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FLOW MODELING IN A LIQUID - SOLID FLUIDIZED BED REACTOR

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Abstract. The objective of the proposed work is to present a simple and efficient method of simulating the flow in a solid/liquid fluidized bed reactor The model system considered concerns the fluidization of glass beads (2 mm in diameter and 1554 kg . m^{-3} of density) by water in a glass column (High/internal diameter = 91/2 cm/cm. The experiments were carried out with repetitions for three flow rates: 5, 7 and 8.2 cm s^{-1} corresponding to the porosities of: 0.55, 0.64 and 0.69. The height of the fixed bed was set at 10 cm, and the water temperature was maintained at 20 ± 2°C. The shape of the measured RTD was characterized by three flow patterns: the Tanks in series model (TSM), the combination: plug flow reactor and stirred tank reactor in series (PFR + CSTR), and the dispersed plug flow model (DPFM). The comparison of the average residence time with the space time confirms the absence of dead zones in the fluidized bed. Based on the lowest Root Mean Square Error (RMSE), the dispersed plug flow model seems more representative in the case of low flow velocity, while at higher flow rates, the combined model is more suitable. The results also showed that the mixing was maximum for the porosity of 0.64, corresponding to the intermediate speed of 7cm s^{-1}.

Keywords: Fluidized bed, RTD, axial dispersion model, tanks in series model, variance, tracer.

INTRODUCTION

Currently, the fluidization technique is experiencing a resurgence of interest in the material transformation, energy production, and environmental industries (Kader Gaid et al.,2019). The interface between the particles and the fluid, made up of all the particle surfaces of a fluidized bed, represents a considerable contact area, which makes heat and material transfer operations very efficient. This feature is at the origin of the success of fluidized beds in many industries (Tsheno Nirina, 2005). Fluidized bed reactors are widely used in many unit operations such as; chemical industry, petroleum, pharmaceutical...etc. (Linda Brakchi, 2015). In a real reactor, the molecules remain in the reactive volume for residence times that depend in particular on the hydrodynamic profile, and the reactor geometry. These times can deviate significantly from the theoretical residence time, so there is a residence time distribution (RTD), and this dispersion has an influence on the chemical performance of the reactor (Moumtez Bensouici, 2007). In this work, we modeled the residence time distribution in a solid-liquid fluidized bed reactor using the tanks in series model, the dispersed plug flow model, and the combined model based on the series combination of a plug flow reactor and a mixed flow reactor.

NOMENCLATURE

Adimensional numbers

Re_p Reynolds' particle number
Pe Peclet number

Latin letters

CSTR: continuous stirred tank reactor
PFR plug flow reactor
RTD residence time distribution
RMSE Root Mean Square Error
N number of tanks in series
T temperature (°C)
Q_0 volumetric feed rate ($cm^3.s^{-1}$)
U flow velocity ($cm.s^{-1}$)
E residence time distribution function (s^{-1})
E_θ normalized residence time distribution function
C concentration measured at time t (g/L).
Ci concentration measured at time t_i (g/L).
H bed height at rest or fixed bed (m).
$\bar{t}$ mean residence time (s)
$\bar{t}_{tsm}$ mean residence time for tanks in series models (s)
$\bar{t}_{dm}$ mean residence time for dispersed plug flow model (s)
d_p mean particle diameter (mm)
m_p mass of glass particles (kg)
$\hat{y}_i$ predicted value of the response
y_i observed value
n number of observations

Greek letters

τ space time (s)
ρ_p density of glass particles ($kg.m^{-3}$)
ρ_ℓ density of water ($kg.m^{-3}$)
θ dimensionless residence time
μ dynamic viscosity (Pa.s)
ε porosity
σ_t^2 variance (s^2)
σ_θ^2 dimensionless variance
π pi

THEORY

Residence time distribution function

The residence time distribution function from the measured tracer response for the impulse signal is (Octave Levenspiel, 1999).

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$$E(t) = \frac{C(t)}{\int_0^\infty C(t)dt} \quad (1)$$

In discrete form, it is expressed as :

$$E(t) = \frac{C_i}{\sum C_i \Delta t_i} \quad (2)$$

Mean residence time

The mean residence time $\bar{t}$ which is the first moment, is given by :

$$\bar{t} = \frac{\int_0^\infty tC(t)dt}{\int_0^\infty C(t)dt} = \int_0^\infty t\, E\,(t)dt \quad (3)$$

In discrete form :

$$\bar{t} = \frac{\sum t_i C_i \Delta t_i}{\sum C_i \Delta t_i} = \sum t_i\, E(t_i)\Delta t_i \quad (4)$$

The dimensionless time θ is :

$$\theta = \frac{t}{\bar{t}} \quad (5)$$

Normalized RTD function E(θ)

The relationship between E(t) and E(θ) is found from the basis that both represent the same physical entity, the fraction of exit fluid with age θ. Thus, E(θ)dθ = E(t)dt. Therefore (Abel Kayode Coker, 2001).

$$\mathrm{E}(\theta) = \bar{t}E(t) \quad (6)$$

Variance

The second moment is taken about the mean and is referred to as the variance σ_t^2 defined as :

$$\sigma_t^2 = \int_0^\infty (t-\bar{t})^2 E(t)dt = \int_0^\infty t^2 E(t)dt - \bar{t}^2 \quad (7)$$

In discrete form σ_t^2 is :

$$\sigma_t^2 = \sum (t_i - \bar{t})^2\, E(t_i)\Delta t_i = \sum t_i^2\, E(t_i)\Delta t_i - \bar{t}^2 \quad (8)$$

The dimensionless variance σ_θ^2 is expressed as :

$$\sigma_\theta^2 = \frac{\sigma_t^2}{\bar{t}^2} \quad (9)$$

Dispersed plug flow model (DPFM)

This model is usually used to describe nonideal PERs. For small extents of dispersion(1/Pe) < 0.01) the system emulates plug flow and E_θ is expressed as (Chuntian Hu, 2021):

$$E(\theta) = 0.5\left(\frac{Pe}{\pi}\right)^{0,5} e^{\left(-\frac{Pe(1-\theta)^2}{4}\right)} \quad (10)$$

The variance σ_θ^2 is calculated by :

$${\sigma_\theta}^2 = \frac{2}{Pe} - \frac{2}{Pe^2}(1 - e^{-Pe}) \quad (11)$$

Accordingly, when (1/Pe) > 0.01, the system is open and far from plug flow, and E_θ and its variance are expressed as :

$$E(\theta) = 0.5\left(\frac{Pe}{\theta\pi}\right)^{0,5} e^{\left(-\frac{Pe(1-\theta)^2}{4.\theta}\right)} \quad (12)$$

$${\sigma_\theta}^2 = \frac{2}{Pe} + 8\left(\frac{1}{Pe}\right)^2 \quad (13)$$

Tanks in series model (TSM)

A model frequently employed to simulate the behavior of an actual reactor is a series of ideal stirred tank reactors. The actual reactor can be replaced by N identical stirred tank reactors whose total volume is the same as that of the actual reactor. For a series of CSTRs, the RTD for CSTR in series E(t) is (Gilbert F. Froment et al., 1979) :

$$E(t) = \left(\frac{N}{\bar{t}}\right)^N \frac{t^{N-1}}{(N-1)!} e^{\left(-\frac{Nt}{\bar{t}}\right)} \quad (14)$$

The function E (θ) is given by :

$$\mathrm{E}(\theta) = N\frac{(N\theta)^{N-1}e^{-N\theta}}{(N-1)!} \quad (15)$$

The number of tanks in series is :

$$\mathrm{N} = \frac{1}{\sigma_\theta^2} = \frac{\bar{t}^2}{\sigma_t^2} \quad (16)$$

Combined model based on the serial combination of a plug flow reactor and a continuous stirred–tank reactor (PFR+CSTR)

The RTD function is given by the following equation (Marie Louise Bouchard, 2011) :

$$E(t) = \begin{cases} 0 & t < \tau_p \\ \dfrac{e^{\frac{-(t-\tau_p)}{\tau c}}}{\tau_c} & t \geq \tau_p \end{cases} \quad (17)$$

In the absence of dead zones in the reactor, the average residence time $\bar{t}$ is written :

$$\bar{t} = \tau_p + \tau_c \quad (18)$$

EXPERIMENTAL PROCEDURE

The flow parameters for this study are grouped in the table.1

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Table 1.

Flow parameters

T (°C)	ρ_p (kg.m^{-3})	dp (mm)	m_p (kg)	H (m)	ρ_ℓ (kg.m^{-3})	μ (Pa.s)	U (cm.s^{-1})	R_{ep}	ε
20±2	1554	2	0.077	0.01	1000	0.001	5.00	100	0.55
							7.00	140	0.64
							8.20	170	0.69

Measurements of residence time distribution (RTD) in the liquid phase were carried out using the saline tracing technique. The method chosen consists of injecting (pulse type injection) at the base of the glass column (2cm internal diameter and 91 cm high) 10 ml of a sodium chloride (NaCl) solution of concentration 100 g/L, and following the evolution over time of its concentration at the outlet of the column. The calibration curves (conductivity-NaCl concentration), established from several standard solutions prepared with tap water, made it possible to obtain the concentration of the NaCl solution from the measured conductivity. For each flow, we monitored by video recording the change in conductivity over time of a layer of glass particles fluidized by tap water. Data acquisition was done with a time step of 3 seconds.

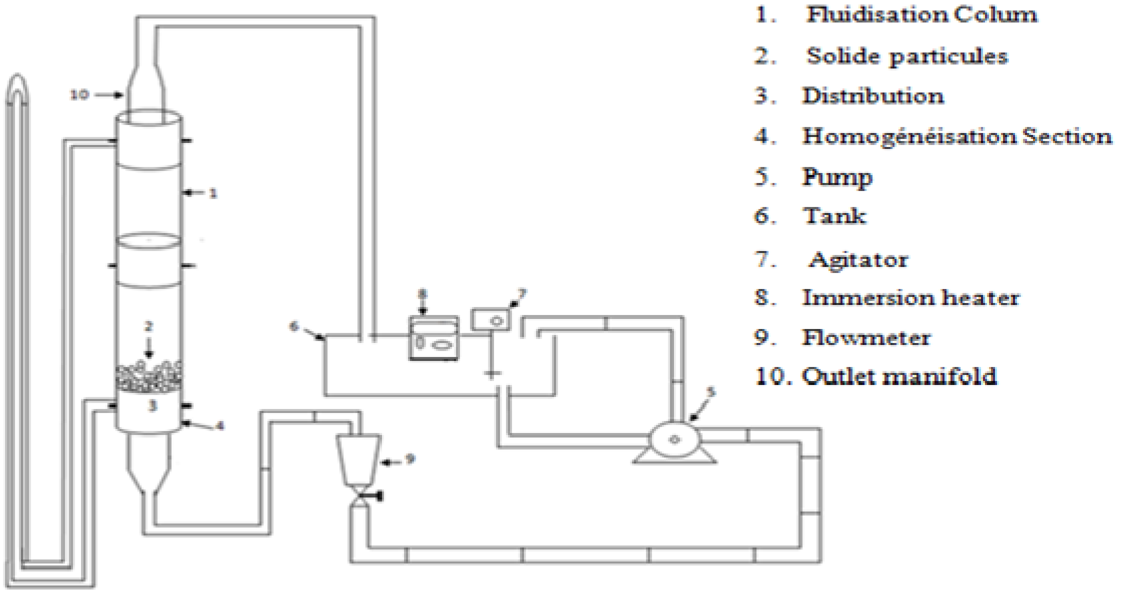

Fig. 1. Experimental apparatus.

RESULTS AND INTERPRETATIONS

Reproducibility test

In order to test the reproducibility of our experiments, each test is repeated twice. Fig. 2 shows the evolution of the experimental RTD as a function of time for the fluidization speeds corresponding to 5 cm.s^{-1}, 7 cm.s^{-1} and 8.2 cm.s^{-1}. These flow speeds correspond respectively to the following porosities: 0.55; 0.64 and 0.69. Note that the experimental RTD values are reproducible, except those obtained for a porosity of 0.64 (case (b)). This is confirmed by the calculation of the differences between the two tests of the dimensionless variance σ_θ^2. Where the relative difference does not exceed 10% for the porosities 0.55 and 0.69 reaches 21% for 0.64 (see Table. 2).

Table 2.

Experimental values of the variances for each experimental test

Porosity	Test	$\bar{t}$	σ_t^2	σ_θ^2	Relative error (%)
ε = 0.55	1	237.71	25491.24	0.451	9.8
	2	220.4	24273.3	0.500	
ε = 0.64	1	186	16289.07	0.471	21.23
	2	204	15451.16	0.371	
ε = 0.69	1	161.57	12331.46	0.472	9.74
	2	144.78	7186.4	0.426	

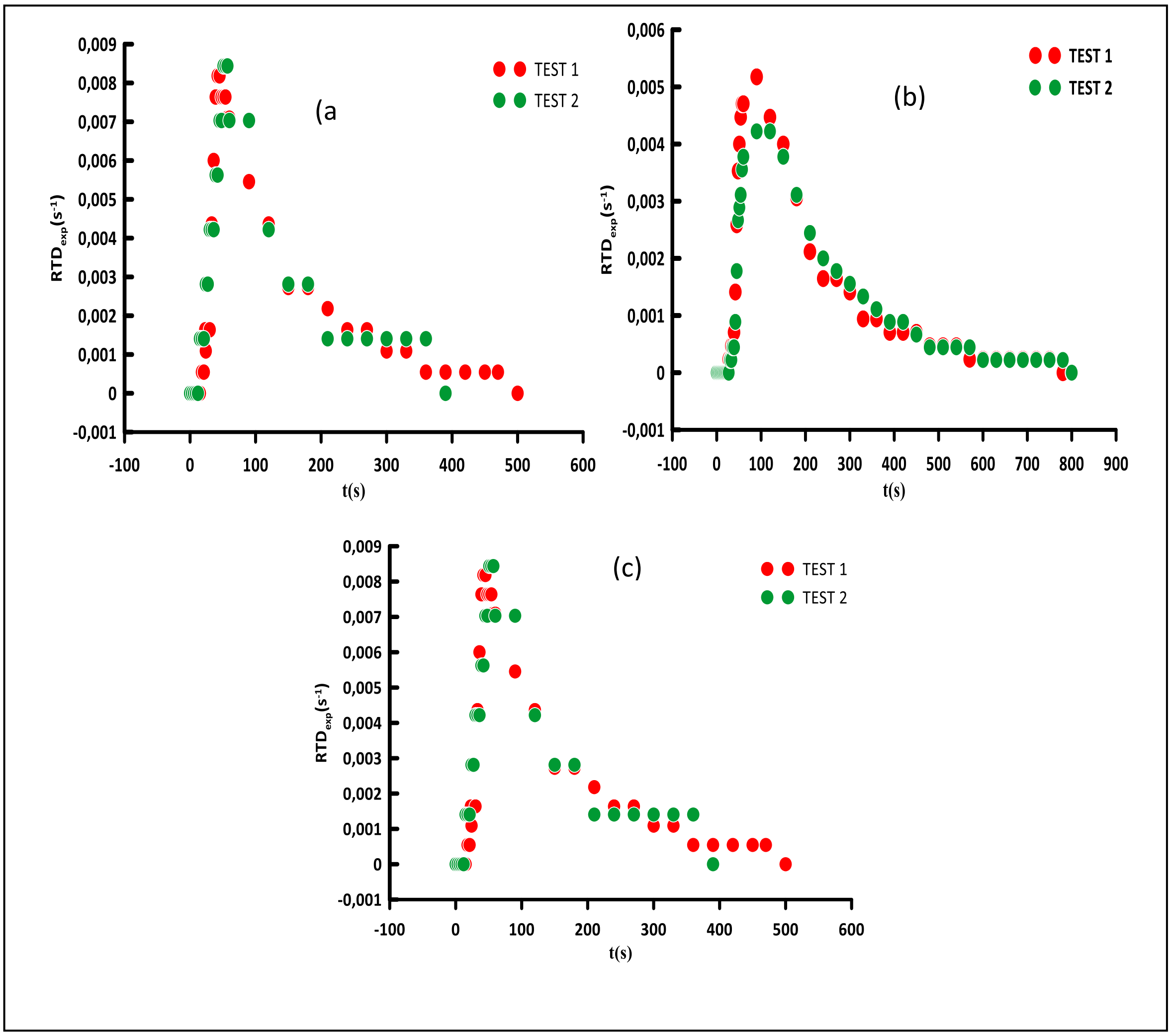

Fig. 2. Evolution of the experimental RTD as a function of time.
(a) : U= 5cm.s^{-1} ; ε = 0.55, (b) : U= 7cm.s^{-1} ; ε = 0.64, (c) : U= 8.2 cm.s^{-1} ; ε = 0.69.

Evolution of RTD as a function of porosity and flow velocity

The curves in Fig.3 show a peak between 50 and 90 seconds, this peak is followed by an exponential decrease, which is spread over a time equivalent to 50 times the passage time. The appearance of these curves is typical of a response intermediate between that of a tubular laminar flow reactor and that of a series association of PFR and CSTR.

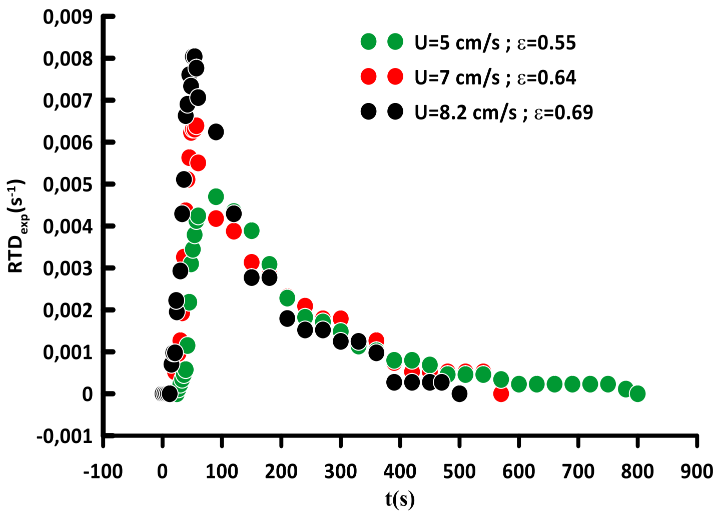

Fig. 3. Effect of porosity and flow velocity on RTD.

Detection of dead space or bypass

One of the main uses of RTD analysis is to determine if there are any dead zones or preferential paths in the reactor under study. For this purpose, we compared the average experimental residence time (this is the average experimental residence time calculated from the two tests for each operating condition) with the passage time, τ calculated for each operating condition (see table. 3). It can be seen that the average residence time is much greater than the passage time τ, which indicates a priori the absence of dead volume (or stagnant zones) in the mixing zone. This may indicate the existence of preferential passages or a short circuit. Table 3 shows that the mean residence time $\bar{t}$ in the reactor decreases with increasing porosity. It should also be noted that with increasing fluidization speed, the peak height increases and the peak is shifted to shorter times, so the RTD becomes narrower and narrower.

Table 3.

Passage time values for each operating condition

ε	U(cm.s^{-1})	$\bar{t}$(s)	Q$_0$(cm^3.s^{-1})	τ (s)
0.55	5	229.06	15.84	16.69
0.64	7	195.00	21.88	12.14
0.69	8.2	153.24	25.82	10.27

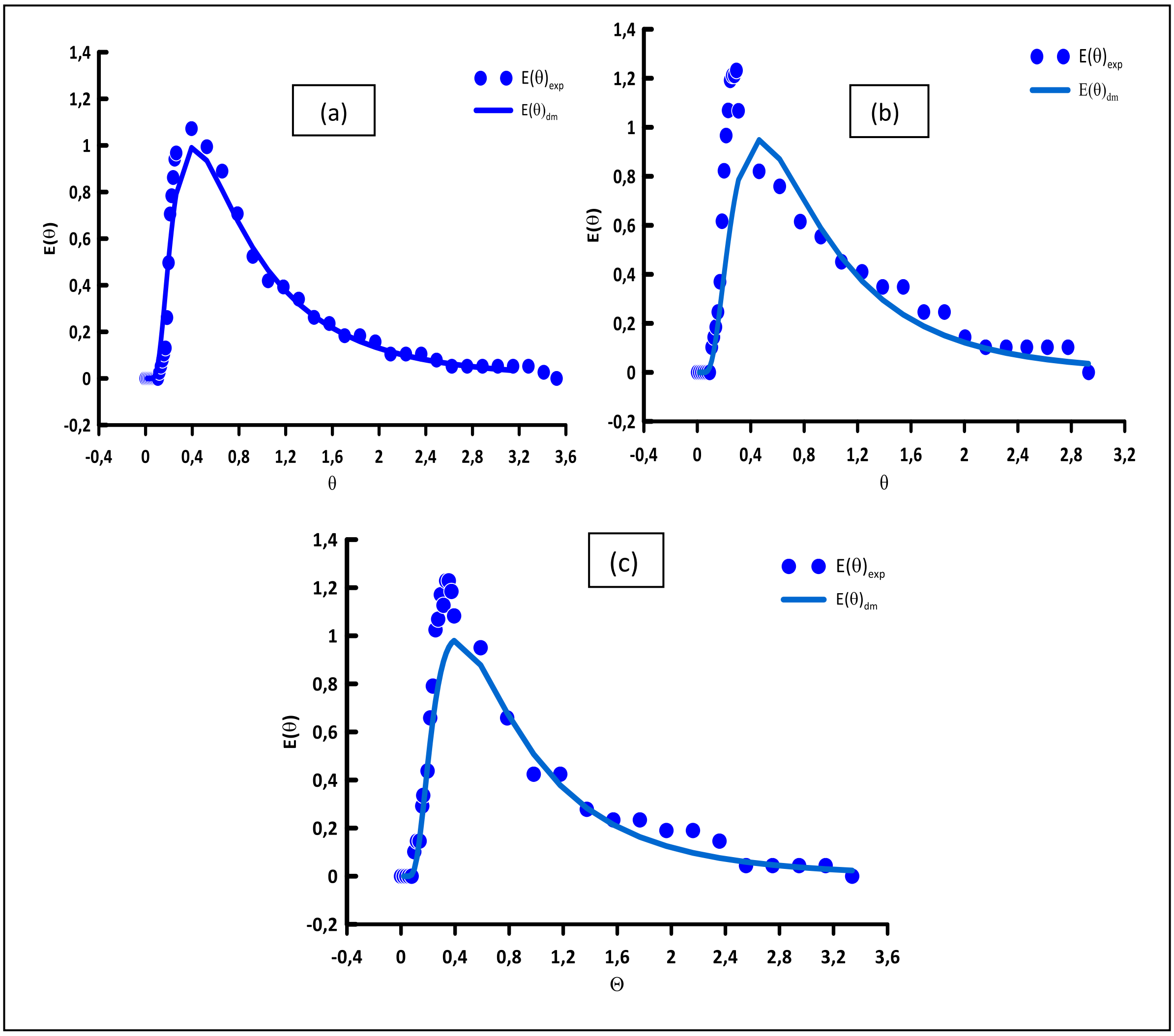

Fig. 4. Evolution of E(θ) as a function of θ and comparison. with the DPFM:(a) : U= 5 cm.s^{-1} ; ε = 0.55 (b) : U= 7 cm.s^{-1} ; ε = 0.64, (c) : U= 8.2 cm.s^{-1} ; ε = 0.69

Comparison of experimental results with flow models

- **With the dispersed plug flow model (dpfm)**

For each operating condition, we calculated the dimensionless Peclet number (Pe) using Solver. The values thus found are grouped in Table .4.

In Fig. 4, the results of the comparison of the various measurements of experimental RTD with those obtained in a closed system are presented. We notice that the values predicted by the model coincide with the experimental values for the fluidization speed of 5cm.s^{-1} corresponding to a porosity of 0.55. This is not the case for the other two fluidization speeds: 7 cm.s^{-1}

and 8.2 cm.s^{-1} corresponding respectively to the porosity of the bed of 0.64 and 0.69. In Table 5, we see that the average residence time calculated from the model is very close to the experimental one for porosity 0.55 with a relative difference of around 4%; on the other hand, for porosities 0.64 and 0.69, the relative difference is 12%.

Table 4.

Experimental values Pe

ε	U (cm.s^{-1})	$\bar{t}$(s)	σ_t^2 (s^2)	σ_θ^2	Pe
0.55	5	229.06	24957.16	0.474	2.807
0.64	7	195.00	15951.14	0.417	3.450
0.69	8.2	153.24	9758.93	0.416	3.462

Table 5.

Comparison of the experimental mean residence time and that predicted by DPFM

ε	U (cm.s^{-1})	$\bar{t}$(s)	$\bar{t}_{dm}$ (s)	Relative error (%)
0.55	5.0	229.06	219.61	4.13
0.64	7.0	195.00	171.23	12.19
0.69	8.2	153.24	134.12	12.48

- **With the tanks in series model (tsm)**

In Table 6, we note that the number of stirred tanks calculated from the model is the same for the three fluidization speeds studied. The values of N show that our fluidized bed reactor can be assimilated to a cascade of two perfectly stirred open reactors of identical volumes. The values of the average residence time calculated for all the operating conditions show that the relative deviation is very small for U = 5 cm.s^{-1} (3.5%), which corresponds to a porosity of 0.55, whereas it is 15.54% for U = 7 cm.s^{-1} (ε = 0.64) and 20.69% for U = 8.2 cm.s^{-1} (ε = 0.69). Thus, we can conclude that this model corresponds only in the case of the low flow velocity(U = 5 cm.s^{-1}). Fig. 5, illustrating the comparison of the experimental RTD values to those obtained by the TSM, clearly shows that this model is not representative of the experimental values for high flow velocities.

Table 6.

Results of the TSM analysis.

ε	U(cm.s^{-1})	$\bar{t}$(s)	$\bar{t}_{tsm}$ (s)	σ_θ^2	N	Relative error (%)
0.55	5.0	229.06	220.96	0.474	2.11	3.5
0.64	7.0	195.00	164.70	0.417	2.40	15.54
0.69	8.2	153.24	121.53	0.416	2.40	20.69

- **With the combined model based on the series association of a PFR and CSTR**

In Fig. 6, we notice that for higher flow speeds (7 and 8.2 cm.s^{-1}) relating to the large porosities of the fluidized bed (0.64 and 0.69), the curves of the RTD predicted by the model coincide with the experimental ones. In Table. 7, where the results of the analysis of the combined model are shown, we notice that when the fluidization speed goes from 5 cm.s^{-1} (ε = 0.55) to 7 cm.s^{-1} (ε = 0.64), there is a strong mixing and a weak behavior of the piston flow (see the ratios τ_c / τ) and (τ_P / τ)). On the other hand, it is the reverse which occurs when the fluidization speed passes from 7 cm.s^{-1} (ε = 0.64) to 8.2 cm.s^{-1} (ε = 0.69). From these results, we see that the mixing is maximum at the porosity of 0.64. This result should be compared to certain works in the literature, in particular those of Ken-Ichi Kikuchi et al., 1984) who measured the coefficient of axial dispersion, for polystyrene and glass particles whose diameter varies from 0.2 to 1.7 mm fluidized by water in a cylindrical column 1.7 m high and 3, 4 cm internal diameter (Nassima Kechroud *et al.*, 2010). These authors have shown that the coefficient of axial dispersion of the liquid phase increases with increasing porosity and reaches a maximum at a porosity between 0.7 and 0.8.

Table 7.

Results of the combined model analysis

ε	U (cm.s^{-1})	$\bar{t}$	τ	τ_p	τ_c	$\frac{\tau_c}{\tau}$	$\frac{\tau_p}{\tau}$
0.55	5	229.06	233.58	90	136.06	0.58	0.38
0.64	7	195.00	172.61	51	144	0.83	0.29
0.69	8.2	153.24	145.50	51	102	0.70	0.35

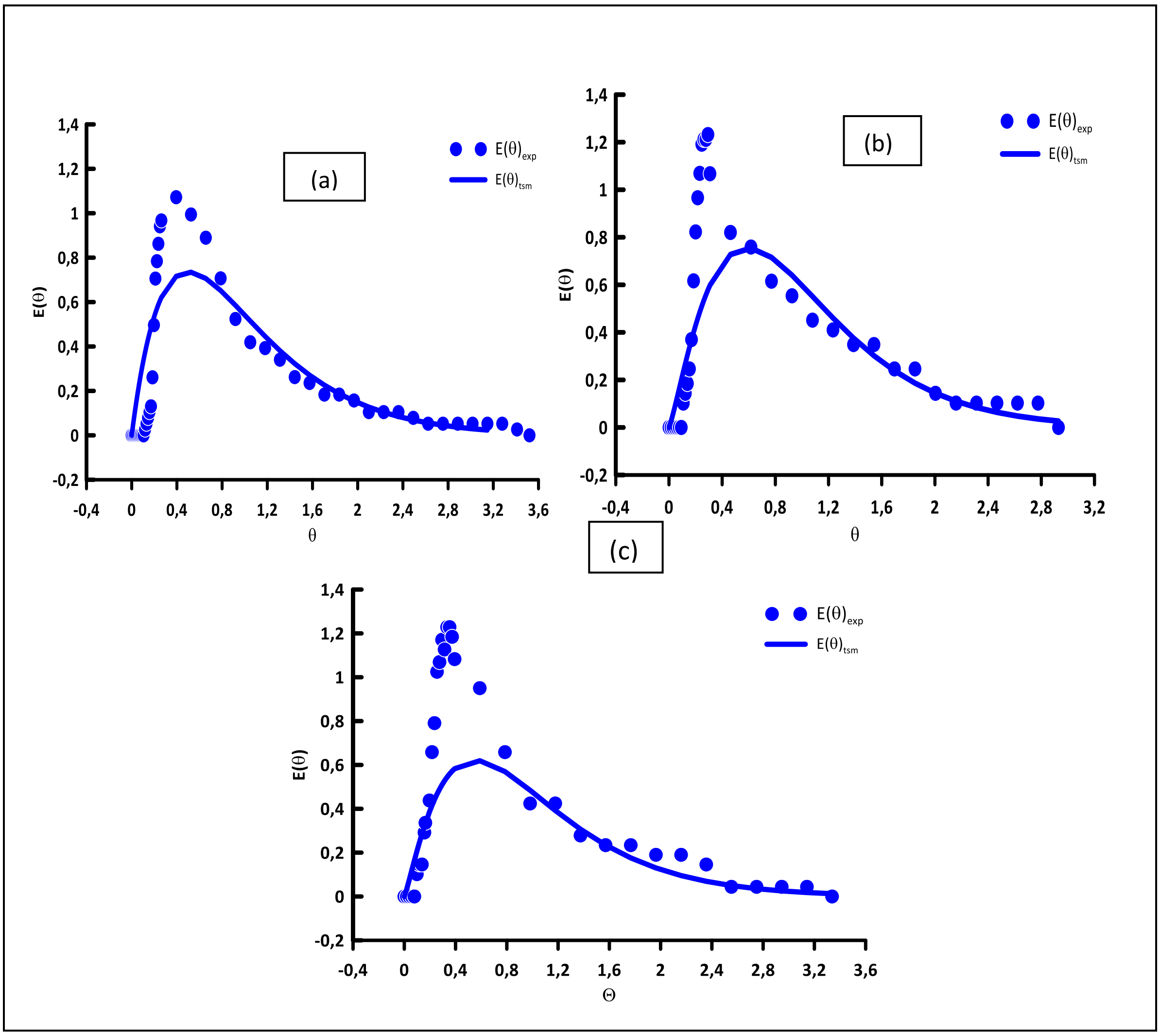

Fig.5. Evolution of the E(θ) as a function of θ and comparison with the TSM:(a) : U= 5 cm.s^{-1} ; ε = 0.55 (b) : U= 7 cm.s^{-1} ; ε = 0.64, (c) : U= 8.2 cm.s^{-1} ; ε = 0.69

COMPARISON OF THE THREE MODELS TESTED

To define the most suitable model, we calculated the Root Mean Square Error (RMSE) defined as follows Simon P. Neill et al., 2018) :

$$\mathrm{RMSE} = \left(\sum_{i=1}^{n}\left(\frac{(\hat{y}_i - y_i)^2}{n}\right)\right)^{\frac{1}{2}} \qquad (19)$$

From the RMSE values shown in the Table. 8, it can be seen that the DPFM and that of STM seem more representative for predicting the values of the RTD in the case of a low fluidization speed (U = 5 cm.s^{-1}). For higher fluidization speeds (7 and 8.2 cm.s^{-1}), the most suitable model is that of the serial combination of a PFR followed by a CSTR.

Table 8.

RMSE values for all models tested

U(cm.s^{-1})	ε	RMSE		
		DPFM	STM	PFR+CSTR
5	0.55	0.013	0.048	0.24
7	0.64	0.11	0.15	0.012
8.2	0.69	0.13	0.16	0.0094

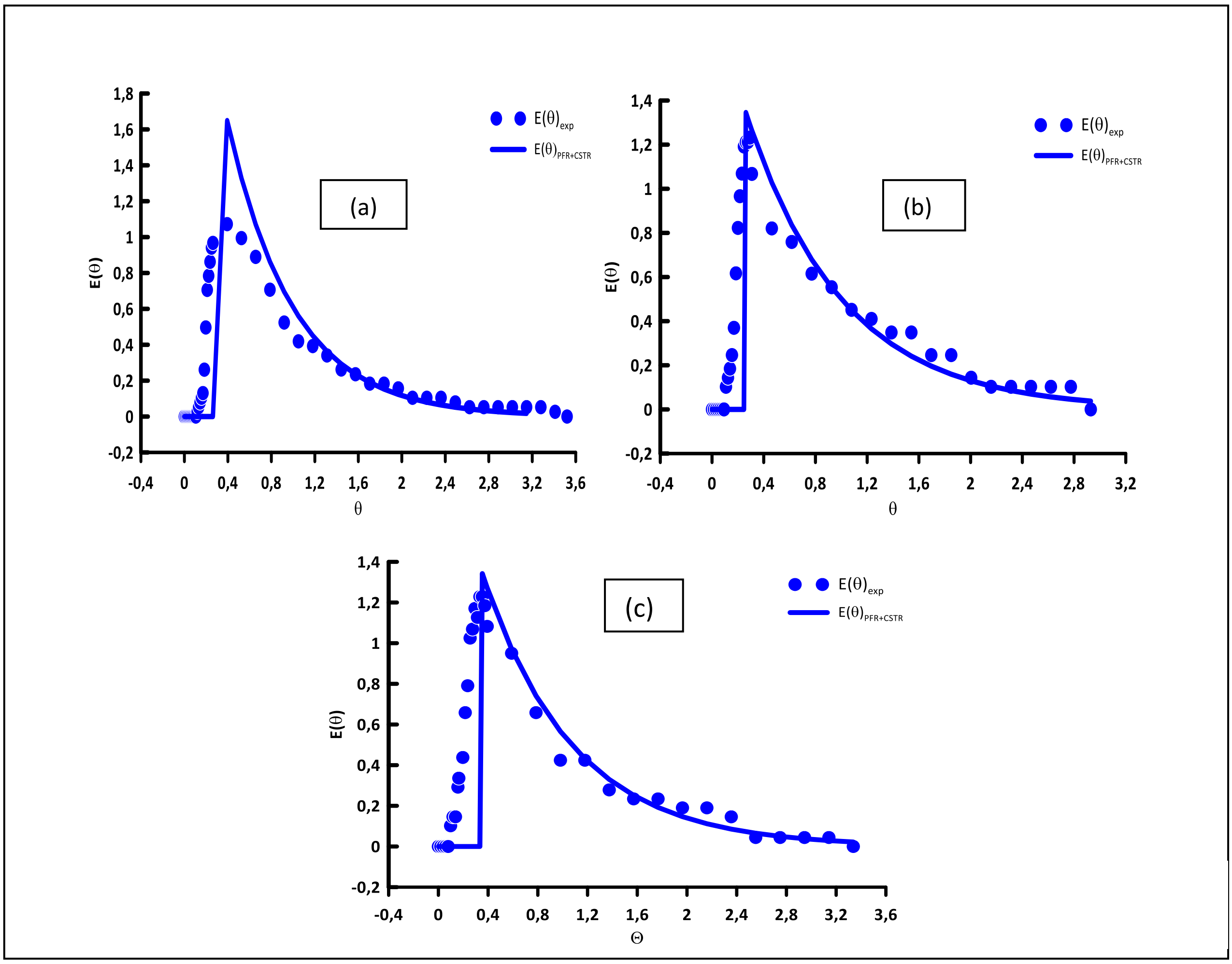

Fig.6. Evolution of E (θ) as a function of θ and comparison with the combined model. U= 5 cm.s^{-1} ; ε = 0.55 (b) : U= 7 cm.s^{-1} ; ε = 0.64, (c) : U= 8.2 cm.s^{-1} ; ε = 0.69

CONCLUSION

The results of this study allowed us to deduce that :

- The experimental RTD values are reproducible, except those obtained for a porosity of 0.64.
- The shape of the RTD curves is typical of a response intermediate between that of a laminar flow tubular reactor and that of a serial combination of a PFR followed by a CSTR
- The mean residence time is much greater than the breakthrough time, which indicates a priori the absence of dead volume (or stagnant zones) in the mixing zone.
- The mean residence time $\bar{t}$ in the reactor decreases with the increase in surface speed (or porosity). It should also be noted that as the fluidization speed increases, the height of the peak increases and the peak is shifted to shorter times, the more the RTD becomes narrower and narrower.

Comparison of experimental results with flow models shows that:

- The number of stirred tanks calculated from the model with N tanks is identical for all the porosities studied (N≈2).
- For lower flow velocity (U = 5m.s^{-1}), the RTD is best represented by the piston dispersion model and that of N stirred tanks. On the other hand, for speeds 7 and 8.2 cm.s^{-1}, the combined model is more representative, which is confirmed by the calculation of the RMSE.
- The mixing is maximum at a porosity of 0.64, this result is close to certain works in the literature

AUTHORS CONTRIBUTIONS

Conceptualization, NHM, HB, NK and HT; methodology, NHM, NK; data collection NHM, OD , NK; data validation, NHM, BH, data processing NHM, HB; writing—original draft preparation,NK, NHM; writing—review and editing, NHM.

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CONFLICT OF INTEREST

The authors declare no conflict of interest.

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