous reactant results.

Conversely, at very low values of  $Da_H$  the concentration of the dissolved gas right after the initial saturation falls sharply, following the decreasing values of the local equilibrium composition, due to the important pressure drop along the TTR (see Fig. 4.a). This fall becomes larger, as the Froude number increases. Indeed, the larger the convective transport rate, the sooner the TTR will enter the critical flow region.

## **Conclusion**

The strong interrelationships between pressure drop and axial variation of the holdups are a distinctive feature of the heterogeneous catalytic TTR. These holdups are proper to the multiphase vertical flow combined with significant gas holdup changes owing to the gas consumption by chemical reaction. Eventually, its operation in the desirable dispersed bubble flow pattern may be overridden by an excessive expansion of the gas phase, which will invariably happen if an upflow configuration is used, and also under chemical reaction rate-controlled regime if an A-configuration is employed.

Complete conversion of the liquid reactant for either D- or A-configurations can be achieved if the ratio between chemical reaction and gas-slurry mass transfer rates is made high enough and, in that case, a very substantial length of the reactor will operate under gas-slurry mass transfer rate-controlled regime. In the absence of selectivity constraints the most efficient gas-slurry contactor will also be the best TTR, provided that the catalytic activity and/or the catalyst loading are high.

## **Acknowledgments**

Support from the Consejo Nacional de Investigaciones Científicas y Técnicas (CONICET) and from the Universidad Nacional del Litoral of Argentina (UNL) is gratefully acknowledged.

## Nomenclature

$a$	= gas-liquid interfacial area referred to gas phase	$[\text{m}^2 \cdot \text{m}^{-3}]$
$B_0$	= Bond number ( $B_0 = \rho_L g D^2 \sigma^{-1}$ )	$[-]$
$D$	= tube diameter	$[\text{m}]$
$Da$	= Damköhler number, Eqs. (9) and (10)	$[-]$
$E_H$	= turbulent energy dissipation rate per unit mass of slurry phase ( $2 f_L Fr \phi_L^2 V_m / \rho_m$ )	$[\text{W} \cdot \text{kg}^{-1}]$
$E_{H\min}$	= minimum turbulent energy dissipation rate per unit mass of slurry phase ( $0.556 g^{1.5} D^{0.5} B_0^{-0.25} F_C^{2.5}$ )	$[\text{W} \cdot \text{kg}^{-1}]$
$f_L$	= liquid-phase friction factor ( $4.6 \times 10^{-2} Re_L - 0.2$ )	$[-]$
$F_C$	= coalescence factor ( $7.25 \times 10^{-1} + 4.15 \varepsilon_G^{0.5}$ )	$[-]$
$Fr$	= Froude number, Eq. (12)	$[-]$
$Fr_{\min}$	= minimum Froude number needed to achieve dispersed bubble pattern ( $E_{H\min} / (2 f_L \phi_L^2 V_m / \rho_m)$ )	$[-]$
$g$	= acceleration of gravity	$[\text{m} \cdot \text{s}^{-2}]$
$G(\xi / \bar{\xi})$	= Green's function, Eqs. (5)-(8)	$[-]$
$G_G$	= gas superficial mass velocity	$[\text{kg} \cdot \text{s}^{-1} \cdot \text{m}^{-2}]$
$G_L$	= slurry superficial mass velocity	$[\text{kg} \cdot \text{s}^{-1} \cdot \text{m}^{-2}]$
$Ge$	= geometrical parameter; Eq. (13)	$[-]$
$k_C$	= second-order kinetic constant based on mass fraction	$[\text{s}^{-1}]$
$k_L$	= mass transfer coefficient	$[\text{m} \cdot \text{s}^{-1}]$
$L$	= reactor length	$[\text{m}]$
$M_H$	= molecular weight of reactant $\mathbf{H}_2$	$[\text{kg} \cdot \text{kmol}^{-1}]$

$M_A$	= molecular weight of reactant <b>A</b>	[kg•kmol <sup>-1</sup> ]
$P$	= local pressure	[Pa]
$St$	= Stanton number, Eq. (11)	[—]
$U_D$	= characteristic velocity ( $g^{1/2}D^{1/2}$ )	[m•s <sup>-1</sup> ]
$U_L$	= mean superficial velocity of slurry phase	[m•s <sup>-1</sup> ]
$V_m$	= mass mean velocity ( $(G_L + G_G)/\rho_m$ )	[m•s <sup>-1</sup> ]
$w_A^0$	= reference mass fraction of <b>A</b>	[—]
$w_C$	= mass fraction of catalyst into slurry phase	[—]
$w_H$	= mass fraction of the dissolved gas <b>H<sub>2</sub></b>	[—]
$w_H^*$	= local equilibrium mass fraction of <b>H<sub>2</sub></b>	[—]
$w_H^{*0}$	= reference mass fraction of <b>H<sub>2</sub></b>	[—]
$x_A$	= conversion of <b>A</b>	[—]
$Y$	= unconverted mass fraction of gas phase	[—]

### *Greeks*

$\varepsilon_G$	= local gas holdup,	[m <sup>3</sup> •m <sup>-3</sup> ]
$\varepsilon_L$	= local slurry holdup	[m <sup>3</sup> •m <sup>-3</sup> ]
$\eta$	= local effectiveness factor for gas-liquid mass transfer, Eq. (1)	[—]
$\xi$	= dimensionless axial coordinate	[—]
$\rho_G, \rho_L$	= density of gas and slurry phases	[kg•m <sup>-3</sup> ]
$\rho_m$	= mean density of multiphase mixture ( $\rho_L(1 - \varepsilon_G) + \rho_G \varepsilon_G$ )	[kg•m <sup>-3</sup> ]

$\phi_L$	= multiphase flow correction for friction factor $(1 - \varepsilon_G)^2$	[—]
$\sigma$	= interfacial tension	[N•m <sup>-1</sup> ]
$\Theta$	= step function	[—]
$\theta$	= flow inclination angle	[rad]

## Literature Cited

- Cantero H. J.; "Análisis del Modelado de un Reactor Tubo para la Hidrogenación Selectiva de Aceites Vegetales," Doctoral Dissertation, Universidad Nacional del Litoral, Santa Fe, Argentina (1993)
- Cantero, H.J., R. J. Grau and M. A. Baltanás; "Modelling Gas-liquid Turbulent Tube Reactors: A Model Incorporating Variable Hold-up and Compressible Flow with Multiple Gas Injections and Recycling of Slurry," *Trans IChemE*, **74**, part A, 635-648 (1996)
- Cantero, H.J., M.A. Baltanás and R.J. Grau, "Interaction Between Hydrodynamics and Chemical Reaction Rate in a Vertical Gas-Slurry Cocurrent Turbulent Tube Reactor. I. Performance Indicators," *J. Chem. Eng. Japan*, **33**, 00-00 (2000)
- Calderbank, P.H. and M. Moo-Young; "The Continuous Phase Heat and Mass Transfer Properties of Dispersions," *Chem. Eng. Sci.*, **16**, 39-54 (1961)
- Calderbank P. H.; "Gas Absorption from Bubbles," *Chem. Engng.*, **45**, 209-233 (1967)
- Coombes, W. A., R.A. Zavada, J.E. Hansen, R. King, R. Robert and W.A. Singleton; "Continuous Hydrogenation of Fatty Oils," US Patent 3,792,067 (1974)
- Deckwer, W. and A. Schumpe; "A. Bubble Columns-The State of the Art and Current Trends," *Int. Chem. Eng.*, **27**, 405-411 (1987)
- Hastert, R.C.; "Practical Aspects of Hydrogenation and Soybean Salad Oil Manufacture," *J. Am. Oil Chem. Soc.*, **58**, 169-174 (1981)
- Käbisch, G.H; *Chemische Technologie*, Vol I, R. Winnacker Ed., p.529-559, Rheinfelder, Baden, Germany (1969)
- Nagel, O., H. Kurten and B. Hegner; "Criteria for the Selection and Design of Gas Liquid Reactors," *Int. Chem. Eng.*, **21**, 161-172 (1981)
- Wallis, G.B.; *One Dimensional Two Phase Flow*, Mc Graw-Hill, New York, USA (1969)

Table 1: Performance indicators expressed in terms of integral equations (Cantero *et al.*, 1999).

Gas-slurry mass transfer effectiveness factor		Green's functions
$\eta(\xi) = \int_0^1 d\bar{\xi} G_H(\xi/\bar{\xi}) \varepsilon_G(\bar{\xi}) St w_H^*(\bar{\xi})/w_H^*(\xi) \quad (1)$		$G_H(\xi/\bar{\xi}) = \exp\left\{-\int_{\bar{\xi}}^{\xi} d\hat{\xi} \varepsilon_G(\hat{\xi}) St \eta(\hat{\xi})^{-1}\right\} \quad (5)$
Local conversion of the liquid reactant		
$x_A(\xi) = \int_0^1 d\bar{\xi} G_A(\xi/\bar{\xi}) Da_H \varepsilon_L(\bar{\xi}) w_H^*(\bar{\xi}) \eta(\bar{\xi}) [1 - x_A(\bar{\xi})] \quad (2)$		$G_A(\xi/\bar{\xi}) = \Theta(\xi - \bar{\xi}) \quad (6)$
Local unconverted mass fraction of gas phase		
$Y(\xi) = Y(0) - \int_0^1 d\bar{\xi} G_Y(\xi/\bar{\xi}) \varepsilon_L(\bar{\xi}) Da_A w_A(\bar{\xi}) w_H^*(\bar{\xi}) \eta(\bar{\xi}) \quad (3)$		$G_Y(\xi/\bar{\xi}) = \Theta(\xi - \bar{\xi}) \quad (7)$
Pressure drop		
$\Delta P(\xi) = \Delta P^T - \int_0^1 d\bar{\xi} G_P(\xi/\bar{\xi}) Ge [2f_L Fr \phi_L(\bar{\xi}) - \varepsilon_L(\bar{\xi}) \cos \theta(\bar{\xi})] \quad (4)$		$G_P(\xi/\bar{\xi}) = \Theta(\xi - \bar{\xi}) \quad (8)$
Dimensionless numbers		
$Da_H = (k_C w_C w_A^0 D / U_D) (M_H / M_A) Ge Fr^{-0.5} \quad (9)$		$St = (k_L a w_H^{*0} D / U_D) Ge Fr^{-0.5} \quad (11)$
$Da_A = (k_C w_C w_H^{*0} D / U_D) Ge Fr^{-0.5} \quad (10)$		$Fr = U_L^2 / g D \quad (12)$
		$Ge = D / L \quad (13)$



## Figure captions

Figure 1: Energy dissipation normalized with minimum energy required to achieve DBP vs. dimensionless axial position of TTR: (a) downflow and (b) upflow configurations at extreme values of chemical reaction rate ( $Da = 10^{-5}$ , fine lines; and  $Da = 10^0$ , full lines), for three levels of mean superficial velocity of the slurry phase ( $Fr = 10, 20$  and  $30$ ). Numerical values correspond to  $L/D = 2000$  and  $St = 3.8 \times 10^{-6}$ . Set of initial conditions is indicated in text

Figure 2: Local effectiveness factor for gas-slurry mass transfer,  $\eta(\xi)$ , vs. dimensionless axial position of TTR: (a) downflow and (b) upflow configurations, for a full range of values of chemical reaction and gas-slurry mass transfer rates ratio,  $Da/St$ , and intermediate values of mean superficial velocity of the slurry phase ( $Fr = 20$ ). Numerical values correspond to  $L/D = 2000$  and  $St = 3.8 \times 10^{-6}$ . Same initial conditions as in Fig. 1 are used

Figure 3: Conversion of liquid reactant,  $x_A$  (full lines) and local unconverted mass fraction of gas phase,  $Y$  (fine lines) vs. dimensionless axial position of TTR: (a) downflow and (b) upflow configurations, for full range of values of chemical reaction rate ( $10^{-5} \leq Da \leq 10^0$ ), and intermediate values of mean superficial velocity of slurry phase ( $Fr=20$ ). Numerical values correspond to  $L/D = 2000$  and  $St = 3.8 \times 10^{-6}$  for same initial conditions as in Fig. 1.

Figure 4: (a) Local dimensionless pressure drop and (b) local effectiveness factor along consecutive section,  $N$ , of TTR in alternating (upflow-downflow) configuration, at extreme values of  $Da/St$  ratio ( $Fr = 20$ ). Upflow and downflow stretches are of equal length, with  $L/D = 200$ . Numerical values correspond to  $D = 0.05$  m and  $St = 3.04 \times 10^{-6}$

Figure 5: Conversion of liquid reactant,  $x_A$  (full lines) and local unconverted mass fraction of gas phase,  $Y$  (fine lines) inside the TTR: (a) at exit from consecutive sections,  $N$ , in alternating flow configuration, for full range of values of chemical reaction rate and (b) at extreme values of  $Da$  number in upflow (**U**), downflow (**D**) and alternating flow (**A**) configurations, with  $Fr = 20$  and  $L/D = 200$ . Numerical values correspond to 8 pairs of stretches or equivalent length of straight tube ( $L/D= 1600$ ) for **U** and **D** configurations, with  $D = 0.05$  m and  $St = 3.04 \times 10^{-6}$  in each case. Same initial conditions as in Fig. 1 are used

Figure 6: Combined effect of convective transport and chemical reaction rates, for TTR operating in alternating flow: (a) Conversion of liquid reactant,  $x_A$ , and (b) concentration of dissolved gas,  $w_H = w_H^* \eta(\xi)$ , at the end of consecutive sections of reactor, for  $Fr = 10, 20$  and  $30$ , and  $Da = 10^0$  and  $10^{-5}$ , with  $L/D = 200$ ,  $D = 0.05$  m and  $St = 3.04 \times 10^{-6}$  when  $Fr = 20$ . Same initial conditions as in Fig. 1 are used