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Axial Dispersion and Back-mixing of Gas Phase in Pebble Bed Reactor

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Abstract

Despite the worldwide attended of pebble bed reactors (PBRs), there is a lack of fundamental understanding of the complex flow pattern. In this work, the non-ideal flow behavior of the gas phase which is used for cooling has been investigated experimentally in a 0.3 m diameter pebble bed. The extent of mixing and dispersion of the gas phase has been qualified. The effect of gas velocity on the axial dispersion has been investigated with range from 0.05 to 0.6 m/s covering both the laminar and turbulent flow regimes. Glass bead particles of 1.2 cm diameter and 2.5 gm/cm^3 which is randomly and closely packed have been used to mimic the pebbles. An advanced gas tracer technique was applied to measure the residence time distribution (RTD) of gas phase using impulse tracer. The axial dispersion coefficients of gas phase in the studied pebble bed have been estimated using the axial dispersion model (ADM). It was found that the flow pattern of the gas phase deviates from plug flow depending on the superficial gas velocity. The results showed that the dispersion of the gas reduces as the gas velocity and Reynolds numbers increased.

Keywords: Pebble bed; gas phase mixing, axial dispersion; packed bed; dispersion flow model

التشتت المحوري و المزج الرجوعي للطور الغازي في مفاعل الوسادة البلورية

الخلاصة

على الرغم من التوجه العالمي نحو استخدام المفاعلات ذات حشوة البلورات، فان الصورة غير مكتملة تماما لفهم تعقيدات اطوار الجريان في ذلك المفاعل. سيتم في هذا البحث سيتم دراسة سلوك الجريان غير المثالي مختبريا في الطور الغازي و الذي يستخدم للتبريد و سيتم هذا في مفاعل ذات حشوة بلورية بقطر 0.3 متر. سيتم تقدير مدى الانتشار و التشتت في الطور الغازي. و تم دراسة تأثير سرعة الغاز على التشتت المحوري في مد 0.6 -0.05 متر / ثانية في نمطين من الجريان الطباقي و الاضطرابي. تم استخدام حشوات زجاجية بقطر 1.2 سم كثافة 2.5 غم/ سم3 بحيث يتم املاؤها في العمود و تحاكي البلورات. تم استخدام تقنية الكشاف الغازي لقياس زمن الاستبقاء للطور الغازي باستخدام نبضة الغاز الكشاف. تم قياس معامل التشتت المحوري للطور الغازي في المفاعل ذور الحشوة البلورية باستخدام نموذج التشتت المحوري. وجد ان نمط الجريان في الطور الغازي يحيد عن الجريان المضغوط بالاعتماد على السرعة السطحية. أثبتت النتائج ان تشتت الغاز يل بازدياد السرعة السطحية و عدد رينولد.

الكلمات الدالة : حشوة بلورية، المزج في الطور الغازي ، التشتت المحوري، العمود ذو الحشوة، نموذج الجريان المتشتت.

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Abbreviations

ADM axial dispersion model

CFD computational fluid dynamics

CSTR continuous stirred tank reactor

HTGR high temperature gas-cooled reactor
PBR pebble bed reactor
PFR plug flow reactor
RTD residence time distribution
TCD thermal conductivity detector
VHTRS very high temperature reactors

Nomenclatures

C_g concentration of the tracer in the gas phase, mol/m^3
C_{inj} concentration of the injection tracer, mol/m^3
C_{in} dimensionless tracer concentration in the gas phase at the plenum outlet
C_{in}^* dimensionless convoluted tracer concentration in the gas phase at plenum outlet
C_{out} dimensionless tracer concentration in the gas phase at the reactor outlet
C_{out}^* dimensionless convoluted tracer concentration in the gas phase at reactor outlet
D_c column diameter, m
D_g axial dispersion coefficient of the gas phase, m^2/s
Pe Peclet number, dimensionless
Re Reynolds number, dimensionless
u interstitial gas velocity, m/s
U_g superficial gas velocity, m/s
t time, s
t_m mean residence time of the bed, s
Z axial distance along the column, m

Greek letters

τ space time of the bed, s
τ_o residence time in the plenum, s
ε voidage of bed (porosity)

Introduction

Pebble bed reactor (PBR) is one type of very high temperature reactors (VHTRs) for fourth generation reactor core. It is a gas-cooled, graphite-moderated high-temperature reactor that is continuously fueled with spherical fuel elements. Due to its high conversion efficiency, inherent safety performance, characterized as environmentally benign and low power density design, the high temperature gas-cooled reactor (HTGR) attracts a lot of attention worldwide. The core of a commercial HTGR normally has a cylindrical shape with a conical bottom and contains a huge number of pebbles, and has a 'double-zone' configuration, i.e. the central column zone consists of graphite pebbles as the moderator, which is surrounded by an annular zone of fuel pebbles. Both the fuel and graphite pebbles are almost the same in terms of shape and average density except that the fuel pebbles contain minute amounts of sand-sized uranium fuel substance [1].

In PBR thousands of coated particles (~900-950 micron) which moved downwards, called tristructural isotropic (TRISO) fuel particles are imbedded within a graphite matrix. These pebbles are continuously circulated until they are spent. It consists of a fuel kernel composed of UOx (sometimes UC or UCO) in the center, coated with four layers of three isotropic materials. The four layers are a porous buffer layer made of carbon, followed by a dense inner layer of Pyrolytic Carbon (PyC), followed by a ceramic layer of SiC to retain fission products at elevated temperatures and to give the TRISO particle more structural integrity, followed by a dense outer layer of PyC. TRISO fuel particles are designed not to crack due to the stresses from processes at temperatures beyond 1600°C. Due to their high surface/volume ratio, TRISO easily transfer heat from fuel to matrix graphite. Graphite is the moderator in the core, and can at the same time be utilized as a structure material. Helium (He) is chosen as coolant in VHTRs because it is hardly absorbs neutrons, not activated by neutrons, chemically inert, does not undergo a phase change, has good heat-exchange properties, and is naturally available in sufficient quantities [2]. As the helium gas flows under a high Reynolds number flow conditions through the reactor core and over the heated, randomly and closely distributed pebbles, the gas attains a temperature of 900 ^{o}C. Physically, these pebbles within the PBR core are in contact with each other [3]. Due to this interaction between the flowing gas phase and the heat generating pebbles within randomly and closely packed geometry the flow and heat transport characteristics are very complicated. Hence, the detailed information and understanding of such complex phenomenon within the bed are needed. On the other hand, mixing and the dispersion of the gas phase in PBRs directly affect the amount of heat removal from the reactor. Also, it can have negative impact on the temperature gradient of the bed and the physical properties of the system. In addition, the high local temperature gradients and hot spots should be avoided in the core of PBR for proper design and safe operation. Hence, a hydrodynamic study related to investigation of gas phase dispersion and the extent of back mixing is very crucial for PBRs. Also, the efficiency of the reactor is dependent upon how the gas flowing through the bed is distributed; therefore, the ability to measure the gas distribution in a PBR is practically very useful in designing and operating these reactors.

Hassan and Dominguez-Ontiveros[4], measured local velocity field with particle tracking velocimetry (PTV) technique in a small sized (3cm x 3cm x 35 cm) packed bed using refractive index matching liquid. They packed the column randomly with in 4.7 mm beads. The authors conclude that the obtained data would be useful for enhancing the understanding of flow through packed bed and to be utilized in the computational fluid dynamics code validation.

In the study of Lee and Lee[5], flow field measurements were taken in a two-dimensional wind tunnel by particle image velocity (PIV) technique in the very narrow flow channel between the pebbles. Also, small sized (170mm x 170mm x 505 mm) pebble bed test section has been equipped. Even these two attempted, it still hard to say that the hydrodynamics analysis is completely understood experimentally.

In computational fluid dynamics (CFD) analysis, it is not practical to create mesh for total flow field[2], because a huge number of grids is needed to resolve the flow structure around the spheres that require huge central processing unit (CPU) computation time and memory [4].

Generally, in open literatures there are no detailed experimental measurements, knowledge and quantification of the gas dynamics and extent mixing of the gas phase in PBR. Also, most of pervious experimental studies were restricted to understand the global parameters such as pressure drop and overall voidage of the bed. In addition, the non-ideality of complex flow structure is completely understood experimentally.

However, in proceed, there are some studies related to the dispersion in packed bed have been done for both gas and liquid phases [6,7,8,9, 10,11,1]. Also, starting from the investigations of Danckwerts (1953)[12] , Bischoff and McCracken (1966) [13] and Barjaktarovic et al., (2003)[14], the back-mixing in packed columns has been extensively studied. Delgado, (2006)[15] has been summarized and reviewed the phenomenon of dispersion (longitudinal and transverse) in packed beds for a great deal of information from the literatures. The author stated that there are several variables that must be considered, in the analysis of dispersion in packed beds, like the length of the packed column, viscosity and density of the fluid, ratio of column diameter to particle diameter, ratio of column length to particle diameter, particle size distribution, particle shape, effect of fluid velocity and effect of temperature (or Schmidt number). In spite of the large number of studies in packed beds, the different techniques were not advanced in terms of time frequency of measurements, on-line conductmetric measurements and unequal pulse injection time. However, none of them have been accounted the non-ideal tracer injection. In addition, packed beds with a very low aspect ratio (tube-to-particle-diameter ratio), between 1.0 and 2.0 have been used for investigation with large wall effects.

Therefore, this study is focusing on the quantification of the gas dynamics and its extent of mixing and dispersion in the pebble bed of 0.3 m of diameter by using an advanced gaseous tracer technique. The residence time distribution (RTD) of gas phase has been characterized well to predict the pebble bed performance. Also, the non-ideal flow in the pebble bed is described in one-dimensional axial dispersion model (ADM) of one adjusted parameter. Finally, The effect of superficial gas velocity on the axial dispersion coefficient is investigated with wide range of Reynolds number (from 5 to 1100) to cover both laminar and turbulent flow regimes.

Experimental Setup and Measurements

Experiments were performed in a Plexiglas column of 0.3 m in diameter and 0.92 m in height. A schematic diagram of the gas dynamics experimental set-up containing fixed pebble bed along with gaseous tracer technique components is shown in Figure (1).

Oil-free compressed air was used as the gas phase under down flow mode, while glass bead particles of 1.2 cm diameter were used to mimic the pebbles in the bed. In PBR, helium gas flows at a very high velocity as compared to velocity of pebbles. Hence, the entire pebble bed can be assumed stationary (fixed bed) relative to the flowing gas phase. Therefore, pebbles are kept stationary (not moving) in the gas dynamics experiments representing the fixed bed of solids. Air enters the bed from the top and leaves from the bottom. The flow rate of the filtered dry air was adjusted by a pressure regulator and rotameters system, which consists of two rotameters connected in parallel. The superficial gas velocity (U_g) was varied within the range of 0.08 m/s to 0.60 m/s which covers both laminar and turbulent flow regimes based on particle Reynolds number. A plenum is required at the top to distribute evenly the supply of gas phase to the bed. Cone type plenum with 0.3m opening and 0.1 m height has been used. A plenum offering effective back mixing and less in volume is desirable. The gas distributor used is a perforated plate having 140 holes of 3 mm diameter. These holes are arranged in a square grid of 2.25 cm pitch. The opening area is 1.09% of total area. The bottom of the pebble bed consists of a plastic cone shape with an angle of 60^o horizontally and 5 cm opening.

It belongs to setup; a well designed gaseous tracer technique was used to measure the extent of the gas mixing. The gaseous tracer unit consists of gas analyzers, gas pump, and PC with data acquisition software. The gas analyze are of binary analyzer type which contains a thermal conductivity detector (TCD). Here helium gas will be used as a tracer in the air stream, TCD was found to be suitable for helium. A vacuum pump is used to draw the gas sample out of the reactor through one of the detector. The response of the detector is then amplified, converted to digital signals, and recorded as time-series data at a sampling frequency of 10 Hz. The technique is similar to one developed by Han[16] in the study carried out on slurry bubble columns. This method offers an advantage over other gas tracer techniques since it yields an accurate estimation of the RTDs of the gas phase as it accounts for the extra dispersion that occur due to the non-ideal tracer injection and extra dispersion in the plenum and sampling and analysis system which causes significant measurements errors. The tracer injection at the gas distributor, which is the input boundary for the reactor model, does not make a delta function since the gas phase undergoes mixing in the plenum. Similarity, due to the extra dispersion caused by sampling lines and analytical components, response measured by the gas detection system do not exactly represent the actual tracer response at the reactor outlet. In order to compensate for the extra dispersion effects in the distributor and plenum zone, and sampling/analytical system a convolution method was applied (Levenspiel, 1999[17]; Fogler, 2005[18]) by which the extra dispersion is added to the model predictions.

The developed tracer technique involves two injecting ports and thre sampling ports as shown in Figure (1). The tracer was injected at the center of the inlet gas line (I1) and at the bottom conical cone of bed outlet (I2), while the sampling was done at: 1) the gas inlet (S1, view A, Figure 1) close to port I1, 2) the pores of the gas distributor under plenum (S2), and 3) the neck of conical bottom cone (S3). Using the pre-mentioned injection and sampling ports, four measurements (i-iv) were conducted at each experimental condition. Table (1) shows the different ports of trace injection and gas sampling used for the four measurements and the gas dispersion effects associated with each measurement. The obtained response curves were normalized by the maximum value in each curve. Finally, gas phase axial dispersion was determined by model fitting and convolution method as discussed in the next section.

Mathematical Formulation

The experimentally obtained RTD was analyzed using one dimensional (1-D) axial dispersion model (ADM) to estimate the value of axial dispersion coefficient (D_g). The estimated value of D_g provides quantification of the extent of the gas phase mixing. It should be noted that in this work, the tracer pulse to the plenum is not considered as a delta input to the ADM. The input to the ADM, at the gas distributor boundary is the output response of the plenum to the pulse input of the tracer. The details of the model formulation are shown below:

A. Estimation of the gas mixing in the plenum and distributor zone

The gas mixing occurring in the plenum and distributor is assumed to be of continuous stirred tank reactor (CSTR) type. This will be used to provide the input for the reactor model. The impulse injection in the plenum can be expressed based on CSTR assumption as follows:

$$\frac{dC_g}{dt} = -\frac{1}{\tau_0} C_g \qquad \text{......................(1)}$$

Where τ_0 = the residence time in the plenum.
C_g= concentration of tracer in the gas phase.
The initial condition for was given by: $C=C_{inj}$ at t=0, where C_{inj} is the tracer concentration in the plenum immediately after the injection. The solution of Eqn. 1 gives the plenum output at the gas distributor as the input to the reactor model. The plenum output in a dimensionless form (C_{in}) is defined as (C/C_{inj}) which is given as below:

$$C_{in} = \frac{C_g}{C_{inj}} = e^{-\frac{t}{\tau_0}} \quad \text{……………………. (2)}$$

The unknown quantity τ_0 for CSTR model was estimated by a regression based analysis.

Measurements (i) and (ii) of Table 1, respectively, represents the dispersion occurring in the top sampling/analytical system and the total dispersion in the plenum section plus the top sampling/analytical system. For measurement (i), the gaseous tracer input profile can be considered as an ideal pulse function. This is a reasonable assumption; as the sampling tube for port S1 is placed close to the injection nozzle (Figure 1, View A). The flow of air removes the gas tracer around the nozzle almost instantaneously.

Design of sampling lines from ports S1 and S2 is such that their length and diameter are equal. This will ensure that same external volume will be offered for measurements (i) and (iii). Hence the dispersion from S1 and S2 can be considered identical and measurement (i) ($C_{(i)}$) can be used to convolute the plenum CSTR predictions (C_{in}). Then the convoluted plenum CSTR prediction (C_{in}*) was compared against the response of the measurement (ii) ($C_{(ii)}$) , where τ_0 was estimated by minimizing the averaged squared error in the time domain. As shown in Figures (2 a) and (2b), there is an acceptance fit between $C_{(ii)}$ and C_{in}* in both the laminar and turbulent flow regimes, respectively, also confirming that the plenum can be modeled as a CSTR.

B. Estimation of the axial dispersion of gas in the reactor zone

A mass balance around a differential segment of the bed, in absence of chemical reaction and radial variations yield the 1-D axial dispersion model represented by:

$$\frac{\partial C_g}{\partial t} = D_g \frac{\partial^2 C_g}{\partial z^2} - \frac{U_g}{\varepsilon}\frac{\partial C_g}{\partial z} \quad \text{………...(3)}$$

Where:
D_g = axial gas dispersion coefficient, which required to account the mixing phenomena created from a non-ideal flow.
ε = porosity of the bed (voidage)
Since there was a sufficient pressure drop across the gas distributor, and the conical bottom cone covers all the reactor cross-section at the outlet, Danckwerts boundary conditions were used for the closed-closed boundaries as:
B.C.1:

$$t > 0, \quad z = 0, \quad U_g C_{in} = U_g C_{in}\big|_{z=0} - D_g \frac{\partial C_g}{\partial z}\Big|_{z=0} \quad \text{……….. (4a)}$$

B.C.2:

$$t > 0, \quad z = L, \quad \frac{\partial C_g}{\partial z}\Big|_{z=L} = 0 \quad \text{…………………. (4b)}$$

The initial condition is given by:

$$\text{I.C} \quad t = 0, \quad 0 \le z \le L, \quad C_g = 0 \quad \text{………………………………(4c)}$$

Here C_{in} was calculated using Equation 2 with a fitted τ_0 as discussed earlier. The superficial gas velocity (U_g) is known from the pre-set flow rate and the bed voidage (ε) was measured by balance method.

The dispersion in the sampling/analytical system from port S3 was obtained by measurement (iii) (Table 1). The response of the whole system was obtained by measurement (iv). Using C_{in} obtained from Eqn. 2 as an input tracer profile, the reactor model yield an output profile (C_{out}) at the bottom level. The output profile (C_{out}) is then convoluted by $C_{(iii)}$ to yield the convoluted reactor model predictions (C_{out}^{*}) . Then convoluted reactor model predictions (C_{out}^{*}) was compared against the response of the whole system ($C_{(iv)}$), where D_g was estimated by minimizing the averaged squared error in the time domain. Figures 3a and 3b show the model fit of $C_{(iv)}$ and C_{out}* in both the laminar and turbulent flow regimes, respectively.

The calculations of D_g are repeated using a delta function as an input to ADM instead of C_{in}. This is to check the effect of extra dispersion occurring in the plenum on values of D_g. This predicts higher D_g values and suggests that ignoring the extra dispersion occurring in the plenum introduces significant error in the prediction of D_g.

Results and Discussion

As explained in previous section and shown in Figures (2) a and b, the continuous stirred tank reactor (CSTR) model has been successfully validated for the plenum and the distributor zone over a wide range of superficial gas velocities (from 0.08 m/s to 0.6 m/s). Based on the particle Reynolds number ($Re = \rho U_g d_p/\varepsilon\mu$, where, d_p is the pebble diameter, ε is the bed voidage) values, this range covers both laminar and turbulent flow regimes. According to the estimated of the plenum residence time (τ_0) obtained by minimizing the averaged squared error in the time domain, there is an acceptance fit between $C_{(ii)}$ and C_{in}* in both the laminar and turbulent flow regimes. Also, this small error between $C_{(ii)}$ and C_{in}* confirms that the plenum and distributor zone can be modeled as a CSTR. The gas phase dispersion occurring in the pebble bed has been mathematically represented by one-dimensional (1-D) axial dispersion model (ADM) over the same range of gas velocities.

In the analysis methods outlined in the previous section; the degree of mixing of the gas phase in pebble bed has been quantified using axial dispersion coefficient (D_g) which is estimated using ADM at different gas flow rates. Figures (3a) and (3b) show the model fit of $C_{(iv)}$ and C_{out}* in both the laminar and turbulent flow regimes, respectively. By minimizing the averaged squared error between $C_{(iv)}$ and C_{out}* in the time domain, axial dispersion coefficient (D_g) has been estimated. Also, Figures (5a) and (5b) illustrates the effect of the gas velocity on the gas phase axial dispersion coefficient (D_g). Peak width decreased with increasing the gas velocity in pebble bed, where the increase of gas velocity leads to an increase in the pressure drop along the bed. This leads to uniform distribution of the gas phase and hence reduction in its dispersion and back-mixing. Therefore, at high gas Reynolds number (turbulent flow), the turbulent mixing becomes the main mechanism of the dispersion in the bed compared to the molecular diffusion. The results indicate that at high Reynolds numbers deviation from the idealized plug flow is reduced in pebble beds. Hence, ADM can be used to mathematically represent the dispersion occurring in pebble bed at turbulent flow conditions. At high Reynolds numbers (turbulent flow) conditions, ADM will be suitable only for small deviation from ideal plug flow pattern. For low Reynolds number (laminar flow) conditions where the gas dispersion is relatively larger, ADM model is used at this time to characterize the dispersion by its coefficient (D_g). This will be assessed against the analysis of moments which are outlined in next section. In addition to that, mathematical approach based on stirred tanks-in-series model will be used to represent the pebble bed response where the gas phase dispersion will be accounted for by number of mixed tanks in series. This approach could be applied for conditions of low to high Reynolds numbers.

In this work, also the degree of longitudinal gas mixing in the pebble bed is described by the dimensionless particle Peclet number ($Pe=U_g d_p/\varepsilon D_g$), where, d_p is the pebble diameter, ε is the bed voidage. It has a strong influence on the performance of pebble bed that can be reached for a given mean residence time of gas coolant. Figures (4) through (6) show the variation of the axial dispersion coefficient, Peclet number and reciprocal of Peclet number (called as dispersion number) with the superficial gas velocity and with particle Reynolds number.

These Figures demonstrate that the axial dispersion decreases noticeably with superficial gas velocity and with particle Reynolds number at low range of velocities. At higher range of velocity, the decrease in the coefficient reduces with respect to the gas velocity and Reynolds number as less dispersion is encountered at these velocities. With increasing Reynolds number, the radial velocity profiles in the voids of randomly packed bed become more uniform and possibly spatially presence stagnant zones reduces. This gives rise to a small deviation from the ideal plug flow in the pebble bed at high superficial gas velocities.

Conclusions

Quantification of the gas phase mixing and dispersion in terms of axial dispersion coefficients and Peclet numbers has been done for different gas velocities. The non-uniformity of gas flow in the pebble bed has been described successfully by one-dimensional ADM at different Reynolds numbers. Also, the present work provided better understanding of the complicated dependence between non-uniformities of flow and the coefficient of axial dispersion in pebble bed.

References

1) Lee, J., Park, G., Kim, K., Lee, W., “Numerical treatment of pebble contact in the flow and heat transfer analysis of a pebble bed reactor core” Nuclear Engineering and Design 237, 2183–2196 (2007).
2) Huda, M. Q., Obara, T., “Development and testing of analytical models for the pebble bed type HTRs” Annals of Nuclear Energy, 35, 1994-2005 (2008).
3) Yanjie, L., Yong, X., Shengyao, J.,. “DEM simulations and experiments of pebble flow with mono-sized spheres” Powder Technology, 193, 312-318 (2009).
4) Hassan, Y. A., Dominguez-Ontiveros, E.E., “Flow visualization in a pebble bed reactor experiment using PIV and refractive index matching techniques” Nuclear Engineering and Design , 238, 3080-3085 (2008).
5) Lee, J-Y., Lee, S-Y., “Flow visualization in the scaled up pebble bed of high temperature gas-cooled reactor using particle image velocimetry method” J. Eng. Gas

Turbines Power, 131, 064502-064506, (2009)

6) Levenspiel, O., Smith, W.K., “Notes on the diffusion-type model for the longitudinal mixing of fluids in flow” Chem. Eng. Sci. 6, 227–233 (1957).
7) Chao, R., Hoelscher, H.E., “Simultaneous axial dispersion and adsorption in packed beds” AIChE Jornal, 12, 271–278 (1966).
8) Edwards, M.F., Richardson, J.F., “Gas dispersion in packed beds” Chem. Eng. Sci. 23,109–123 (1968).
9) Gunn, D.J., “Theory of axial and radial dispersion in packed beds” Trans IChemE. 47, T351–T359, (1969).
10) Tsotsas, E., Schlunder, E., “On axial dispersion in packed beds with fluid flow” Chem. Eng. Process. 24,15–31(1988).
11) Gunn, D.J., “An analysis of convective dispersion and reaction in the fixed-bed reactor” International Journal of Heat and Mass Transfer, 47, 2861-2875, (2004).
12) Danckwerts, P. V., “Continuous flow systems: distribution of residence times”, Chem. Eng. Sci., Vol. 2, 1-13 (1953).
13) Bischoff, K. B., McCracken, E. A., “Tracer tests in flow systems” Ind. Eng. Chem. 58, 18–31(1966).
14) Barjaktarovic, B., Sovilj, M., Popovic, S., “Hydrodynamics and axial mixing in a packed gas-liquid column” BIBLID, 34, 33-48 (2003).
15) Delgado, J., “A critical review of dispersion in packed beds” Heat & Mass Transfer, 42, 279–310 (2006).
16) Han, L., “Slurry bubble column hydrodynamics”, Washington University, St. Louis, D.Sc. Thesis (2007).
17) Levenspiel, O., Chemical Reaction Engineering, Third ed. John Wiley & Sons, New York (1999).
18) Fogler, H.S., “Elements of chemical reaction engineering” Fourth ed. John Wiley & Sons, New York (2005).

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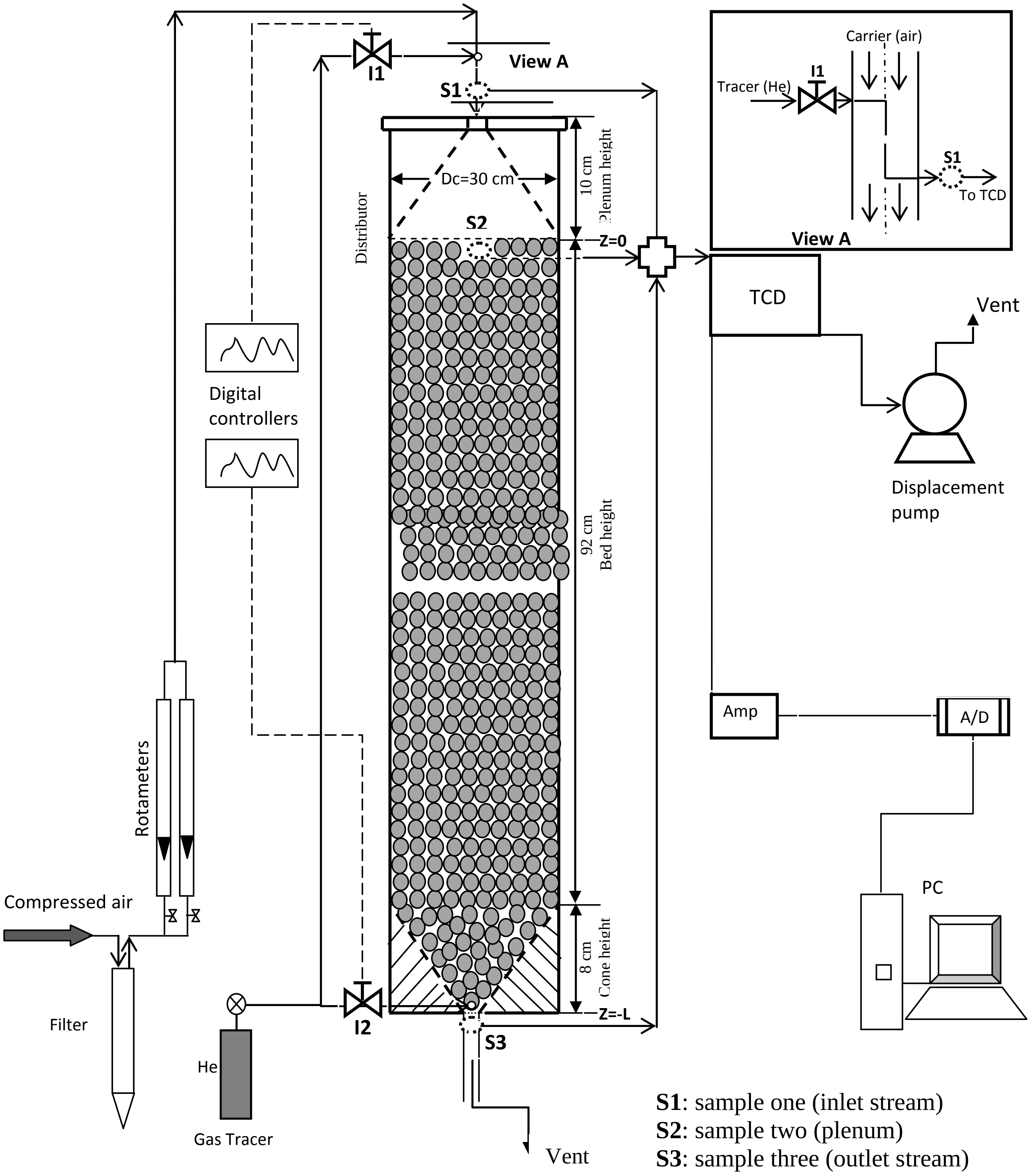

S1: sample one (inlet stream)
S2: sample two (plenum)
S3: sample three (outlet stream)
I1: injection one (inlet)
I2: injection two (outlet)

Figure (1): Schematic diagram of the advanced gas dynamics experimental set-up

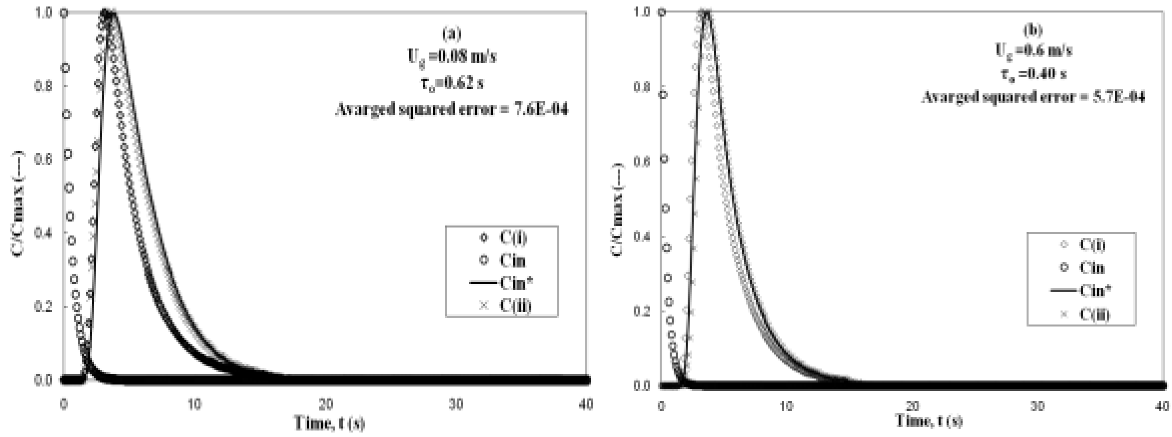

Figure (2): Dynamic of the gas tracer concentration at the plenum and distributor with CSTR model fit; a) laminar flow, and b) turbulent flow

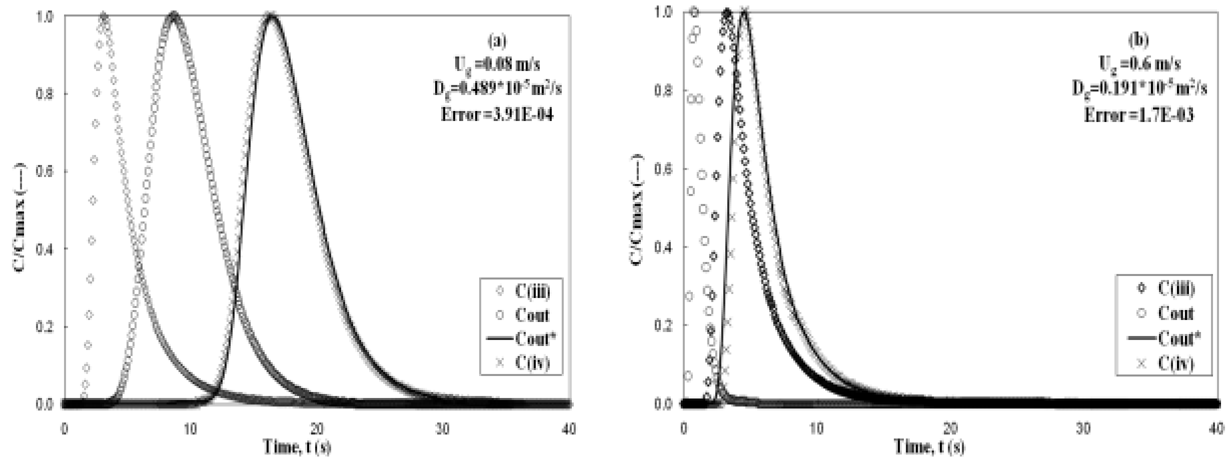

Figure (3): Dynamic of the gas tracer concentration at the reactor outlet with ADM fit; a) laminar flow, and b) turbulent flow

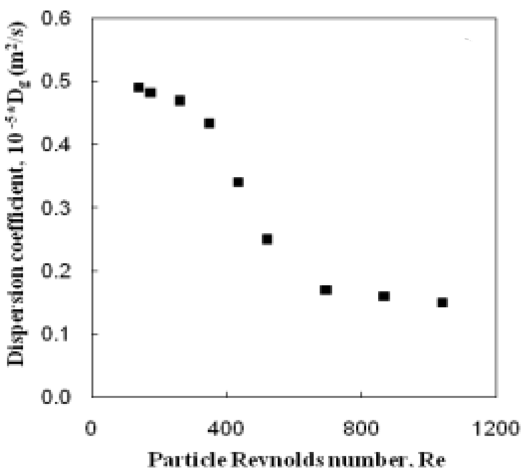

Figure (4): Variation of the axial dispersion coefficient (D_g) with particle Reynolds number

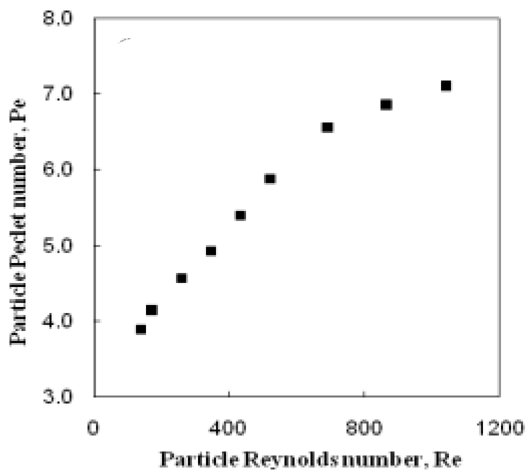

Figure (5): Variation of particle Peclet number (Pe) with particle Reynolds number

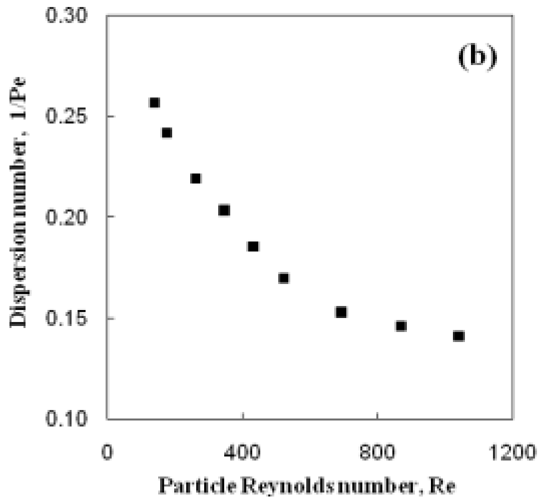

Figure (6): Variation of dispersion number ($\varepsilon D_g/U_g d_p$)with particle Reynolds number (Re)

Table (1): The designed four measurements for the gaseous tracer technique

Measurement	Tracer injection	Sampling location	Tracer concentration	Dispersion zones measured
(i)	I1	S1	$C_{(i)}$	sampling/analytical system from S1
(ii)	I1	S2	$C_{(ii)}$	plenum and distributor zone + sampling/analytical from S2
(iii)	I2	S3	$C_{(iii)}$	sampling/analytical system from S3
(iv)	I1	S3	$C_{(iv)}$	plenum and distributor zone + reactor zone +sampling/analytical system from S3

I1, **I2**: Injection ports; **S1**, **S2**, **S3**: sampling ports. All locations indicated in Figure (1)