



HOKKAIDO UNIVERSITY

Title	Motion of Solid Particles in a Riser of Circulating Fluidized Bed Systems
Author(s)	Kuramoto, Koji
Degree Grantor	北海道大学
Degree Name	博士(工学)
Dissertation Number	甲第4404号
Issue Date	1998-03-25
DOI	https://doi.org/10.11501/3137120
Doc URL	https://hdl.handle.net/2115/51472
Type	doctoral thesis
File Information	000000322310.pdf



Motion of Solid Particles in a Riser
of Circulating Fluidized Bed Systems

Koji KURAMOTO

1998

①

Doctoral Thesis

Motion of Solid Particles in a Riser of Circulating Fluidized Bed Systems

Koji KURAMOTO
March 1998

Contents

	Page
GENERAL INTRODUCTION	1
1 Fluidization	1
2 Thesis Summary	2
 Chapter 1 REVIEW OF FUNDAMENTAL STUDIES ON HYDRODYNAMICS IN CIRCULATING FLUIDIZED BEDS	 4
1.1 Gas Fluidization	4
1.1.1 Fluidization Phenomena	4
1.1.2 Stability of Fluidization	6
1.2 Circulating Gas-Fluidized Bed (CGFB) Systems	8
1.2.1 Features	8
1.2.2 Fluidization States in Riser of CGFB System	11
1.2.3 Effects of System Configuration	12
1.2.4 Slip Velocity in CGFB Riser	21
1.3 Circulating Liquid-Fluidized Bed (CLFB) System	21
1.4 Scope of Present Study	25
 Literature Cited	 26
 Chapter 2 EFFECTS OF RISER HEIGHT AND DOWNCOMER SOLID INVENTORY ON ABD DISTRIBUTION IN CFB RISERS	 28
2.1 Introduction	28
2.2 Experimental	29
2.3 Results and Discussion	31
2.3.1 Pressure Distribution in Circulating System	31
2.3.2 Change in Local Bed Density	31
2.3.3 Change in Height of Dense Region	36
2.3.4 Effect of Riser Height on Dense-Region Height	41
2.4 Conclusion	41

Literature Cited	42
------------------	----

Chapter 3 UNSTEADY STATE BEHAVIOR OF A CIRCULATING GAS-FLUIDIZED BED SYSTEM	43
--	-----------

3.1	Introduction	43
3.2	Experimental	44
3.3	Results and Discussion	44
3.3.1	Transitional Change of Static Pressures at Riser and Downcomer Bottoms	44
3.3.2	Effect of Downcomer-Bottom Pressure on ABD Distribution in Riser	56
3.4	Conclusion	59

Literature Cited	59
------------------	----

Chapter 4 HIGH-VELOCITY FLUIDIZATION OF SOLID PARTICLES IN A CIRCULATING LIQUID-FLUIDIZED BED RISER	61
--	-----------

4.1	Introduction	61
4.2	Experimental	62
4.2.1	Circulating Liquid-Fluidized Bed System	62
4.2.2	Experimental Procedure	64
4.2.3	Circulating Operation	67
4.3	Results and Discussion	67
4.3.1	Bed Expansion Characteristics	68
4.3.2	Liquid-Solid Slip Velocity	68
4.3.3	Transition of Fluidization State	73
4.4	Conclusion	81

Literature Cited	82
------------------	----

Chapter 5 CHAOTIC CHARACTERISTICS OF LIQUID-SOLID FLOW IN A CIRCULATING LIQUID-FLUIDIZED BED RISER	84
---	-----------

5.1	Introduction	84
5.2	Experimental	85
5.2.1	Probes	85
5.2.2	Data Analysis	90
5.3	Results and Discussion	93
5.3.1	Variation of Local Bed Voidage and Heat Transfer	93
5.3.2	Chaotic Analysis	99
5.4	Conclusion	108

Literature Cited	109
------------------	-----

Chapter 6 GENERAL CONCLUSION	111
-------------------------------------	-----

Nomenclature	114
--------------	-----

Acknowledgments	116
-----------------	-----

GENERAL INTRODUCTION

1 Fluidization

When fluid, i.e. gas or liquid, is supplied into a bed of solid particles supported by a horizontal porous, perforated or gauze plate in a vertical column, all the particles experience an upward drag force exerted by the fluid stream as well as the gravitational and buoyant forces. The drag force increases with the fluid flow rate leaving the latter two forces unchanged and the increase is practically observed as an increase in the pressure drop through the bed. When the pressure drop becomes sufficient to support the weight of the bed in the fluid per unit cross-section, the particles are released from gravitation and are said to be *incipiently fluidized*. At this condition, the bed gains a fluid-like properties and hence is called as a *fluidized bed*. Terminologically, a bed fluidized by gas is referred to a *gas-fluidized bed* while that fluidized by liquid as a *liquid-fluidized bed*. Further, a bed fluidized by co-current flow of gas and liquid is called as a *gas-liquid-solid fluidized* or *three-phase fluidized bed*. Such a fluid operation to fluidize a bed of particles is generally called *Fluidization*.

So far, fluidization has widely been utilized as one of the promising techniques to bring solid particles and fluid into continuous contact in a reactor. It offers a good contact of them promoting heat and mass transfers between them. In addition, continuous feed and discharge of solid particles can easily be achieved in a fluidized bed. From these advantages, fluidization has been extensively adopted in various industrial processes for catalytic as well as non-catalytic gas-solid reactions, calcination, combustion, adsorption, drying and so on. When solid particles such as catalyst and coal are continuously fed into and discharged from a fluidized bed reactor having a finite volume, the particles in the reactor can generally be assumed to be perfectly mixed. This is because rate of solid mixing is usually much higher than those of gas-solid reaction in the above mentioned processes. This rapid mixing sometimes makes the reaction operation inefficient due to a wide distribution of solid residence time and consequently a wide distribution in the extent of reaction. Circulation of particles elutriated from a reactor at high gas velocities back to the reactor has offered a practical solution to narrow the widely distributed solid residence time. The circulating system is now termed as a circulating fluidized bed (CFB) system. The main components of CFB system are generally a riser, where the solid particles are

fluidized and entrained upward at fluid velocities much higher than the terminal velocity of a single particle, and a downcomer, where the elutriated particles from the riser are stored and recycled to the riser. Though industrial applications of CFB started in 1940s and were rapidly expanded in the last two decades stimulating fundamental research, our knowledge of the motion of particles fluidized at such high fluid velocities in the riser has still been far from sufficient for reliable design of the system. In particular, some inconsistent macroscopic flow patterns have been reported by different investigators who worked with different experimental systems under different operating conditions. It is clear from literature that the flow pattern strongly depends on the scale and configurations of CFB system as well as on the operating variables. Thus, the purpose of the present study is to elucidate the solid flow patterns as a function of the above variables.

2 Thesis Summary

This thesis consists of six chapters which are summarized as follows:

In **Chapter 1**, a critical review is given for existing investigations on hydrodynamic characteristics in CFB risers to make clear the background and purpose of the present study.

Chapter 2 deals with effects of the riser height and the downcomer solid inventory in a circulating gas-fluidized bed (CGFB) system on the axial bed density (ABD) distribution in the riser and on the stability of solid circulation. Experiments were carried out using two different-height risers (5.5 m and 10 m) having an inner diameter of 0.1 m. Particles were fed from the downcomer bottom to the riser bottom through a screw feeder at various values of the solid mass flux regulated independently of the gas velocity in the riser and of the downcomer solid inventory. It is shown that the ABD distribution strongly depends on the downcomer solid inventory when the solid flux takes a critical value at a given gas velocity. Furthermore, at the critical solid mass flux and gas velocity, a significant variation of the static pressure with time has been detected at the bottom of the downcomer, which is always accompanied by that of the ABD distribution.

Hence, **Chapter 3** describes variations of the static pressure at the downcomer bottom with time, which were observed at the above-mentioned critical conditions. Their correspondence is examined from simultaneous observations of transitional changes in the ABD distribution, static pressure at the riser bottom and circulating solid mass flux in the CGFB system. It is demonstrated that the circulating system experiences a long unsteady state after all operating variables are fixed. During the unsteady state, both ABD distribution and solid mass flux are shown to vary with the difference between pressures at

the riser and the downcomer bottoms. When the pressure difference between the riser and downcomer bottoms becomes settled, a steady solid mass flux is indicated to be attainable. Even at the steady operating conditions, an appreciable change in ABD distribution with time is observed, which is attributable to the changes with the static pressure at the downcomer bottom.

As described in the preceding chapters, solid movements in the CGFB systems are affected not only by the operating variables but by the system configuration. In particular, the ABD distribution exhibits a very complicated response to the operational variables, forming various sizes of solid clusters in the lower part of the riser. To simplify such a solid movement, a circulating liquid-fluidized bed (CLFB) system was applied. The systematic observations were performed for the static pressure distribution in the riser at fairly high relative velocities of liquid to solid and the results are summarized in **Chapter 4**. Based on systematic measurements of the static pressures along the height of a riser (4.0 m high and 0.049 m i.d.) at various superficial liquid velocity and circulating solid mass flux, effects of the operating variables on variations in the bed voidage and the liquid-solid slip velocity were examined. An image analysis of photographs taken through the riser wall was also carried out to quantify fluidization states of the solid particles in terms of the power spectral density distribution and the fractal dimension.

It is demonstrated that the particles fluidized at velocities much higher than the terminal velocity of single particle exhibit to an aggregative fluidization state where the local bed voidage and the particle movements vigorously fluctuate with time. Thus, in **Chapter 5**, the complex fluctuations in the local bed voidage and the heat transfer were simultaneously measured using an optical fiber probe and a heat transfer probe and the time series data was analyzed on the basis of a deterministic chaos theory. As a result, the complexity of the voidage and heat transfer is shown to be characterized by the correlation dimension and Kolmogorov entropy for the attractor.

Chapter 6 gives general conclusions from the present study with some indications for further work needed in future.

Chapter 1

REVIEW OF FUNDAMENTAL STUDIES ON HYDRODYNAMICS IN CIRCULATING FLUIDIZED BEDS

1.1 Gas Fluidization

1.1.1 Fluidization Phenomena

The flow state of gas-fluidized bed in a vertical column considerably varies with the superficial gas velocity, u_{go} , as depicted in **Fig.1.1**. When u_{go} is fairly low, gas simply percolates through the interstice among stationary particles and the particle bed is called a *fixed bed*. At a certain value of u_{go} which gives the upward drag force equal to the downward gravitational force on the particles, the particles are suspended dynamically in the gas flow. The velocity is referred as the *minimum or incipient fluidization velocity*, u_{mf} , and the bed at u_{mf} is referred as the *minimum or incipiently fluidized bed*. For beds consisting of fine particles which belong to the so-called Geldart A group (Geldart; 1973), they expand homogeneously without forming bubbles (*particulate fluidization* or *homogeneous fluidization*) when u_{go} is increased slightly over u_{mf} . With a further increase of u_{go} , bubbles form above the distributor and rise through the bed. The velocity at which the bubbles start to appear in the bed is referred as the *minimum bubbling velocity*, u_{mb} . For the Geldart B and D group particles, the particulate fluidization state is seldom observed, i.e., u_{mb} usually coincides with u_{mf} . The size and rising velocity of bubbles depend on operating variables and particle properties as investigated by many workers. A bed is called a *slugging fluidized bed* when bubbles grow in size to the bed diameter. In these bubbling and slugging fluidized beds, solid particles and interstitial gas form a continuous emulsion phase while gas forms a dispersed phase of bubbles and slugs. A further increase of u_{go} brings about a transition from the bubbling or slugging state to a *turbulent fluidization* state. In the latter state, the boundary of bubble or slug becomes indistinct and the emulsion phase is seen to be no more continuous phase. Furthermore, the particles irregularly move upward and downward in a highly turbulent manner and vigorous pressure fluctuations are observed as a consequence of turbulent interaction between gas and particles.

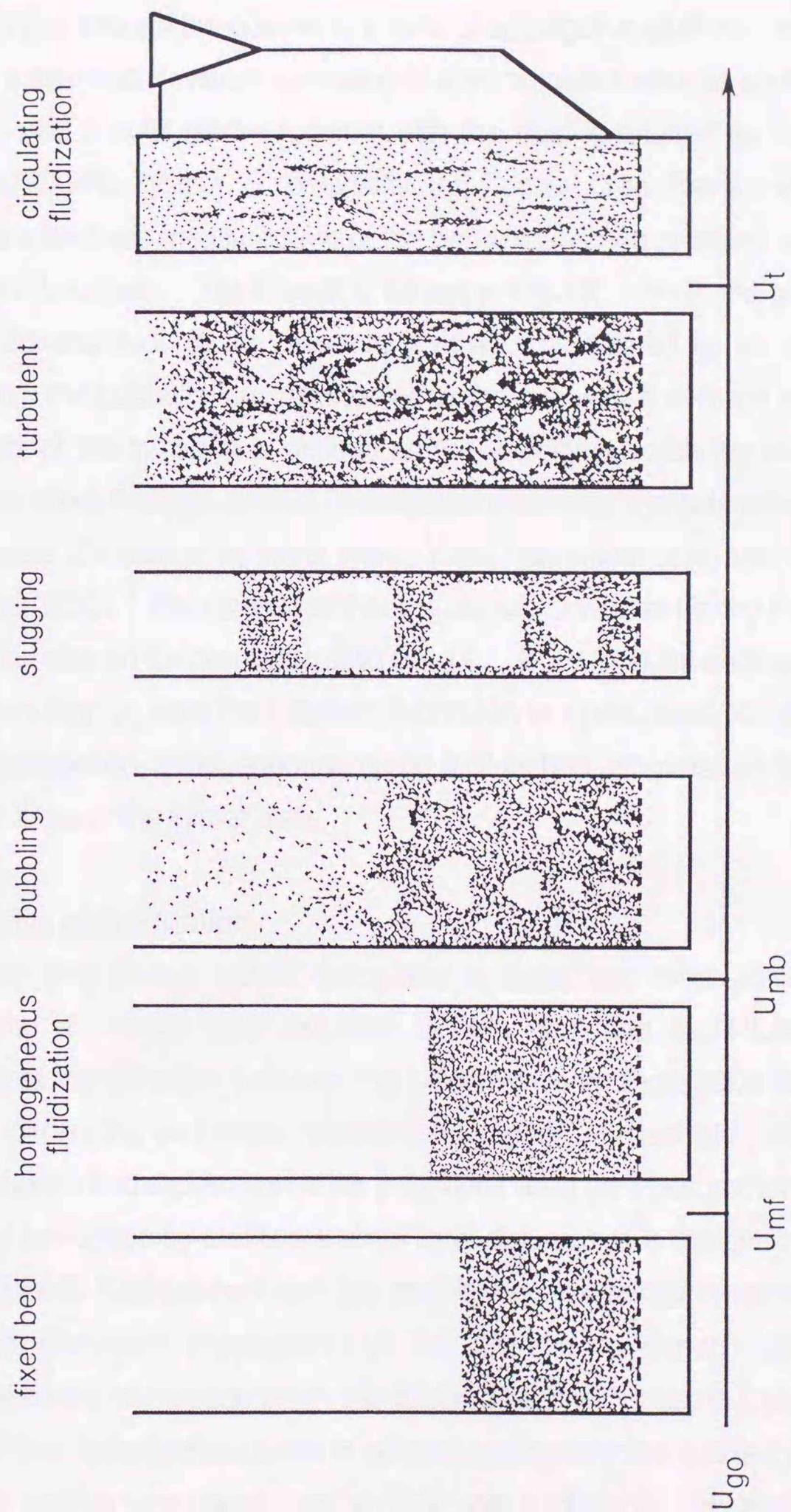


Fig.1.1 Variation of fluidization state with gas velocity

At u_{go} higher than the particle terminal velocity, u_t , most of the particles are blown away from the column and particle fluidization of the particle bed in the column becomes impossible. However, if the particles elutriated from the column are steadily recycled to its bottom, they rise through the column in a form of aggregative clusters. Here, the column is called as a riser and a system consisting of solid circulation devices such as a cyclone, a downcomer and a solid feeder together with the riser is referred as a *circulating gas-fluidized bed* (CGFB) system. Yerushalmi and Cancurt (Yerushalmi and Cancurt; 1979) first made a systematic measurement of the bed voidage and pressure fluctuations in the riser of a CGFB system. Their result is shown in **Fig.1.2**. Here, the gas velocity giving the maximum amplitude in the pressure fluctuation is defined as an onset velocity of transition from the bubbling fluidization state to the turbulent fluidization state. Also, the offset velocity of the transition is defined as the velocity at which the amplitude falls off. Following the above findings, several investigators examined characteristics of the turbulent fluidization state (Canada et al; 1978, Horio; 1986, Hashimoto et al; 1988, Nakajima et al; 1990, Bi et al; 1993). They indicated that the particle movement in the state depends not only on u_{go} but also on the circulating solid flux, G_s . In addition, another transition takes place at a very high u_{go} from the turbulent fluidization to a *pneumatic transport* state, where particles are dispersed homogeneously giving a dilute bed with pressure fluctuations much smaller than those in the former state.

1.1.2 Stability of Fluidization

Wilhelm and Kwauk (1948) conducted a systematic measurement of the bed pressure drop at various fluid velocities in both gas- and liquid-fluidized beds and suggested that the transition between the particulate and aggregative fluidization states takes place depending on Froude number ($= u_{go}^2 g^{-1} d_p^{-1}$). Jackson (1963) first made a theoretical approach to explanation of the transition based on a perturbation analysis. The analysis was performed by exerting a small local disturbance in the particle concentration onto an incipiently fluidized bed and the stability was examined in terms of the rate of upward and downward propagation of the disturbance through the particle bed. Comparing the rates of propagation in gas-fluidized beds with those in liquid-fluidized beds, he indicated that liquid-fluidized beds in general are much more stable than gas-fluidized beds so that particle beds usually exhibit particulate fluidization. He also suggested that even the liquid-fluidized beds possibly exhibit an aggregative fluidization state which is similar to the fluidization characteristics in gas-fluidized beds when heavy particles such as lead shot or copper particles are employed as a bed material. Recently, Gibilaro et al

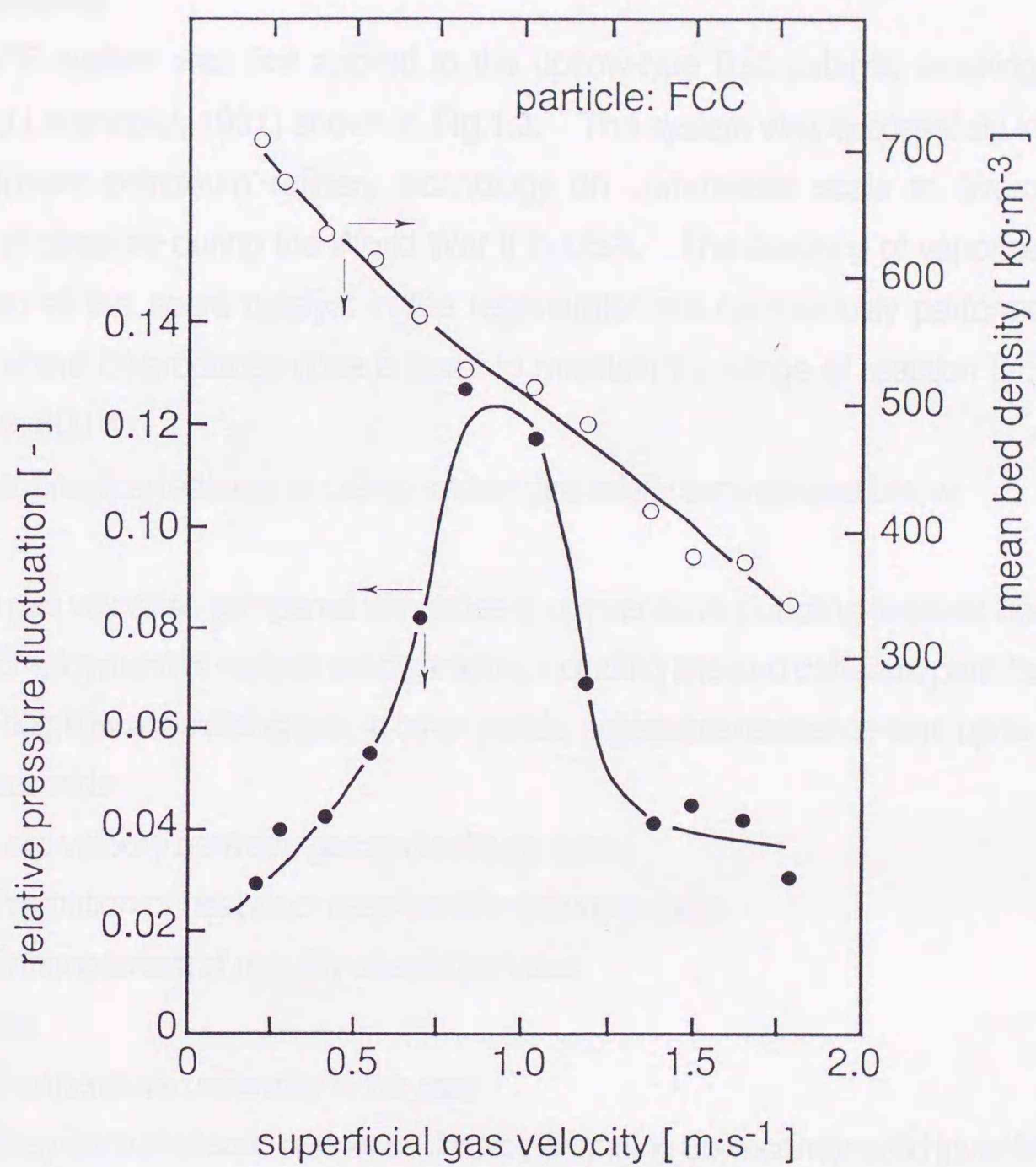


Fig.1.2 Variation of amplitude of pressure fluctuation and mean bed voidage with superficial gas velocity

(1986) semi-empirically derived a transition criteria for the fluidization states based on a dynamic wave theory (Wallis; 1969) and showed that beds consisting of fine spherical particles always exhibit the particulate fluidization.

1.2 Circulating Gas-Fluidized Bed (CGFB) Systems

1.2.1 Features

CGFB system was first applied to the upflow-type fluid catalytic cracking process (Kunii and Levenspiel; 1991) shown in **Fig.1.3**. The system was successfully introduced as an efficient petroleum refinery technology on commercial scale to overcome the shortage of gasoline during the World War II in USA. The cracking of vaporized oil and reactivation of the spent catalyst in the regenerator are continuously performed in the process, where by-produced coke is burnt to maintain the range of reaction temperature from 500 to 800 K.

Here, practical features of CGFB system are briefly summarized below:

- 1) High gas velocities compared with those in conventional bubbling fluidized beds:
 - a) Employment of various solid particles, including fine and coherent particles
 - b) Plug flow of fluidizing gas, in other words, adjustable residence time up to several seconds
- 2) High slip velocity between gas and solid particles:
 - a) Promotion of heat and mass transfer between them
 - b) Enhancement of mobility of solid particles
- 3) Others:
 - a) Temperature uniformity in the riser
 - b) Easy control of solid residence time by changing downcomer solid inventory
 - c) Staged addition of gaseous reagents at various locations
 - d) High solid loading operation

Hence, CGFB system has been extensively adopted in various industrial processes such as separation, drying, adsorption, coating, granulation, calcination, combustion and so on. The current status of industrial applications of the system is summarized in **Table 1.1**.

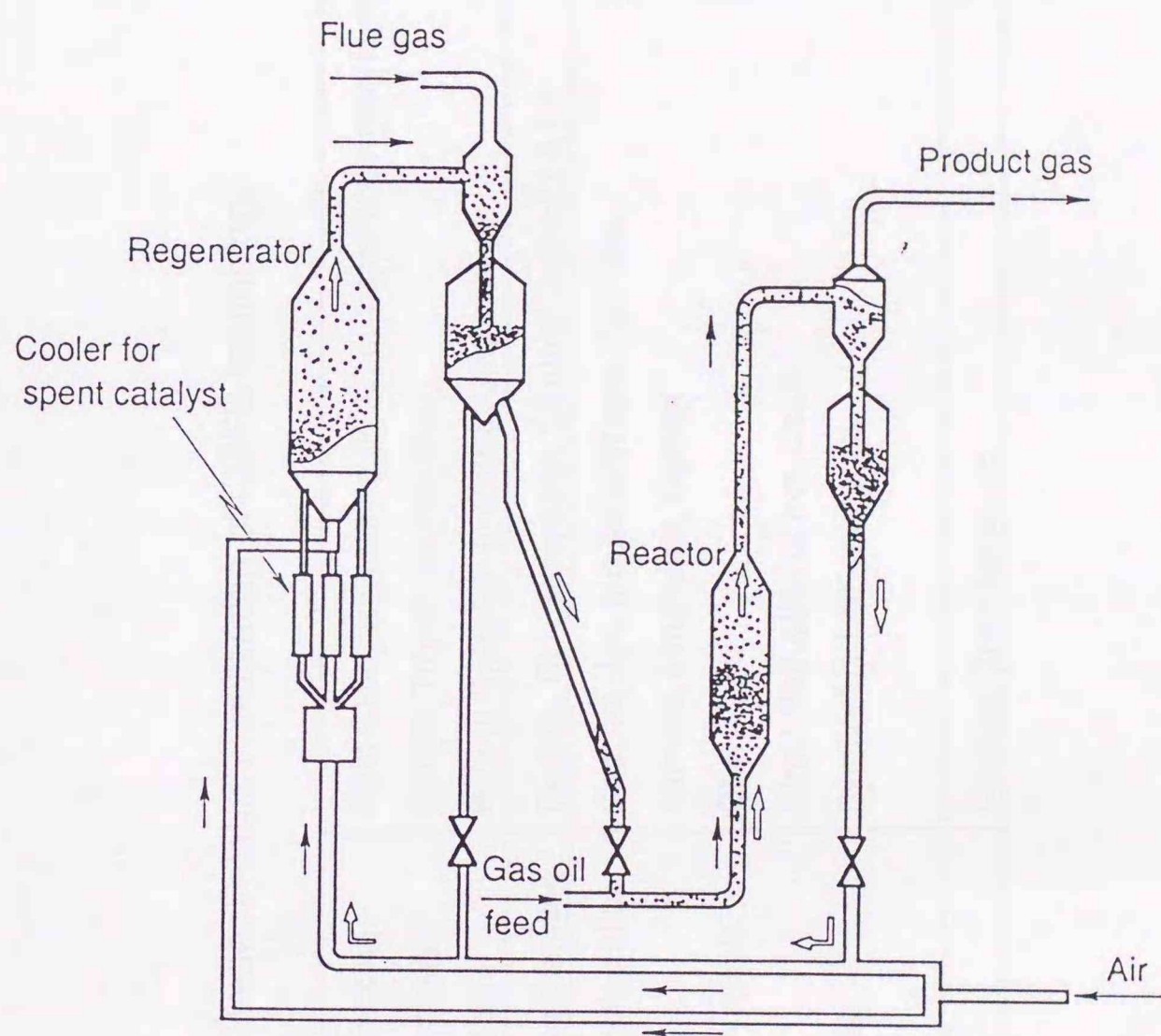


Fig.1.3 First CFB-type pilot plant for catalytic cracking (SOD Model I)

Table 1.1 Present status in utilization of CGFB technology

Gas Treatment	Catalytic Reaction	Cracking Synthesis Others	Fluid Catalytic Cracking (FCC) of Vaporized Heavy Hydrocarbon Fischer-Tropsch Synthesis etc. Butane Oxidation to Maleic Anhydride etc.
Solid Treatment	Exhausted Gas Cleaning	De-SO _x , CO ₂ Separation, Elimination of HCl & HF	
	Combustion/Incineration	Coal, Sludge, Industrial Waste, RDF etc.	
	Calcination	Alumina, Limestone, Cement	
	Gasification/Pyrolysis	Char, Coke,	
	Iron Making Material Making	Direct Reduction of Iron Oxide Synthesis of AlF ₃ and SiCl ₃	
Other Possible Utilization			Coating, Granulation etc.

In development of these processes, however, the following disadvantages have been indicated:

- a) Too highly sensitive change in fluidization states in the riser to operating variables and properties of solid and gas
- b) Necessity of high risers to give a sufficient residence time for gas and solid particles
- c) Need of efficient separation of fine particles and dust from gas flow
- d) Unknown and undesirable dependency of stable and flexible operation on system configuration

Most of the above advantages and disadvantages would reflect the complicated hydrodynamic characteristics of gas and solid particles in the riser, which are arisen from their turbulent interactions at gas velocities much higher than the terminal velocity of the single particle. However, in spite of increasing in number of industrial applications, existing methods of design, control and scale-up are purely empirical and almost covered by engineering know-how at various stages. This is due to a lack of fundamental understanding of unstable and heterogeneous nature of gas and solid flow in CGFB system.

1.2.2 Fluidization States in Riser of CGFB System

As briefly described in 1.1, solid particles are always fluidized aggregatively by gas. At low gas velocities, gas bubbles form above the distributor and rise through the bed growing by coalescence, which eventually result in bulk solid circulation and rapid solid mixing. In the riser of CGFB system, where the gas velocity is usually much higher than the particle terminal velocity, particle clusters appear near its bottom and rise at fairly high velocities. The clusters, depending on the superficial gas velocity, u_{go} , and the circulating solid mass flux, G_s , sometimes disappear along the height forming a dilute phase and consequently giving an S-shaped bed density distribution. This sort of flow pattern is affected in a highly complex manner by configuration of CGFB system as well as the above operating variables. In the last two decades, numerous investigations have been carried out to characterize the flow pattern in the riser from macroscopic and microscopic observations with various experimental methods. They are briefly reviewed in the following.

Macroscopic flow pattern in CGFB riser is to a large extent affected by u_{go} and G_s , which in turn offers a wide flexibility in handling of solid particles (Yerushalmi; 1986). So

far, several investigations (Yerushalmi and Cancurt; 1976, Li and Kwauk; 1980, Rhodes; 1989, Takeuchi et al; 1986, Horio et al; 1986) have attempted to classify the flow regime for better understanding of hydrodynamic characteristics in CGFB risers. Yerushalmi and Cancurt (1979) investigated variation of the static pressure gradient, dP/dz , with G_s at a certain height in the riser for fixed gas velocities. Typical variations are shown in **Fig.1.4**. As can be seen, a rather smooth increase of dP/dz with G_s is seen at high gas velocities while an abrupt increase appears at relatively low gas velocities. This abrupt increase reflects occurrence of transition of the fluidization state in the riser from a dilute-phase transport state to a dense-phase transport state and becomes indistinguishable at u_{go} higher than a critical velocity. This critical velocity was defined as the transport velocity, u_{tr} . Furthermore, the state of fluidization at u_{go} higher than u_{tr} was named as *fast fluidization*, where the gas-solid slip velocity varies widely with u_{go} and G_s . Li and Kwauk (1980) defined the fast fluidization state differently from the above as a state that exhibits an S-shaped axial bed-density distribution where lower-dense and upper-dilute regions are formed in the riser. Similar definitions have so far been proposed by other investigators (Rhodes and Geldart; 1986, Takeuchi et al; 1986).

Based on the above definitions of the fast fluidization state, flow regime classification maps have been constructed (Yerushalmi and Cancurt; 1979, Li and Kwauk; 1980, Takeuchi et al; 1986, Horio et al; 1986, Rhodes; 1989). They, as a matter of course, differ one another due to their inconsistent definitions of the fast fluidization state. **Figures 1.5 and 1.6** represent the classification maps proposed by Yerushalmi and Cancurt (1979) and Rhodes (1989), respectively. In the former map, u_{tr} is employed to give a boundary between the turbulent and fast fluidization states and the region of fast fluidization is classified in a range of gas velocity from u_{tr} to a certain u_{go} at which the maximum slip velocity is obtained. In the latter, on the other hand, the region is classified between the dense-phase transport and dilute-phase transport states with solid reflux.

1.2.3 Effects of System Configuration

As mentioned above, the flow patterns in the riser of CGFB system are known to depend on system configuration. Nonetheless, little attention has been paid to the dependency and there have been a limited number of investigations of its effects on the axial bed-density (ABD) distribution (Brereton et al; 1994, Zheng and Zhang; 1994, Hirama et al; 1992). **Figure 1.7** shows a representative example for effect of the shape of riser exit on the ABD distribution (Brereton et al; 1994). It is apparent from the figure that the distribution over the whole riser is considerably affected by the exit shape.

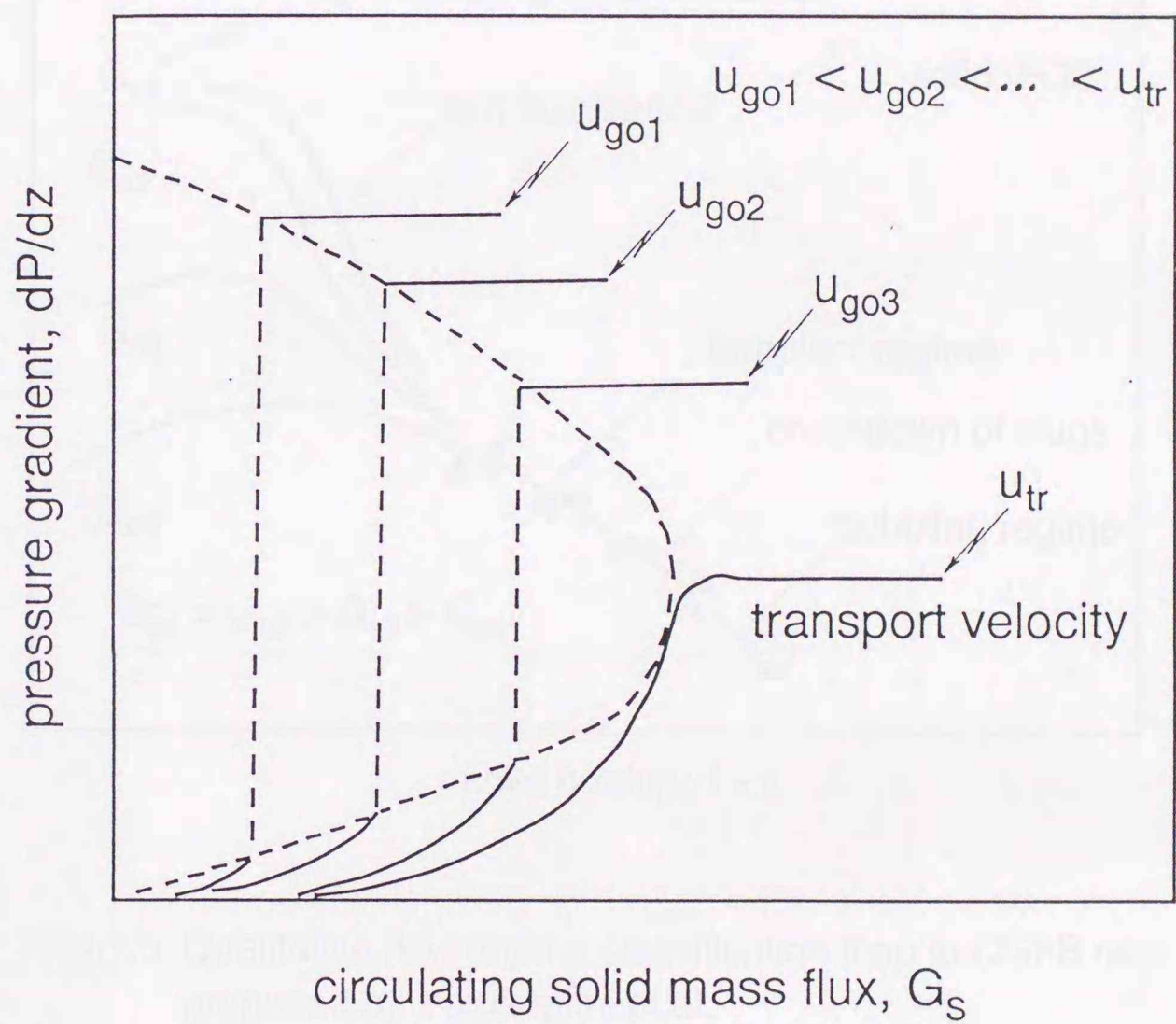


Fig.1.4 Change of pressure gradient, dP/dz , in CGFB riser with G_s for various fixed velocities

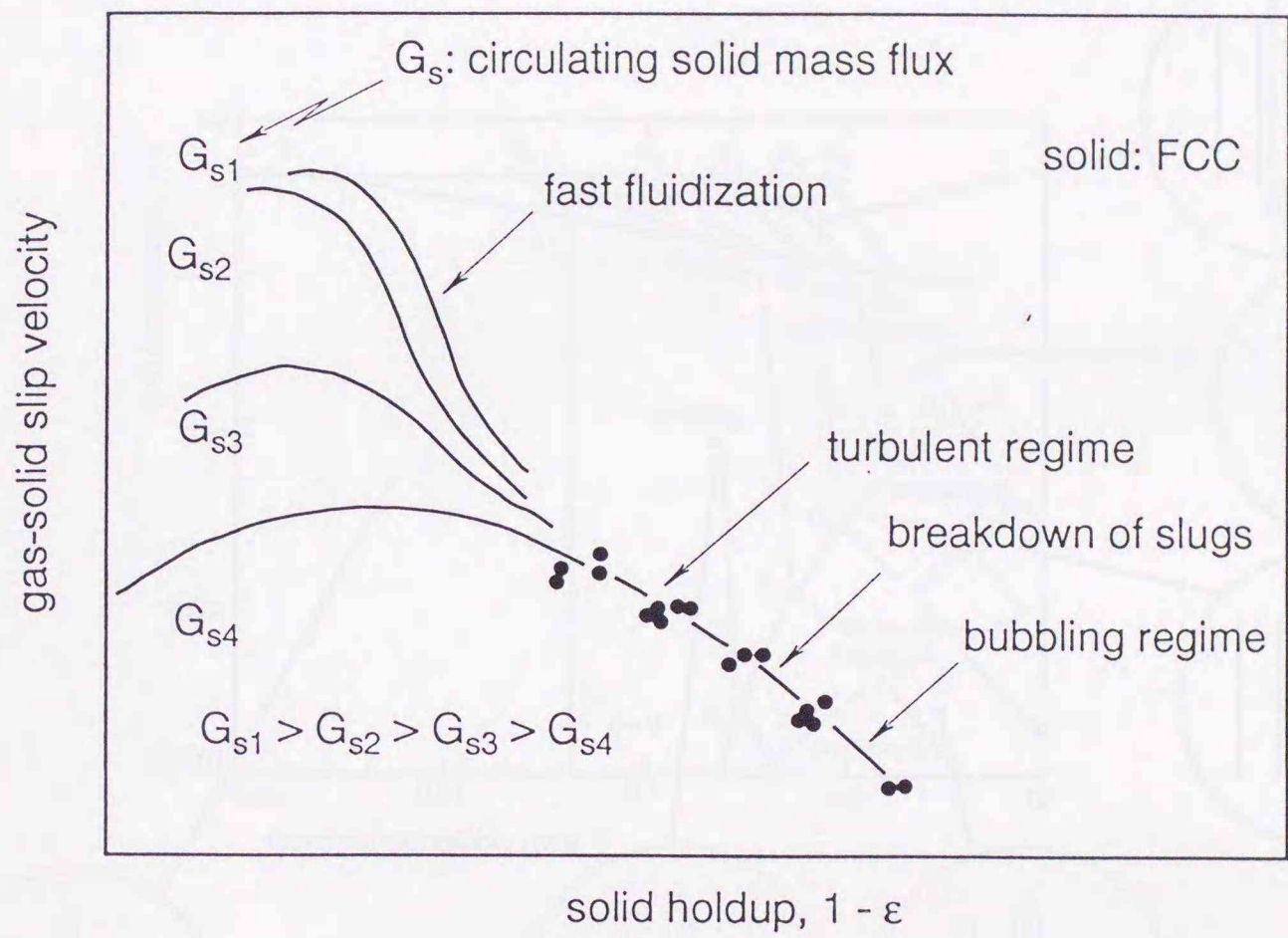


Fig.1.5 Qualitative flow regime classification map in CGFB riser proposed by Yerushalmi et al.

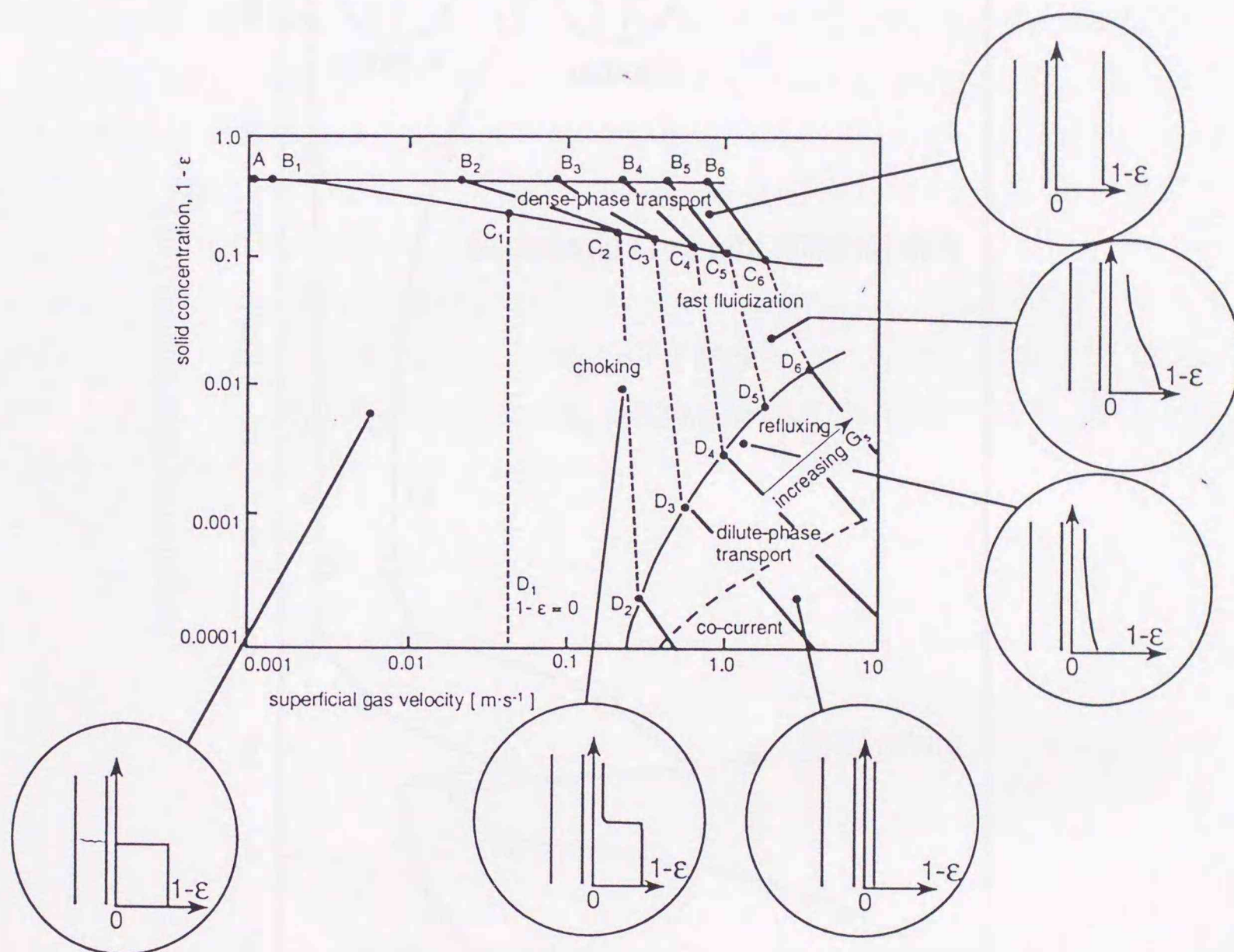


Fig.1.6 Qualitative flow regime classification map in CGFB riser proposed by Rhodes.

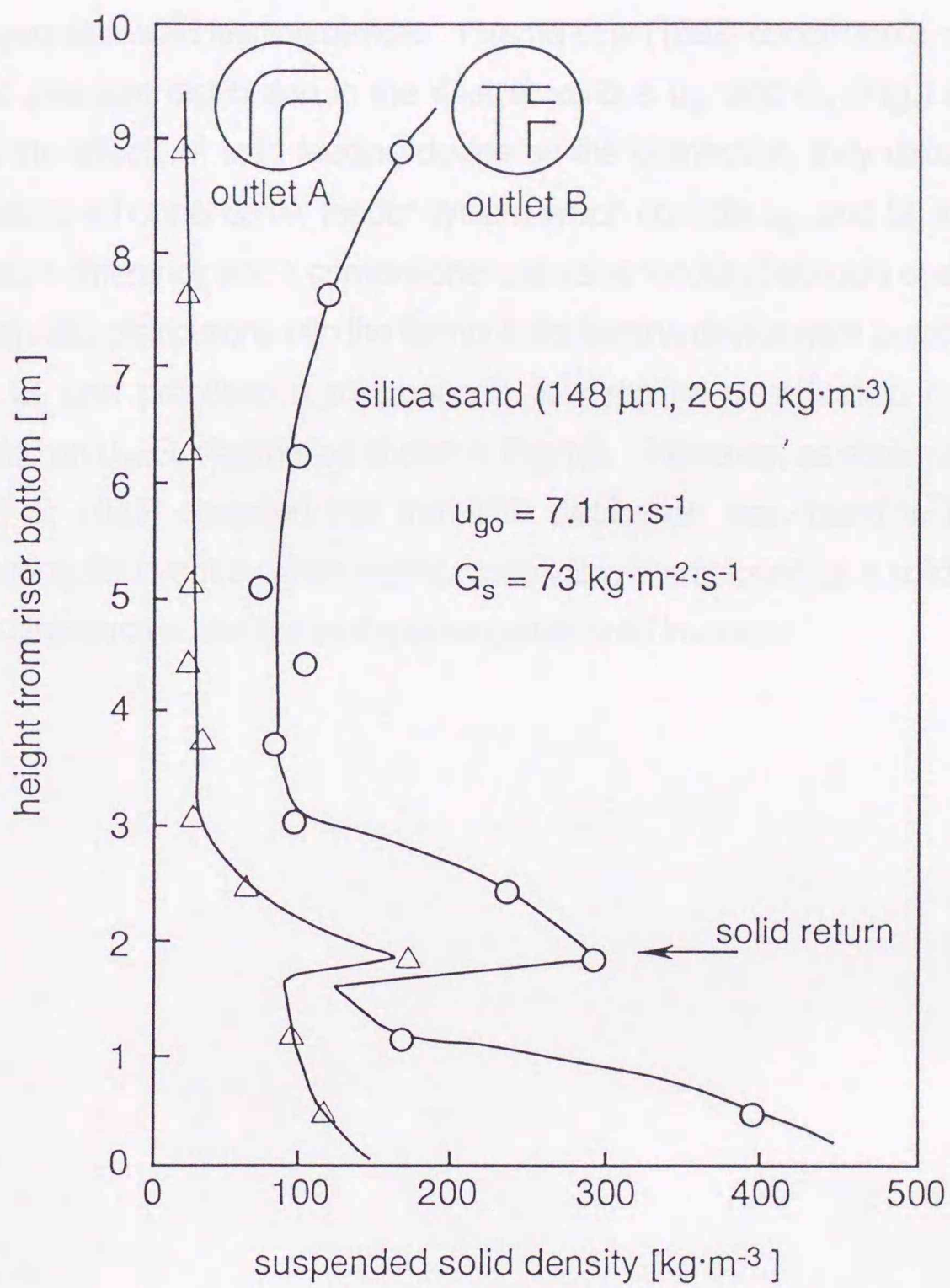


Fig.1.7 Effect of outlet configuration of CGFB riser on ABD distribution

It is also known that pressure difference between the riser and downcomer bottoms has an appreciable influence on stability of solid circulation in CGFB system (Takeuchi et al; 1986, Bi and Zhu; 1993) when a pneumatic valve like an L-shaped or a loop-seal valve is employed as a solid feeding device. Hiramata et al (1992) conducted a measurement of the static pressure distribution in the riser at various u_{go} and G_s (**Fig.1.8**). In order to evaluate the effects of solid feeding device on the distribution, they utilized two different solid feeders; a hopper-screw feeder system which controls u_{go} and G_s independently of the pressure difference and a conventional ball valve feeder (Takeuchi et al; 1986). They found that ABD distributions with the former solid feeding device were purely determined by u_{go} and G_s and proposed a macroscopic flow regime classification map, which was constructed on u_{go} - G_s diagram as shown in **Fig.1.9**. However, as shown in **Fig.1.10**, Li et al. (Li et al; 1988) observed that the ABD distribution was found to depend on the downcomer solid inventory when a pneumatic valve is employed as a solid feeder and the pressure difference is affected by the downcomer solid inventory.

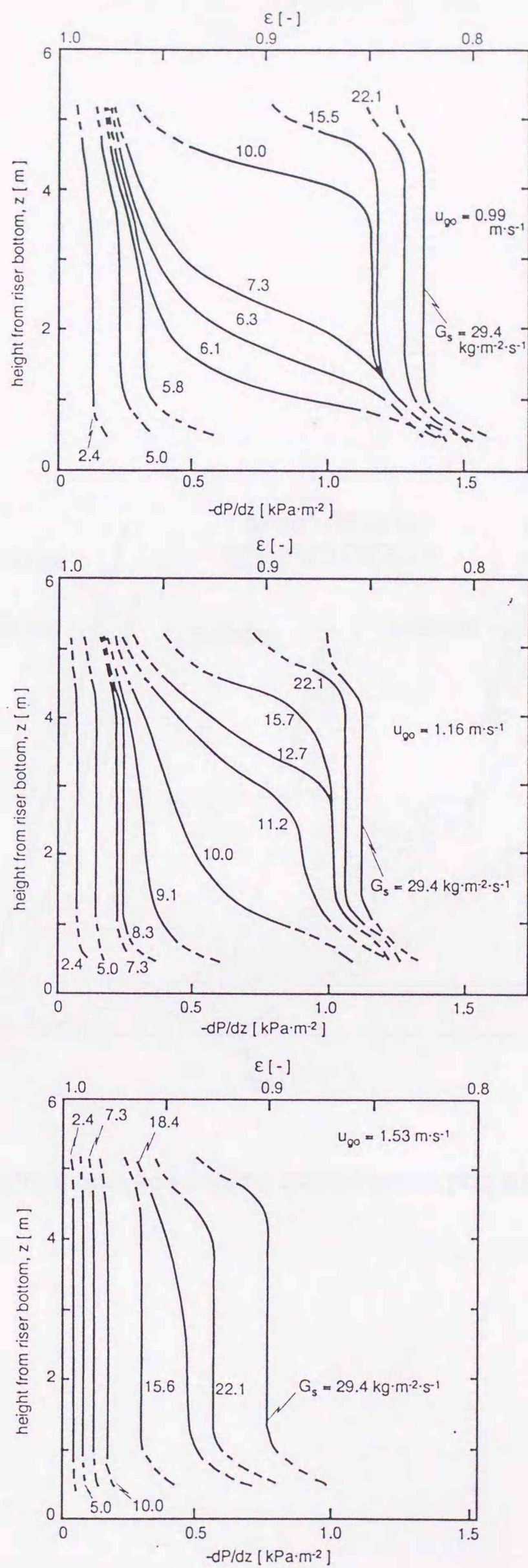


Fig.1.8 Axial distributions of static pressure gradient at various operating conditions in CGFB riser with hopper-screw feeder system

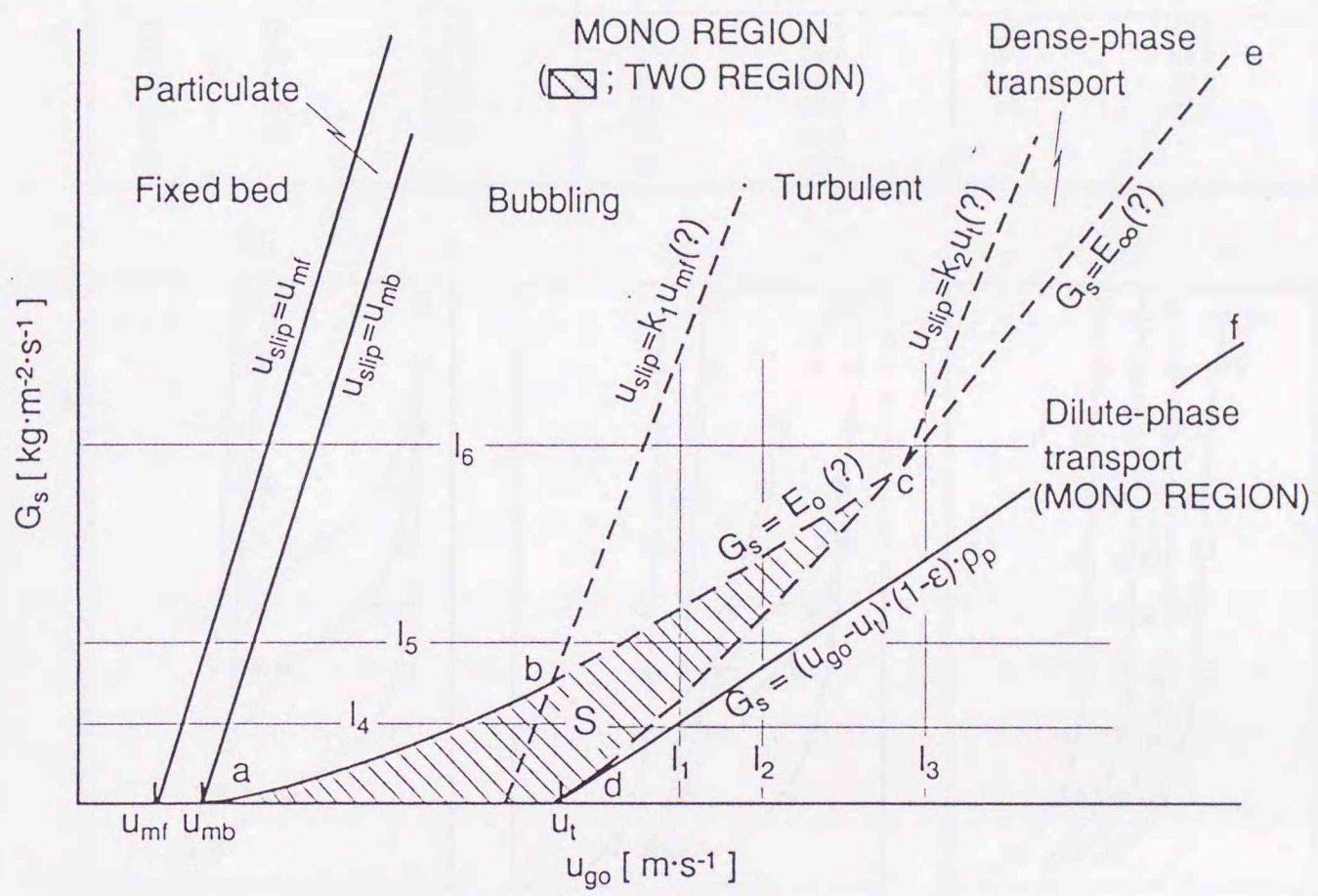


Fig.1.9 Macroscopic flow regime classification proposed by Hirama et al.

	(a) $u_{\infty} = 1.52 \text{ m}\cdot\text{s}^{-1}$		(b) $u_{\infty} = 2.10 \text{ m}\cdot\text{s}^{-1}$		(c) $u_{\infty} = 2.60 \text{ m}\cdot\text{s}^{-1}$	
	$I \text{ [kg]}$	$G_s \text{ [kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}]$	$I \text{ [kg]}$	$G_s \text{ [kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}]$	$I \text{ [kg]}$	$G_s \text{ [kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}]$
○	15	14.3	15	24.1	25	42.8
●	20	14.3	20	24.1	30	64.2
◐	25	15.2	22	24.1	35	96.3
◑	35	26.6	25	32.0	40	192.7
◊	40	48.2	35	64.2		
			40	96.3		

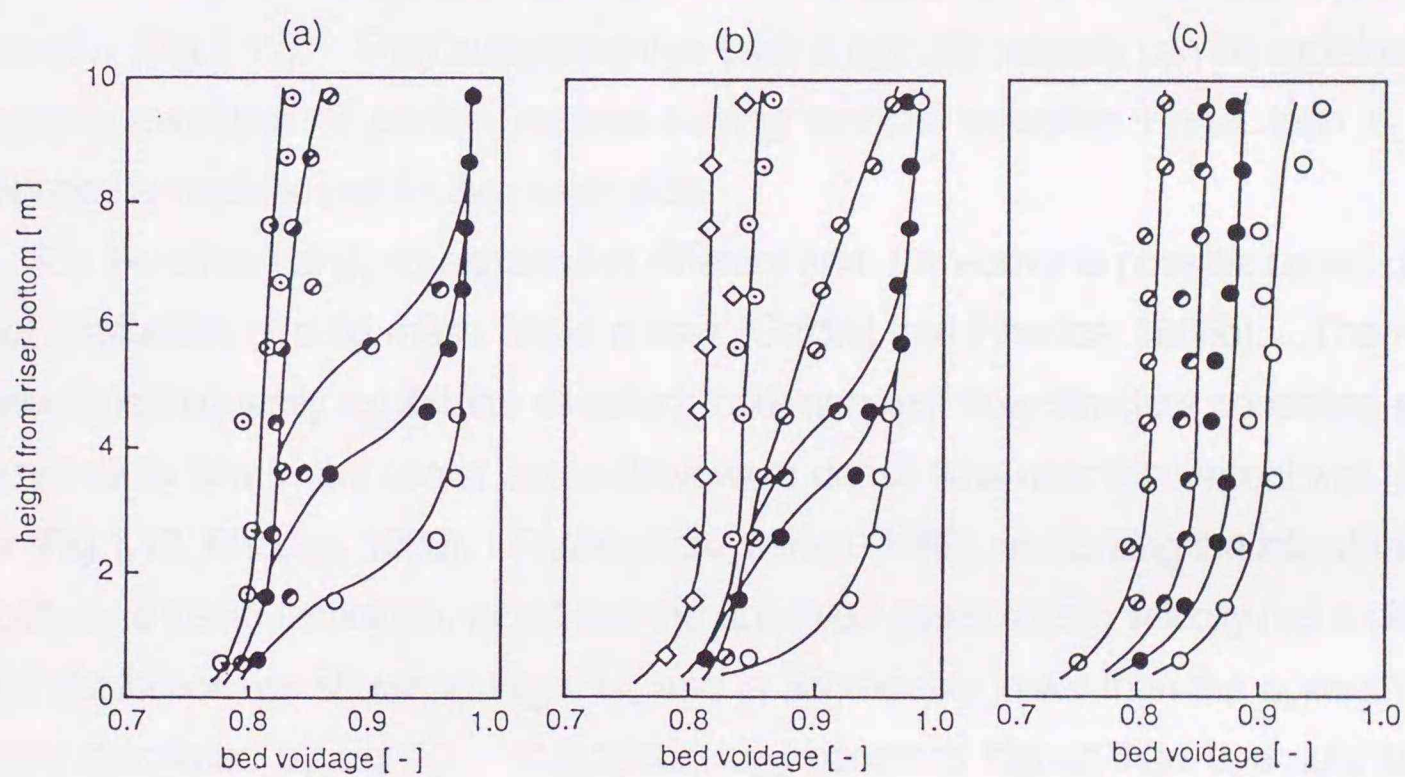


Fig.1.10 Effect of solid inventory on ABD distribution

1.2.4 Slip Velocity in CGFB Riser

One of the characteristic features of the particle flow in a CGFB riser is a high slip velocity between fluidizing gas and particles. Here, the apparent gas-solid slip velocity, u_{slip} , is expressed as:

$$u_{slip} = u_{go} / \varepsilon - G_s / \rho_p (1 - \varepsilon) \quad (1-1)$$

where ρ_p and ε respectively denote the particle density and the mean bed voidage. Lewis et al (1949) first reported u_{slip} observed at various operating conditions in a CGFB riser and showed that u_{slip} is considerably higher than the terminal velocity of single particle, u_t , when fine particles are employed as a bed material. Yerushalmi et al (1979) examined u_{slip} under a wider range of u_{go} and G_s , and found that u_{slip} possibly attains more than thirty times of u_t (**Fig.1.11**). They suggested that such a high slip velocity can be explained by assuming formation of particle clusters having terminal velocities higher than u_t and developed a model to predict the cluster size.

On the other hand, an explanation different from the above is possible based of the radial distribution of solid mass flux in a riser (Geldart and Rhodes; 1986b). The radial distributions frequently exhibit the so-called "core-annulus" flow structure consisting of an upward dilute flow in the center and a downward dense flow near the vertical wall of the riser (**Fig.1.12**, Rhodes; 1990). Furthermore, Mineo (1989), measuring the interstitial gas velocity by a tracer technique, found that the observed gas-solid slip velocity has a distinct radial distribution, as shown in **Fig.1.13**, and is appreciably lower than the apparent slip velocity calculated by Eq.1-1. In addition, u_{slip} becomes higher than u_t at any lateral position, except near the boundary between the core and annulus regions. Thus, there has been no general agreement for explanation of the high slip velocities observed in CGFB risers.

1.3 Circulating Liquid-Fluidized Bed (CLFB) System

In comparison with CGFB systems, little has been clarified of flow characteristics of solid particles fluidized with liquid at velocities higher than the terminal velocity of single particle and most of the existing findings have been obtained in batch liquid-fluidized bed system.

As mentioned in 1.1.2, solid particles are particulate fluidized by liquids unless the density difference between the solid and liquid is not so big. Particles are homogeneously suspended in liquid stream and the particle bed expands with an increase of the liquid

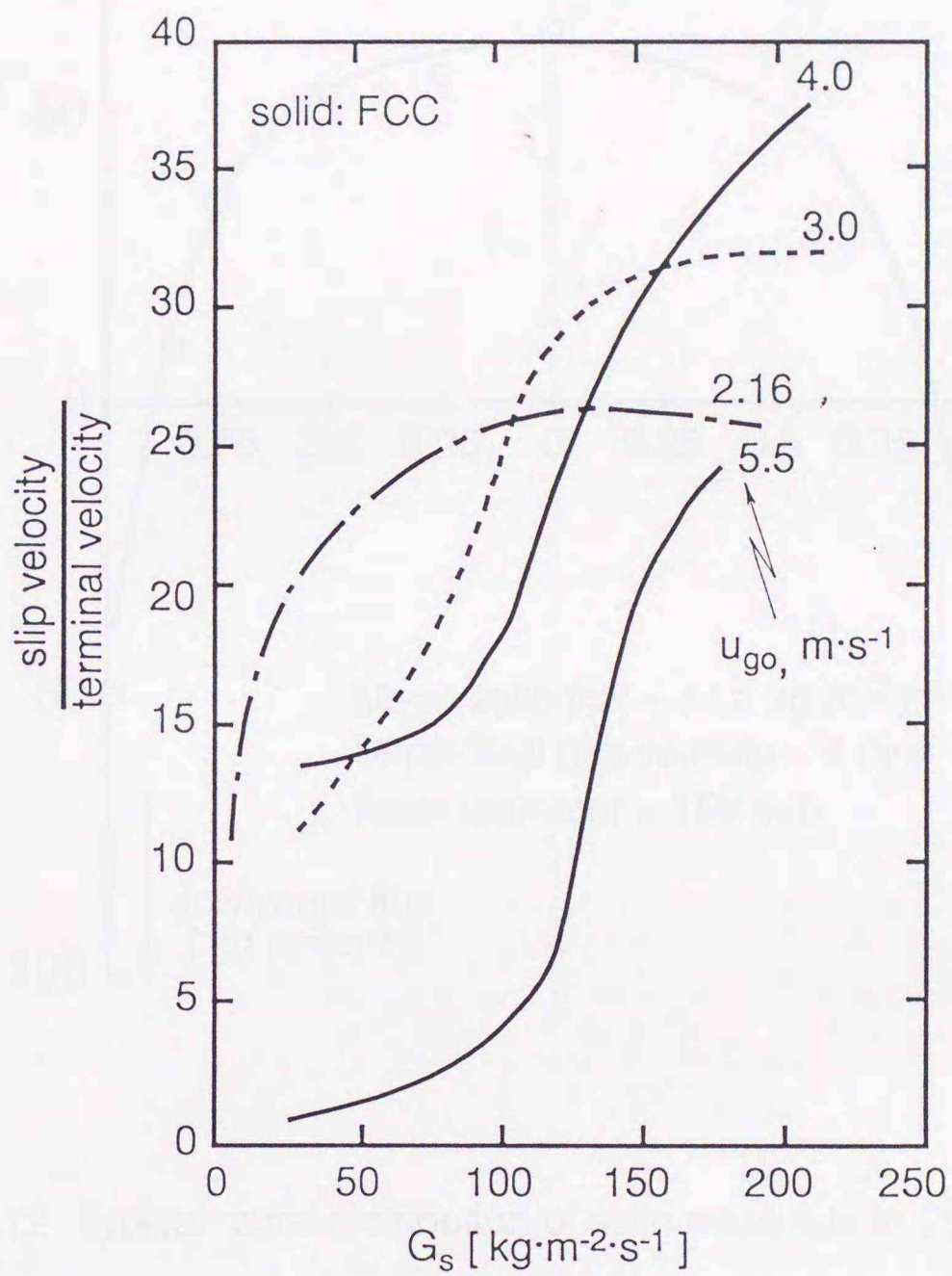


Fig.1.11 Variation of gas-solid slip velocity with u_{g0} and G_s

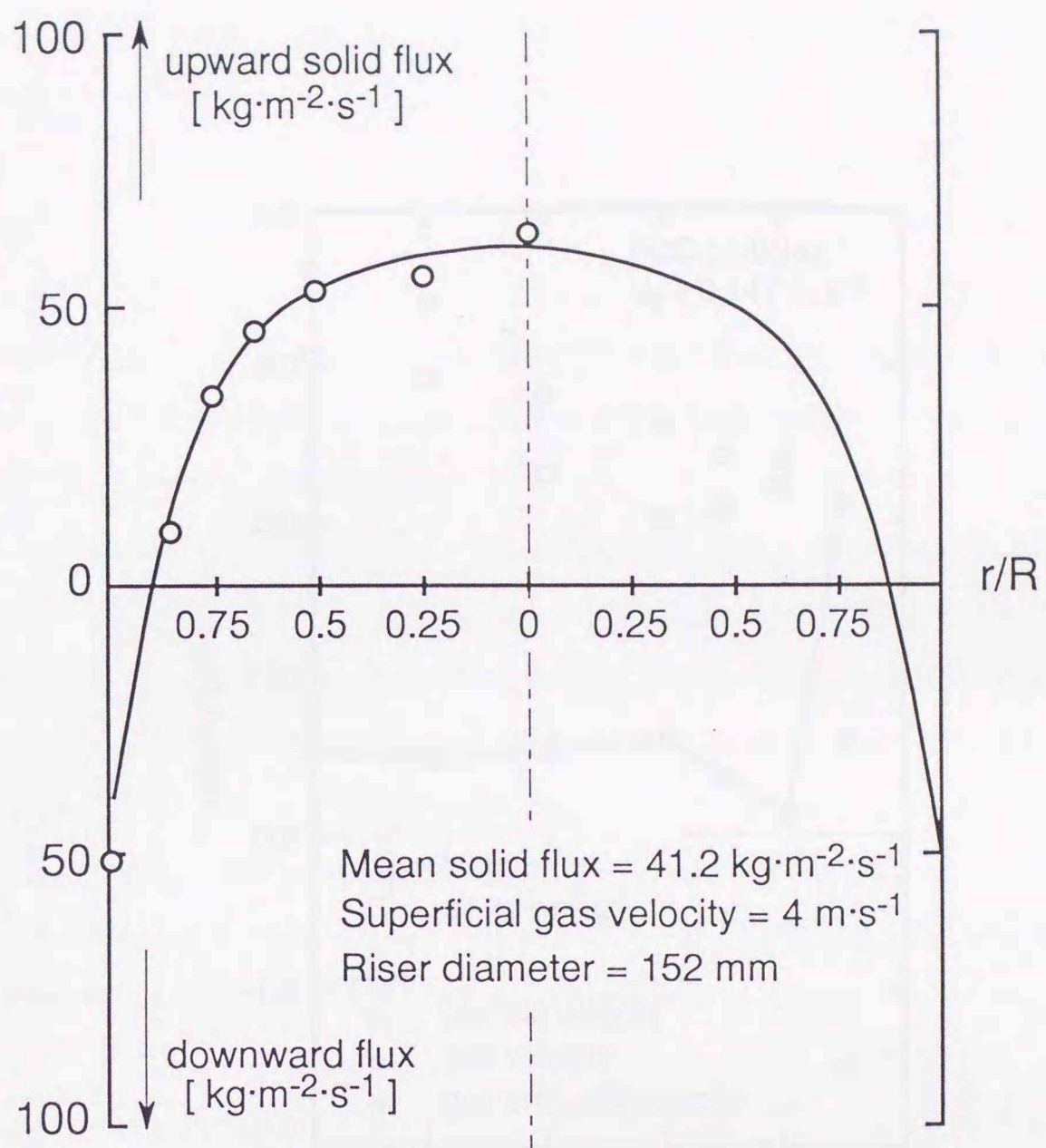


Fig. 1.12 Typical radial distribution of solid mass flux in CGFB riser

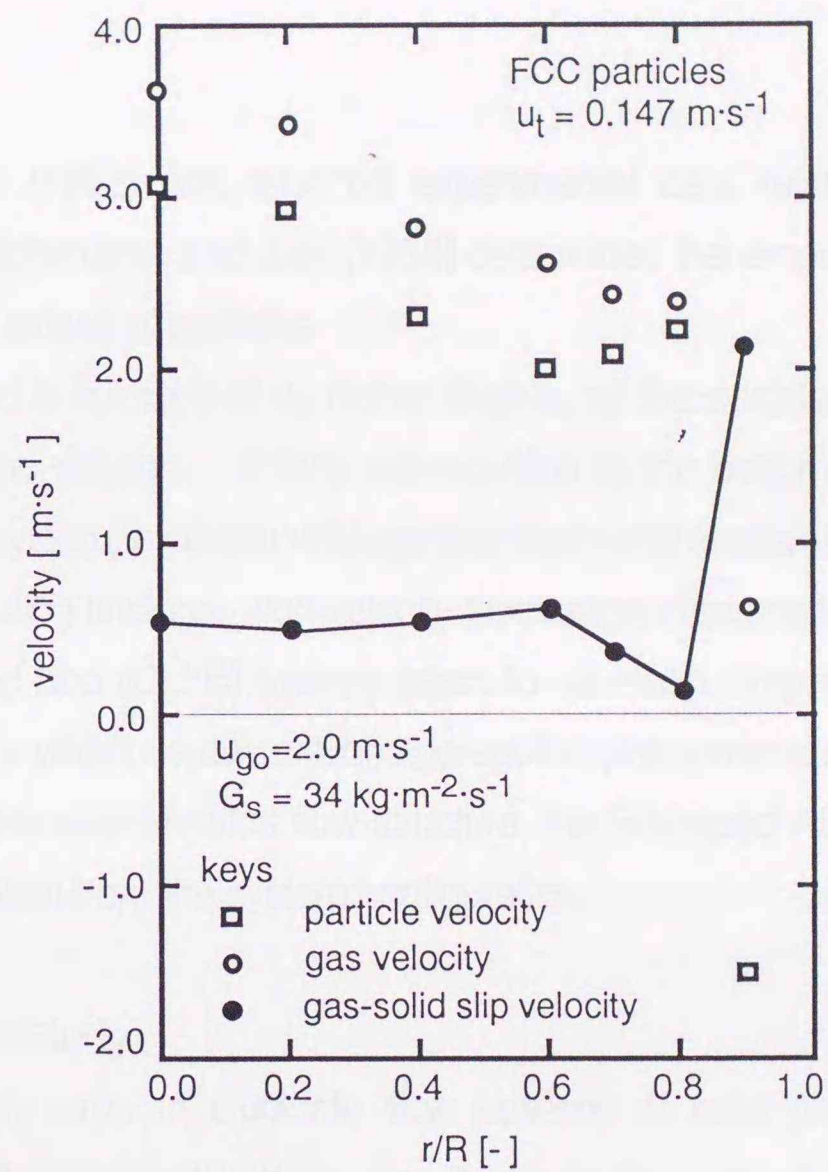


Fig. 1.13 Radial distribution of gas-solid slip velocity at $u_{g0} = 2.0 \text{ m}\cdot\text{s}^{-1}$
and $G_s = 34 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$

velocity. If the logarithms of the bed voidage, ϵ , are plotted against those of the superficial liquid velocity, u_o , a linear relationship is obtainable between them, which can be extrapolated to $\epsilon = 1$ at u_o approximately equal to u_t and generally described by the following equation:

$$u_o / u_t = \epsilon^n \quad (1-2)$$

Lewis and Bowerman (1952) first reported experimental data relating to the above relationship and later Richardson and Zaki (1954) determined the empirical constant, n , in the above equation for various of particles.

When particle bed is fluidized at u_o higher than u_t , all the particles are entrained by liquid from the top of the column. If they are recycled to the bottom of the column as usually seen in CGFB system, a certain voidage less than unity is attainable in the column. According to a few existing findings, high-velocity fluidization characteristics in the riser of circulating liquid-fluidized bed (CLFB) system seem to be much simpler and more stable than those in CGFB riser which usually exhibit aggregative phenomena such as the bubble and cluster formation, the core-annulus flow structure, the S-shaped ABD distribution and the system instability arisen from the system configuration.

1.4 Scope of Present Study

The present study aims to elucidate flow patterns of solid particles fluidized at velocities higher than the terminal velocity, i.e., those in the riser of CGFB and CLFB systems. Attention has consistently been focused to effects of operating variables such as the riser height and the initial solid inventory in the downcomer on ABD distributions, which have been obtained from measurements of the static pressures along the height of the riser and downcomer in a CGFB system having 5.5 m and 10 m high risers with an inner diameter of 0.1 m. In order to simplify the complicated and unstable flow characteristics in the CGFB system, a circulating liquid-fluidized bed (CLFB) system with a 4.0 m high and 0.049 m i.d. riser were employed and the systematic measurements were carried out for the axial static pressure distribution through the riser as well as fluctuations in the local bed voidage and conductive heat flux, all of which strongly reflect particle flow patterns in the riser.

Literature Cited

- Bi, H. and J. Zhu, "Static Instability Analysis of Circulating Fluidized Beds and Concept of High-Density Risers", *AIChE J.*, 39, 1272-1280 (1993)
- Brereton, C. M. H. and J. R. Grace, "End effects in circulating fluidized bed hydrodynamics", *Circulating Fluidized Bed Technology IV* (edited by A. A. Avidan), *AIChE*, p.137-144 (1994)
- Canada, G. S., M. H. McLaughlin and F. W. Staub, "Flow regime and void fraction distribution in gas fluidization of large particles in beds without tube banks", *AIChE symp. Ser.*, 74, 176, 14-26 (1978)
- Geldart, D., "Types of Gas fluidization", *Powder Tech.*, 7, 285-292 (1973)
- Gibilaro, L. G., I. Hossain and P. U. Foscolo, "Aggregative Behaviour of Liquid Fluidised Beds", *Can. J. Chem. Eng.*, 64, 931-938 (1986)
- Hashimoto, O., T. Haruta, K. Mochizuki, W. Matsutani, S. Mori, S. Hiraoka, I. Yamada, T. Hiramatsu, T. H., Takeuchi and T. Chiba, "Regime classification of macroscopic gas-solid flow in a circulating fluidized bed riser", *Powder Tech.*, 70, 215-222 (1992)
- Horio, M., "High velocity operation of fluidized beds", *J. Soc. Powder Tech. Japan*, 23, 80-90 (1986)
- Jackson, R., "The Mechanics of Fluidised Beds: Part I: The Stability of The State of Uniform Fluidisation", *Trans. Instn. Chem. Engrs.*, 41, 13-21 (1963)
- Kunii, D. and O. Levenspiel, "Fluidization Engineering", 2nd ed., Butterworth - Heineman, Stoneham, U.S.A., Chapter 2 (1991)
- Lewis, E. W. and E. W. Bowerman, "Fluidization of Solid Particles in liquids", *Chem. Eng. Prog.*, 48, 603-610 (1952)
- Lewis, W. K., E. R. Gilliland and W. C. Bauer, "Characteristics of Fluidized Particles", *Ind. Eng. Chem.*, 41, 1104-1117 (1949)
- Li, Y. and M. Kwauk, "The Dynamics of Fast Fluidization", in *Fluidization* (edited by J. R. Grace and J. M. Matsen), Plenum Press, p.537-543 (1980)
- Li, J., Y. Tung and M. Kwauk, "Axial Voidage Profiles of Fast Fluidized Beds in Different Regions", in *Circulating Fluidized Bed Technology II*, (edited by P. Basu and J. F. Large), p.193-203, Pergamon Press, Oxford, U.K. (1988)
- Mineo, H., "Research on Fast Fluidization" in *Doctoral Thesis of University of Tokyo*, Chapter 3 (1989)
- Nakajima, M., M. Harada, R. Yamazaki and G. Jimbo, "Bubble Fraction and Voidage in an Emulsion Phase in the Transition to a Turbulent Fluidized Bed", *Kagaku Kougaku Ronbunshu*, 16, 322-328 (1990)

- Richardson, J. F. and W. N. Zaki, "The Sedimentation of a Suspension of Uniform Spheres under Conditions of Viscous Flow", Chem. Eng. Sci., 3, 65-73 (1954)
- Rhodes, M., "Modeling the Flow Structure of Upward-Flowing Gas-Solids Suspension," Powder Tech., 60, 27-38 (1990)
- Rhodes, M. J., "The Upward Flow of Gas/Solid Suspensions Part 1: A Model of the Circulating Fluidized Bed Incorporating Dual Level Gas entry in to the riser", Chem., Eng., Res., Des., 67, 20-29 (1989)
- Rhodes, M. J., "The Upward Flow of Gas/Solid Suspensions Part 2: A Practical Quantitative Flow Regime Diagram for the Upward Flow of Gas/Solid Suspensions", Chem., Eng., Res., Des., 67, 30-37 (1989)
- Rhodes, M. J. and D. Geldart, "The Hydrodynamics of Re-Circulating Fluidized Bed", (edited by P. Basu), p. 193-200, Pergamon Press, Ontario, Canada (1986)
- Takeuchi, H., T. Hirama, T. Chiba, J. Biswas and L. S. Leung, "A Quantitative Definition and Flow Regime Diagram for Fast Fluidization", Powder Tech., 47, 195-199 (1986)
- Wallis, G. B., One Dimensional Two Phase Flow, McGraw-Hill, New York, U.S.A. (1969)
- Wilhelm, R. H. and M. Kwauk, "Fluidisation of Solid Particles", Chem. Engng. Prog., 44, 201-218 (1948)
- Yerushalmi, J., "High Velocity Fluidized Beds," in Gas Fluidization Technology (edited by D. Geldart), Chapter 7 (1986)
- Yerushalmi, J. and N. T. Cankurt, "Further Studies of the Regimes of Fluidization", Powder Tech., 24, 187-205 (1979)
- Yerushalmi, J., D. H. Turner and A. M. Squires, "The Fast Fluidized bed", Ind. Eng. Chem. Process Des. Dev., 15, 47-53 (1976)
- Zheng, Q. Y. and H. Zhang, "Experimental Study of the Effect of Exit (End Effect) Geometric Configuration on Internal Recycling of Bed Material in CFB Combustor", in Circulating Fluidized Bed Technology IV (edited by A. A. Avidan), AIChE, New York, p.145-151 (1994)

Chapter 2

EFFECTS OF RISER HEIGHT AND DOWNCOMER SOLID INVENTORY ON ABD DISTRIBUTION IN CGFB RISERS

2.1 Introduction

As introduced in the previous chapter, axial bed density (ABD) distributions in a CGFB riser have been measured by many investigators and the flow regime classification maps have been proposed on the basis of the variation of ABD distributions with u_{go} and G_s . This is because ABD distributions is obtained as a result of macroscopic flow characteristics which directly reflect fundamental characteristics of fluidization state of solid particles at high gas velocities. ABD distributions have been found to depend not only on operating variables but also on configuration of the solid circulation system. However, in existing investigations, little attention has been focused to the latter effect.

Most of ABD distributions in literature were obtained in CGFB risers where the downcomer bottom is simply connected to the riser bottom with L-shaped or loop seal valve which usually has a pressure drop comparable to or smaller than that through the riser and the downcomer. Thus, with such a solid feeding device, pressure interaction would inevitably be induced between the riser and downcomer, resulting in ABD distributions which more or less depend on experimental system configuration. For instance, some investigators reported that ABD distributions vary appreciably with the weight of solid initially loaded in the downcomer (downcomer solid inventory) even at a given set of the gas velocity, u_{go} , and the circulating solid mass flux, G_s , (Weinstein et al; 1983, Li et al; 1988, Mori et al; 1989). However, it has never been made clear the correspondence between the distribution and the pressure interaction.

Hirama et al (1992) made a systematic measurement of ABD distribution in an experimental CGFB riser having a hopper-screw feeder with which u_{go} and G_s were independently controlled suppressing pressure interaction between the riser and the downcomer bottoms. The results evidenced from the measurement are as follows:

- 1) S-shaped ABD distributions as a result of formation of the lower-dense and upper-

dilute regions are obtained at relatively low u_{go} and G_s but are never observed when u_{go} and G_s are higher than the so-called "transport velocity" and corresponding circulating solid mass flux.

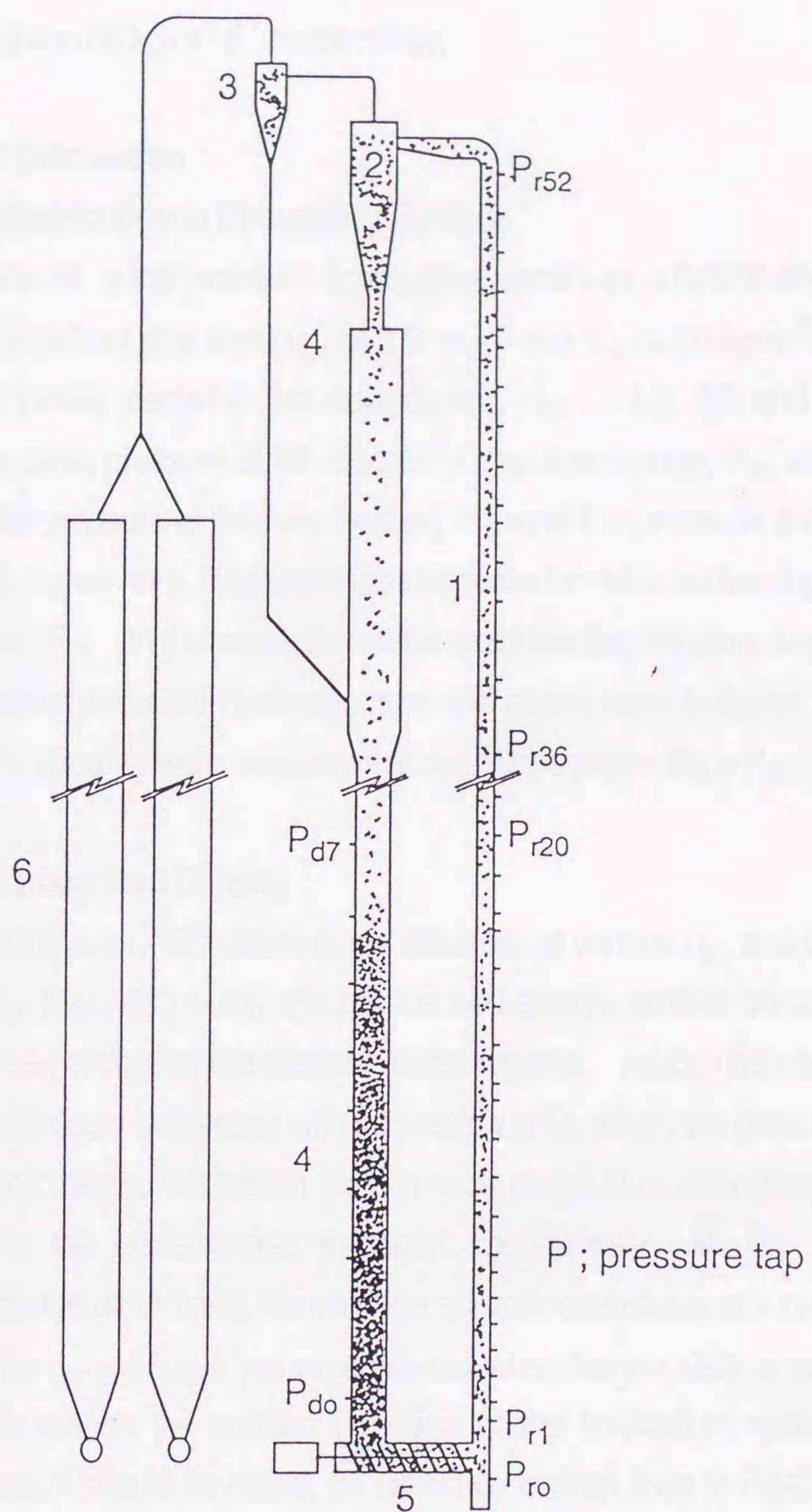
- 2) ABD distribution is essentially independent of the downcomer solid inventory when no pressure interaction exists between the riser and the downcomer.

Based on these findings together with existing models for solid flow in the freeboard of bubbling fluidized beds, a conceptual flow regime diagram was constructed simply as a function of u_{go} and G_s . However, their ABD distributions were measured at a fixed downcomer solid inventory and its effects on their diagram were not examined. Thus, in this chapter, the effect of the pressure interaction on ABD distribution is examined on the basis of systematic observations for a range of the downcomer solid inventory.

2.2 Experimental

Figure 2.1 illustrates a schematic diagram of CGFB system employed here. Two different-height risers, CGFB-55 and -100, were employed which have respective heights of 5.5 m and 10.0 m. Both risers were made of a transparent PMMA-resin column having an inner diameter of 0.10 m. Pressure taps were attached vertically on the riser wall at intervals from 0.15 to 0.55 m for measurements of the static pressure distribution in the riser. As a moving-bed downcomer a 0.15 m-dia. column was used for CGFB-55 system while for CGFB-100 system a 0.20 m-dia. column was used. Both of the columns have pressure taps attached on their wall. Each downcomer bottom was simply connected with the riser bottom by a screw feeder. The feeder controls the circulating mass flux of solid particles by adjusting revolution speed of the screw. The fluctuation in the solid feeding rate per revolution was less than 1 % of the average rates. Differential pressure between two different vertical positions in the riser were measured by water manometers. The measurements of the pressure distribution were conducted at steady state which was confirmed by measuring the total pressure drop through the riser by an electric pressure transducer.

All experiments were carried out at ambient conditions using air as a fluidizing fluid. The bed material used was FCC particles with a median of 54 μm in the particle size distribution and with a mercury-porosimetric density of approximately 810 $\text{kg}\cdot\text{m}^{-3}$. The superficial gas velocity and the circulating solid flux in the present experiments ranged from



- | | |
|--|-----------------|
| 1. riser (5.5 & 10m height, 0.1m i.d.) | 4. downcomer |
| 2. primary cyclone | 5. screw feeder |
| 3. secondary cyclone | 6. bag filter |

Fig.2.1 Schematic diagram of experimental circulating gas-fluidized bed system

1.3 to 3.0 m·s⁻¹ and 4 to 60 kg·m⁻²·s⁻¹, respectively.

2.3 Results and Discussion

2.3.1 Pressure Distribution in Circulating System

Figure 2.2 shows typical pressure distributions observed in CGFB-55 system. The distributions were obtained at a fixed u_{go} of 1.3 m·s⁻¹ and G_s of 20 kg·m⁻²·s⁻¹ for different static bed heights initially loaded in the downcomer, H_{sd} , of 1.8, 2.3 and 2.7 m. With increasing H_{sd} , the static pressure at the bottom of the downcomer, P_{db} , simply increases without affecting the pressure at the riser bottom, P_{rb} , and the pressure distributions in the riser. Though not shown here, fluctuation was observed in the solid feeding rate when P_{db} became higher than P_{rb} . In this case, the stable solid feeding became impossible due to the significant pressure interaction between riser and downcomer bottoms. Hence, ABD distributions shown hereafter were measured at conditions where $P_{rb} > P_{db}$.

2.3.2 Change in Local Bed Density

Figure 2.3 compares ABD distributions obtained at various u_{go} and G_s in CGFB-55 riser. At higher u_{go} (Fig.2.3(c) & (d)), the particle bed always exhibits homogeneous ABD distribution in the riser, except for the inlet and outlet regions. As can be seen in the figure, bed voidage monotonically decreases with increasing in G_s within the present range of G_s . This clearly indicates that the fluidization state in such range of u_b changes from the dilute-phase transport to the dense-phase transport continuously with G_s . At lower u_{go} (Fig.2.3(a) & (b)), on the other hand, the voidage abruptly decreases at a particular G_s (e.g. $G_s = 20$ kg·m⁻²·s⁻¹ for $u_{go} = 1.5$ m·s⁻¹) at which so-called S-shaped ABD distribution appears in the riser. This reflects the sudden transition of the fluidization state occurs at the particular G_s . Here, it should be noted, as drawn by broken lines in Fig.2.3(a), that ABD distribution transitively changed with time even at fixed u_{go} and G_s when G_s takes a particular value and the unsteady phenomena have been discussed later. Figure 2.4 shows ABD distribution in CGFB-100 which were obtained at the same operating conditions for CGFB-55 (Fig.2.3). It becomes clear from the figure that the ABD distribution becomes homogeneous at higher u_{go} while the fluidization state suddenly changes with increasing in G_s at lower u_{go} , as observed in the lower riser. The transition in the macroscopic fluidization states is further examined from the variation of the local pressure gradient at a certain height of the riser. In Fig. 2.5, the local bed density in terms of static pressure gradient, $-dP/dz$, at a certain height, z , of the risers, i.e., $z = 2.4$ m for

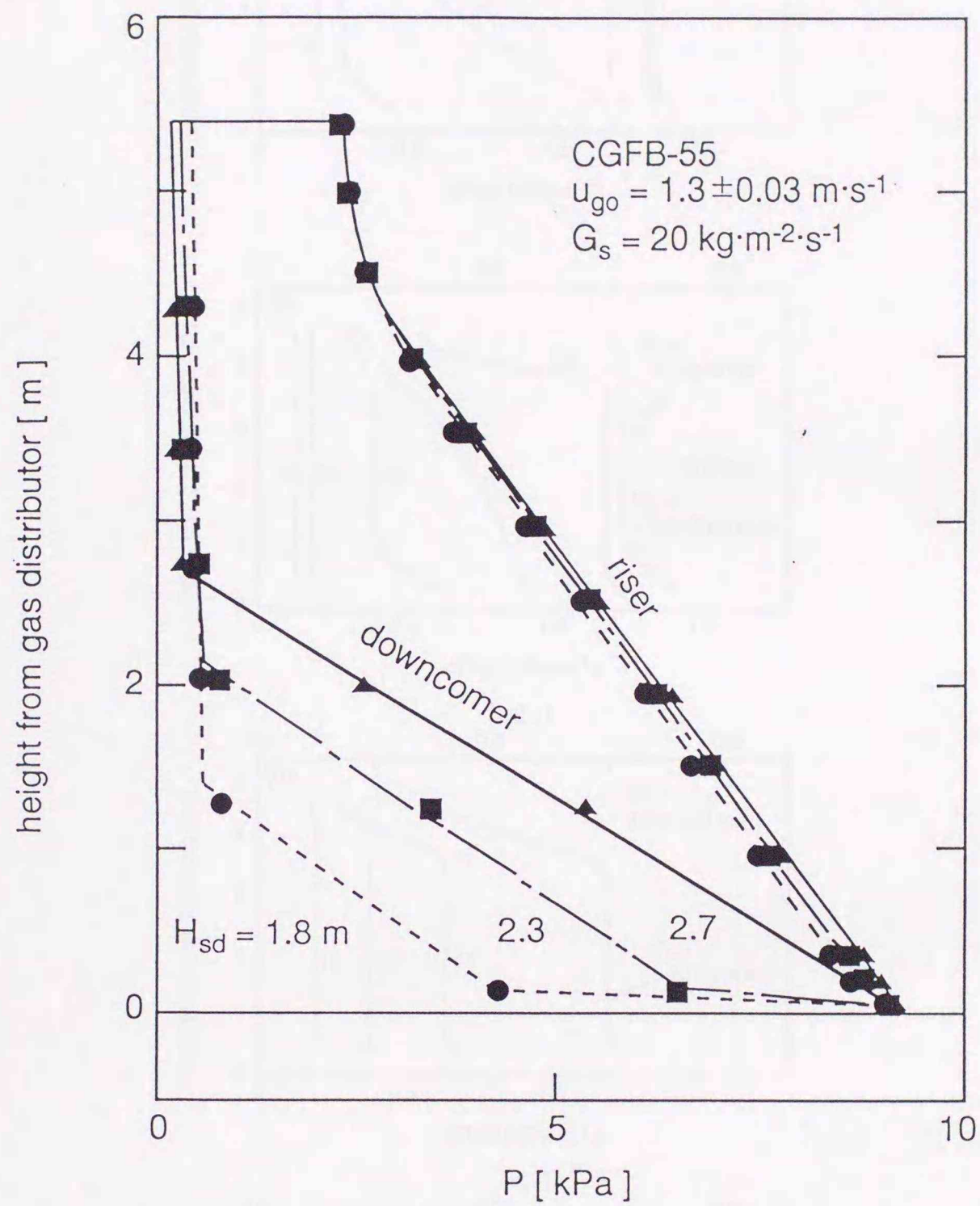


Fig. 2.2 Typical pressure distribution at $u_{l0} = 1.3 \text{ m} \cdot \text{s}^{-1}$
and $G_s = 20 \text{ kg} \cdot \text{m}^{-2} \cdot \text{s}^{-1}$ for different H_{sd}

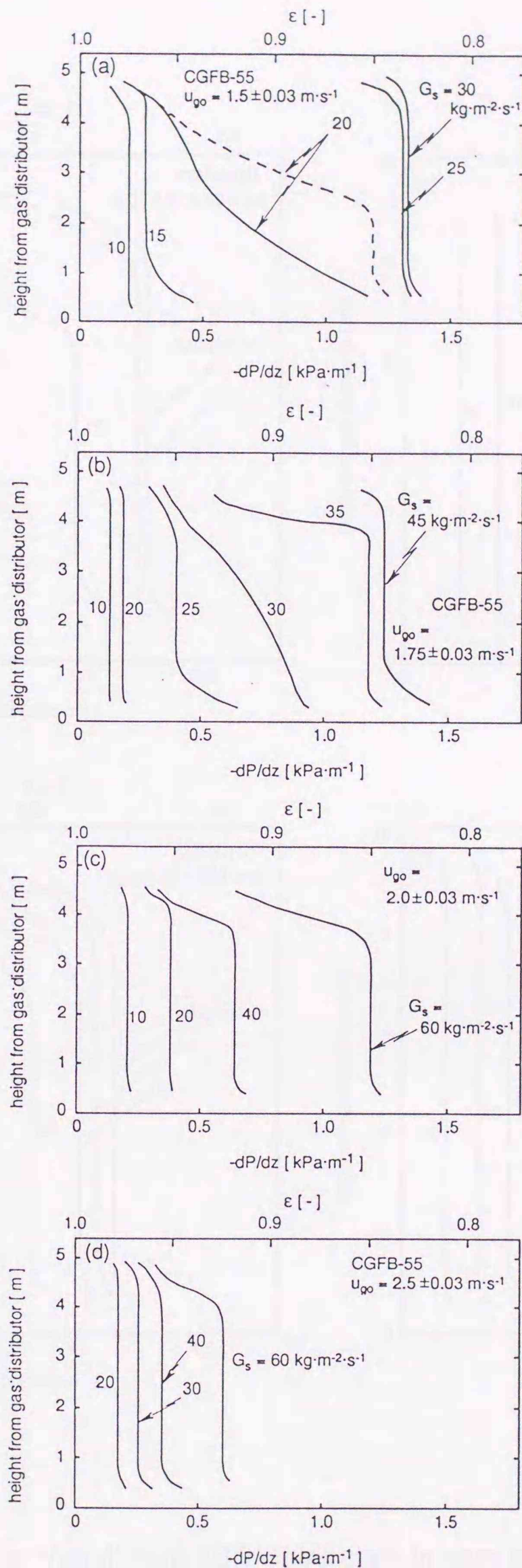


Fig. 2.3 Variation of ABD distribution in riser of CGFB-55 at various operating conditions

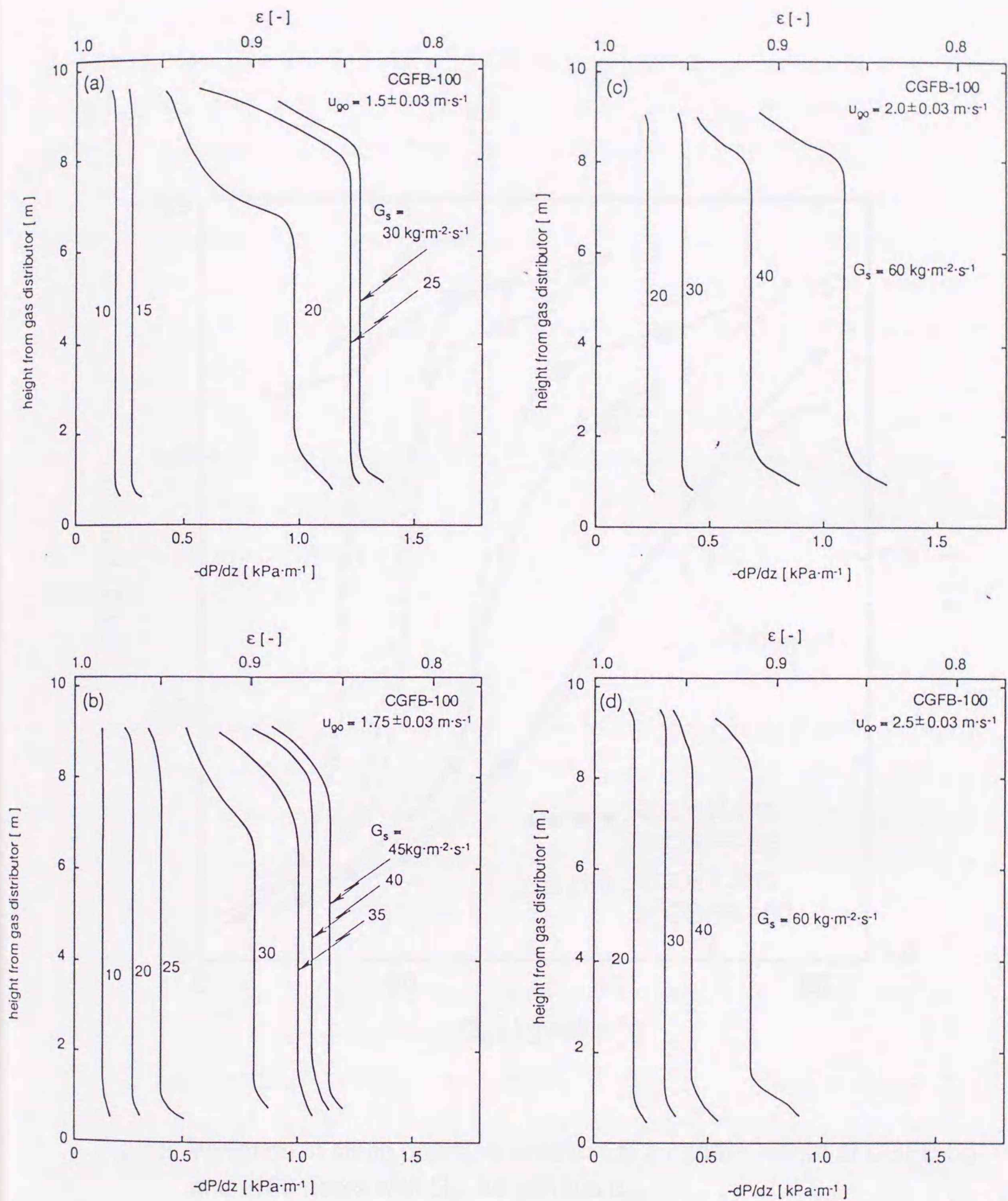


Fig. 2.4 Variation of ABD distribution in riser of CGFB-100 at various operating conditions

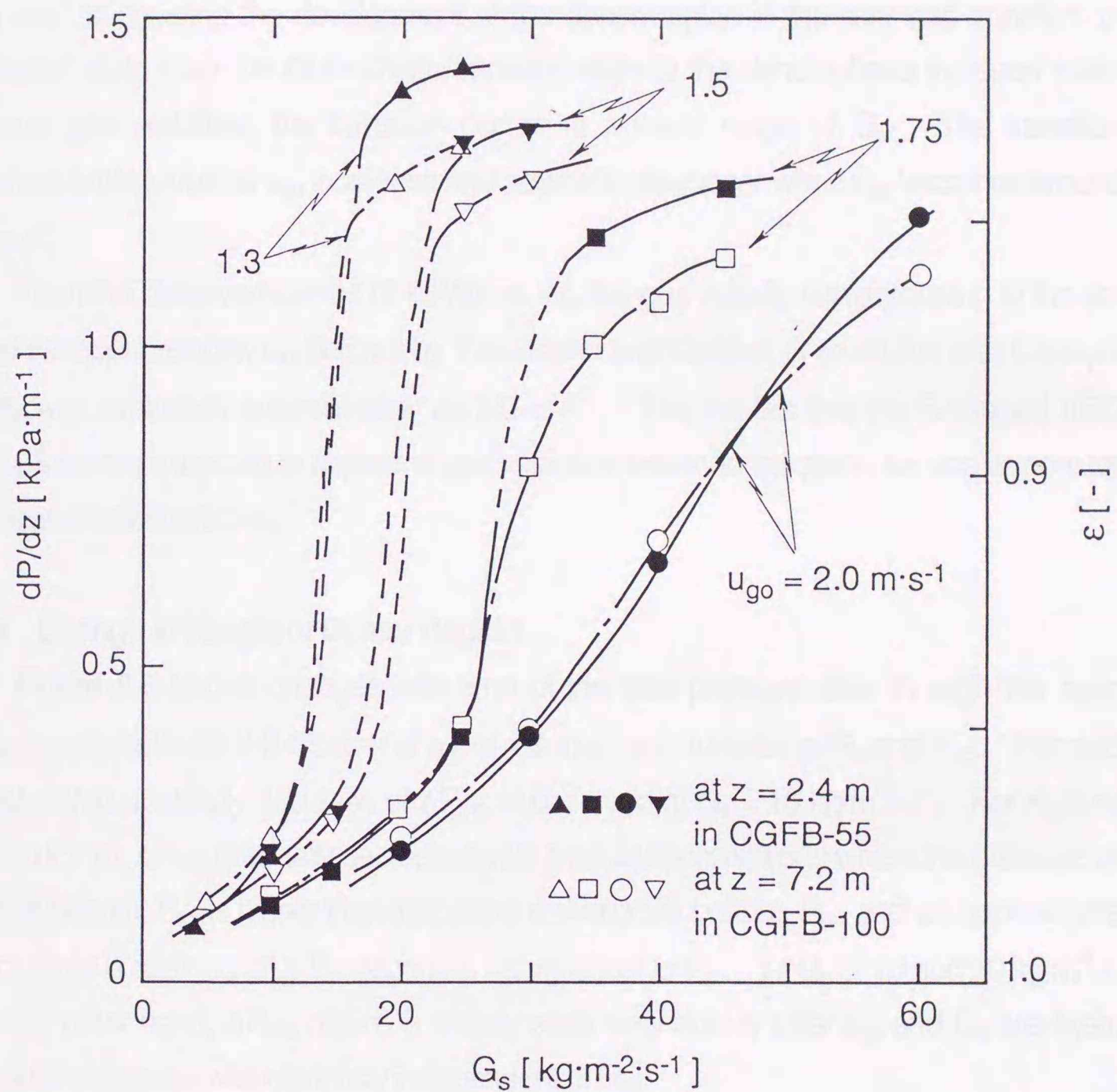


Fig.2.5 Variation in static pressure gradient at a certain height of CGFB-55 and -100 risers with G_s for various u_{go}

CGFB-55 and $z = 7.2$ m for CGFB-100, respectively, was plotted against G_s at various gas velocities. For all cases the bed density increases with G_s . No differences are seen between the bed densities in CGFB-55 and -100 risers particularly when G_s is small and even the differences seen at higher G_s are not essential within the present experimental accuracy. As G_s increases, an abrupt change in the bed density occurs at u_{go} lower than $1.75 \text{ m}\cdot\text{s}^{-1}$ suggesting the development of the dense region in the riser and transition of fluidization state from the dilute-phase transport state to the dense-phase transport state. At lower gas velocities, the transition occurs at a lower range of G_s . The transition becomes ambiguous as u_{go} increases and seems to disappear when u_{go} becomes around $2.0 \text{ m}\cdot\text{s}^{-1}$.

From the observed curves of $-dP/dz$ vs. G_s , the gas velocity corresponding to the so-called *transport velocity*, u_{tr} , defined by Yerushalmi and Cancurt (Yerushalmi and Cancurt; 1979), was evaluated approximately as $2.0 \text{ m}\cdot\text{s}^{-1}$. This implies that the S-shaped ABD distributions are expected to appear at gas velocities less than this particular gas velocity for the present solid particles.

2.3.3 Change in Height of Dense Region

Figure 2.6 shows changes with time of the total pressure drop through the riser, ΔP_{total} , measured in CGFB-55 riser at u_{go} of $1.3 \text{ m}\cdot\text{s}^{-1}$ as a function of G_s and H_{sd} . It should be noted that unsteady decrease of ΔP_{total} is observed at $G_s = 15 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$. For H_{sd} 's of 1.8 and 2.1 m, ΔP_{total} gradually decreases with time under conditions where the pressure at the riser bottom, P_{ro} , is higher than that at the downcomer bottom, P_{do} , and an approximate steady-state is attained after P_{ro} becomes nearly equal to P_{do} . At G_s of 10 and $20 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$, on the other hand, ΔP_{total} attains a steady state fairly quickly after u_{go} and G_s are fixed. This quick response was obviously independent of H_{sd} .

The unsteady change in ΔP_{total} corresponded not only to the drastic increase in the pressure gradient with G_s described in Fig.2.5 but also to significant changes in height of the dense region height which was simultaneously observed at the conditions indicated as a, b, c, d and e in Fig.2.6. The results are shown in Fig.2.7 as ABD distributions. At $G_s = 15 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$, as drawn by broken lines, the S-shaped ABD distributions are seen at conditions of a, b, c, d and e. It is also seen that the height of the lower dense region decreases with ΔP_{total} or with H_{sd} as described in Fig.2.6 even at fixed u_{go} and G_s . From the results shown in Figs. from 2.5 to 2.7, it is evident that the effect of H_{sd} is observed only at a particular condition of G_s and u_{go} where the S-shaped ABD distribution is obtained in the riser.

The above-described experimental results could be explained by an analogy to solid

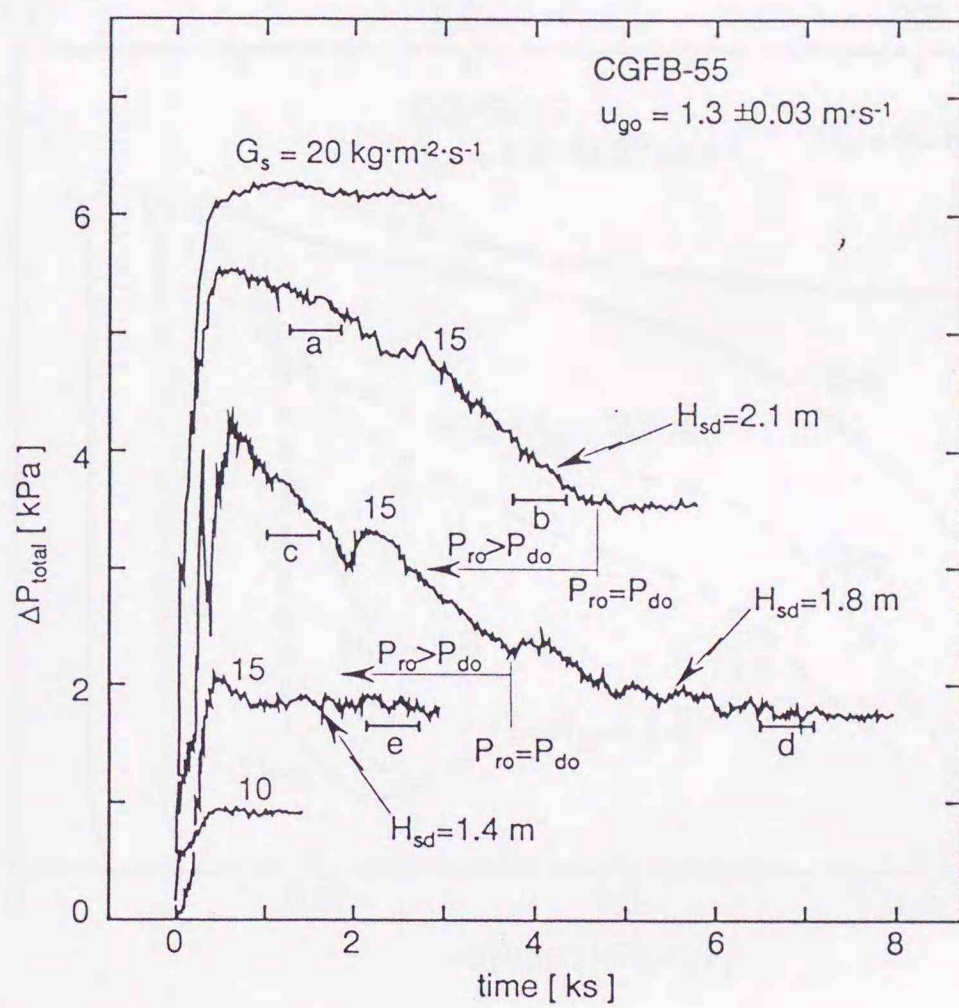


Fig.2.6 Change in total pressure drop, ΔP_{total} with time at $u_{go} = 1.3 \text{ m}\cdot\text{s}^{-1}$ for different H_{sd}

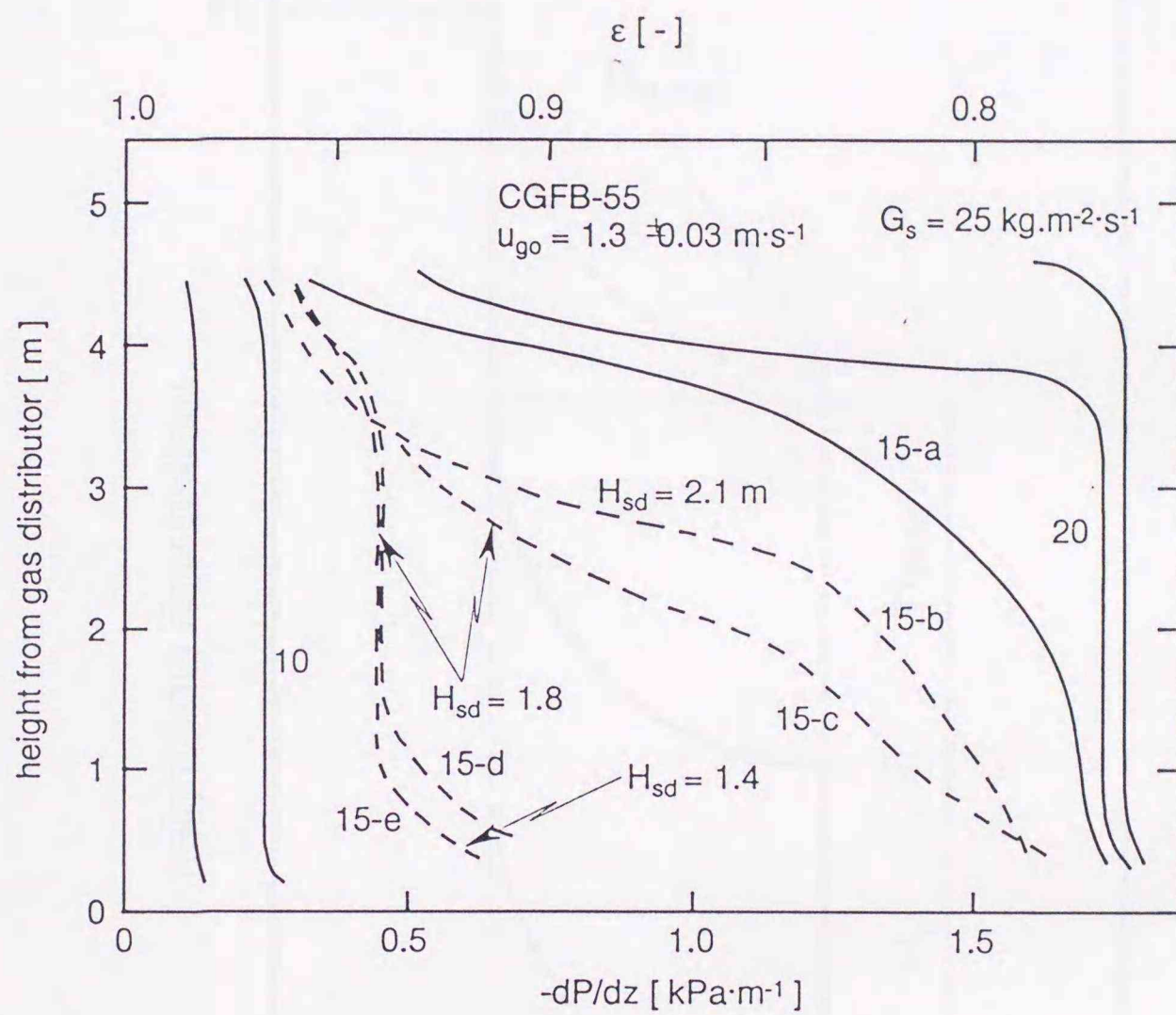


Fig.2.7 ABD distribution in CGFB riser at $u_{l0} = 1.3 \text{ m}\cdot\text{s}^{-1}$ and $G_s = 15 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ for different H_{sd}

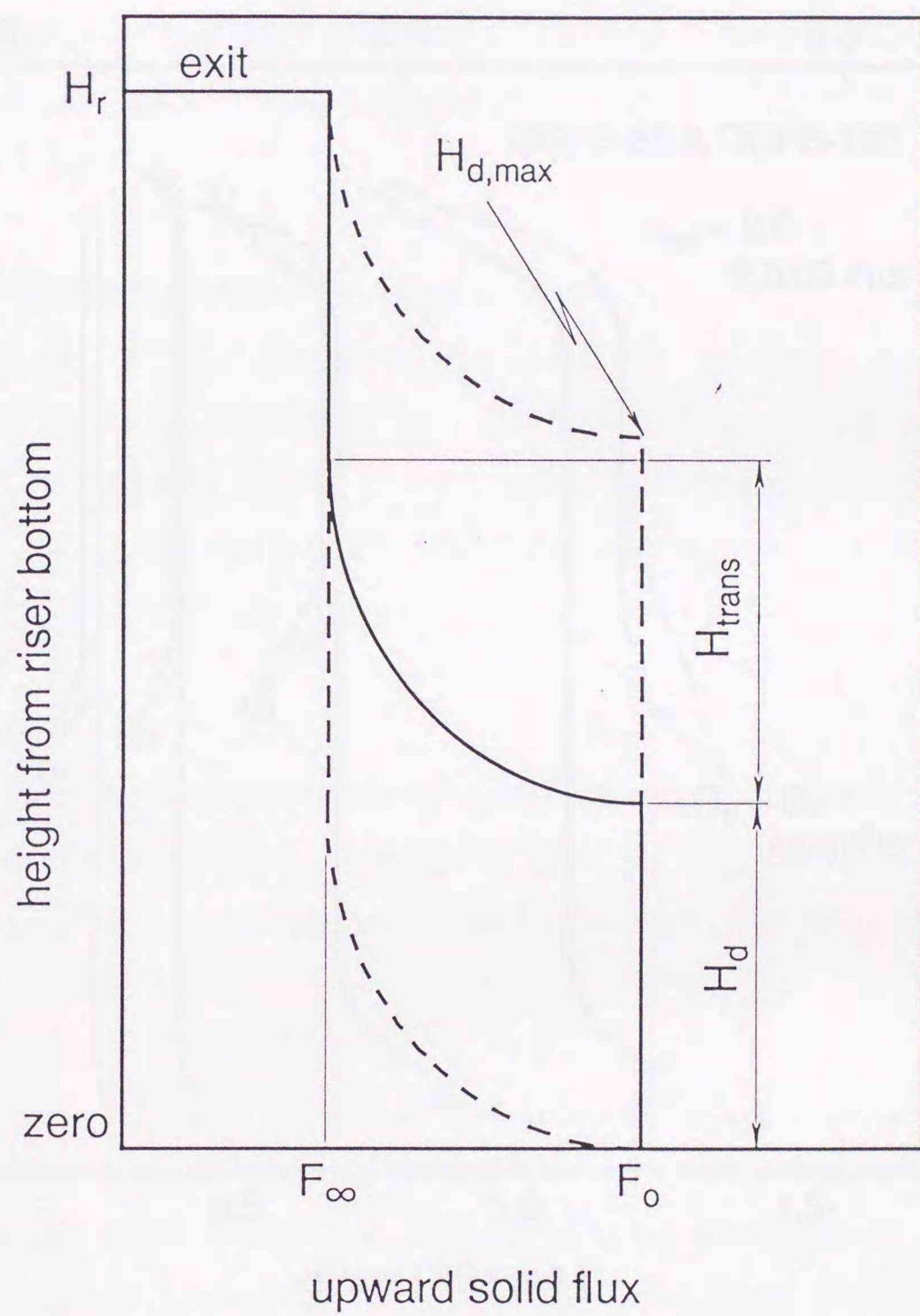


Fig.2.8 Schematic illustration of typical distribution of upward solid flux in CGFB riser

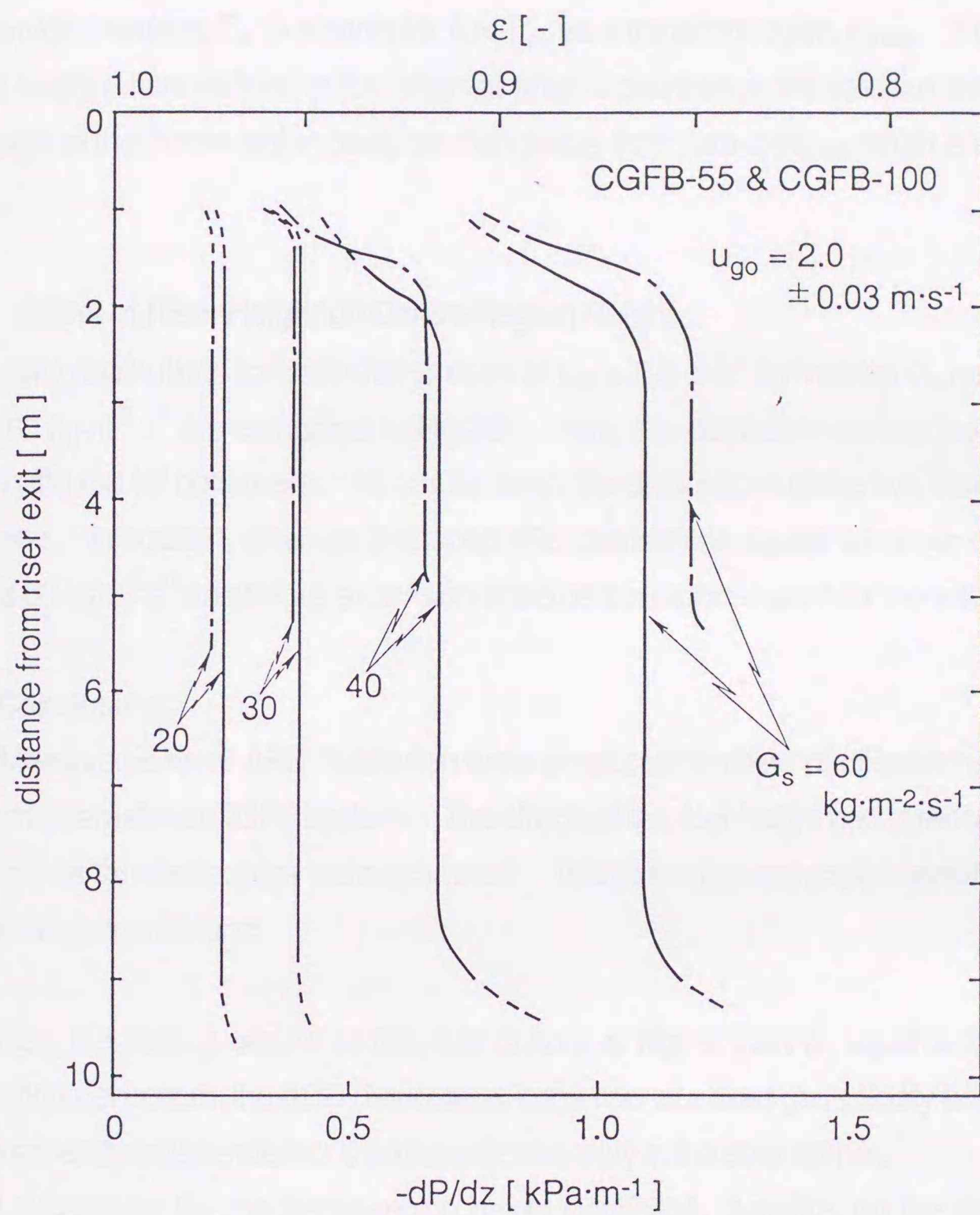


Fig.2.9 A comparison of ABD distribution curves obtained in CGFB-55 and -100 risers at $u_{go}=2.0 \text{ m}\cdot\text{s}^{-1}$

entrainment in the freeboard of the bubbling fluidized beds (Kunii and Levenspiel; 1990, Rhodes; 1989, Wen and Chen; 1972, Zenz and Othmer; 1960). **Figure 2.8** illustrates typical distributions of upward solid flux along the height from the bottom of a riser with a height, H_r . The solid mass flux decreases with height exponentially from a flux at the dense-region surface, F_0 , to a particular flux, F_∞ , via a transition region, H_{trans} . If the dense-region height is here defined as the height at which a decrease in the solid flux commences, the height of the dense-region could be changeable from zero to $H_{d,max}$ which is equal to $H_r - H_{trans}$.

2.3.4 Effect of Riser Height on Dense-Region Height

ABD distributions for both CGFB risers at $u_{go} = 2.0 \text{ m}\cdot\text{s}^{-1}$ for various G_s ranged from 10 to $60 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ are compared in **Fig.2.9**. Here, the distribution curves are drawn by leveling the exit for both risers. As can be seen, the distributions of the two risers seem to be similar. In addition, when an S-shaped ABD distributions appear as observed at $G_s = 40$ and $60 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ an obvious expansion of the dense-region is seen for the taller riser.

2.4 Conclusion

Measurements of ABD distribution were conducted in risers of different heights with the same diameter of CGFB system. The effects of the riser height and downcomer solid inventory on the distributions were examined. Within the present experimental condition, the followings were found;

- 1) When the static pressure at the riser bottom is higher than or equal to that at the downcomer bottom, the ABD distributions in the riser at a fixed gas velocity and solid flux are generally independent of the total solid inventory in the downcomer.
- 2) At a particular G_s , the dense-region height significantly depends on the downcomer solid inventory. The result can be explained by an analogy to existing models of the solid elutriation in the freeboard of the bubbling fluidized bed.
- 3) Dense-region height increases with the riser height even under conditions where all the other operating variables are the same. Hence, an apparent overlap is obtained for the ABD distribution curves for risers of the different heights, when the distributions are drawn by leveling the riser exit.

Literature Cited

- Hirama, T., H., Takeuchi and T. Chiba, "Regime Classification of Macroscopic Gas-Solid Flow in a Circulating Fluidized Bed Riser", Powder Tech., 70, 215-222 (1992)
- Horio, M., "High Velocity Operation of Fluidized Beds", J. Soc. Powder Tech. Japan, 23, 80-90 (1986)
- Kunii, D. and O. Levenspiel, "Entrainment of Solids from Fluidized Beds. 1. Holdup of Solids in the Freeboard Regions. 2. Operation of Fast Fluidized Beds", Powder Tech., 61, 193-206 (1990)
- Li, J., Y. Tung and M. Kwauk, "Axial Voidage Profiles of Fast Fluidized Beds in Different Operating Regions", in Circulating Fluidized Bed Technology II, P. Basu and J.F. Large, Eds., p.193-203, Pergamon Press, Oxford, U.K.(1988)
- Mori, S., Y. Yan, T. Haruta, K. Yamaguchi and I. Yamada, "Dense Bed Height and Voidages in Circulating Fluidized Bed", in Proc. 3rd SCEJ Symp. on Circulating Fluidized Beds, Tokyo, p.29-36 (1989)
- Rhodes, M. J., "The Upward Flow of Gas/Solid Suspensions Part 2: A Practical Quantitative Flow Regime Diagram for the Upward Flow of Gas/Solid Suspensions", Chem., Eng., Res., Des., 67,30-37 (1989)
- Takeuchi, H., T. Hirama, T. Chiba, J. Biswas and L. S. Leung, "A Quantitative Definition and Flow Regime Diagram for Fast Fluidization", Powder Tech., 47, 195-199 (1986)
- Weinstein, H., R. A. Graf, M. Meller and M. J. Shao, "The Influence of The Imposed Pressure Drop across a Fast Fluidized Bed," in "Fluidization," D. Kunii and R. Toei, Eds., p.299-306, Engineering Foundation, New York, U.S.A. (1983)
- Wen, C. Y. and L. H. Chen, "Fluidized Bed Freeboard Phenomena: Entrainment and Elutriation", AIChE J. 28, 117-128(1972)
- Yerushalmi, J. and N. T. Cankurt, "Further studies of the regimes of fluidization", Powder Tech., 24, 187-205 (1979)
- Zenz, F. A. and L. H. Othmer, Fluidization and Fluid-Particle Systems, Reinhold, pp.374, New York (1960)

Chapter 3

UNSTEADY STATE BEHAVIOR OF A CIRCULATING GAS-FLUIDIZED BED SYSTEM

3.1 Introduction

A circulating gas-fluidized bed (CGFB) system basically consists of a riser, a downcomer, cyclones and a solid feeding device. Particles elutriated from the riser top are collected by the cyclones and fed to the downcomer. Then, the particles descend through the downcomer and continuously fed through the solid feeding device to the riser bottom at a steady mass flux. The stable solid circulation is known to depend on the pressure drop through the feeding device, ΔP_{rd} . Most of the axial bed density (ABD) distributions in the riser reported in literature were observed in experimental CGFB systems where the downcomer bottom was simply connected to the riser bottom by an L-shaped or a loop-seal valve. In the case of CGFB system using such a valve, the observed steady-state ABD distributions were found to be dependent more or less on the initial downcomer solid inventory even at fixed gas velocity, u_{go} , and circulating solid flux, G_s (Weinstein et al, 1983, Li et al, 1988, Mori et al, 1989). This is attributed to small ΔP_{rd} of the pneumatic valve. In fact, solid circulation is sensitive to ΔP_{rd} , taking a longer period for attainment of a steady state due to a pressure interaction between the riser and downcomer bottoms. It has also been observed that an appreciable fluctuation in the ABD distribution is brought about during the transitional period for attainment of the pressure balance in the whole system. However, in the existing investigations the circulating solid mass flux has never continuously been measured and details have not been clarified yet of the unsteady-state characteristics of the CGFB system, such as G_s , ΔP_{rd} and the solid holdup in the riser and the downcomer.

In the preceding chapter, a screw feeder having ΔP_{rd} much larger than pneumatic valves was employed to minimize effect of the variation of the downcomer bottom pressure on ABD distributions (Hirama et al, 1996). From measurements of ABD distributions for different heights of the static bed of particles initially loaded in the downcomer, H_{sd} , it was found that the ABD distribution depends on H_{sd} only at conditions where the dense region appears in lower part of the riser. However, again, details are still in question concerning

the solid flow characteristics in the circulating system, in particular, the change of the static pressures with time at the riser and downcomer bottoms which gives rise to unsteady-state ABD distributions in the riser. Hence, in this chapter, a further experimental work has been carried out to evaluate the relation of the variation of ΔP_{rd} to the ABD distribution from continuous measurements of the static pressures as well as G_s for various H_{sd} . In addition, the effect of static pressure at the downcomer bottom on steady-state ABD distribution was examined varying intentionally the downcomer bottom pressure by aeration.

3.2 Experimental

Figure 3.1 illustrates a schematic diagram of the experimental CGFB system. Details of the elements in the system and experimental procedure were specified in the preceding chapter. The bottom of a 10.0 m high riser with an inner diameter of 0.1 m was connected to that of a 0.2 m i.d. and 8.2 m high downcomer by a screw feeder which controlled the circulating mass flux of solid particles, G_s , by adjusting a revolution speed of the screw without aeration. Changes with time of the axial distributions of static pressure in the riser and the downcomer were simultaneously measured with electric transducers which were attached on the riser and downcomer walls at respective intervals from 0.45 to 0.60 m and from 0.40 to 1.00 m, respectively. Signals from the transducers were processed by a personal computer. At a condition, about 20 seconds were required to obtain data for determination of ABD distribution in the riser. As shown in **Fig.3.2**, an impact-type solid flow device (Iinoya et al, 1966) was presently employed to monitor the flow rate of solid particles elutriated from the top of the riser. The device was equipped immediately below the primary cyclone. The solid mass flow rate was continuously measured by a load-cell which detects the vertical component of impact force exerted by the solid particles on a declined plate.

The bed material used in the present experiment was FCC particles having a median of 54 μm in the size distribution, a mercury-porosimetric density of about 810 $\text{kg}\cdot\text{m}^{-3}$ and the minimum fluidization velocity of 0.0011 $\text{m}\cdot\text{s}^{-1}$. The particles were fluidized by air at room temperature and a fixed superficial gas velocity of 1.3 $\text{m}\cdot\text{s}^{-1}$ in the present experiment.

3.3 Results and Discussion

3.3.1 Transitional Change of Static Pressures at Riser and Downcomer Bottoms

When the static bed height of particles initially loaded in the downcomer, H_{sd} , was less than 2.7 m, it was impossible to operate the present CFB system steadily due to occurrence of a considerable gas cross flow between the riser and the downcomer through

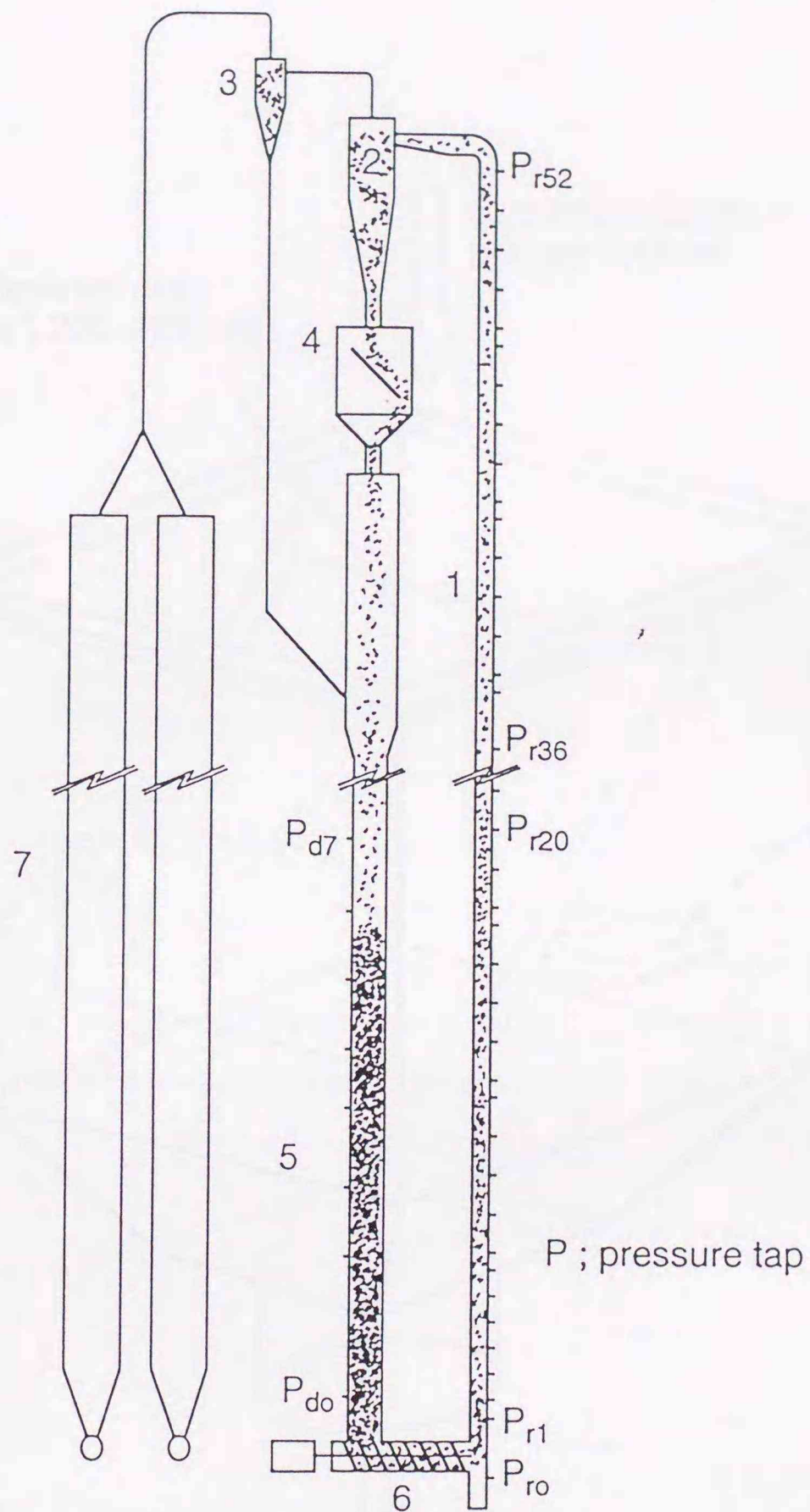


Fig.3.1 Schematic diagram of experimental circulating gas-fluidized bed system

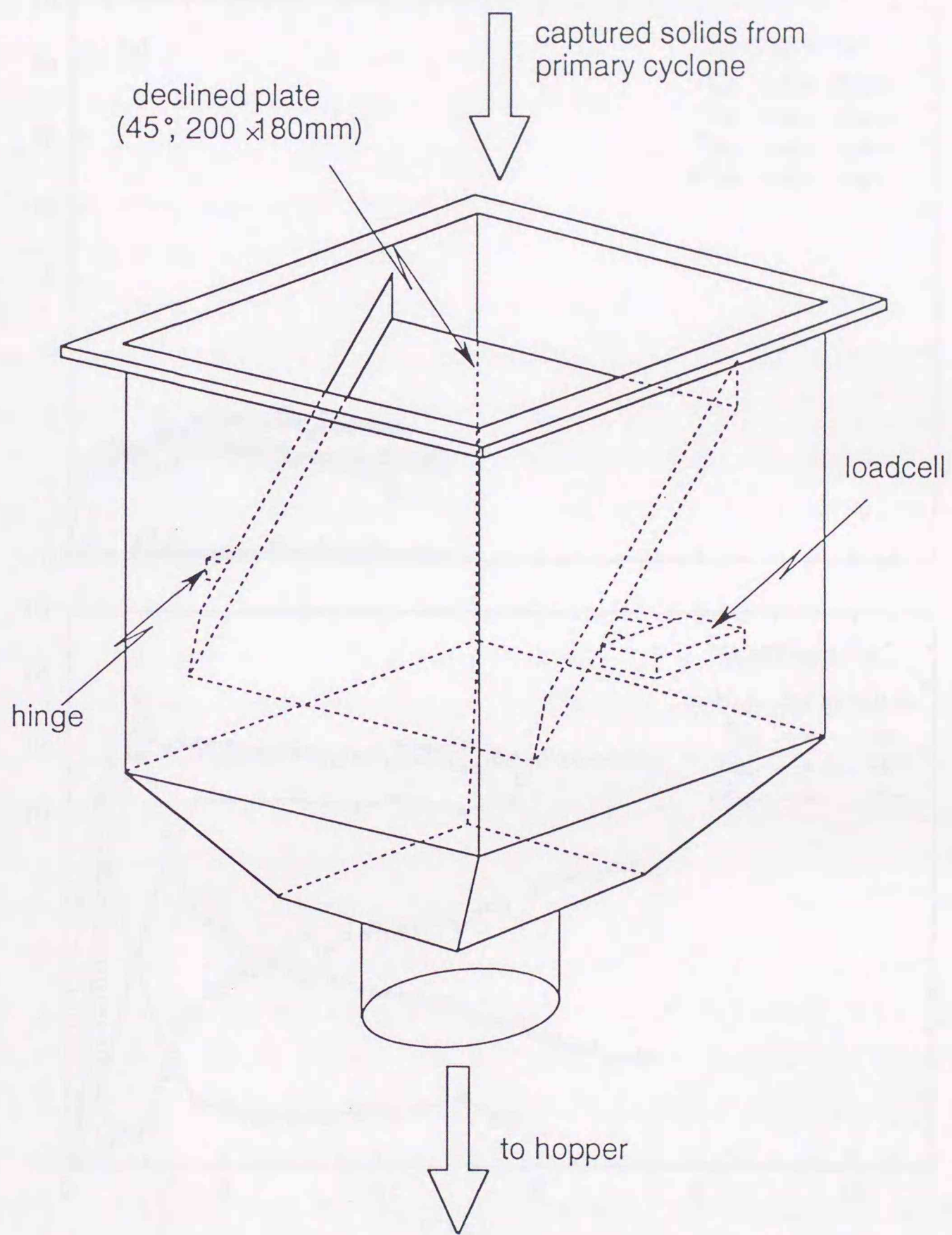


Fig.3.2 Impact-type solid flow device.

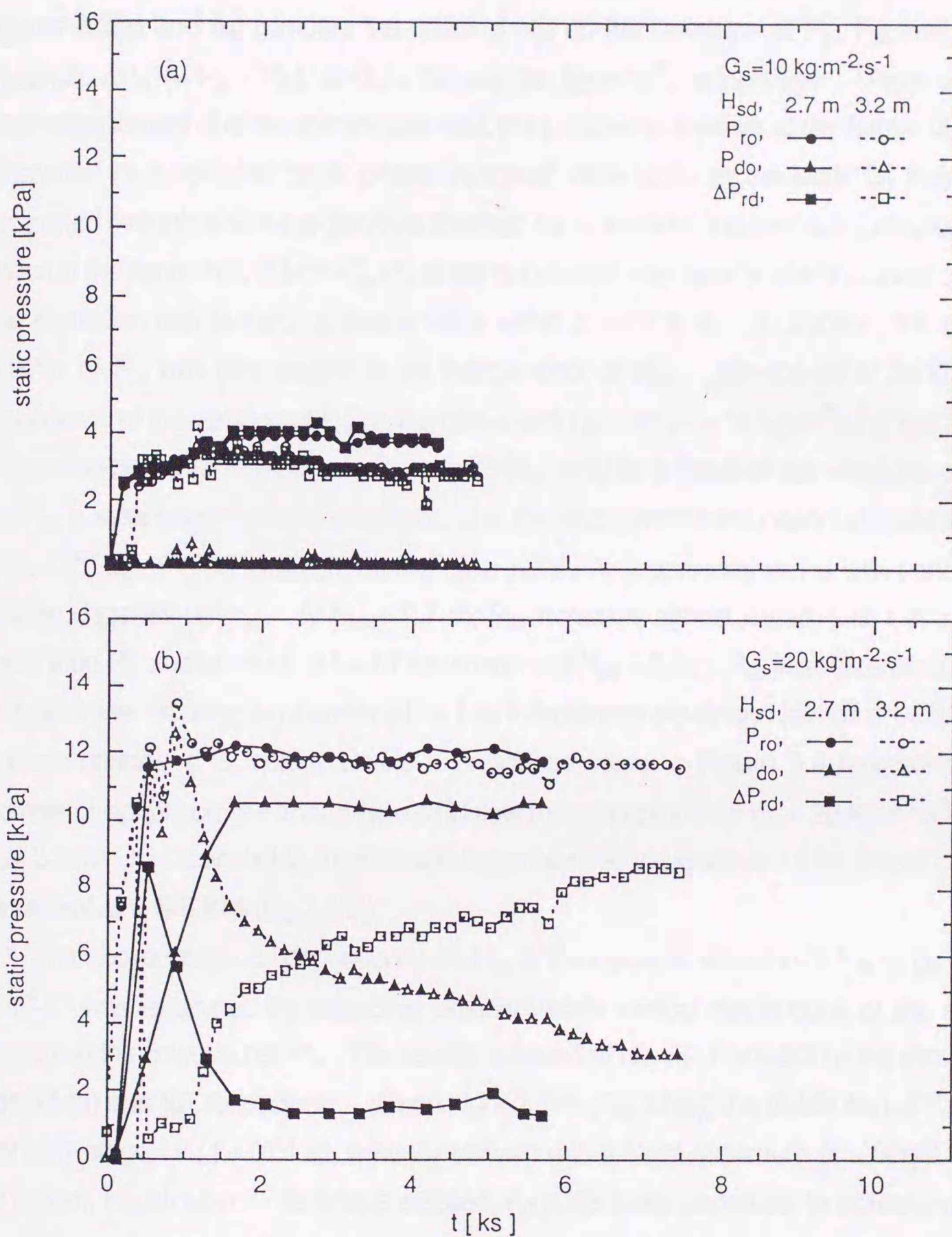


Fig.3.3 Change in P_{ro} and P_{do} with time at different H_{sd} of 2.7 m and 3.2 m:
(a) $G_s = 10 \text{ kg} \cdot \text{m}^{-2} \cdot \text{s}^{-1}$ and (b) $G_s = 20 \text{ kg} \cdot \text{m}^{-2} \cdot \text{s}^{-1}$.

the screw feeder. Variation in the static pressures, P_{ro} and P_{do} , at the riser and downcomer bottoms, respectively, with time, t , elapsed after the onset of the solid circulation ($t = 0$) were measured as a function of the circulating solid flux, G_s , and H_{sd} . **Figures 3.3(a) and (b)** compare the effect of H_{sd} on the variations of P_{ro} , P_{do} and their difference, $\Delta P_{rd} (= P_{ro} - P_{do})$, at $G_s = 10$ and $20 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$, respectively. From visual observation through the transparent riser wall, the particles in the riser at the former G_s are transported in a so-called "dilute-phase transport" state while at the latter G_s they are transported through a dense region over the riser, as indicated in the preceding chapter. It is seen in the figure that, at both G_s , P_{ro} tends to increase very quickly after the onset of the solid circulation and to reach a steady value within a short time. In addition, this quick variation of P_{ro} with time seems to be independent of H_{sd} . On the other hand, the dependency of the variation of P_{do} on H_{sd} differs with G_s . At $G_s = 10 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ (Fig.3.3(a)), no appreciable variation is seen in P_{do} at both H_{sd} and, as a result of the variations of P_{ro} and P_{do} , a steady value of ΔP_{rd} is attained after the same transitional period as that for P_{ro} . At $G_s = 20 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ (Fig.3.3(b)), on the other hand, P_{do} apparently varies with t and the variation depends on H_{sd} . At $H_{sd} = 2.7 \text{ m}$, P_{do} increases almost linearly with t at a rate slower than P_{ro} and levels off at $t = 1.7 \text{ ks}$ whereas at $H_{sd} = 3.2 \text{ m}$, P_{do} increases as quickly as P_{ro} and after attaining a maximum at $t = 1 \text{ ks}$ it decreases slowly and falls off at $t = 6.8 \text{ ks}$. At these conditions G_s quickly attains a prescribed value. **Figure 3.4** represents an example of signal from the impact-type solid flow meter recorded for $G_s = 20 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ and $H_{sd} = 3.2 \text{ m}$. It is seen that G_s increases to the prescribed value within 1.5 ks despite of the increase of ΔP_{rd} till 6.8 ks (Fig.3.3(b)).

The above-described dependency on H_{sd} of the dynamic variation of P_{do} at $G_s = 20 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ was examined by inspecting unsteady-state vertical distributions of the static pressure in the downcomer, P_d . The results obtained at $H_{sd} = 2.7$ and 3.2 m are shown in **Figs.3.5 (a) and (b)**, respectively. When $H_{sd} = 2.7 \text{ m}$ (Fig.3.5(a)) the distribution of P_d at a very early stage, i.e., $t = 0.01 \text{ ks}$, is nearly uniform with a slight maximum at a height from the bottom, H_d , of 1.8 m . As time is elapsed, P_d in the lower part starts to increase with t due to a small amount of gas flow from the riser to the downcomer, which was visually observed through the transparent wall of the tube between the screw feeder and the downcomer bottom. A steady state distribution is attained at $t = 1.6 \text{ ks}$ (Fig.3.3(b)), which is very close to that at incipient fluidization of the particles as is shown by the curve at $t = 3.04 \text{ ks}$ in the figure. When $H_{sd} = 3.2 \text{ m}$ (Fig.3.5(b)) such a distribution is seen at $t = 1.32 \text{ ks}$ but P_d in the lower part of the bed ($H_d < 1.8 \text{ m}$) then decreases leaving the upper part distribution essentially unchanged and the steady state distribution is obtained at $t = 6.8 \text{ ks}$,

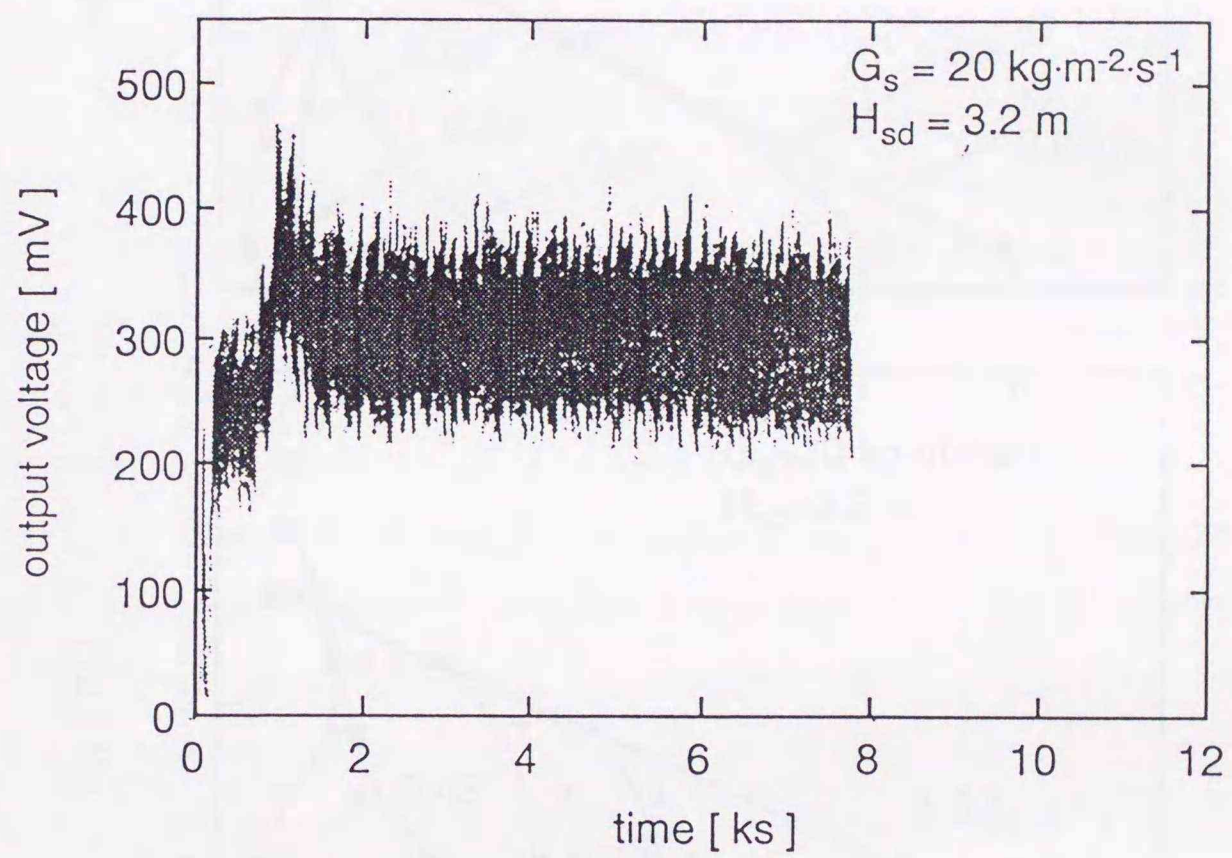


Fig.3.4 Change with time of output voltage from impact-type flow meter at $G_s = 20 \text{ kg} \cdot \text{m}^{-2} \cdot \text{s}^{-1}$ for $H_{sd} = 3.2 \text{ m}$.

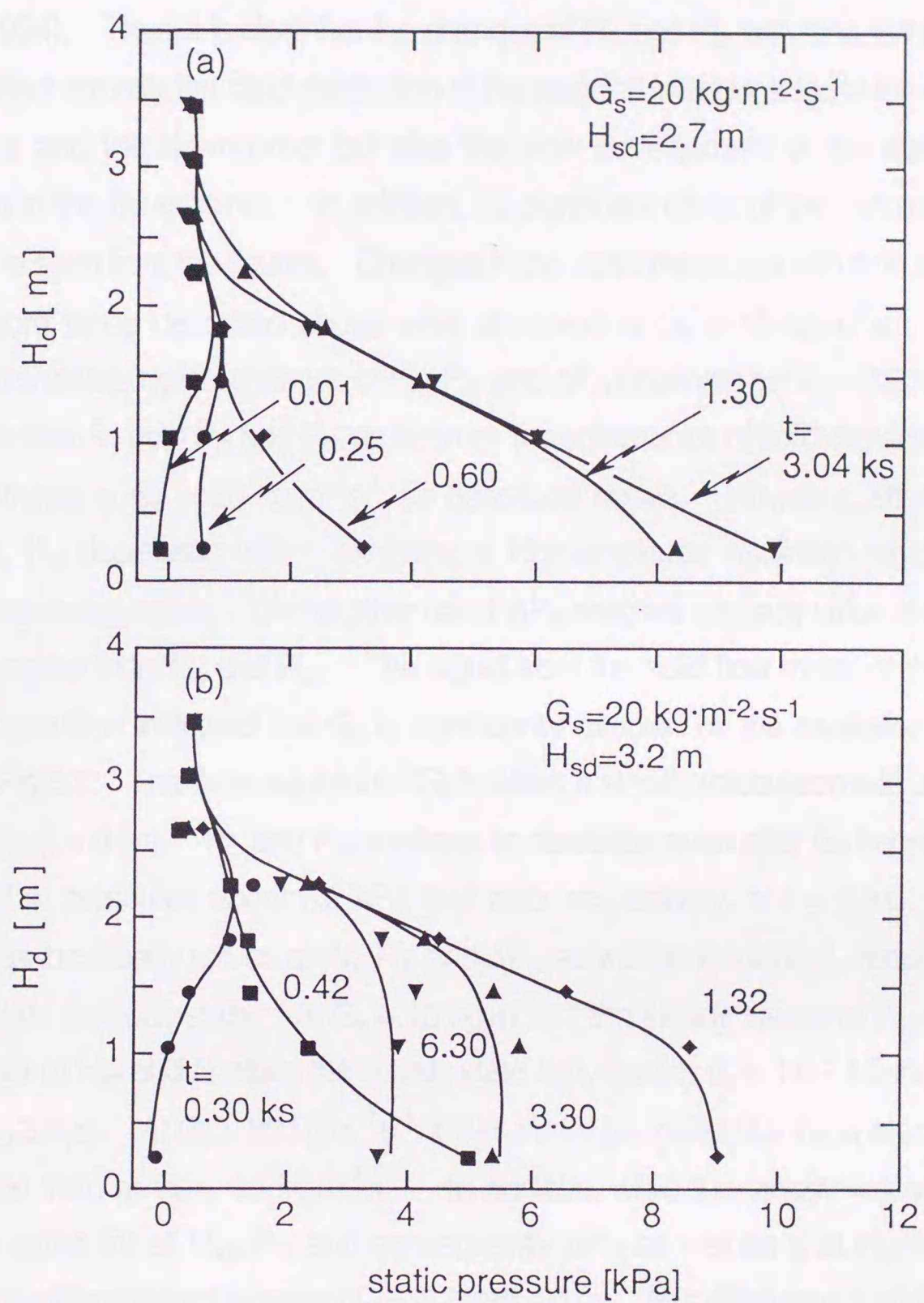


Fig.3.5 Transitional variation of static pressure distribution in downcomer at $G_s = 20 \text{ kg} \cdot \text{m}^{-2} \cdot \text{s}^{-1}$: (a) $H_{sd} = 2.7 \text{ m}$ and (b) $H_{sd} = 3.2 \text{ m}$.

suggesting that the bed in the downcomer consists of the upper fluidized bed and the lower moving bed. The formation of the fluidized bed can be attributed to an upward flow of gas which is accompanied with particles raining down from the cyclones onto the bed surface (Li et al; 1994). Thus, it is clear that the changes of P_{ro} and P_{do} with time in the circulating system reflect not only the rapid distribution of the particles initially loaded in the downcomer to the riser and the downcomer but also the slow development of the static pressure distribution in the downcomer. In addition, no significant effect of the variation of P_{do} on that of P_{ro} is seen from the figures. Changes in the static pressures with time considerably different from those described above were observed at $G_s = 15 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ for all H_{sd} 's. **Figure 3.6** shows a typical changes of P_{ro} , P_{do} and ΔP_{rd} observed at $H_{sd} = 3.2 \text{ m}$. A sharp increase is seen in both P_{ro} and P_{do} as soon as the commence of solid circulation, which is similar to those at $G_s = 20 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$, as described above. However, after attaining a maximum, P_{ro} decreases with t , exhibiting a high-amplitude oscillation which gradually decays to its steady value. On the other hand, ΔP_{rd} reaches a steady value at a time ($t = 4 \text{ ks}$) much earlier than P_{ro} and P_{do} . The signal from the solid flow meter at this particular operating condition indicated that G_s is significantly affected by the oscillation of ΔP_{rd} , as shown in **Fig.3.7**. The time required for G_s to attain a steady value seems to coincide with that for ΔP_{rd} ($t = 4 \text{ ks}$). P_{ro} and P_{do} continue to decrease even after G_s becomes steady and fall off at pressures about 6.2 kPa and zero, respectively, at $t = 8 \text{ ks}$. **Table 3.1** summarizes the steady values of P_{ro} , P_{do} and ΔP_{rd} as well as the time, t_s , required for all of them to attain a steady state. At $G_s = 10 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$, the steady values of P_{do} seem to be independent of H_{sd} and to attain the steady state fairly quickly ($t_s = 1.0 - 1.5 \text{ ks}$) as can be seen in Fig.3.3(a). At $G_s = 20 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$, it takes a longer period for the system to attain a steady state than at $G_s = 10 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$. In addition, while the steady values of P_{ro} are nearly the same for all H_{sd} , P_{do} and consequently ΔP_{rd} as well as t_s at $H_{sd} = 2.7 \text{ m}$ are considerably different from those at $H_{sd} = 3.2$ and 4.0 m . This difference is attributed to the gas flow from the riser to the downcomer which results in a negligible decrease of the gas velocity in the riser but is sufficient to partially fluidize the solid particles in the downcomer. On the other hand, though the steady values of P_{ro} and P_{do} at $G_s = 15 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ are between those at $G_s = 10$ and $20 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$, t_s is larger than those even at $G_s = 20 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ due to the oscillating slow decrease of P_{ro} and P_{do} described in Fig.3.6.

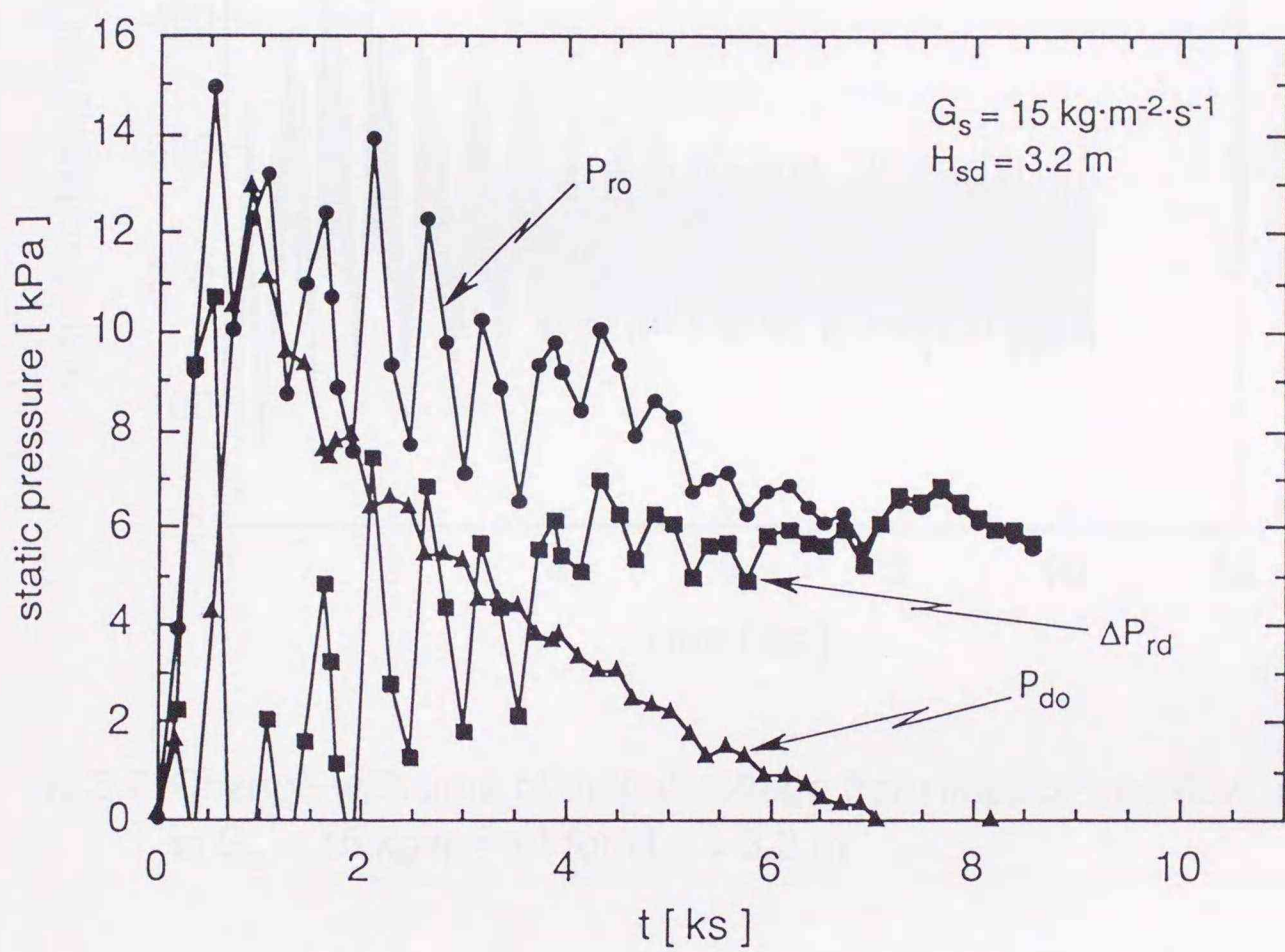


Fig.3.6 Typical variation in P_{ro} and P_{do} with time at $G_s = 15 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$.

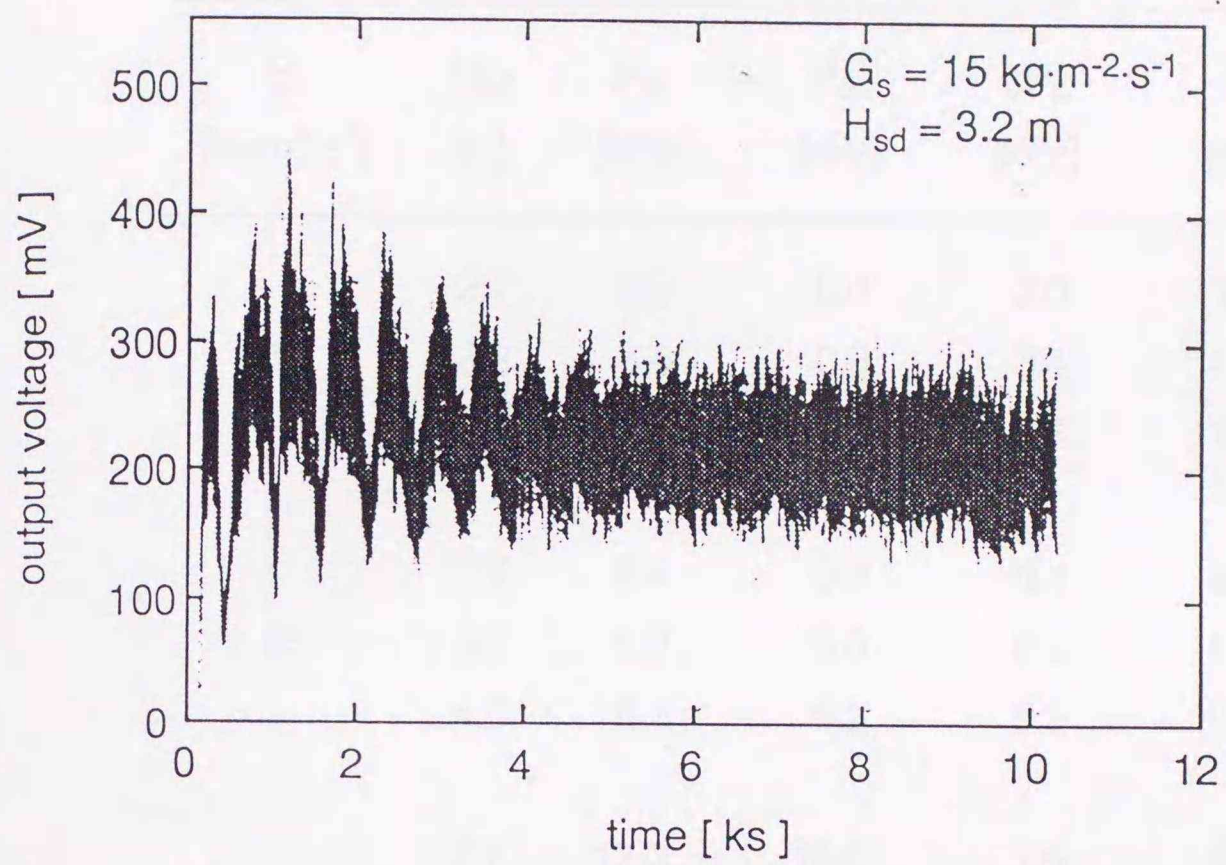


Fig.3.7 Change with time of output voltage from impact-type flow meter
at $G_s = 15 \text{ kg} \cdot \text{m}^{-2} \cdot \text{s}^{-1}$ for $H_{sd} = 3.2 \text{ m}$

Table 3.1 Steady Static Pressures and Time Required for Attainment of Steady State

G_s [kg·m ⁻² ·s ⁻¹]	H_{sd} [m]	P_{ro} [kPa]	P_{do} [kPa]	ΔP_{rd} [kPa]	t_s [ks]
10	2.7	3.0	0.0	3.0	1.5
	3.2	3.4	0.0	3.4	1.0
	4.0	3.2	0.0	3.2	1.0
15	2.7	6.4	0.0	6.4	9.5
	3.2	6.2	0.0	6.2	8.0
	4.0	6.1	0.2	5.9	10.0
20	2.7	12.0	10.5	1.5	2.5
	3.2	11.8	3.0	8.3	7.0
	4.0	12.6	2.6	10.0	8.3

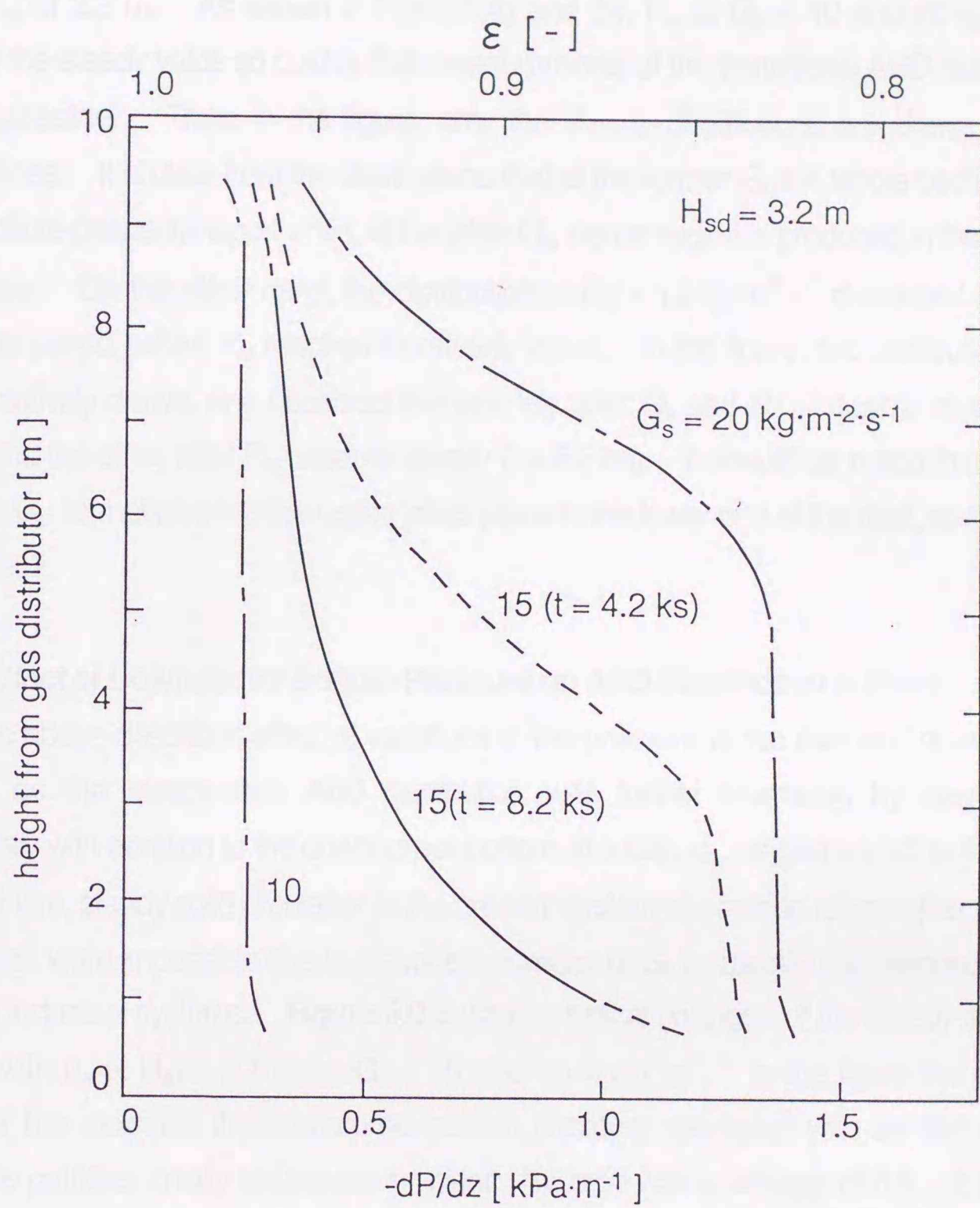


Fig.3.8 Typical ABD distribution observed at $H_{sd} = 3.2 \text{ m}$

The ABD distributions varied with time in accordance with the variations in P_{ro} . The results are shown in **Fig.3.8**, where the ABD distributions at different G_s are compared for the same H_{sd} of 3.2 m. As shown in Figs.3.3(a) and (b), P_{ro} at $G_s = 10$ and $20 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ reached the steady value so quickly that measurements of the transitional ABD distribution were impossible. Thus, in the figure, only the steady distributions are drawn by the dashed lines. It is clear from the distributions that at the former G_s the whole bed is in the state of dilute-phase transport while, at the latter G_s , dense region is produced in the middle of the riser. On the other hand, the distribution at $G_s = 15 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ depended on time during the period before P_{ro} reached the steady value. In the figure, two distributions are representatively drawn; one observed immediately after G_s and ΔP_{rd} became steady ($t = 4.2 \text{ ks}$) and the other after P_{ro} became steady ($t = 8.2 \text{ ks}$). It should be noted that, at this particular G_s , formation of dense region takes place in the lower part of the riser, as reported in Chapter 2.

3.3.2 Effect of Downcomer-Bottom Pressure on ABD Distribution in Riser

The above-described effect of variations in the pressure at the riser and downcomer bottoms on the steady-state ABD distribution was further examined by varying P_{db} intentionally with aeration to the downcomer bottom at a rate, u_a , ranged from 0 to $42 \times 10^{-6} \text{ m}^3\cdot\text{s}^{-1}$. Here, steady solid circulation in the present system at aeration rates higher than $42 \times 10^{-6} \text{ m}^3\cdot\text{s}^{-1}$ were impossible due to a sudden increase of G_s as usually experienced with L-valves in industrial systems. **Figure 3.9** illustrates typical changes of the steady-state P_{ro} and P_{db} with u_a at $H_{sd} = 2.7 \text{ m}$ for $G_s = 10$ and $15 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$. In the figure the dashed horizontal line indicates the downcomer-bottom pressure estimated with an assumption that all the particles initially loaded are incipiently fluidized with a voidage of 0.5. It is seen from the figure that at $G_s = 10 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ P_{db} increases with u_a while P_{ro} remains constant, suggesting that the ABD distribution is invariable with u_a . On the other hand, at $G_s = 15 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$, the increase of u_a results in an increase of P_{ro} as well as P_{db} . This clearly suggests that the ABD distribution varies with P_{db} as observed for the change of P_{db} with H_{sd} . The variation is confirmed in **Fig.3.10** which clearly indicates that the lower dense region develops with P_{db} .

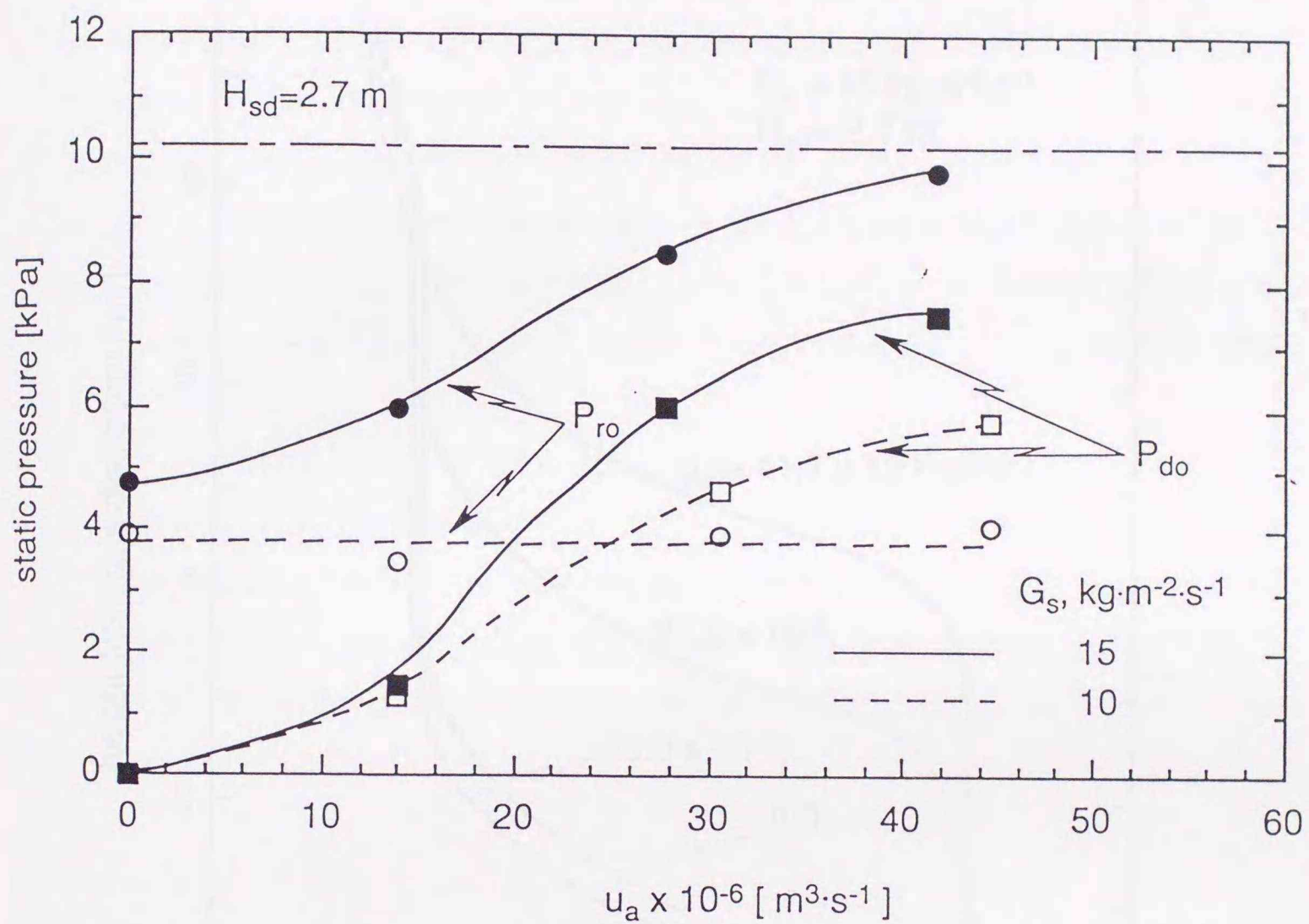


Fig.3.9 Change in P_{ro} and P_{do} with aeration rate at $G_s = 10$ and $15 \text{ kg} \cdot \text{m}^{-2} \cdot \text{s}^{-1}$

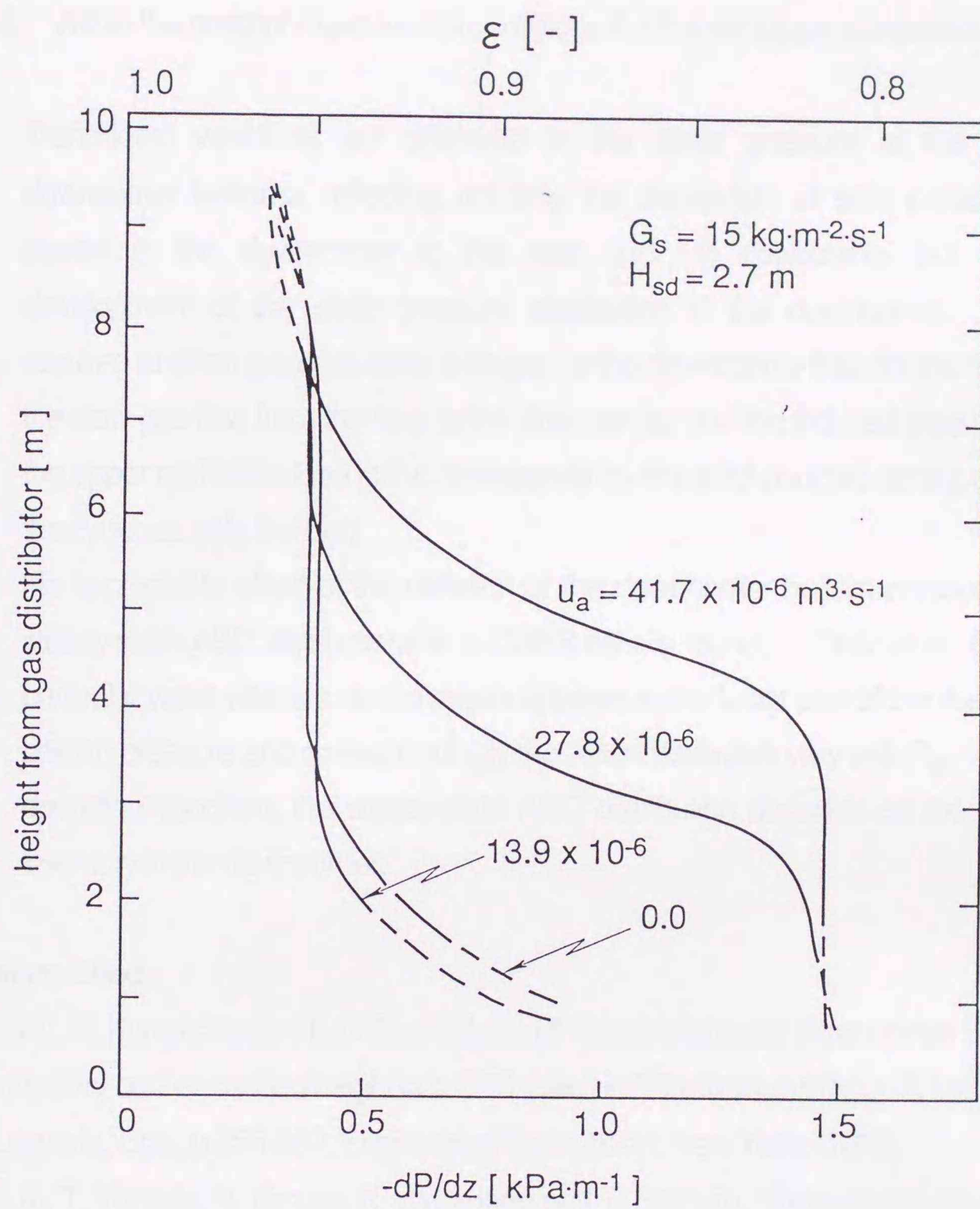


Fig.3.10 ABD distribution at $G_s = 15 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ for various aeration rates

3.4 Conclusion

Measurements were conducted of the static pressures at riser and downcomer bottoms and the ABD distributions in the riser of an experimental CGFB system which employed a screw feeder having a pressure drop much higher than widely used pneumatic valves. Within the present experimental conditions the followings are concluded:

- 1) Transitional variations are observed in the static pressure at the riser and downcomer bottoms, reflecting not only the distribution of solid particles initially loaded in the downcomer to the riser and the downcomer but also slow development of the static pressure distribution in the downcomer. The time required to attain a steady state is longer for the downcomer than for the riser due to the slow gas flow from the riser to the downcomer and the induced slow gas flow in the upper part of the bed in the downcomer by the solid particles raining down from the cyclones onto the bed.
- 2) No appreciable effect of the variation of the downcomer bottom pressure, P_{db} , on steady-state ABD distributions in a CGFB riser is found. Only when G_s takes a particular value where a dense region appears in the lower part of the riser, the riser bottom pressure and correspondingly the ABD distribution vary with P_{db} . At such a operating condition, the steady-state ABD distribution depends on the initial solid inventory in the downcomer.

Literature Cited

- Hirama T., K. Kuramoto and T. Chiba, "Effects of Riser Height and Downcomer Solids Inventory on Dense-Bed height in a CFB riser," in "Fluidization VIII," J. F. Large and C. Laguerie, Eds., p.255-262, Engineering Foundation, New York, (1996)
- Iinoya, K., T. Yoneda, N. Kimura, K. Watanabe and T. Shimizu, "Experiments on a Particle Flow Meter", J. Society of Powder Technology, Japan, 3, 424-431 (1966)
- Li, J., Y. Tung and M. Kwauk, "Axial Voidage Profiles of Fast Fluidized Beds in Different Operating Regions", in "Circulating Fluidized Bed Technology II," P. Basu and J.F. Large, Eds., p.193-203, Pergamon Press, Oxford, U.K.(1988)
- Li, Y., Y. Lu, F. Wang, K. Han, W. Mi, X. Chen and P. Wang, "Behavior of gas-solid flow in the downcomer of CFB reactors", in Proc. of 5th China-Japan Symposium on Fluidization, Beijing, p.61-68(1994)
- Mori, S., Y. Yan, T. Haruta, K. Yamaguchi and I. Yamada, "Dense Bed Height and Voidages in Circulating Fluidized Bed", in Proc. 3rd SCEJ Symp. on Circulating

Fluidized Beds, Tokyo, p.29-36(1989)

Weinstein, H., R. A. Graf, M. Meller and M. J. Shao, "The Influence of The Imposed Pressure Drop across a Fast Fluidized Bed", in "Fluidization," D. Kunii and R. Toei, Eds., p.299-306, Engineering Foundation, New York, U.S.A. (1983)

Chapter 4

HIGH-VELOCITY FLUIDIZATION OF SOLID PARTICLES IN A CIRCULATING LIQUID-FLUIDIZED BED RISER

4.1 Introduction

As introduced in chapter 1, fluidization of solid particles by gas always exhibits to an aggregative fluidization state. On the other hand, when fluidized by liquid a bed of solid particles mostly exhibits particulate fluidization state (Couderc, 1985). This difference between the fluidization states was first examined theoretically by Jackson (1963) as a stability problem of an incipiently fluidized bed. He analyzed the stability based on a perturbation theory and evaluated propagation of a small local disturbance in the solid concentration given to an incipiently fluidized bed. As a result of the analysis, it was clearly shown that growth rate of the disturbance in liquid-fluidized beds is essentially much slower than that in gas-fluidized beds, i.e. the former being much more stable than the latter against the disturbance and thus giving a particulate fluidization state.

As described in the preceding chapters, in circulating gas-fluidized bed risers where solid particles are fluidized at gas velocities much higher than those in bubbling fluidized beds, complicated change in both axial and radial distributions are observed in gas and solid velocities, bed pressure and voidage not only with operating variables but also with the system configuration. In contrast, a batch liquid-fluidized bed expands homogeneously as liquid flow rate increases (e.g. Richardson and Zaki, 1954a). The phenomena observed in liquid-fluidized beds are much simpler than those in gas-fluidized beds. For example, change of the bed voidage, ϵ , with the superficial liquid velocity, u_b , is empirically expressed (Richardson and Zaki, 1954b) by

$$u_{slp} = u_b/\epsilon - u_p = u_t \epsilon^{n-1} \quad (4-1)$$

where u_{slp} and u_p respectively denote the liquid-solid slip velocity and the superficial particle

velocity. For a system consisting of uniform spherical particles, the exponent n in Eq.4-1 is given as a function of the particle Reynolds number at the terminal velocity which is based upon the diameter and the free falling velocity of a single particle, u_t . Practical applicability of the correlation to particulate gas- and liquid-fluidized beds has been experimentally confirmed by many investigators (Garside and Al-Dibouni, 1977; Rowe, 1987; Gibilaro et al, 1988).

On the other hands, Wilhelm and Kwauk (1948) first reported occurrence of aggregative fluidization state in liquid-fluidized beds consisting of high-density particles like lead shot and copper sphere. They observed liquid void bands moving up through the bed and proposed the particle Froude number ($= u_b^2 \cdot g^{-1} \cdot d_p^{-1}$) as a criterion for the transition of the particulate and aggregative fluidization states. The void bands were later confirmed by a number of visual observations (Caims and Prausnitz, 1961; Hassett, 1961; Volpicelli et al, 1966; Kmiec, 1978). Foscolo and Gibilaro (e.g., Foscolo and Gibilaro, 1987; Gibilaro et al, 1986; Foscolo et al, 1986; Foscolo et al, 1984) extensively examined the transition between particulate and aggregative fluidization states based on an elastic wave theory (Wallis, 1969). Based on the results of their semi-empirical analysis, they constructed a classification map which suggests that low-density particles having relatively small diameter (less than 1 mm) always exhibit the particulate fluidization.

All of the knowledge on the macroscopic flow characteristics described above have been confirmed only for a batch liquid-fluidized bed (BLFB) system where u_b is lower than the terminal velocity of a single particle, u_t . When u_b exceeds u_t in BLFB, all particles are swept away from the column and the bed voidage eventually reaches unity. Then, if the elutriated particles were fed back to the bottom of the column, as is usually seen in a circulating gas-fluidized bed (CGFB), a certain value of the bed voidage less than unity can be obtained. In this circulating liquid-fluidized bed (CLFB) system, hydrodynamic phenomena even in the liquid-fluidized bed are expected to become complicated because turbulent interactions dominate between the liquid and solid phases. Hence, in the present study, measurements were conducted on the static pressure distributions through a riser of CLFB at various u_b and circulating mass flux of solid particles, G_s . Variations in u_{sp} and the spatial distribution of particles with u_b and G_s are analyzed to characterize macroscopic liquid-solid flow structures in the CLFB riser.

4.2 Experimental

4.2.1 Circulating Liquid-Fluidized Bed System

Figure 4.1 shows a schematic diagram of experimental apparatus employed in the

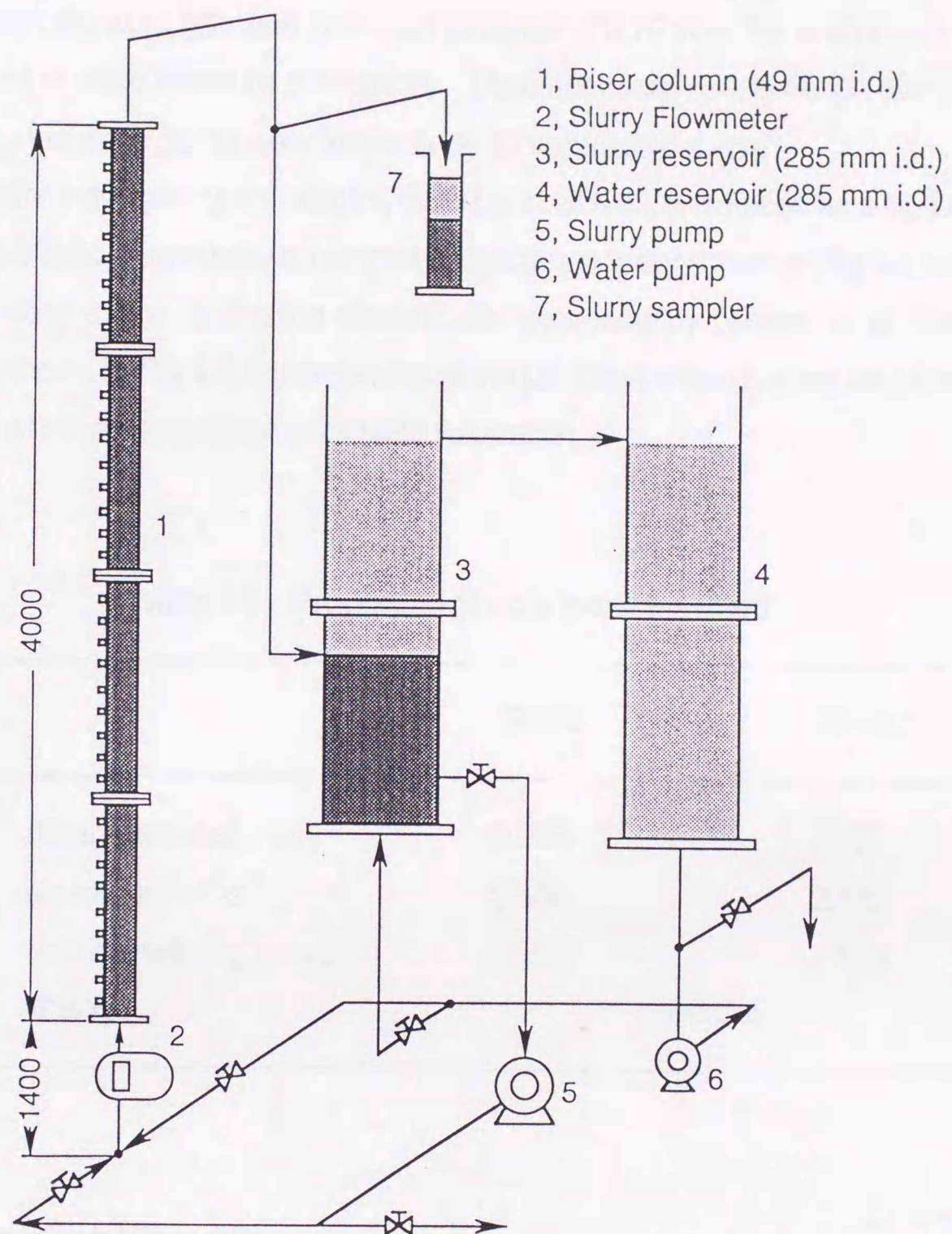


Fig.4.1 Experimental apparatus of circulating liquid-solid fluidized bed system

present study. The circulating liquid-fluidized bed (CLFB) system consists of a 0.049 m i.d. riser, 0.285 m i.d. slurry and water columns and slurry and water pumps. These were connected with pipes having the same inner diameter to that of the riser. The riser was constructed by connecting four 1.0 m high transparent PMMA-resin pipes with one another. Nine pressure taps were attached in an axial distance of 0.10 m on the wall of each riser for measurement of static pressure distribution. Maximum solid concentration and feed rate obtainable by the slurry pump were respectively 30 vol% and $0.4 \text{ kg}\cdot\text{s}^{-1}$.

Two differently sized glass ballotini, GB93 and GB182, were used as a bed material. Their size distributions and relevant properties are respectively shown in **Fig.4.2** and **Table 4.1**. According to the fluidization classification proposed by Gibilaro et al (Gibilaro et al;1986), as shown in **Fig.4.3**, the properties of the particles presently used are classified as the particles which always exhibit particulate fluidization.

Table 4.1 Properties of solid particles used

	GB93	GB182
mean diameter, mm	0.093	2.380
density, $\text{kg}\cdot\text{m}^{-3}$	2.484	2.480
terminal velocity, $\text{mm}\cdot\text{s}^{-1}$	6.5	19.9
shape	spherical	

4.2.2 Experimental Procedure

All experiments were carried out at ambient conditions using tap water as a fluidizing fluid. A weighted amount of the solid particles was loaded in the slurry reservoir(3). The particles are fluidized by water at a velocity required to maintain solid holdup in the reservoir at a certain value less than 0.3, which can be varied by adjusting the water flow rate from the water pump(6). Water overflowed from the slurry reservoir was led to the water reservoir(4) and circulated by the water pump. The solid-water slurry was fed to a bottom of the riser(1) by the slurry pump(5). The concentration of the particles in the riser was controlled by adjusting the solid holdup in the slurry reservoir and diluting the slurry feed with water fed by the water pump immediately below the slurry flow meter(2). All effluent from

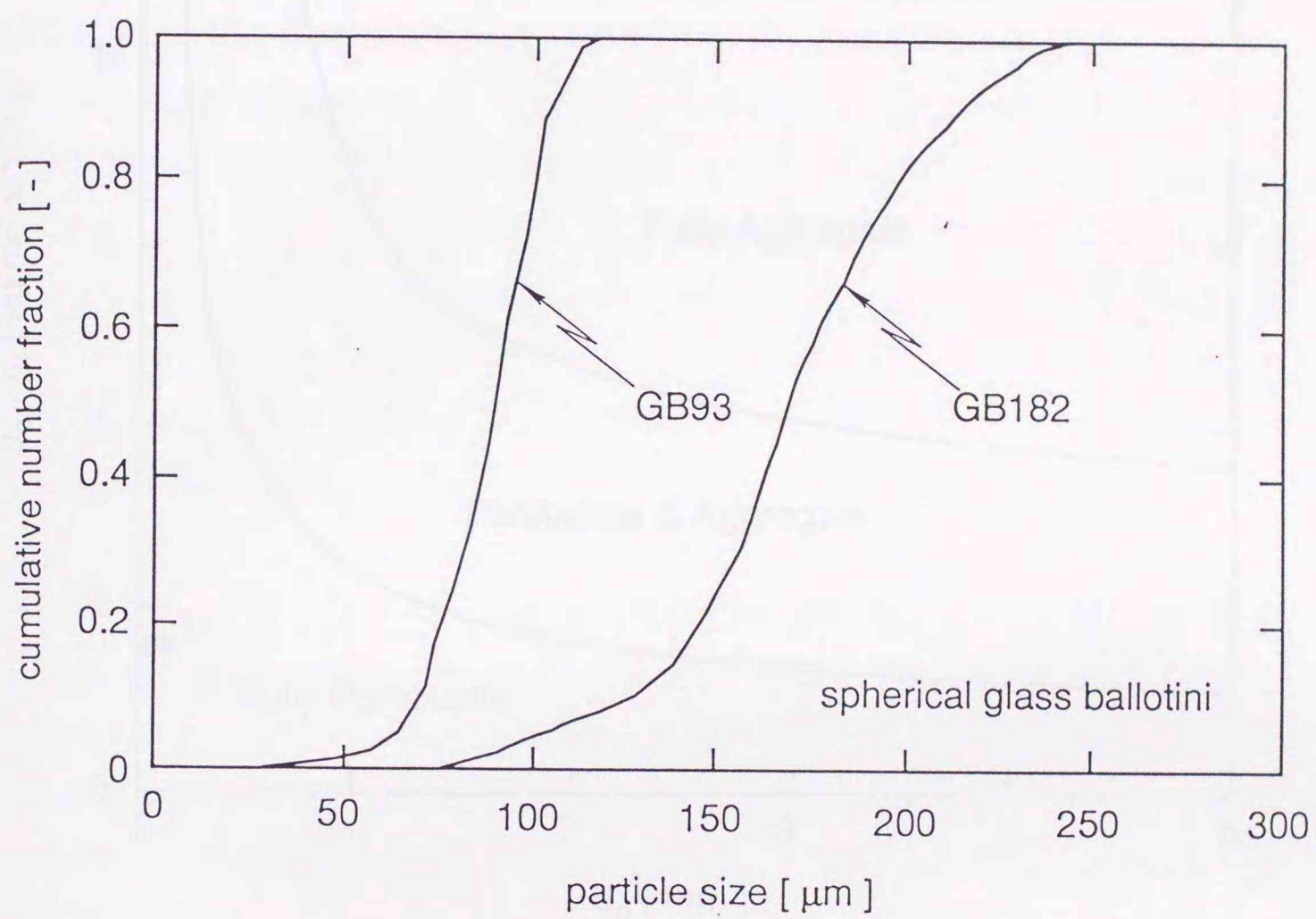


Fig.4.2 Cumulative size distribution of particle used.

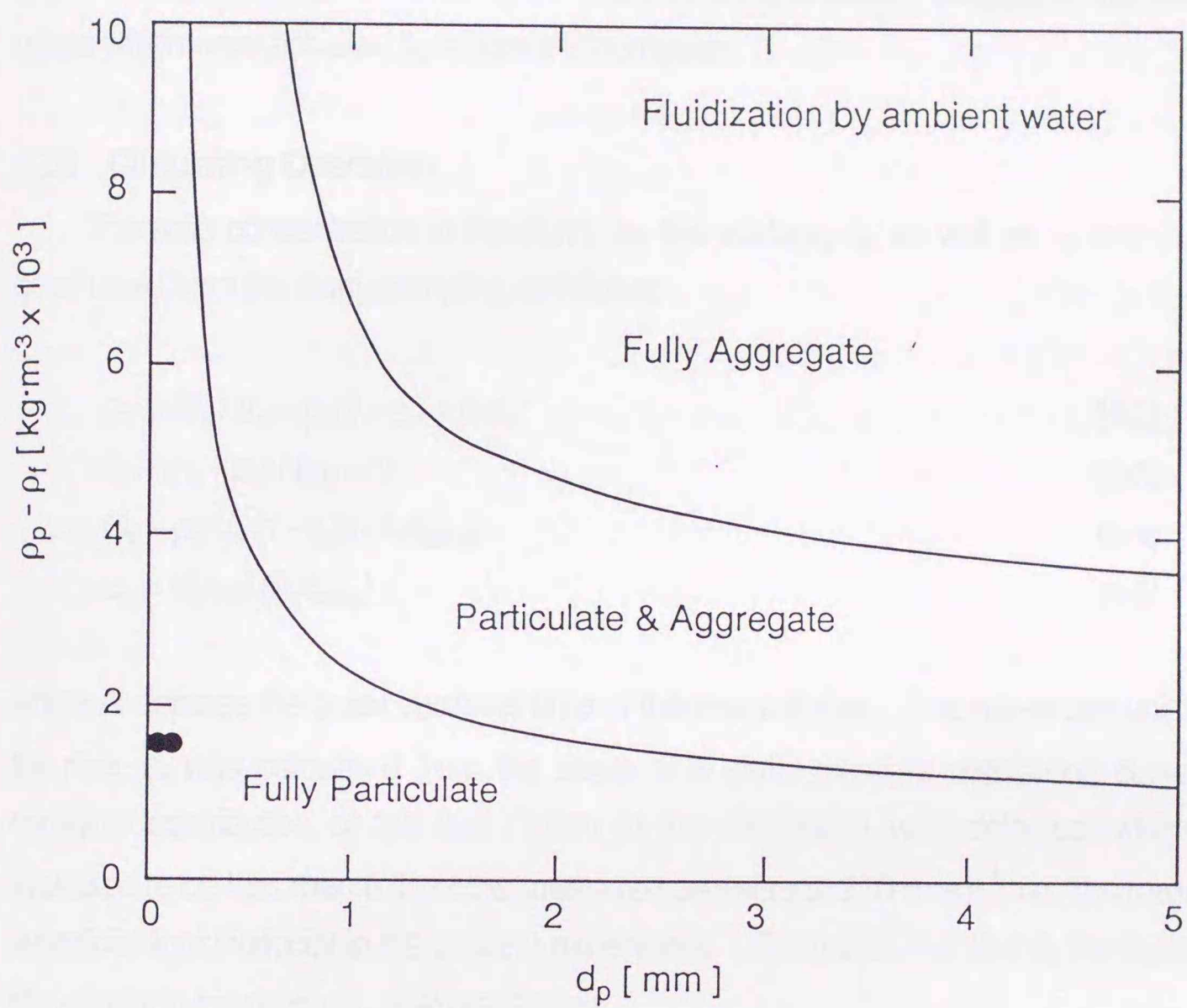


Fig.4.3 Gibilaro et al's criterion for fluidization regimes

the top of the riser was led to the slurry reservoir or to a slurry sampler(7). The operating variables such as u_b and G_s as well as the steady state of the whole system were confirmed by both the static pressure distribution in the riser and the slurry sampling where weight, W_s , and volume, V_s , sampled for a certain sampling time, t_{samp} , were measured. The sampling was repeated more than 4 times for each run. The static pressures at different heights along the riser were measured by a pressure transducer via. solenoid valves which were actuated by a personal computer.

4.2.3 Circulating Operation

The solid concentration in the slurry, ρ_s , the voidage, ϵ_s , as well as u_b and G_s were determined from the slurry sampling as follows;

$$\rho_s = W_s / V_s = \rho_p(1 - \epsilon_s) + \rho \cdot \epsilon_s \quad (4-2)$$

$$\epsilon_s = (\rho_p - \rho_s) / (\rho_p - \rho) \quad (4-3)$$

$$G_s = \rho_p V_s (1 - \epsilon_s) / (A_r t_{\text{samp}}) \quad (4-4)$$

$$u_b = V_s \epsilon_s / (A_r t_{\text{samp}}) \quad (4-5)$$

where A_r denotes the cross sectional area of the riser column. The mean bed voidage in the riser, ϵ , was calculated from the slope of a static pressure distribution curve. A negligible contribution of the wall friction to the distribution was detected, which was evaluated to be less than 5 % of the measured pressure at $370 \text{ mm} \cdot \text{s}^{-1}$, i.e., the maximum superficial liquid velocity in the present experiment. From Eqs. 4-1 to 4-5, the liquid-solid slip velocity in the riser, u_{slip} , is expressed as;

$$u_{\text{slip}} = V_s \{ \epsilon_s / \epsilon - (1 - \epsilon_s) / (1 - \epsilon) \} / (A_r t_{\text{samp}}) \quad (4-6)$$

Since no downflow of the particles near the riser wall was observed at u_b higher than u_t , thus, the following relationship can be obtained from Eq.4-6:

$$\epsilon_s > \epsilon \quad (4-7)$$

The above condition was employed to check if the slurry sampling was correctly carried out or not.

4.3 Results and Discussion

4.3.1 Bed Expansion Characteristics

At superficial liquid velocities, u_b , lower than the terminal velocity, u_t , the beds exhibited particulate fluidization for both GB93 and GB182 particles as expected from Foscolo et al's criterion (Foscolo et al, 1984). In **Fig.4.4** the bed voidage, ϵ , is plotted against u_b on a logarithmic scale. This result was obtained from a preliminary experiment by using a batch liquid-fluidized bed (BLFB) system. For both particles, the relationship between ϵ and u_b can be well represented by a straight line. It is well known that the relationship can be practically described by the so-called Richardson-Zaki (R-Z) equation, Eq.4-1 (Richardson and Zaki, 1954a). The index n for ϵ in Eq.4-1 for each particle bed is respectively found to be 4.5 and 3.8 from the linear relationship.

At u_b higher than u_t with circulation of the elutriated particles, a certain voidage was obtained throughout the riser. **Figure 4.5** shows typical axial distributions of the static pressure in the riser. The distributions were obtained at u_b ranged from 141 to 168 $\text{mm}\cdot\text{s}^{-1}$ for GB182. It is clearly seen that the static pressure decreases linearly with the height throughout the riser. It should be noted that the so-called S-shaped distribution which is frequently observed in CGFB risers is never produced within the present operating conditions. In addition, descent of particles near the riser wall was never observed, which is usually observed in both CGFB and BLFB. **Figure 4.6** shows typical variations of mean bed voidage, ϵ , with G_s for various u_b . It can be seen that ϵ increases with G_s . The rate of decrease of ϵ is first rapid and then gradually deteriorates as G_s increases. Furthermore, it should be noted that a significant increase of ϵ with increasing of u_b . In particular, for the coarser particles, an apparent increase in ϵ can be seen as u_b is varied from 23-29 $\text{mm}\cdot\text{s}^{-1}$ to 151-181 $\text{mm}\cdot\text{s}^{-1}$, which might be brought about by a transition of flow state from particulate fluidization to aggregative.

4.3.2 Liquid-Solid Slip Velocity

The observed ϵ at different u_b is plotted against u_{slp} calculated from Eq.4-6 for GB 92 and GB 182 in **Fig.4.7**. Here, the results obtained in BLFB are also plotted in the figure. It is clearly seen that u_{slp} in CLFB becomes considerably higher than that in BLFB for both particle beds. A sharp and linear increase of ϵ with u_{slp} can be seen at lower u_b of 80-95 $\text{mm}\cdot\text{s}^{-1}$ for GB93 and 23-29 $\text{mm}\cdot\text{s}^{-1}$ for GB 182 respectively, as is seen in BLFB. On the other hand, the increase of ϵ deteriorates gradually as u_b becomes higher. Comparing the existing correlation as drawn by the solid line, R-Z equation is found to be no more applicable to describe the relationship between ϵ and u_{slp} in CLFB riser. Since the fluidizing water flows through interstice among the particles the existing equations for flow

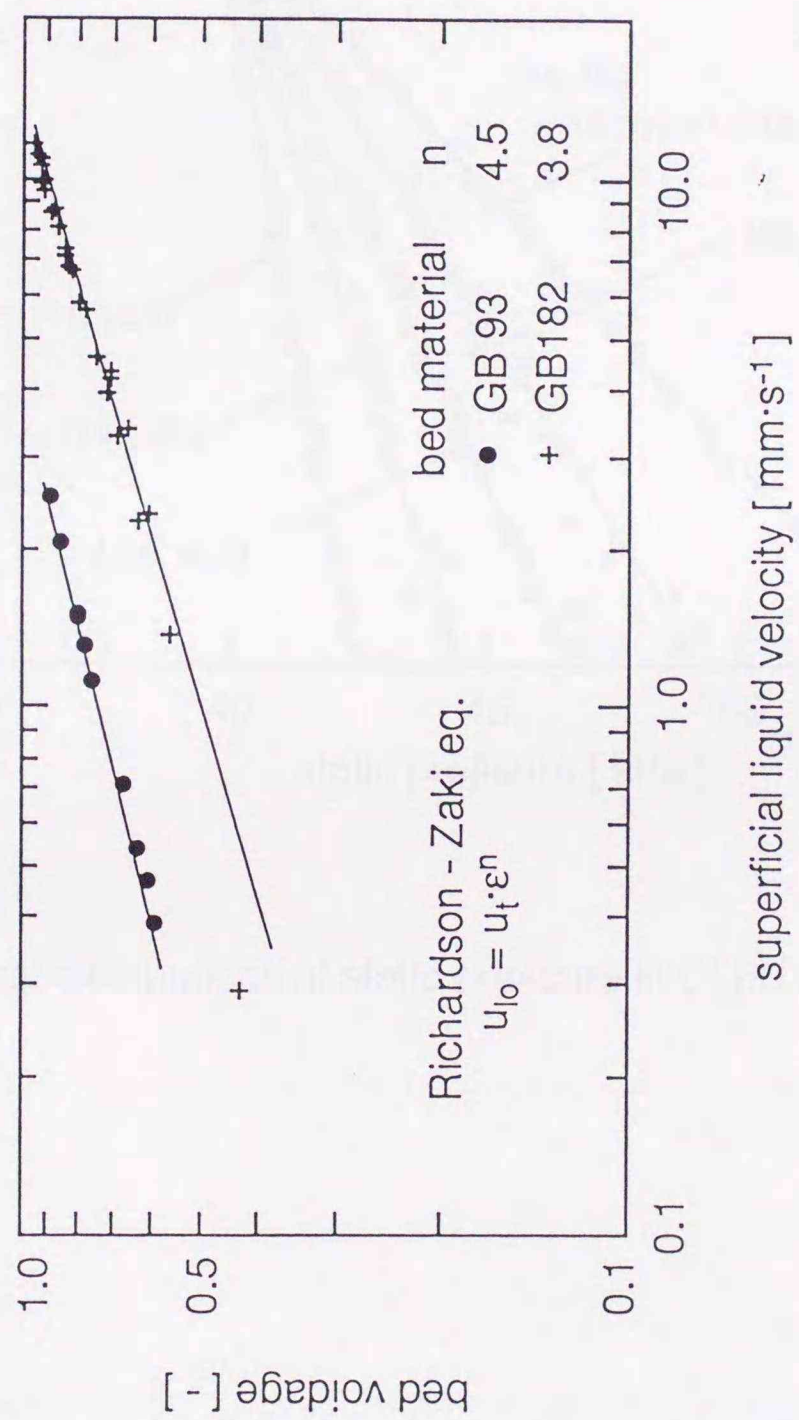


Fig.4.4 Particulate expansion characteristics of GB93 and GB192 beds

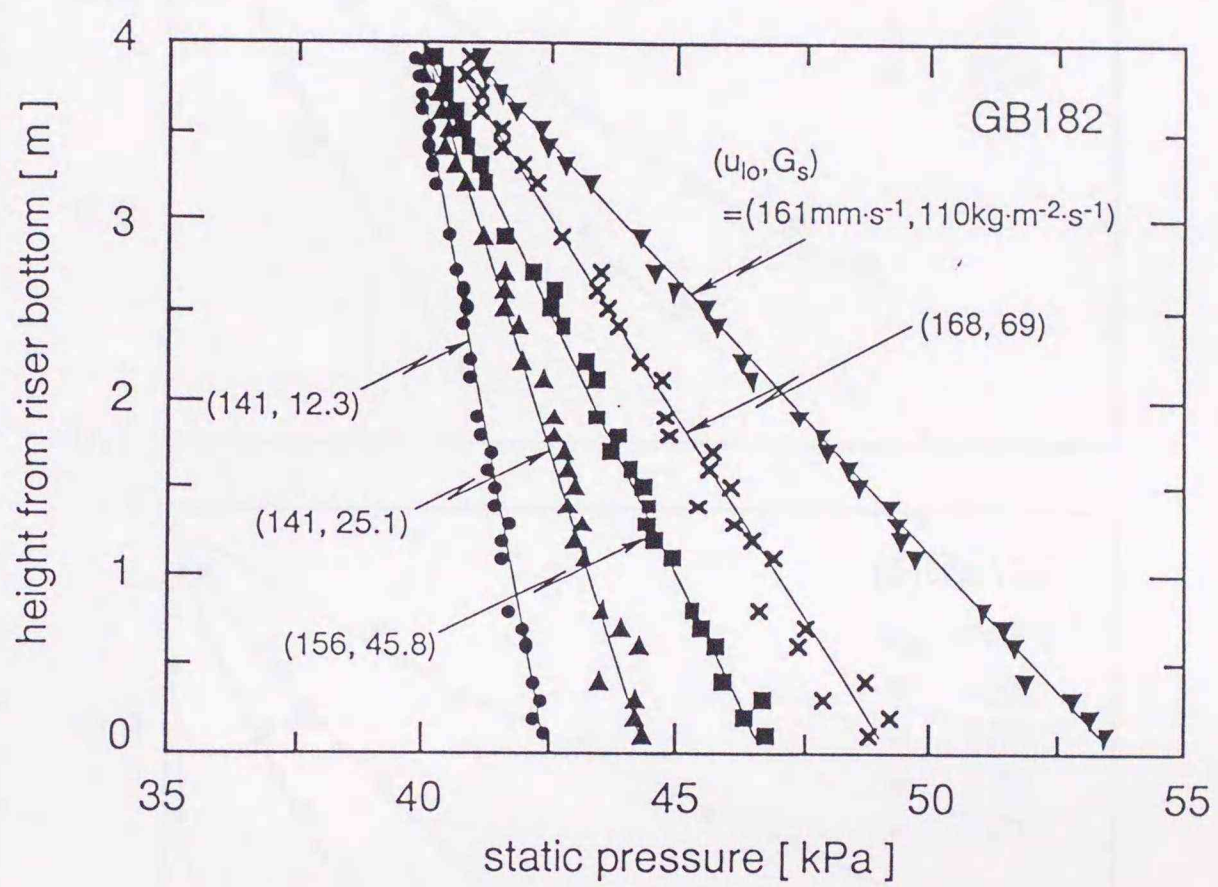


Fig.4.5 Typical distributions of static pressure in CLFB riser for GB182

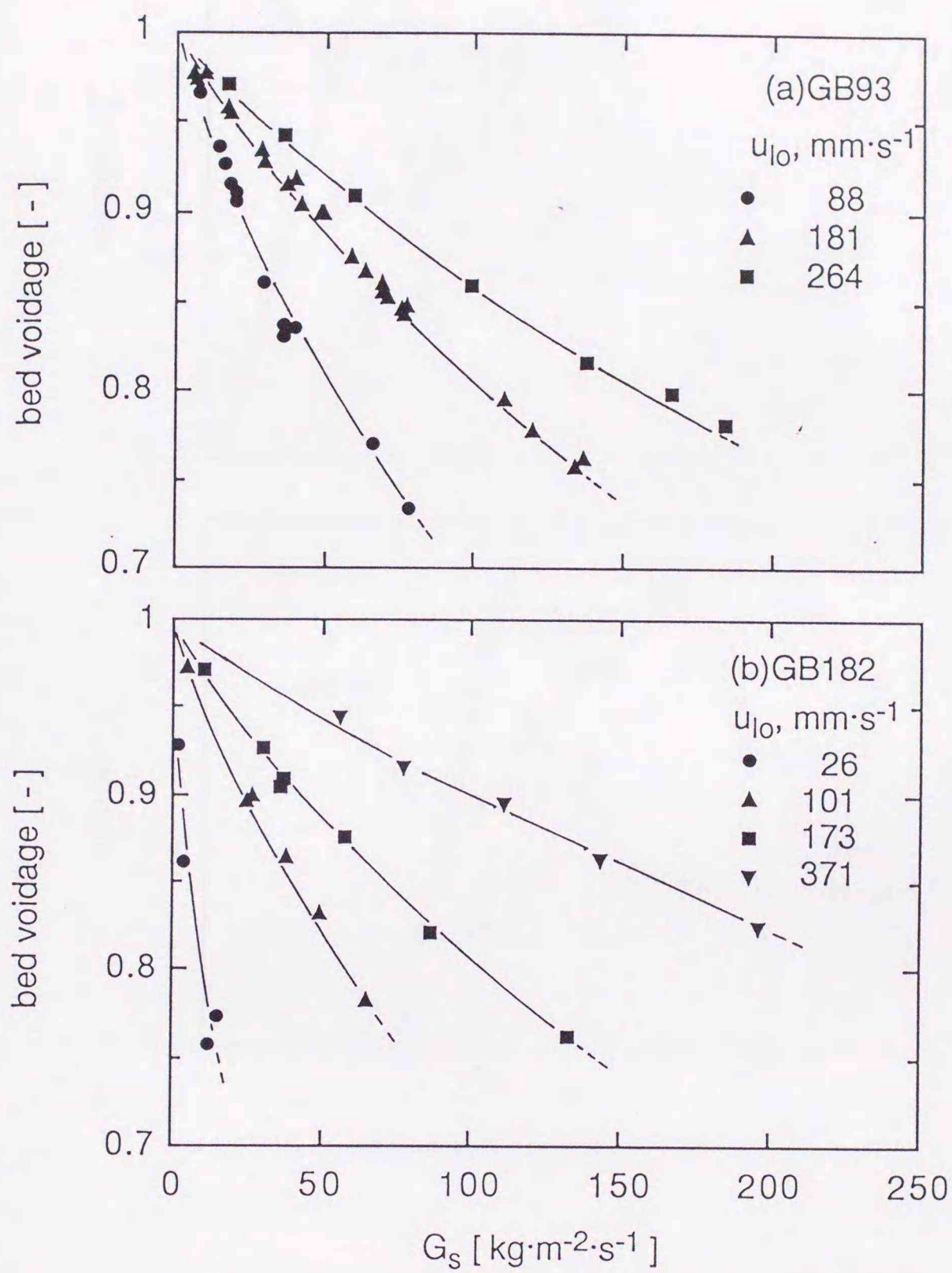


Fig.4.6 Variation of bed voidage with G_s at various u_{lo} in CLFB riser for different particles; (a) GB93 and (b)GB182.

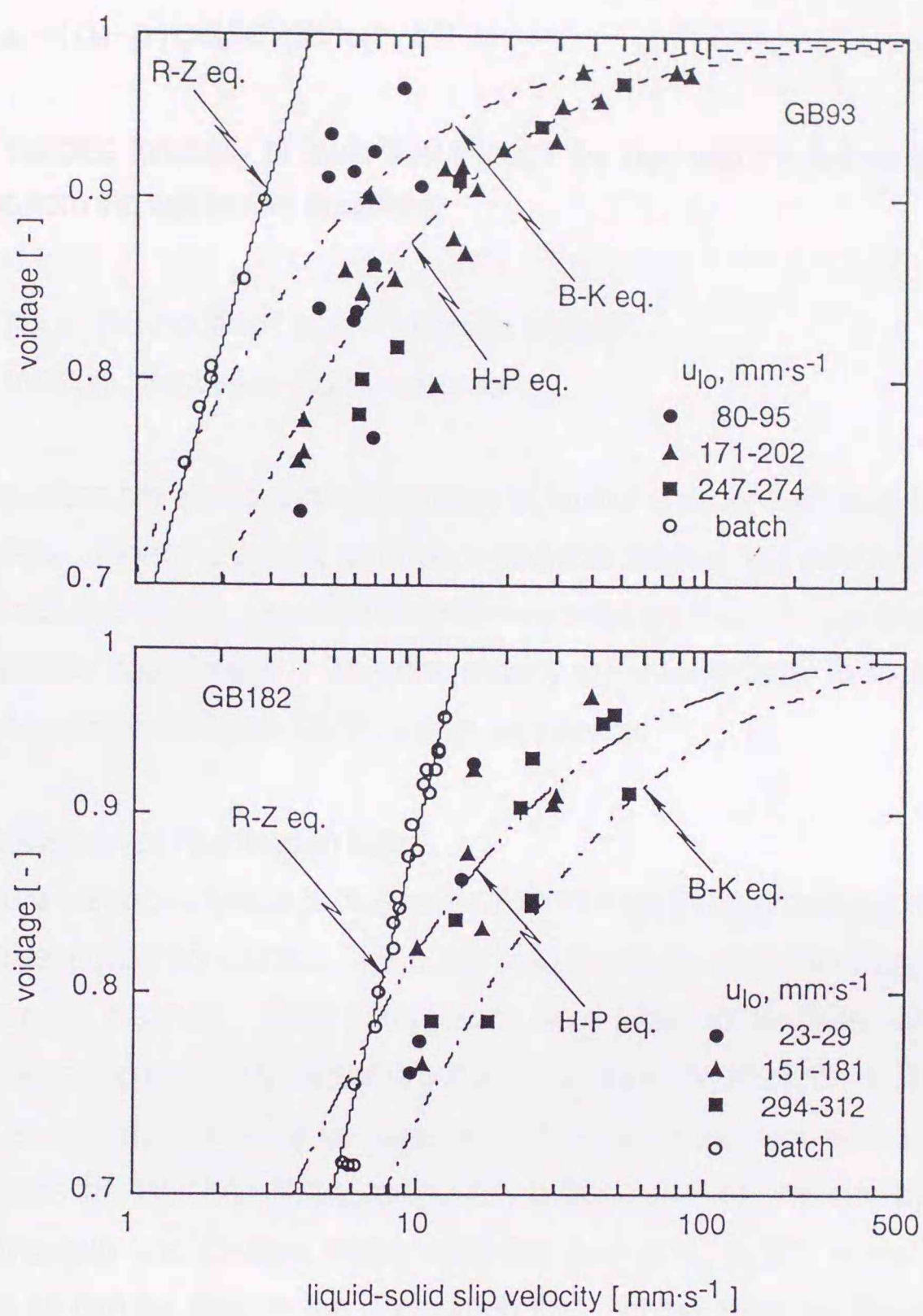


Fig.4.7 Change of bed voidage with liquid-solid slip velocity in BLFB and CLFB;
(a) GB93 and (b) GB182.

through a packed bed (Bird et al, 1960) can rationally be applied for those data:

$$u_{sf} = (\rho_p - \rho) \cdot g \cdot d_p^2 \cdot \epsilon^2 / (36 \cdot \kappa \cdot (1 - \epsilon)) \quad (4-8)$$

Here, κ denotes tortuosity of liquid flow through the bed and the following values were employed from the well-known equations:

$\kappa = 2$; the modified Hagen - Poiseuille equation

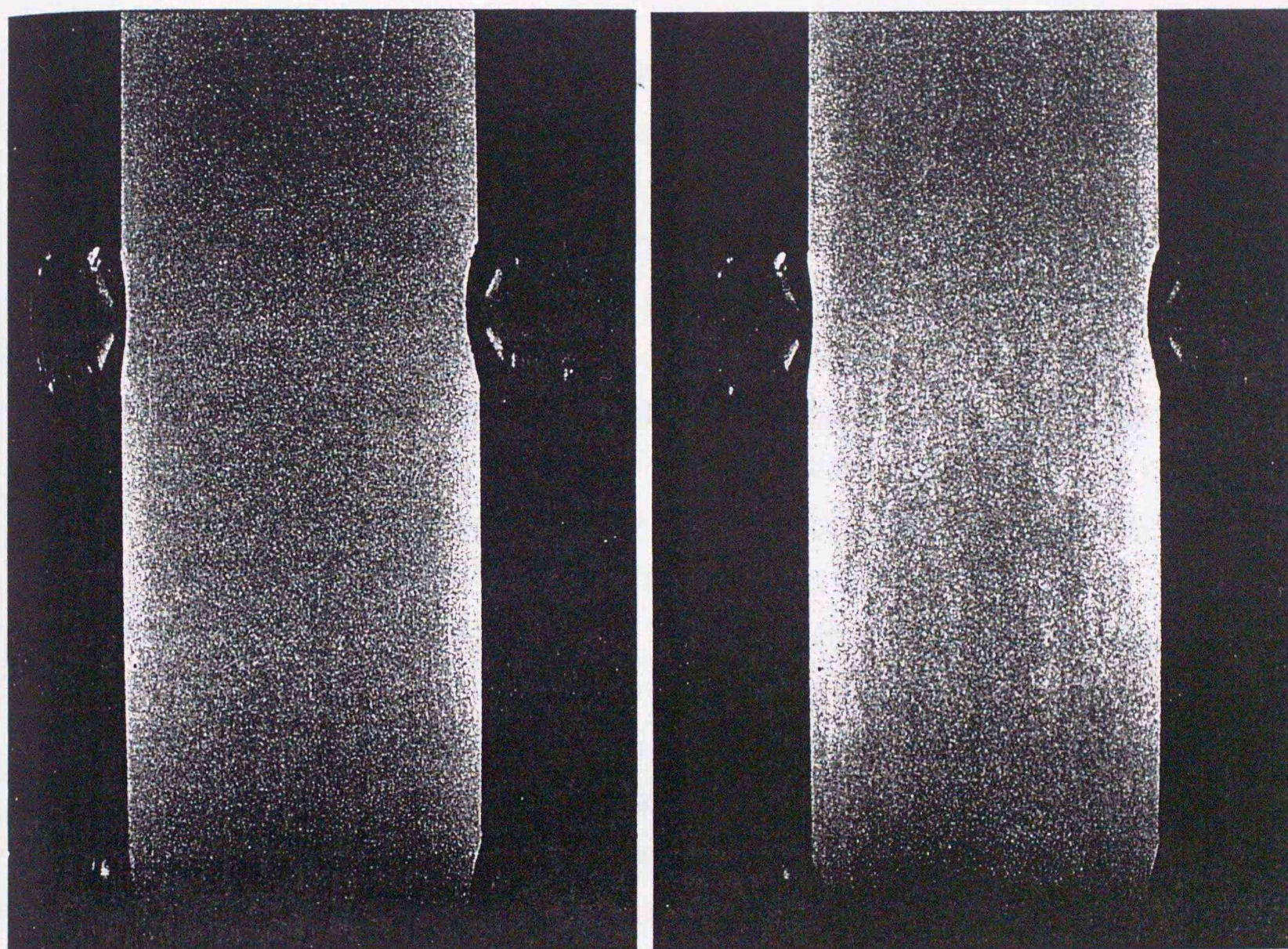
$\kappa = 25/6$; the Blake - Kozeny equation

These equations are also drawn in the figures by broken and dot-dash lines and compared with the data. From the figures, an obvious deviation between the observed data and the predicted results is found, a qualitative agreement between them can be seen at higher u_b for both particle beds though. This discrepancy may be attributed to the unsteady and complex flow structures in the CLFB as discussed below.

4.3.3 Transition of Fluidization State

Figure 4.8 shows typical photographs taken through the riser wall at a height of 1.8 m from the riser bottom for GB182. Here, the solid flow in the riser was visualized by back-lighting through the bed. Both photographs were taken at conditions where the bed voidages were approximately the same but u_b is considerably different. In the photograph (a) taken at $u_b = 101 \text{ mm} \cdot \text{s}^{-1}$, a homogeneous shadow can be seen over the whole riser. This suggests that the bed is in the particulate fluidized state as predicted by Foscolo and Gibilaro (Foscolo and Gibilaro, 1984) while the bed at $u_b = 371 \text{ mm} \cdot \text{s}^{-1}$ is no more particulate as can be seen in the photograph (b). At the lower u_b , the particles were transported upward in a laminar manner. At the higher u_b , on the other hand, the particles tend to form aggregates having irregular shapes in the riser and the fluidizing water passes among the aggregates in the form of streaks. Thus, complex motions of particles and inhomogeneous distribution of the bed voidage were eventually observed through the CLFB riser wall as suggested elsewhere (Grbavcic et al, 1992; Liang et al, 1996).

The inhomogeneity in the spatial distribution of the solid particles is further examined by analyzing the luminous intensity distribution in the photographic images through the following procedure. First, the photographic image was converted to a bit map image consisting of 256 gray scales by digitizing spatial distribution of the luminous intensity over the photograph with an image scanner. The digitized data were stored in a personal



(a) $\varepsilon = 0.900$

(b) $\varepsilon = 0.895$

Fig.4.8 Typical photographs of backlighted bed of GB182 taken at different operating conditions; (a) $u_{l0} = 101 \text{ mm}\cdot\text{s}^{-1}$, $G_s = 26.8 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$ and (b) $u_{l0} = 371 \text{ mm}\cdot\text{s}^{-1}$, $G_s = 111.1 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$

computer. Then, the luminous intensity at each pixel, which corresponds to the partial bed voidage, was transversely scanned along certain horizontal lines chosen arbitrarily on the photograph. Typical raw signal for different u_b obtained from the above procedure is shown in **Fig.4.9**. Here, since base level of transmittancy through the riser is noticeably distorted due to the geometry of the riser column, a quadratic regression curve was analytically established to fit raw signal with correlation coefficients above 0.93. The distortion was equalized by subtracting the data from the regression curve. Then successive deviations of the data from the regression curve can be extracted, as drawn in the same figure. It is clearly found that the spatial fluctuation obtained at the higher u_b (Fig.4.8(b)) can be characterized as a signal exhibiting large amplitude and sharp spikes, compared to that obtained at the lower u_b (Fig.4.8(a)). The fluctuations of the luminous intensity for different u_b were analyzed by the power spectral density (PSD) function. In **Figs.4.10** and **4.11**, typical PSD functions calculated from the extracted fluctuations obtained for different u_b are respectively drawn on normal (Fig.4.10) and logarithmic (Fig.4.11) diagrams. Here, sampling window used for analyzing the PSD contains 4096 points. In both figures, the abscissa denotes the spatial frequency. As can be seen in Fig.4.10, several peaks appear within the range from 0.2 to 0.5 mm^{-1} for higher u_b while such peaks become ambiguous for lower u_b . The result suggests that the spatial scale of the aggregates observed through the riser can be estimated ranged from 2 to 5 mm. In addition, Fig.4.11 shows that the PSD function exhibits a continuous spectrum with many spikes and exponential decay in the higher frequency region for both extreme fluidization states. The appearance of the PSD function is generally attributed to chaotic fluctuation of the signal concerned and may reflect the fractal texture in the spatial distribution of the voidage through the riser.

Thus, diagnosis on the complexity of the luminous intensity fluctuation for different fluidization regimes was conducted on the basis of a fractal analysis. Fractal dimension, D_f , has extensively been used as a quantifier of geometric complexity of an object (Feder, 1988; Hilborn, 1994; Luewisuthichat et al, 1996). The dimension is practically calculated through the box-counting method which is employed in the present work. The fluctuating signal of the luminous intensity was mapped onto a regular mesh of size, s , as shown in **Fig.4.12**, and then the number of boxes required to cover the curve, $N(s)$, was counted. Variations in $N(s)$ with progressively decreasing s were plotted on a logarithmic diagram. On the diagram the linear relationship between $N(s)$ and s is obtained and the slope directly yields the fractal dimension, D_f . **Figure 4.13** shows the typical change in number of the boxes with the segment size from 10 to 100 pixels which corresponds to the length from

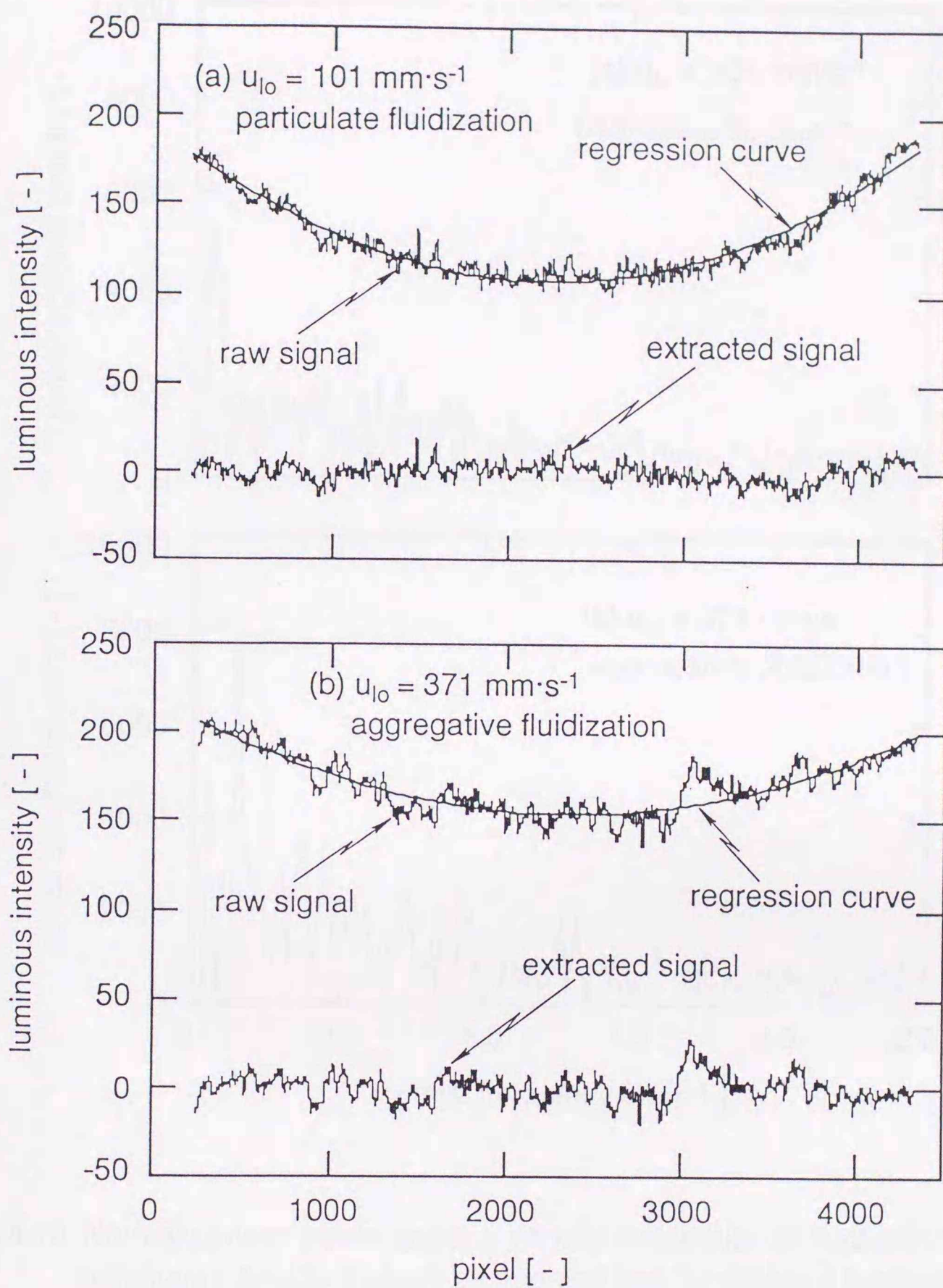


Fig.4.9 Signal processing in image analysis

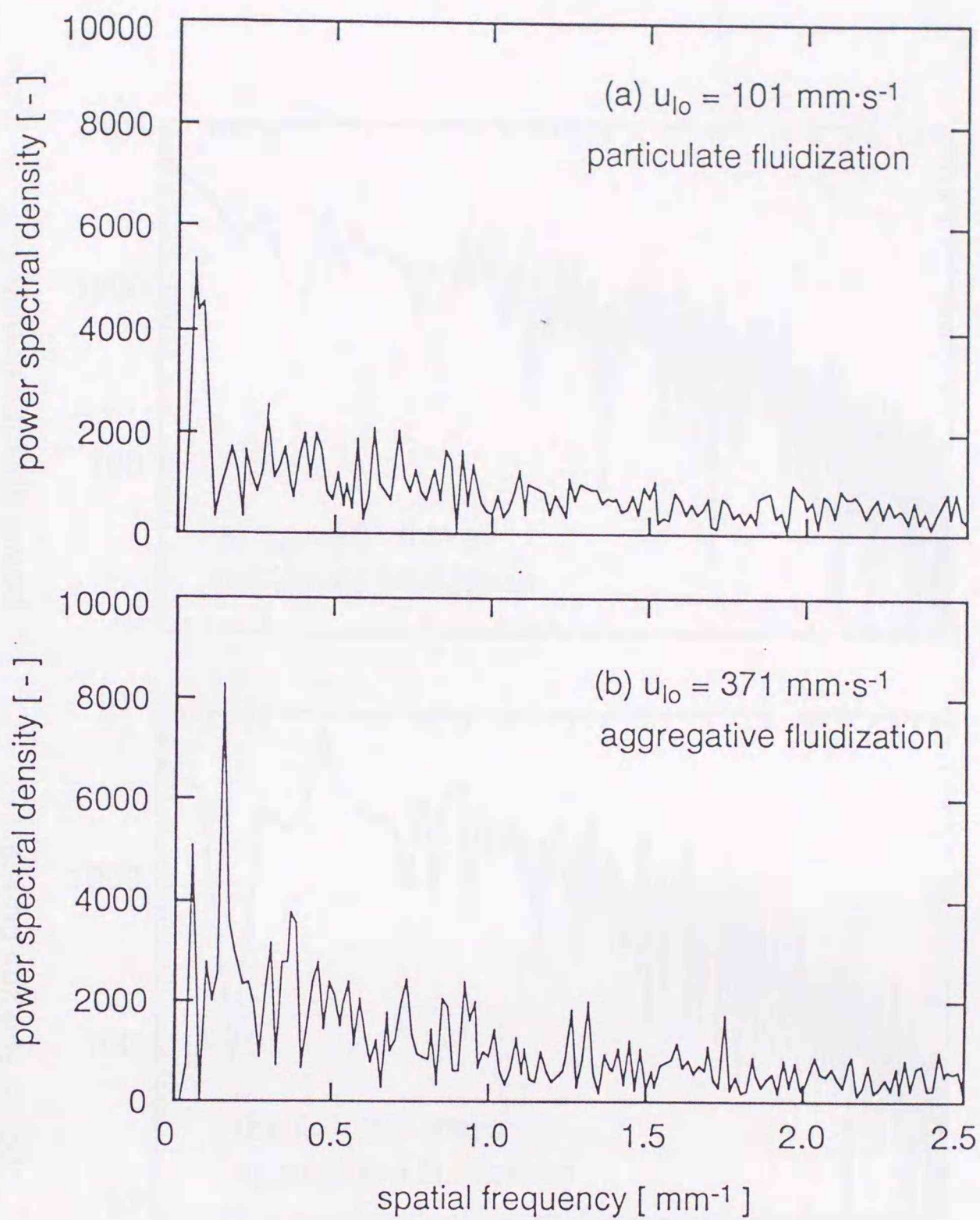


Fig.4.10 Normal plots of power spectral density distribution of fluctuation of luminous intensity through backlighted bed for different fluidization states; (a) $u_{lo} = 101 \text{ mm}\cdot\text{s}^{-1}$ and (b) $u_{lo} = 371 \text{ mm}\cdot\text{s}^{-1}$

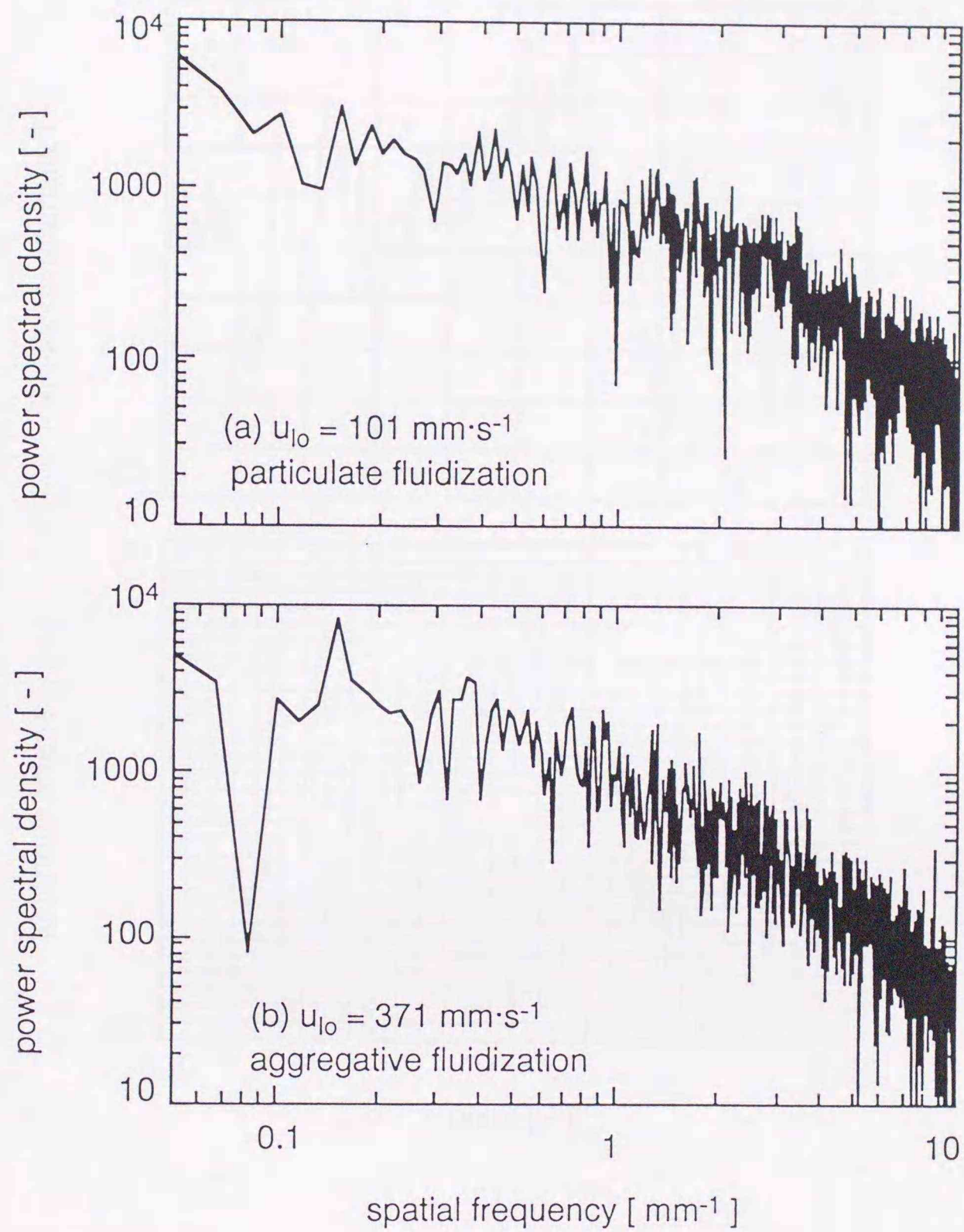


Fig.4.11 Logarithmic plots of power spectral density distribution of fluctuation of luminous intensity through backlighted bed for different fluidization states; (a) $u_{lo} = 101 \text{ mm} \cdot \text{s}^{-1}$ and (b) $u_{lo} = 371 \text{ mm} \cdot \text{s}^{-1}$.

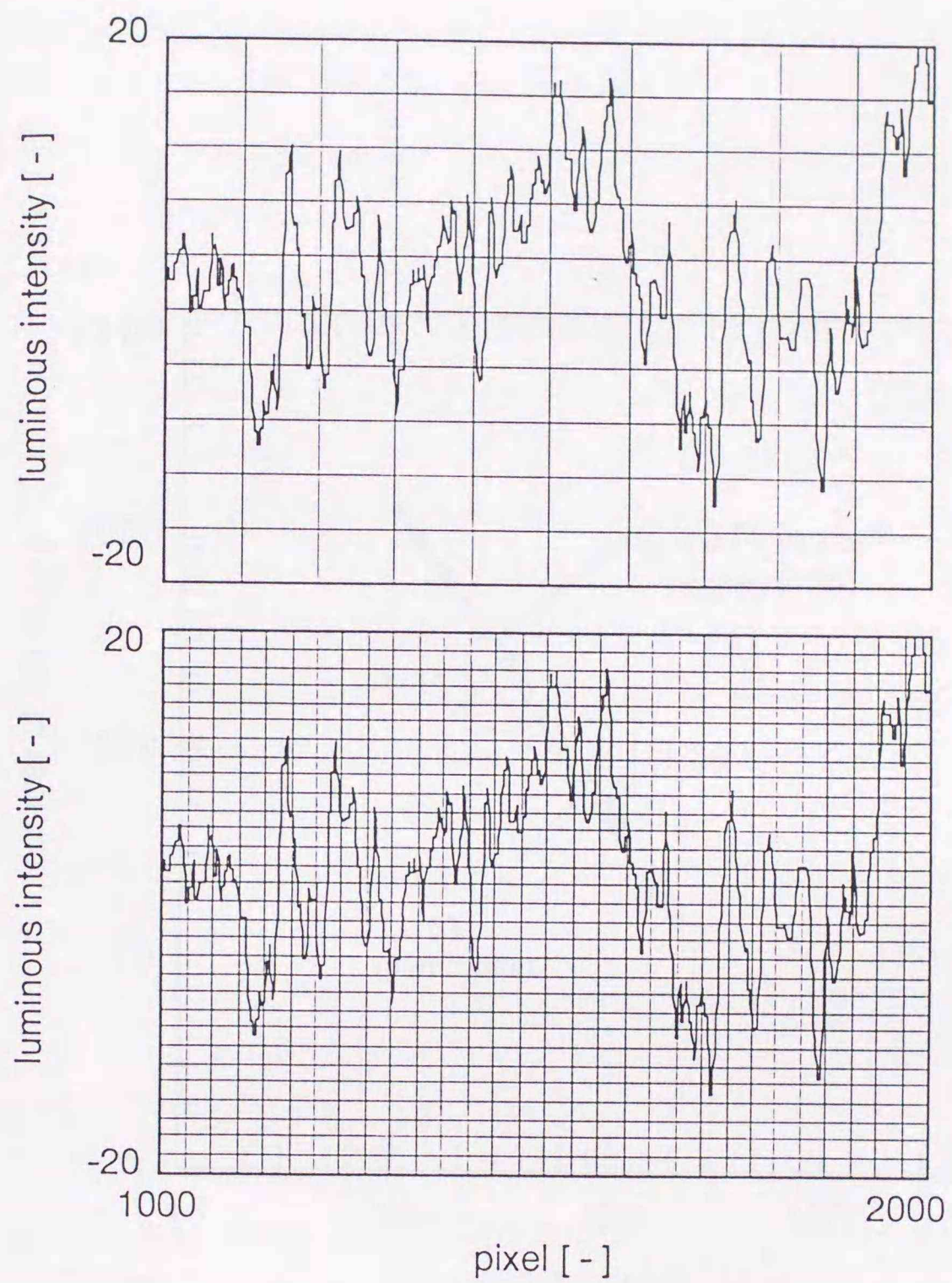


Fig.4.12 Example of partitioning of fluctuating signal with different segment sizes

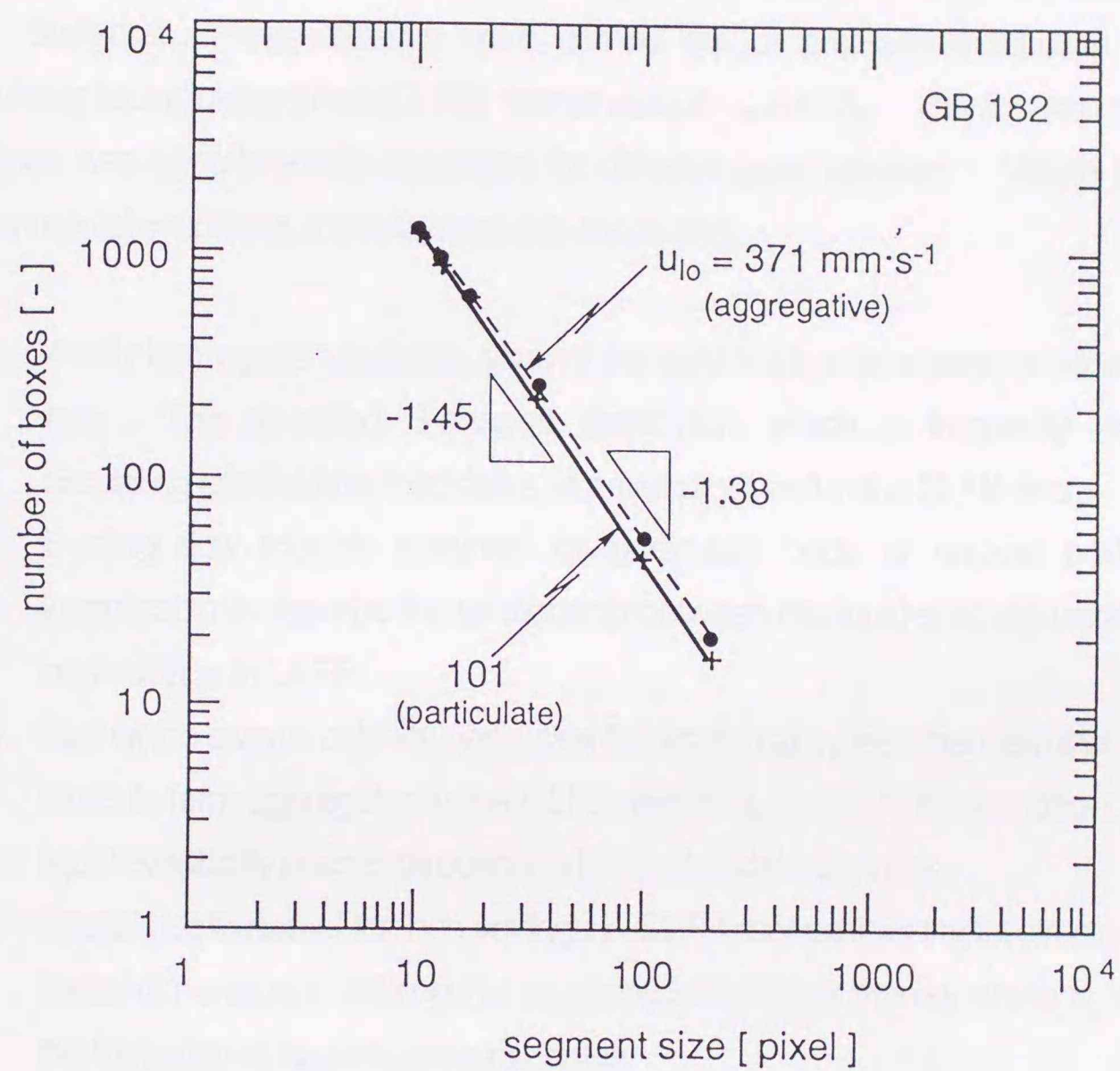


Fig.4.13 Typical change of box number with segment size for fluctuation of luminous intensity obtained at different fluidization states

0.15 to 2.93 mm on the original photography. As can be seen in the figure, the linear relationships between $N(s)$ and s are obtained for both particulate fluidization and aggregative one and provide non-integer dimensions. Averaged D_f for the particulate and aggregative fluidizations is respectively found as 1.45 and 1.39. The slightly lower D_f in the aggregative fluidization state indicates that the degree of complexity of the voidage distribution in the riser decreases due to the formation of the larger scale aggregates.

4.4 Conclusion

Systematic measurements were carried out of pressure distribution through a circulating liquid-fluidized bed (CLFB) riser at various u_b and G_s . The flow structure of solid particles was experimentally examined for different sized particles. Within the present experimental conditions, the followings are concluded:

- 1) Axially homogeneous distributions of the solid holdup are always observed in CLFB riser. The so-called S-shaped distribution, which is frequently observed in circulating gas fluidized bed risers, is never produced in the CLFB riser.
- 2) Existing flow models obtained for particulate beds or packed beds become inapplicable to describe the relationship between the liquid-solid slip velocity and the bed voidage in CLFB.
- 3) Bed which always exhibits particulate fluidization at u_b less than terminal velocity, u_t , tends to form aggregates in the CLFB riser at $u_b > u_t$. At this condition, a fluidizing liquid eventually passes through the bed in the state of streak.
- 4) Spatial distribution of the bed voidage in CLFB riser exhibits fractal structure. Lower fractal dimension is obtained at u_b appreciably higher than u_t , which is attributed to the formation of larger scale aggregates.

Literature Cited

- Bird, R. B., W. E. Stewart and E. N. Lightfoot, TRANSPORT PHENOMENA, 2nd ed., p.196-200, John Wiley & Sons, New York, U.S.A.(1960)
- Couderc, J. P., Incipient Fluidization and Particulate System, Fluidization, 2nd ed. (eds. J. F. Davidson, R. Clift and D Harrison), p.1-46, Academic Press, London, U.K.(1985)
- Foscolo, P. U. and L. G. Gibilaro, "A Fully Predictive Criterion for the Transition between Particulate and Aggregate Fluidization", Chem. Eng. Sci., 39, 1667-1675 (1984)
- Foscolo, P. U., L. G. Gibilaro, S. P. Waldram and R. Di Felice, "The Application of a Particulate Model to Fluidisation Dynamics", Proc. 5th Int. Conf. on FLUIDIZATION, p9-16, Elsinore, U.S.A. (1986)
- Foscolo, P. U. and L. G. Gibilaro, Fluid Dynamics Stability of Fluidised Suspensions: The Particle Bed Model", Chem. Eng. Sci., 42, 1489-1500 (1987)
- Feder, J., "FRACTALS", Plenum, Chap.8, New York, U.S.A. (1988)
- Gibilaro, L. G., R. Di Felice and P. U. Foscolo, On the Minimum Bubbling Voidage and the Geldart Classification for Gas-Fluidised Beds, Powder Tech., 56, 21-29 (1988)
- Gibilaro, L. G., I. Hossain and P. U. Foscolo, Aggregate Behaviour of Liquid Fluidized Beds, Can. J. Chem. Eng., 64, 931-938(1986)
- Grbavcic, Z. B., R. V. Garic, D. V. Vukovic, Dz. E. Hadzismajlovic, H. Littman, M. H. Morgan III and S. Dj. Jovanovic, "Hydrodynamic modeling of vertical liquid-solids flow", Powder Tech., 72, 183-191(1992)
- Hassett, N. J., "Flow Pattern in Particle Beds", Nature, 189, 997-998(1961)
- Hilborn, R. C., Chaos and Nonlinear Dynamics, Chapter 9, Oxford University Press, New York, U.S.A. (1994)
- Jackson, R., "The Mechanics of Fluidised Beds: Part I: The Stability of The State of Uniform Fluidisation", Trans. Instrn. Chem. Engrs., 41, 13-21(1963)
- Khan, A. R. and J. F. Richardson, "Pressure Gradient and Friction Factor for Sedimentation and Fluidisation of Uniform Sphere in Liquids", Chem. Eng. Sci., 45, 255-265(1990)
- Kmiec A, "Particle Distributions and Dynamics of Particle Movement in Solid-Liquid Fluidized Beds", Chem. Eng. J., 15, 1-12(1978)
- Liang W. -G., J. -X. Zhu, Y. Jin, Z. -Q. Yu, Z. -W. Wang and J. Zhou, "Radial Nonuniformity of Flow Structure in a Liquid Solid Circulating Fluidized Bed", Chem. Eng. Sci., 51, 2001-2010(1996)
- Luewisuthichat, W., A. Tsutsumi, K. Yoshida, "Fractal Analysis of Particle Trajectories in Three-Phase Systems", Trans. IChemE., 73A, 222-227(1995)

- Richardson, J. F. and W. N. ZAKI, "The Sedimentation of a Suspension of Uniform Spheres under Conditions of Viscous Flow", Chem. Eng. Sci., 3, 65-73(1954a)
- Richardson, J. F. and W. N. ZAKI, "Sedimentation and Fluidisation: Part I", Trans. Int. Chem. Engrs., 32, 35-53(1954b)
- Rowe, P. N., "A Convenient Empirical Equation for Estimation of Richardson-Zaki Exponent", Chem. Eng. Sci., 43, 2795-2796(1987)
- Volpicelli, G., L. Massimilla and F. A. Zenz, "Nonhomogeneities in Solid-Liquid Fluidization, Chem. Eng. Prog. Symp. Ser., 62, 42-50(1966)
- Wallis, G. B., "One Dimensional Two Phase Flow", McGraw-Hill, New York, U.S.A. (1969)
- Wilhelm, R. H. and M. Kwauk, "Fluidization of Particles, Chem. Eng. Prog., 44, 201-218(1948)

Chapter 5

CHAOTIC CHARACTERISTICS OF LIQUID-SOLID FLOW IN A CIRCULATING LIQUID-FLUIDIZED BED RISER

5.1 Introduction

In the previous chapter, even liquid-fluidized solid particles exhibit an aggregative fluidization state when the particles are fluidized at superficial liquid velocities, u_b , higher than the terminal velocity of the single particle, u_t . The state has never been observed in existing investigations since all experiments were carried out in batch fluidized bed at $u_b < u_t$. In the state, fluctuations with time are observed in most of flow characteristics such as the solid and liquid velocities, the static bed pressure and the local bed voidage (e.g. Luewisuthichat et al, 1995). These fluctuations would reflect turbulent hydrodynamic interactions between the solid particles and liquid which may be induced by the high slip velocity between them.

In the field of chemical reactor engineering, it has been made clear that multiphase reactors frequently exhibit chaotic fluctuations of the above velocities, pressure, voidage etc, as a direct consequence of the turbulent interactions between the phases. Recently, deterministic chaos theory has successfully been adopted to analyze output signals from various probes for the complex phenomena (Hilbon; 1994). On the basis of the theory, complexity of turbulent phenomena is quantified in terms of the correlation dimension and the Kolmogorov entropy. In most of existing investigations, the chaotic flow structure of multiphase system has been characterized by pressure fluctuations in the bed (e.g. Bleek and Schouten; 1993, Bouillard and Miller; 1994, Daw et al; 1990, Huilin et al; 1997).

Variation of the fluidization state with liquid velocity would result in that of heat and mass transfer characteristics in the bed. Usually, time-averaged heat transfer characteristics have been investigated at various operating conditions in batch liquid-fluidized beds for those between the bed and column wall (Fan; 1989, Kato et al; 1981, Richardson and Mitson; 1958, Richardson and Smith; 1962) or for those between the bed and immersed objects (Richardson et al; 1976, Khan et al; 1983). These investigations provide an insight into the steady heat transfer characteristics but never seem to be sufficient to explain instantaneous characteristics which strongly depend on the fluctuations in liquid-solid flow mentioned above.

In this chapter, the deterministic chaos analysis has been applied to time-series data of local voidage and heat transfer fluctuations measured by novel probes developed here for explanation of their variations with operating condition as well as their correspondence to those of flow patterns in a circulating liquid-fluidized bed riser.

5.2 Experimental

5.2.1 Probes

Figure 5.1 shows a schematic diagram of the experimental circulating liquid-fluidized bed system. Operating procedure of the system has already been described in the preceding chapter in detail. In order to examine variation of the local flow pattern of the solid particles and heat transfer rate with time, an optical transmittance probe and heat transfer probe(8) have been employed. Measurement system and the details of the probes presently developed are shown in Fig. 5.2(a) & (b), respectively. The local voidage was detected with the optical transmittance probe. The probe consists of light emitting tube and detecting optical fibers supported by 4.0 mm o.d. stainless tube (Fig.5.2(a)). An aluminum plate with a pinhole was attached at the tip of the emitting tube to narrow the He-Ne laser beam with initial diameter of 2.0 mm to 0.2 mm, which enable to detect a passage of a single particle (Kikuchi et al; 1996). The narrowed beam introduced through the pinhole was received by the detecting optical fiber having a diameter of 0.5 mm, which was faced to the emitting fiber tip with a gap of 4.0 mm. The beam transmitted through the bed was introduced to the phototransistor via the optical fiber and converted into the voltage output. On the other hand, the instantaneous heat transfer rate is measured with a heat transfer probe. The probe is developed based on the principle of the constant-temperature hot-wire anemometry. A thin piece of tungsten wire having a diameter of 10 μm was employed as a detecting filament. The wire was stretched between two tips of stainless supports and welded with a spot welder. A gap between the supports is exactly fixed to 2.0 mm. The wire and the supports were coated with an anti-water varnish to prevent a direct contact of water with the filament. Then, the supports were connected to one of the arms in the Wheatstone's bridge circuit. A block diagram of the electric bridge circuit presently used is shown in Fig. 5.3.

The optical fiber probe was horizontally mounted at a riser height of 3.1 m and the emitting and detecting tips were located at the axis of the riser. The heat transfer probe was also inserted horizontally at the same level of the riser and the tip was located very close to that of the optical probe, as shown in Fig.5.2(b).

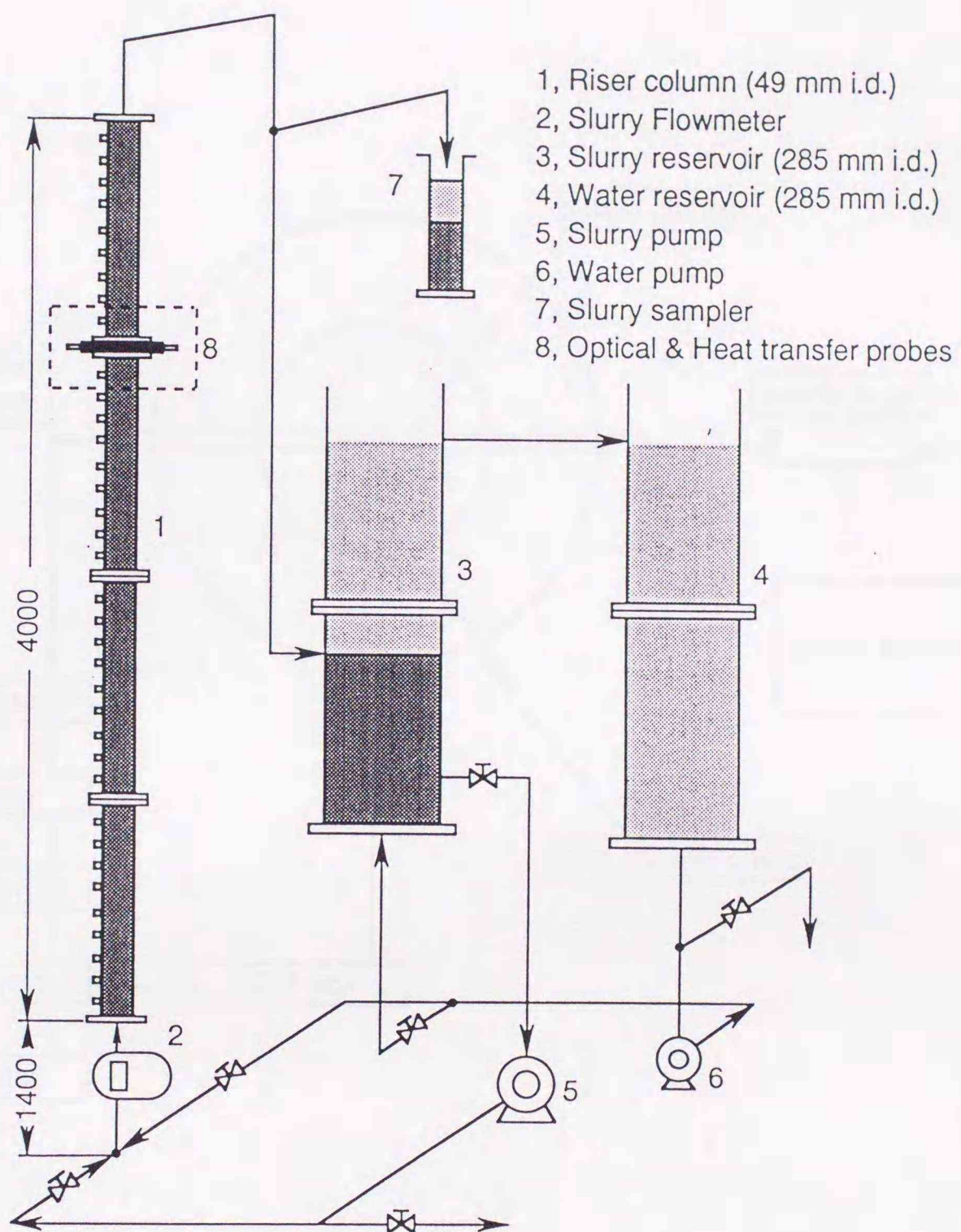


Fig.5.1 Experimental apparatus of circulating liquid-fluidized bed system.

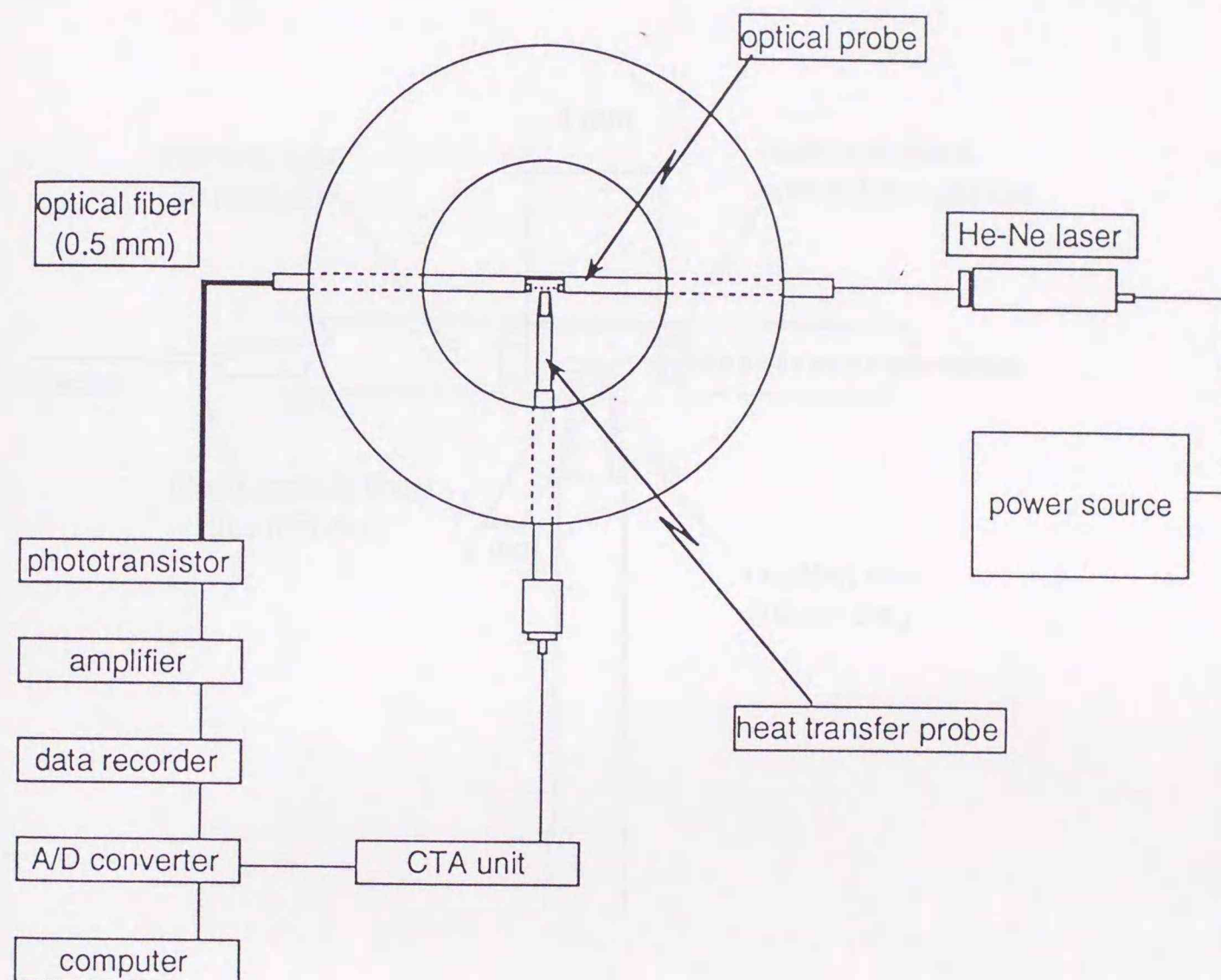


Fig.5.2(a) Schematic diagram of optical and heat transfer probes

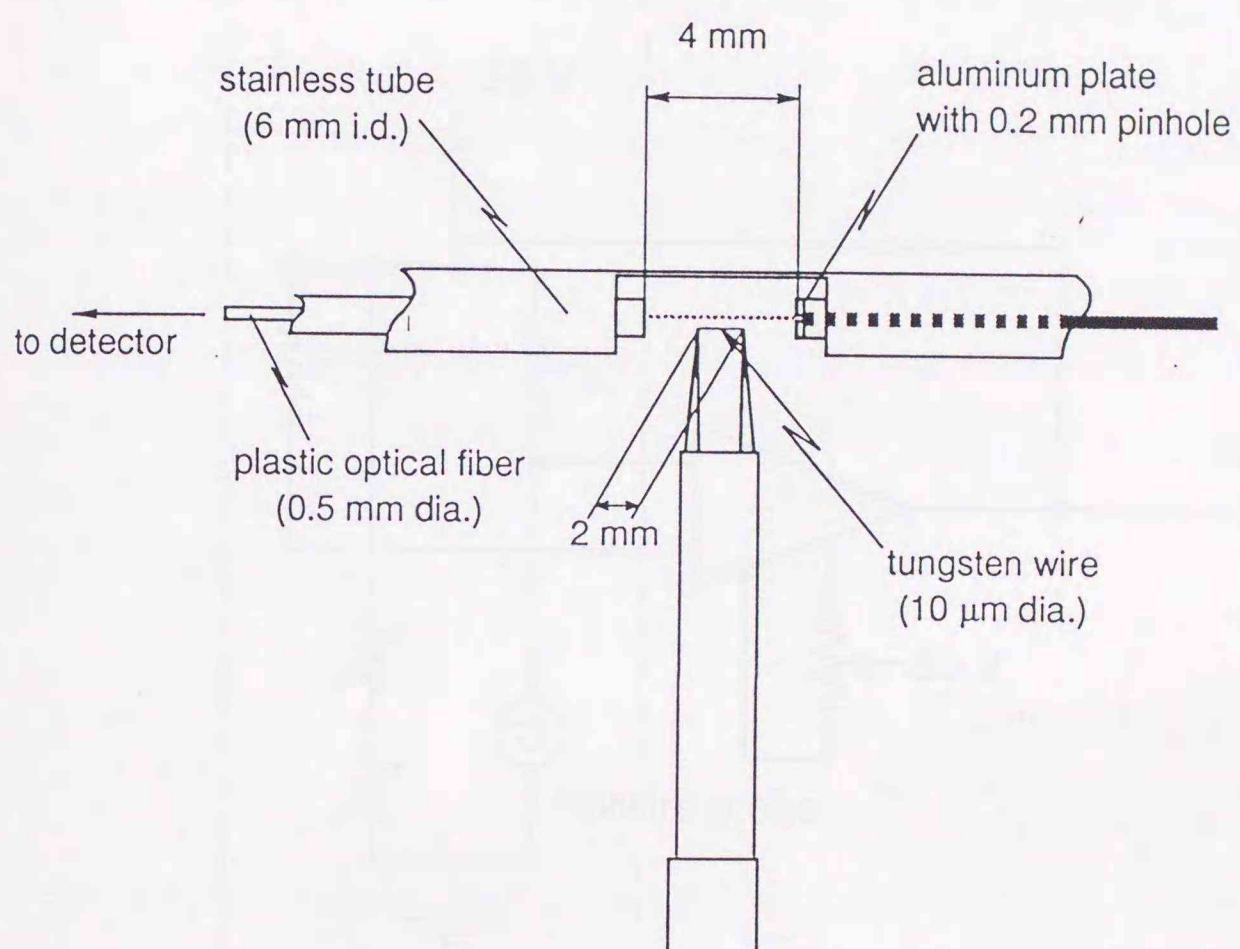


Fig.5.2(b) Configurations of optical and heat transfer probes.

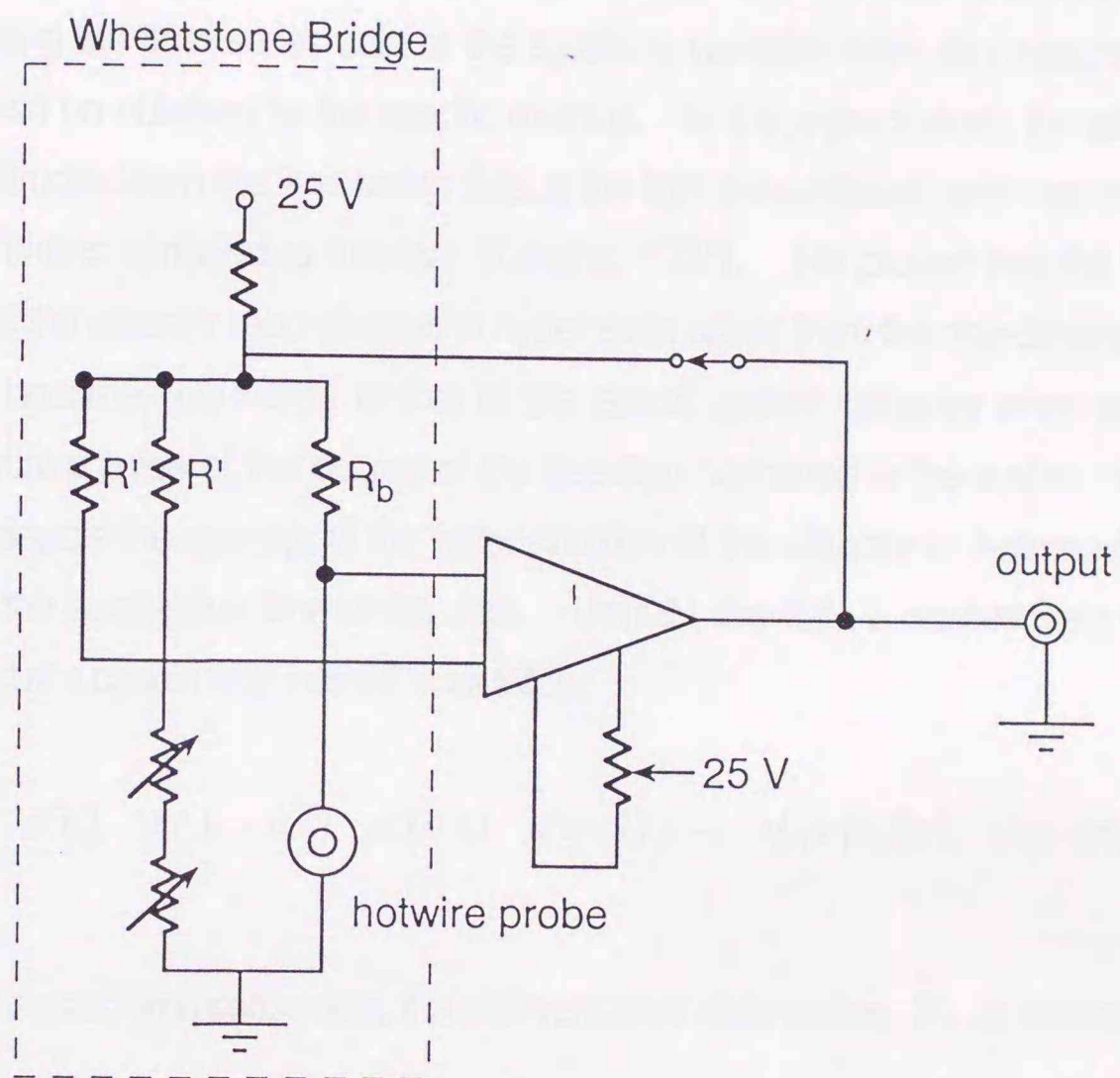


Fig.5.3 Schematic diagram of electric circuit for constant temperature anemometry

The signals obtained by both probes were recorded with a data recorder. The data was digitized with A/D converter at a sampling interval of 1/1000 sec and stored as successive data in a personal computer.

5.2.2 Data Analysis

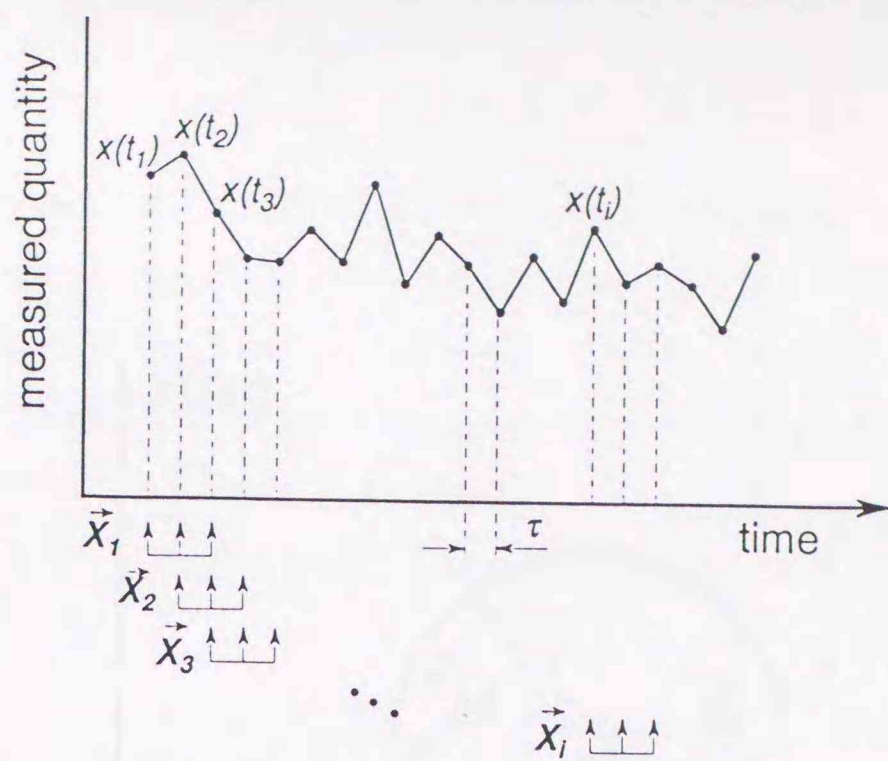
The dynamic flow characteristics in the circulating liquid-fluidized bed riser is represented with an attractor. The attractor is practically reconstructed in the multi-dimensional state space by using the time series data of the system's variables such as pressure, voidage, velocity and so on obtained by the specific method. In the present study, the attractor has been reconstructed from the time series data of the light transmittance and heat transfer rate based on Takens' embedding theorem (Takens; 1981). He proved that the geometric complexity of the attractor reconstructed in hyper state space from the one-dimensional time series data becomes equivalent to that of the actual system behavior when embedding dimension attains twice of the degree of the freedom contained in the system concerned. **Figure 5.4** depicts the concept of the reconstruction of the attractor in 3-dimensional state space from the successive time-series data. Usually, the data is successively measured and recorded at a certain time interval, τ , as follow:

$$x(t_1), x(t_2), x(t_3), \dots, x(t_i), x(t_i + \tau), x(t_i + 2\tau), \dots, x(t_i + (n-1)\tau), x(t_i + n\tau)$$

From the measured time series data, the d -dimensional state vector, $\vec{X}_i$, is reconstructed as following:

$$\vec{X}_i = \{x(t_i), x(t_i + \tau), x(t_i + 2\tau), \dots, x(t_i + (d-2)\tau), x(t_i + (d-1)\tau)\} \quad (5-1)$$

where $x(t_i)$ and d denote the successive data of voidage or heat transfer and the embedding dimension, respectively. Generally, an orbit of a series of the above state vector becomes geometrically simple when the flow in the riser predominantly indicates a periodic behavior. If the system exhibits chaotic fluctuation, the geometric structure of the attractor becomes inevitably complex. The complexity of the reconstructed attractor, which corresponds to the degree of the freedom in the attainable states of the system, is analytically evaluated with the chaos statistics such as correlation dimension and Kolmogorov entropy of the reconstructed attractor through the following procedure.



$$\vec{X}_1 = (x(t_1), x(t_1+\tau), x(t_1+2\tau))$$

$$\vec{X}_2 = (x(t_2), x(t_2+\tau), x(t_2+2\tau))$$

$$\vec{X}_3 = (x(t_3), x(t_3+\tau), x(t_3+2\tau))$$

$$\vdots$$

$$\vec{X}_i = (x(t_i), x(t_i+\tau), x(t_i+2\tau))$$

$$\vdots$$

$$\vec{X}_n = (x(t_n), x(t_n+\tau), x(t_n+2\tau))$$

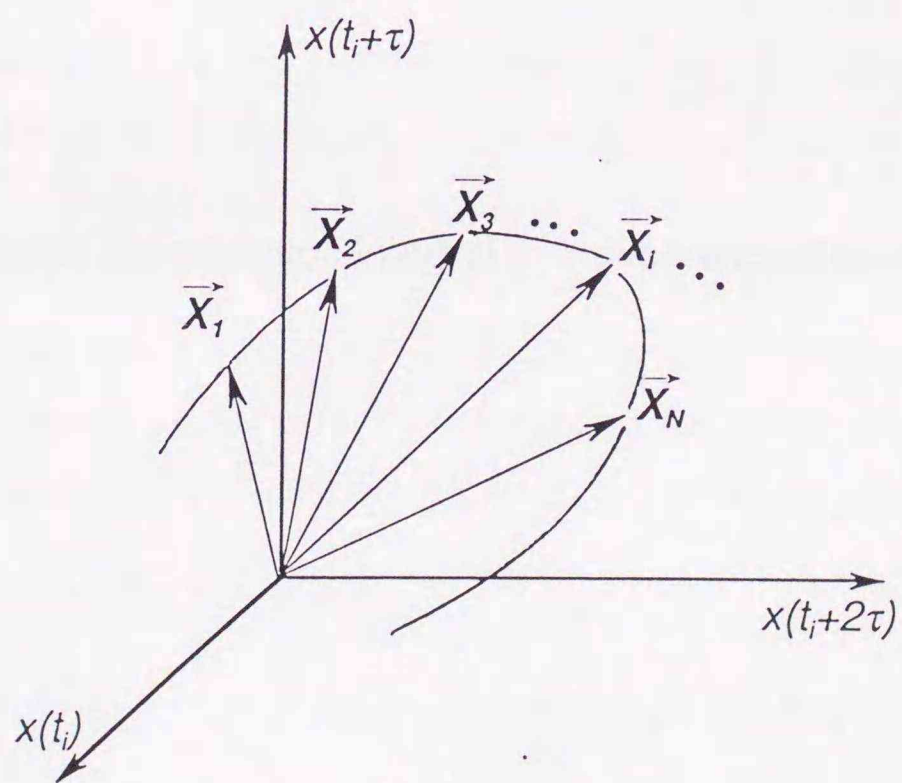


Fig.5.4 Schematic representation of reconstruction of 3-dimensional attractor from time-series data through embedding method

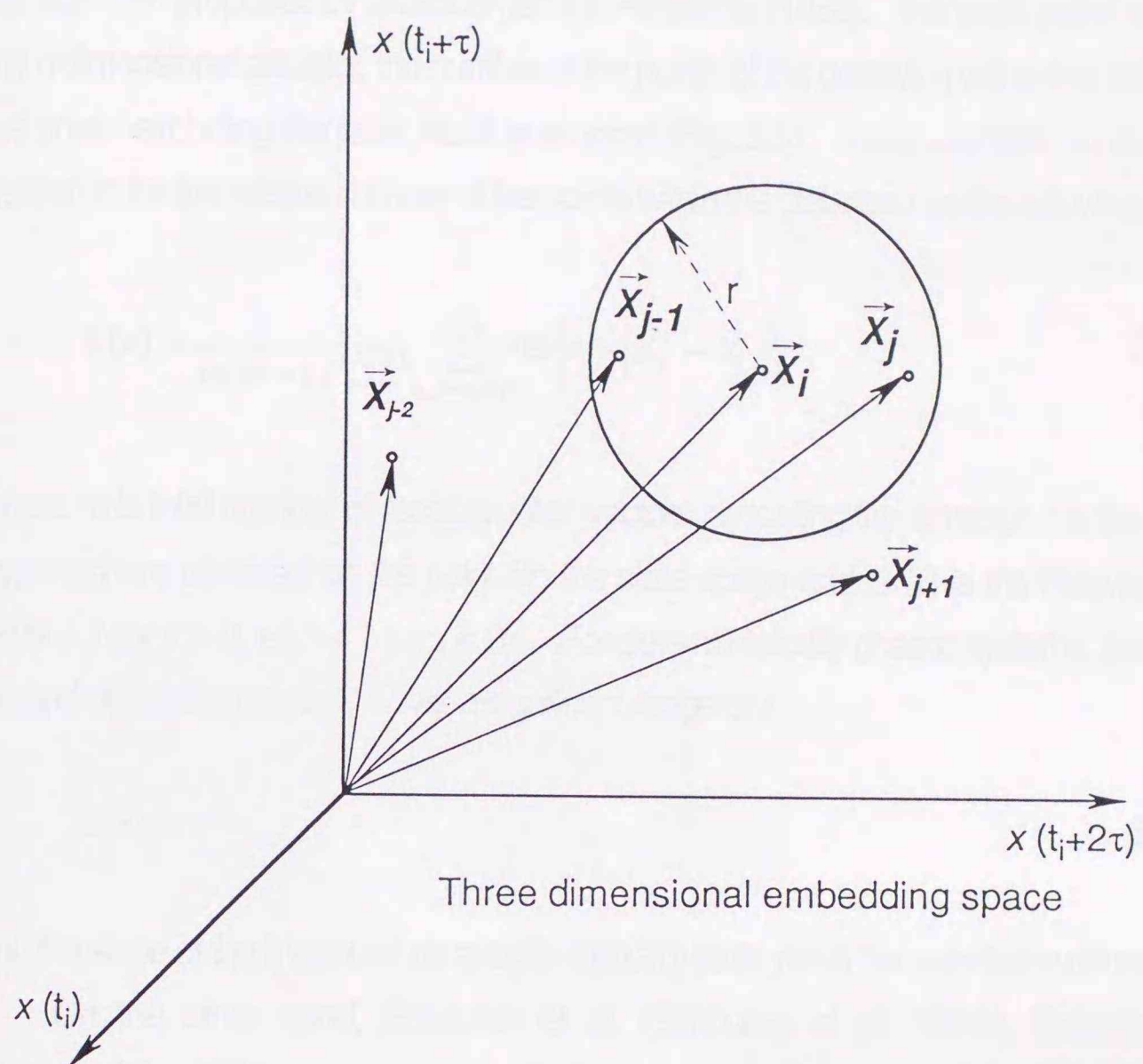


Fig.5.5 Concept of correlation integral in 3-dimensional state space.

In the present study, the correlation dimension, D_c , which can be interpreted as a fractal dimension of the orbit of the attractor in the d -dimensional state space, is determined through the algorithm proposed by Grassberger and Procaccia (1983). For each point i of the orbit of the d -dimensional attractor, the number of the points of the orbit lying within the distance r from the point i excluding the point i itself is counted (**Fig. 5.5**). Here, correlation integral, $C(r)$, is defined to be the relative number of the points within the distance r as the following equation:

$$C(r) = \frac{1}{m(m-1)} \lim_{m \rightarrow \infty} \sum_{i, j=1(i \neq j)}^m \Theta\{r - |\bar{X}_i - \bar{X}_j|\} \quad (5-2)$$

where m is total number of reconstructed vectors consisting the attractor, r is the radius of a hyper sphere centered on the point i in the state space and $\Theta(z)$ is the Heaviside function ($\Theta(z) = 1$ for $z > 0$, $\Theta(z) = 0$ for $z < 0$). For deterministically chaotic systems, the correlation integral obeys the power law over a significant range of r .

$$C(r) \propto r^{D_c} \quad (5-3)$$

and the slope of the linear part on a $\log(r)$ - $\log(C(r))$ plots yields the correlation dimension, D_c .

On the other hand, Schouten et al. (Schouten et al; 1994a, Schouten; 1994b) introduced the Kolmogorov entropy, K , of the reconstructed attractor for estimation of the entropy of the attainable states of the multiphase system based on the maximum-likelihood approach. In this algorithm, the entropy was calculated from the average number of steps requisite for a pair of vectors, which are initially within a specific maximum length, l_0 , to separate until the distance between the pair is larger than l_0 . For all the pairs of points on an attractor, the number of steps was calculated and then the average number of steps was estimated. In this way, the Kolmogorov entropy was defined as follows:

$$K = -\log_2 \left(1 - \frac{1}{\bar{b}} \right) \quad (5-4)$$

where $\bar{b}$ is the average number of steps for which the distance becomes larger than l_0 .

5.3 Results and Discussion

5.3.1 Variation of Local Bed Voidage and Heat Transfer

Figures 5.6 & 5.7 show typical time series output signals simultaneously obtained with the optical transmittance probe and the heat transfer probe for a sequence of circulating solid mass flux at a liquid velocity of $0.107 \text{ m}\cdot\text{s}^{-1}$. Without the circulation of solid particles

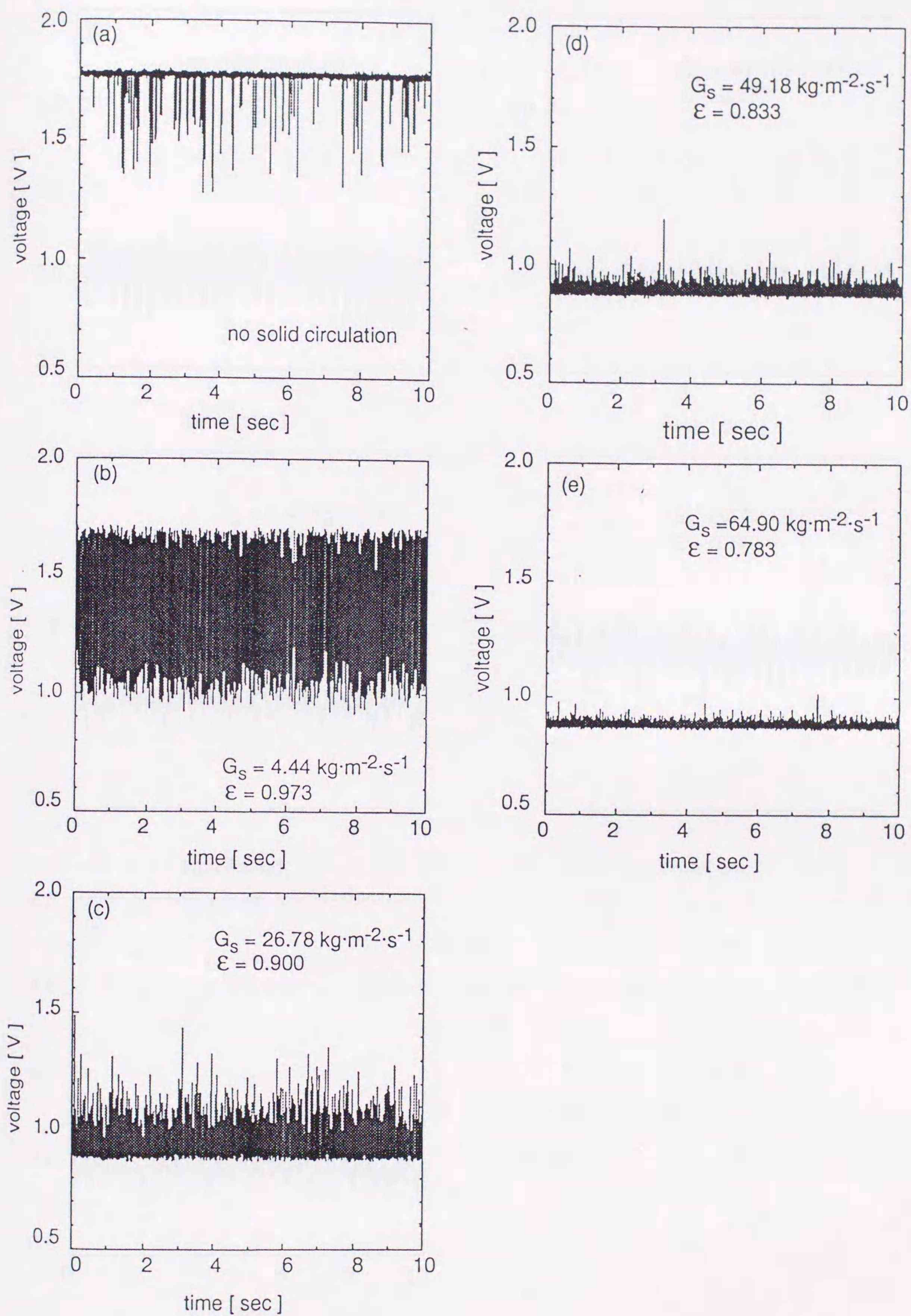


Fig.5.6 Time series of local bed voidage fluctuation measured by optical transmittance probe for various G_s at $u_{l0} = 0.107 \text{ m}\cdot\text{s}^{-1}$

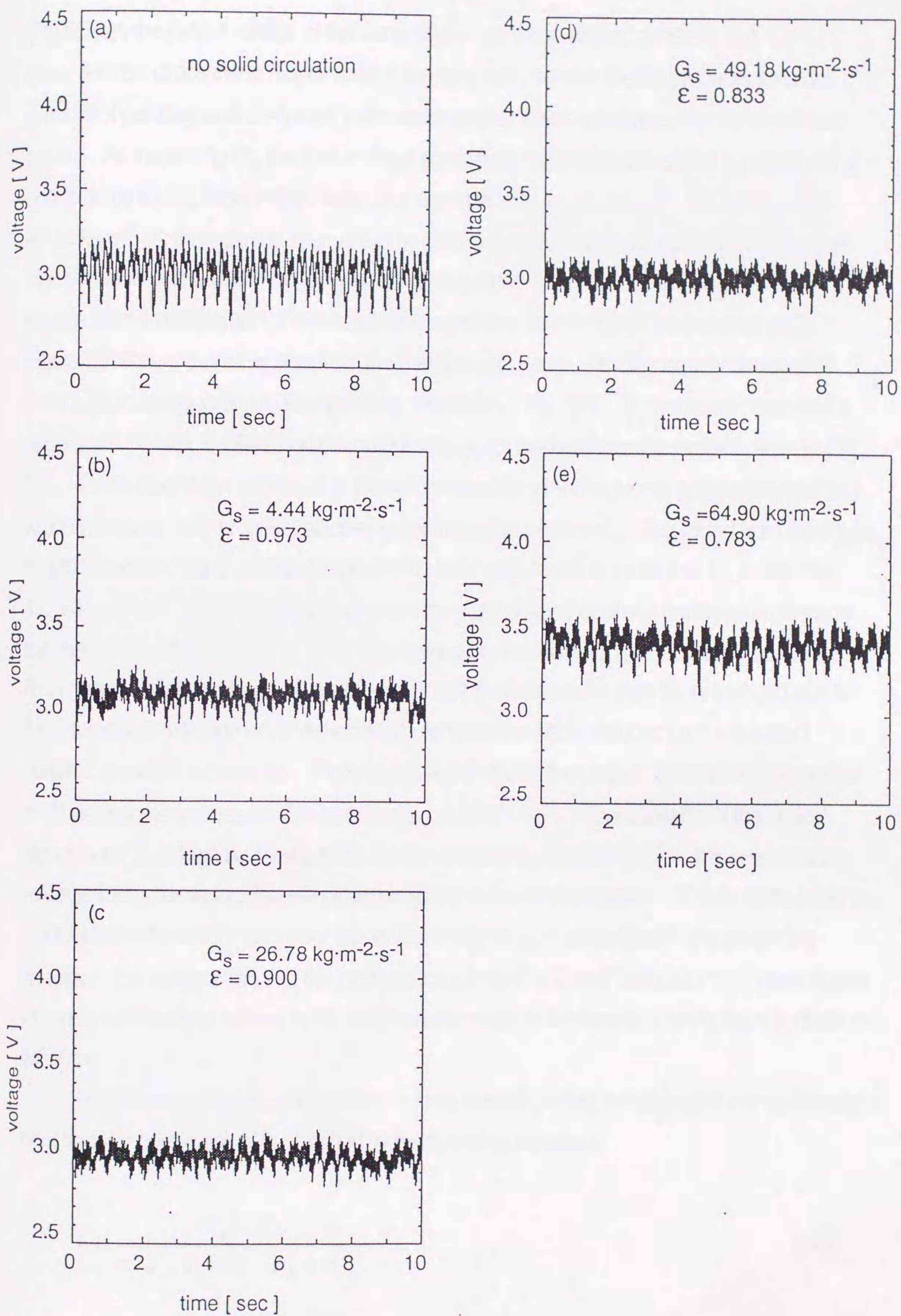


Fig.5.7 Time series of local heat transfer fluctuation measured by heat transfer probe for various G_s at $u_{l0} = 0.107 \text{ m} \cdot \text{s}^{-1}$

(Fig.5.6(a)), the output voltage of the transmittance probe is almost constant, e.g. 1.8 V. Here, sudden drops in the output voltage are frequently seen in the figure because small amount of particles was contained in the water and a particle intercepts the narrowed laser beam. As increasing G_s , the bed voidage correspondingly decreases and eventually time occupied by liquid phase in the measuring region is seen to decrease. At G_s larger than $49.18 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$, the beam is intercepted by solid phase during most of the measuring time. This variation of the optical signal with G_s clearly indicates that the solid phase gradually transits from the dispersed phase to continuous phase as a result of the increase in G_s . Figure 5.7 shows output signals from the heat transfer probe, simultaneously obtained at various G_s corresponding to the operating conditions in Fig. 5.6.. It should be noted that a large-scale periodic oscillation in the output voltage for the heat transfer probe is seen in Fig. 5.7. This is due to the oscillation in the solid-water slurry feed induced by the slurry pump which was seen only at liquid velocities lower than about $0.1 \text{ m}\cdot\text{s}^{-1}$. It is found from the figure that the time-averaged output voltage seems to be constant in the range of G_s lower than $49.18 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$. On the other hand, when the solid phase becomes continuous phase in the riser at $G_s = 64.90 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$, the output signal for the heat transfer obviously increases. From these variations, it can be considered that the heat transfer may be enhanced due to the decrease in voidage because local turbulence is inevitably induced by the fluidized particles around the probe tip. **Figures 5.8 & 5.9** shows the output signals from the optical and heat transfer probes for different G_s at $u_b = 0.320 \text{ m}\cdot\text{s}^{-1}$. The variation of the optical signals with G_s at high u_b seems to be similar to that at $u_b = 0.107 \text{ m}\cdot\text{s}^{-1}$. It is seen that the voidage fluctuates vigorously with time, exhibiting spike-like increases. This is attributable to the liquid streaks which have been observed at higher u_b as described in the preceding chapter. It is clearly seen that the increase in u_b from $0.107 \text{ m}\cdot\text{s}^{-1}$ to $0.320 \text{ m}\cdot\text{s}^{-1}$ gives rise to increase in the output voltage in the heat transfer probe in the bed and to vigorous fluctuations with time.

From these time series data from the heat transfer probe, variation of the time-averaged heat transfer coefficient, h , is obtained by the following equation:

$$h = \frac{V_p^2 \cdot R_t}{\pi \cdot d_w \cdot l_w \cdot \Delta T \cdot (R_b + R_t)^2} \quad (5-5)$$

where R_t is the wire resistance, R_b is the resistance of the bridge of the Wheatstone's bridge in series with R_t and ΔT is the difference in temperature between the wire and the bulk of the bed, T_o . R_t is directly calculated from the wire temperature by the following equation:

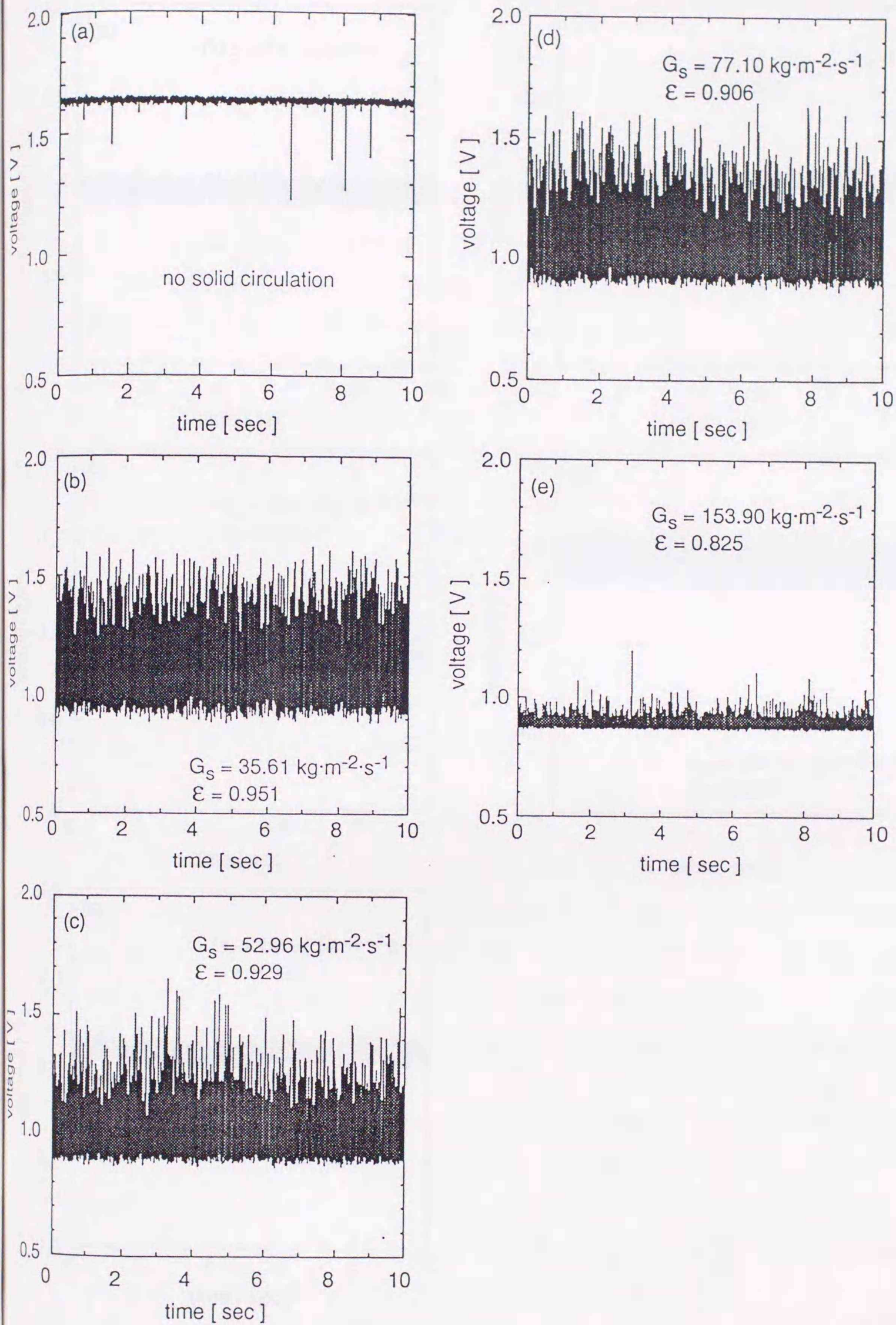


Fig.5.8 Time series of local bed voidage fluctuation measured by optical transmittance probe for various G_s at $u_{l0} = 0.320 \text{ m}\cdot\text{s}^{-1}$

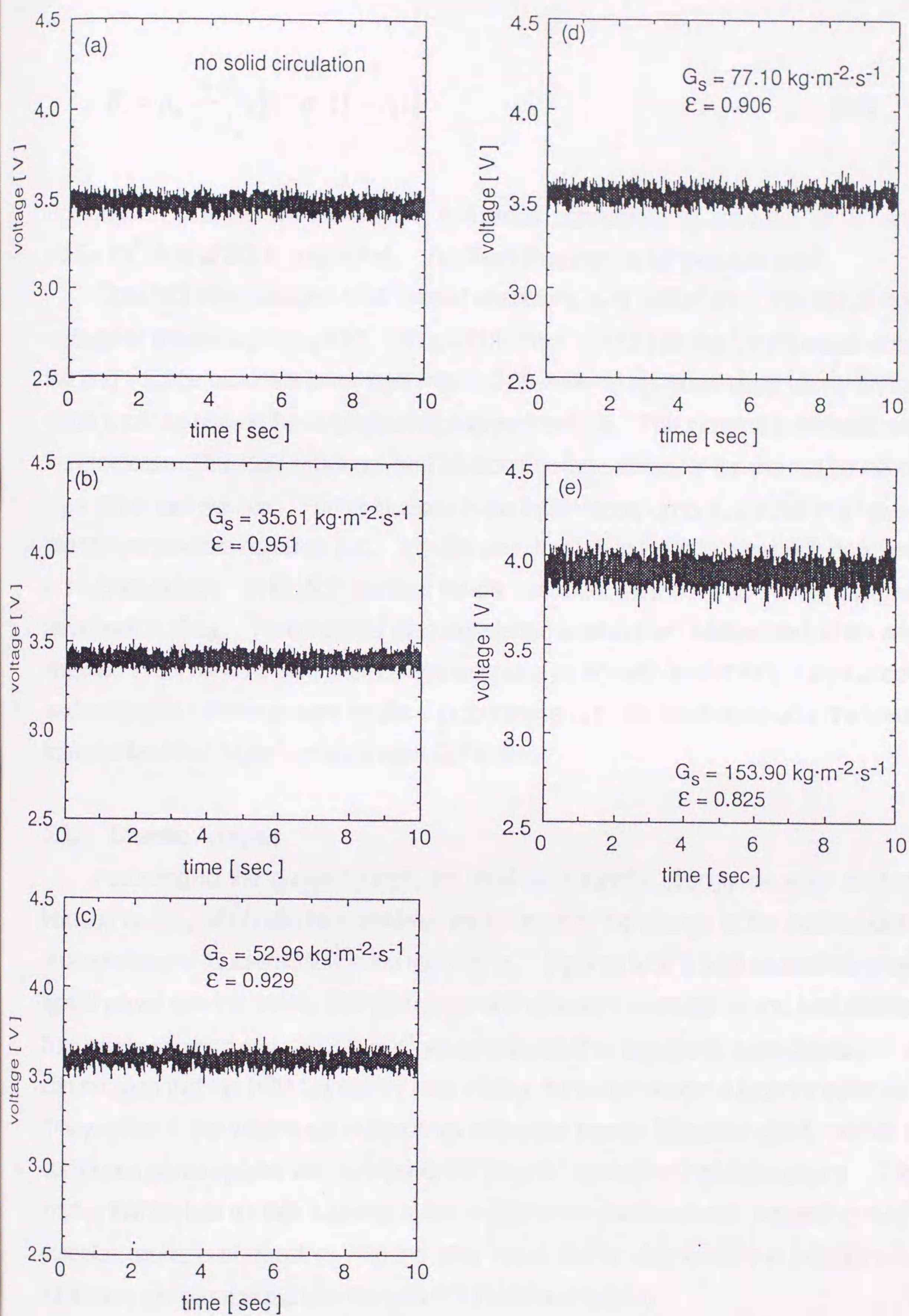


Fig.5.9 Time series of local heat transfer fluctuation measured by heat transfer probe for various G_s at $u_{l0} = 0.320 \text{ m} \cdot \text{s}^{-1}$.

$$R_t = \rho_o \frac{4 \cdot l_w}{\pi \cdot d_w^2} [1 + \sigma \cdot (T - T_o)] \quad (5-6)$$

For tungsten wire, thermal coefficient, σ , and electric conductivity, ρ_o , are $5.3 \times 10^{-3} \text{ K}^{-1}$ and $5.35 \times 10^{-8} \text{ W}\cdot\text{m}$ at 300 K, respectively. R_t is fixed once the wire temperature is set.

Obtained time-averaged heat transfer coefficient, h , is plotted as a function of bed voidage at different u_b in **Fig.5.10**. At $u_b = 0.107 \text{ m}\cdot\text{s}^{-1}$ and $0.139 \text{ m}\cdot\text{s}^{-1}$, h increases when the bed voidage becomes lower than around 0.82 while no significant dependency on the voidage can be seen in the range of voidage larger than 0.85. This increase is attributable to the promotion of the local turbulence and additional surface renewal in the thermal boundary layer at the wire surface. Further increase in the liquid velocity up to $u_b = 0.320 \text{ m}\cdot\text{s}^{-1}$ gives rise to a remarkable increases in h . It is also seen that h is found to increase with decrease in the bed voidage. In **Fig.5.11** the heat transfer coefficient obtained at different u_b is plotted as a function of u_{slp} . Here, heat transfer coefficient in a batch liquid fluidized bed, which was obtained through the hot-wire anemometry technique by Boerefijn et al (1996), is also plotted and compared with the present results. In comparison with the result obtained in the batch liquid fluidized bed, higher h is obtained in CLFB riser.

5.3.2 Chaotic Analysis

According to the above findings, the local heat transfer strongly depends on bed voidage and u_b , which can be considered as the result of the change of the local turbulent characteristics of liquid-solid flow in the riser with u_b . **Figures 5.12 & 5.13** respectively show typical power spectral density (PSD) functions of time series data of optical and heat transfer fluctuations obtained at $u_b = 0.320 \text{ m}\cdot\text{s}^{-1}$, which are plotted on logarithmic scale diagrams. It can be seen that the PSD function for local voidage fluctuation seems a broad spectra with many spikes in the wide range of frequency while heat transfer fluctuation clearly exhibits a continuous power spectra with an exponential decay in the range of higher frequency. This implies that the bed voidage fluctuates in the random or stochastic manner, suggesting much complex structure of attractor. On the other hand, chaotic characteristic in heat transfer fluctuation can be expected from the latter PSD function (Fig.5.13).

Figure 5.14 shows typical result of the logarithmic plots of the correlation integral, $C(r)$, with radius of hyper sphere, r , at various embedding dimensions for the output signal of the optical fiber probe at $u_b = 0.320 \text{ m}\cdot\text{s}^{-1}$ and $G_s = 35.61 \text{ kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}$. For the scaling regions

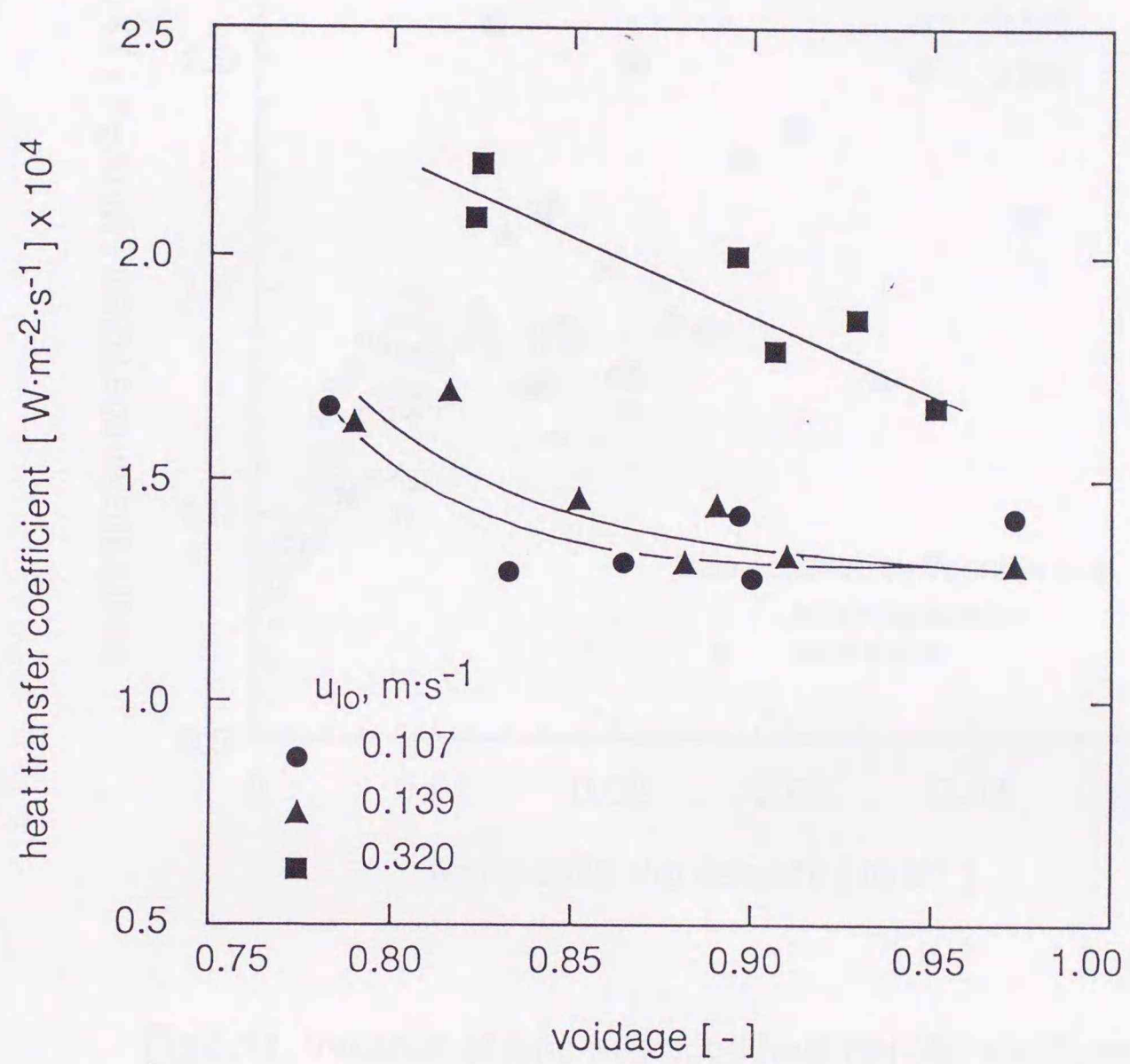


Fig. 5.10 Variation of time-averaged heat transfer coefficient with voidage at various liquid velocity

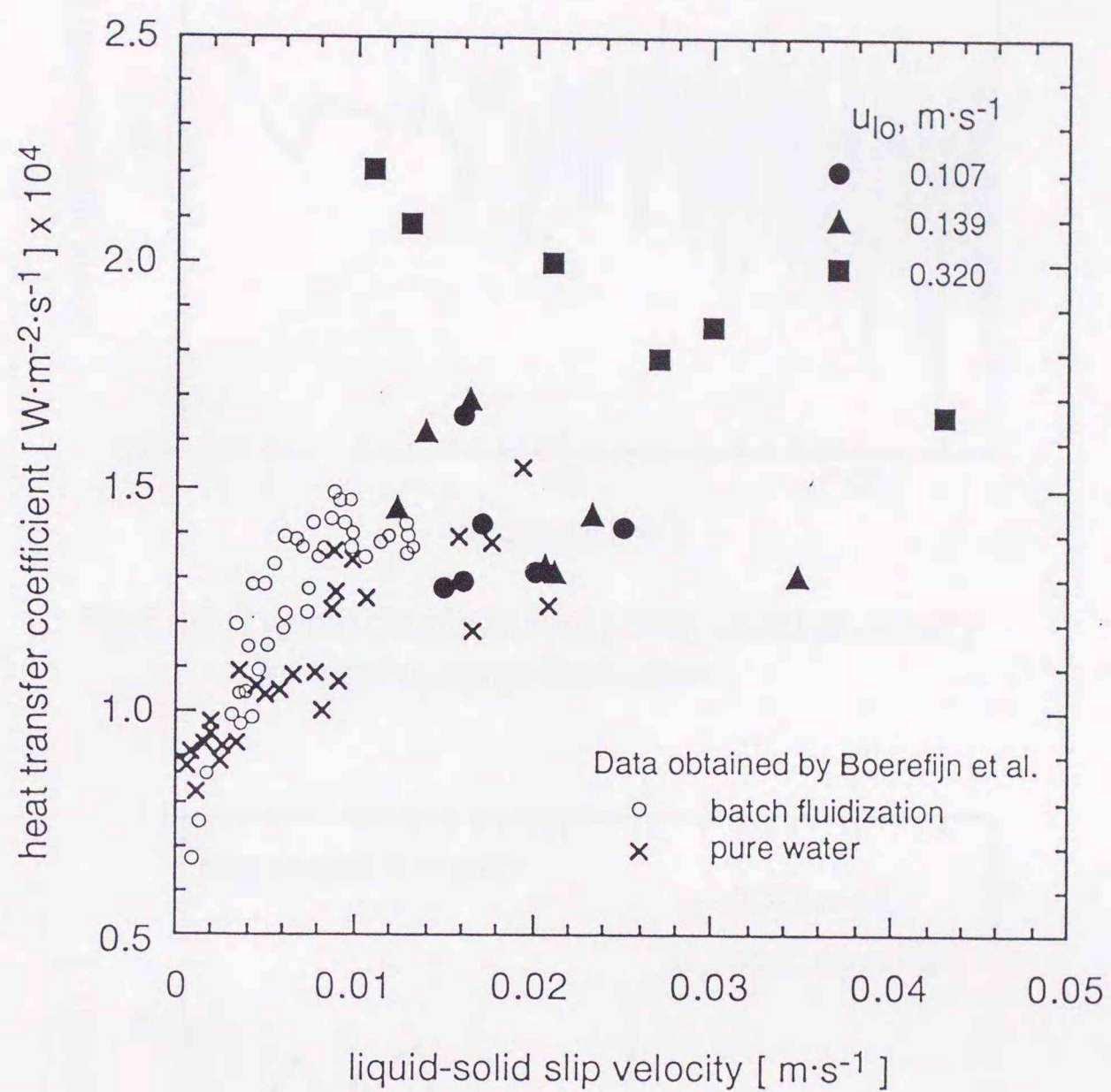


Fig.5.11 Variation of time-averaged heat transfer coefficient with liquid-solid slip velocity.

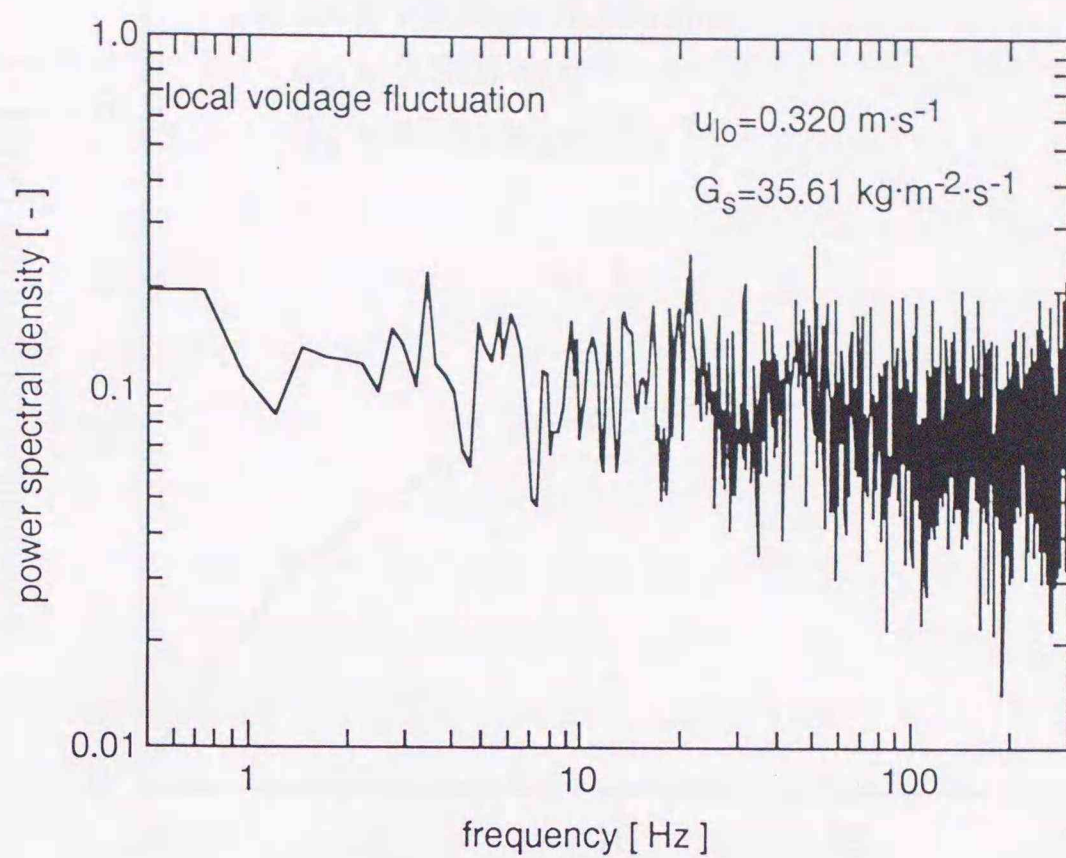


Fig.5.12 Typical distribution of power spectral density for local voidage fluctuation.

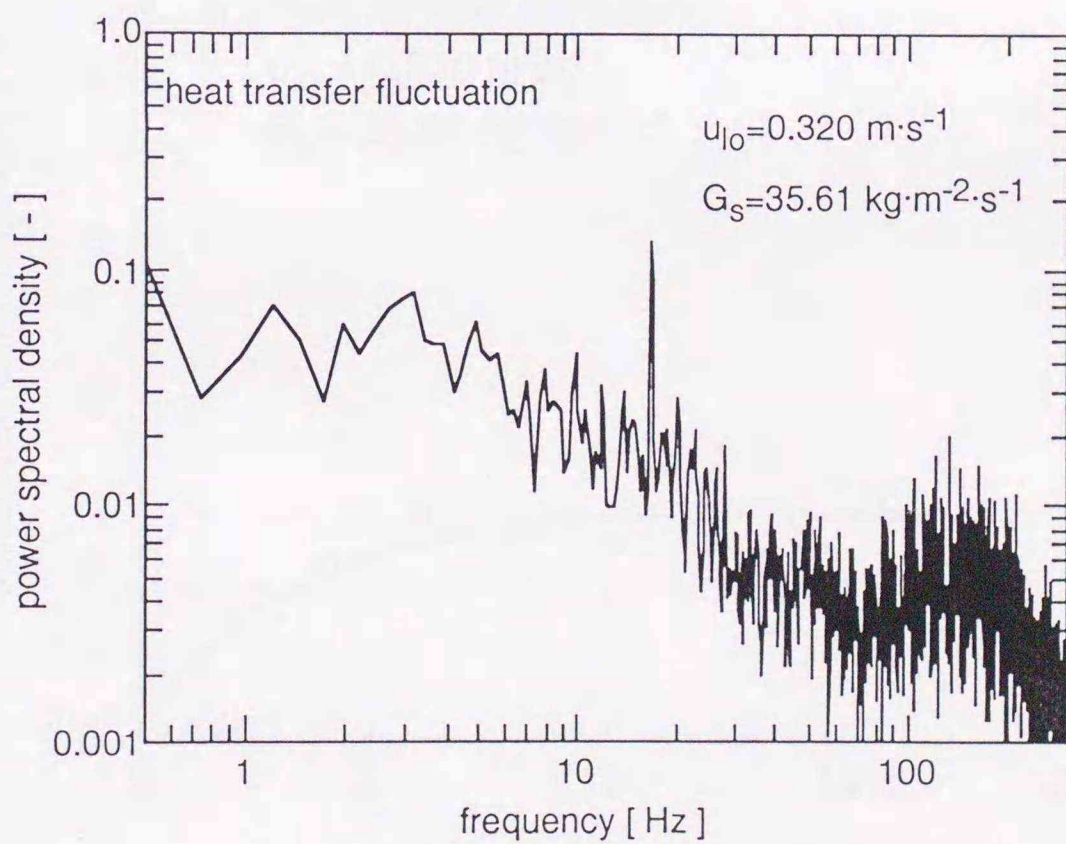


Fig.5.13 Typical distribution of power spectral density for heat transfer fluctuation.

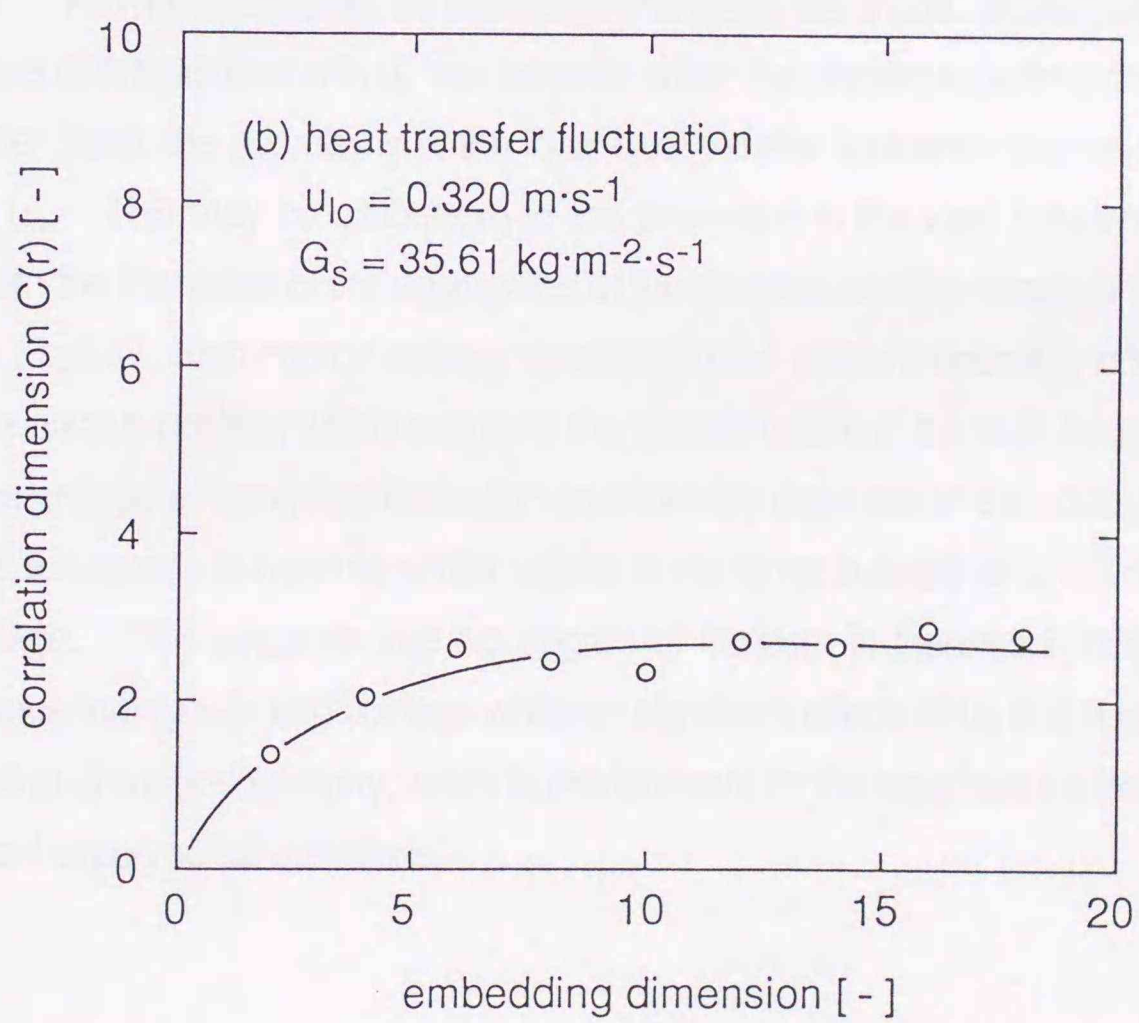
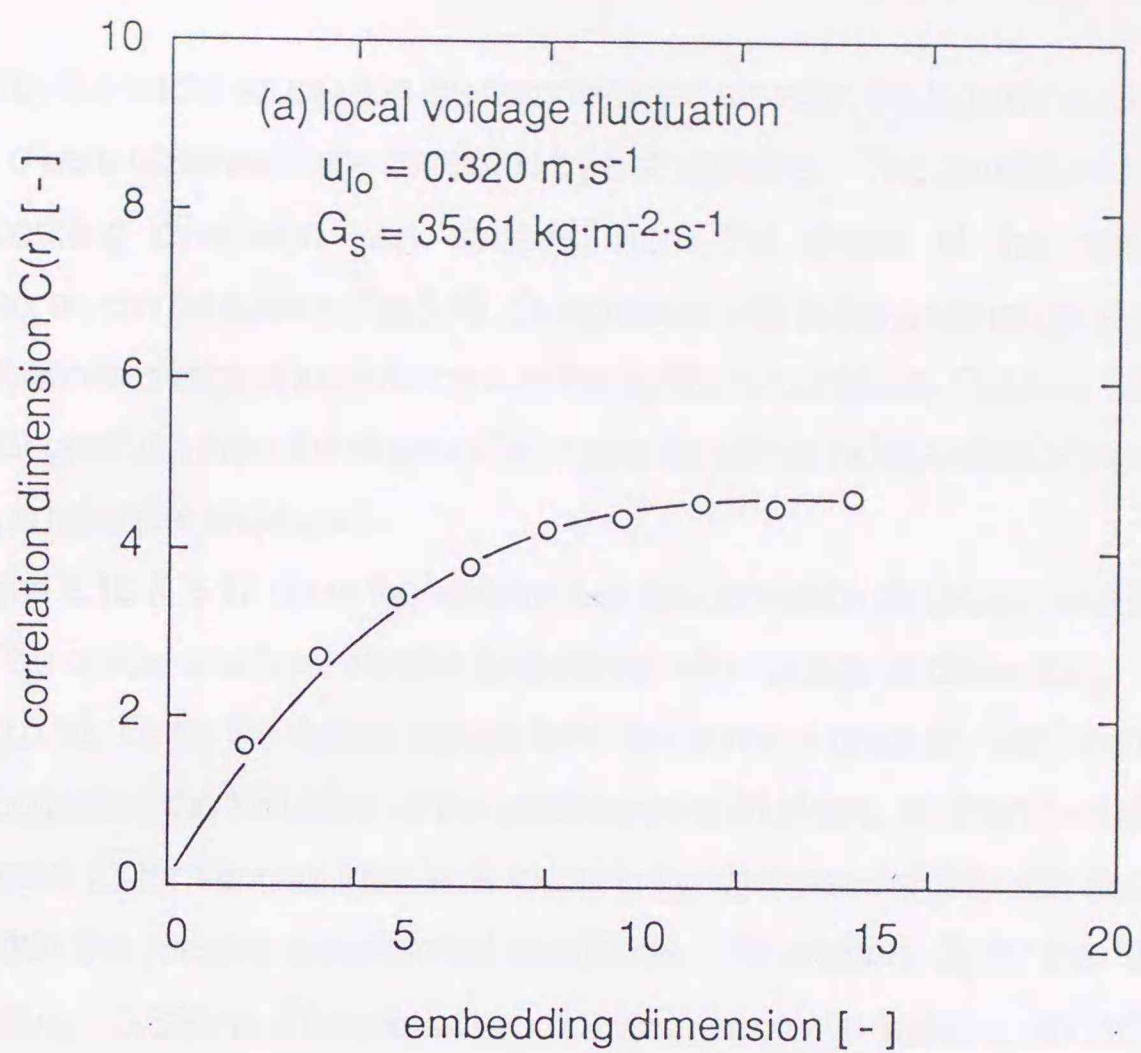


Fig. 5.14 Typical change of correlation dimension for local bed voidage and heat transfer fluctuations with embedding dimension.

dominated by the fractal structure in the reconstructed attractor, the logarithmic plots of $C(r)-r$ for a given d were observed to be correlated by a straight line. The correlation dimension for each embedding dimension was obtained from the slopes of the correlated line. Furthermore, as can be seen in **Fig.5.15**, D_c increases with in the small range of d and levels off when d becomes large enough to capture the system's complexity (Takens; 1981). Thus, D_c for the range of d , where the slope in $C(r)-r$ plots becomes independent of the embedding dimension, is hereafter employed.

Figures 5.16 & 5.17 show the variations of the correlation dimension and Kolmogorov entropy of the optical and heat transfer fluctuations with voidage at different u_b . As can be seen in Fig.5.16, D_c for the optical signals tend to decrease gradually with decrease in the voidage, suggesting the formation of the continuous solid phase, as described above. On the other hand, D_c for the heat transfer is found to be decreased slightly with increase in the voidage within the present experimental conditions. In addition, D_c for the heat transfer fluctuation at $u_b = 0.320 \text{ m}\cdot\text{s}^{-1}$ becomes large in comparison with that at u_b of $0.107 \text{ m}\cdot\text{s}^{-1}$ and $0.139 \text{ m}\cdot\text{s}^{-1}$. From these results, no significant change in the chaotic characteristics of local flow structure of the particles with u_b can be seen within the present experimental conditions. On the other hand, the complexity of the local heat transfer fluctuation becomes higher as increasing u_b . This may be attributable to the promotion of the local turbulence which is induced from the formation of the aggregates of the particles and the streak of the fluidizing liquid. In Fig.5.17, Kolmogorov entropy obtained in the present operating conditions are found to be always positive, which suggests the chaotic nature in the both fluctuations. In addition, the entropy of the optical fluctuation considerably depends on the voidage while that for the heat flux seems to become similar values in the range between 4 to 5.5, at any value of the voidage. This suggests that the degree of freedom in the motion of the fluidized particles varies mainly with bed voidage while no significant effects of u_b and voidage on the chaotic motion of water-solid slurry, which is predominant for the local heat transfer, are seen in the present experimental conditions.

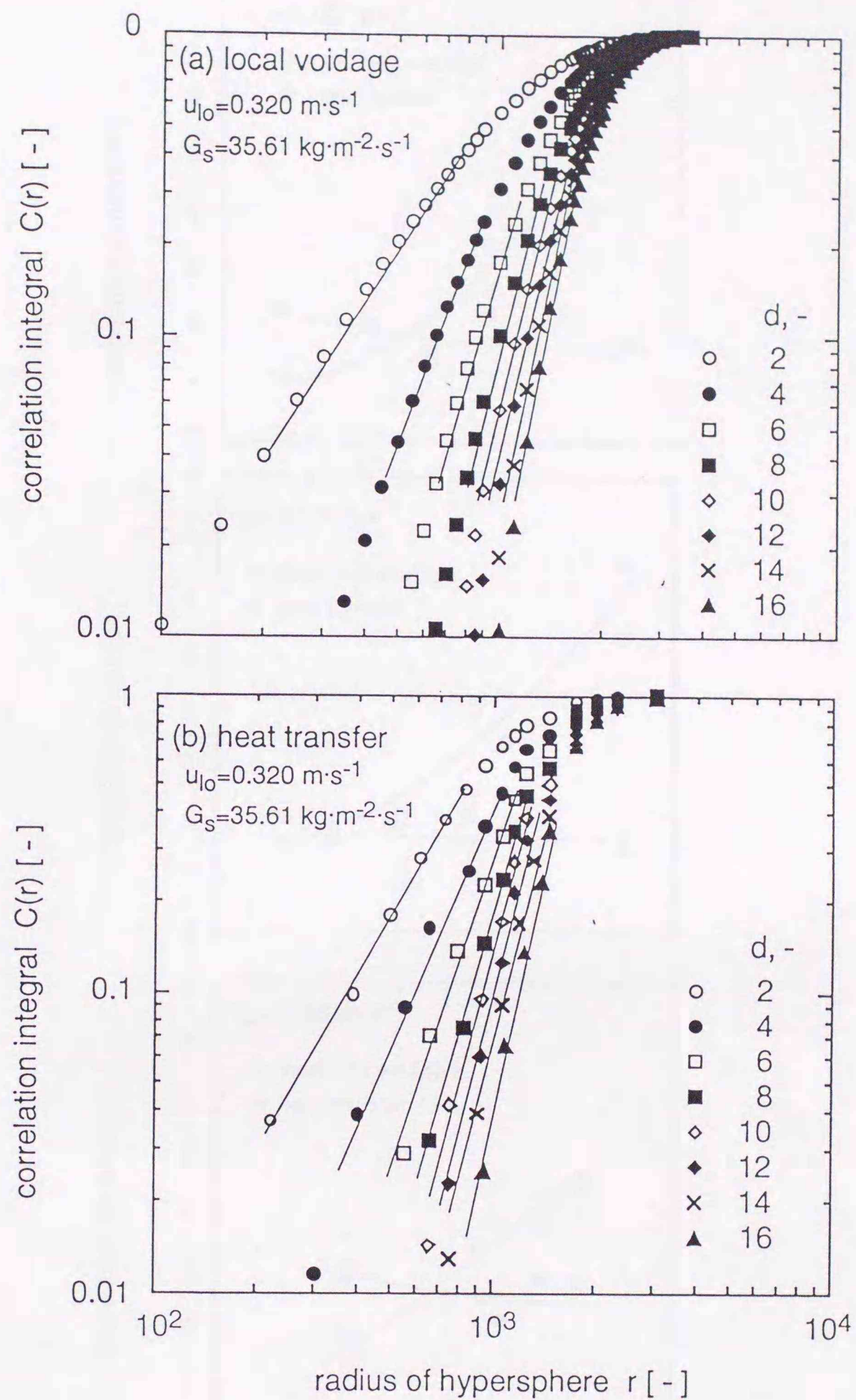


Fig.5.15 Logarithmic plots of $C(r)$ against radius of hyper sphere, r , for various embedding dimensions; (a) for local voidage fluctuation and (b) for heat transfer fluctuation

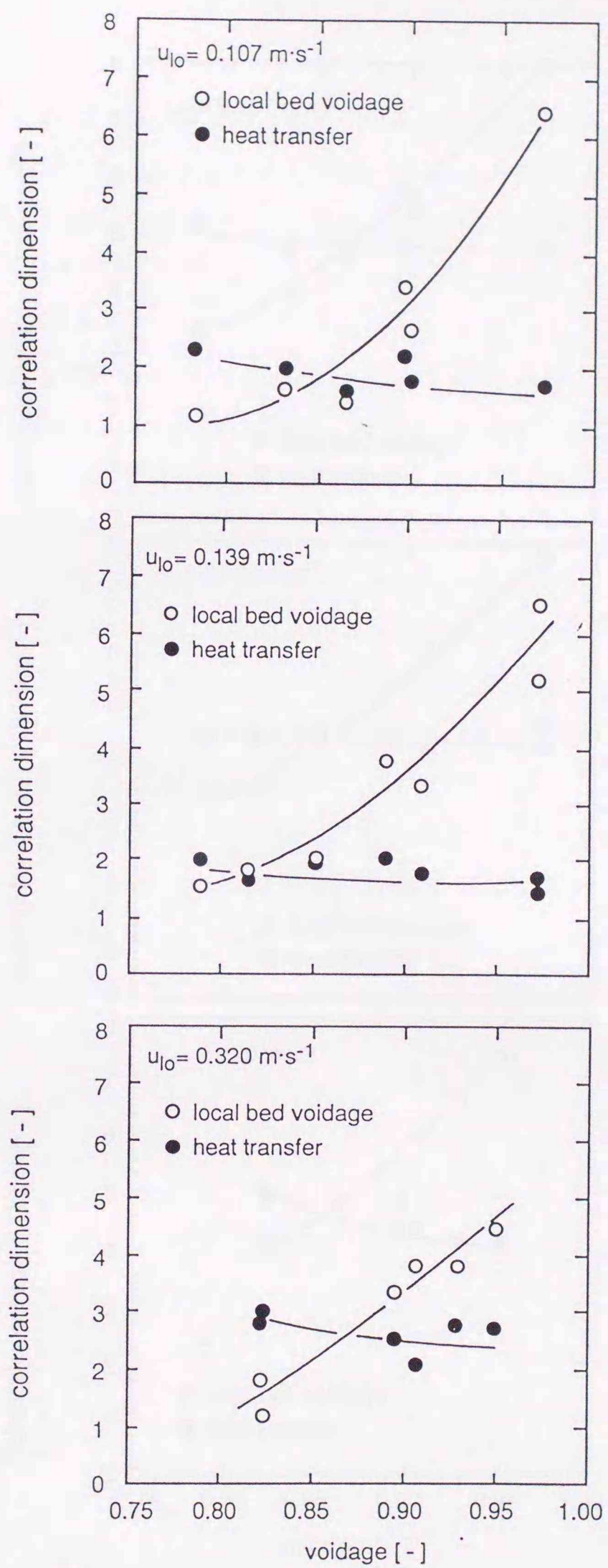


Fig.5.16 Comparison of variation in correlation dimension for local voidage and heat transfer fluctuations as a function of voidage, obtained at different liquid velocities

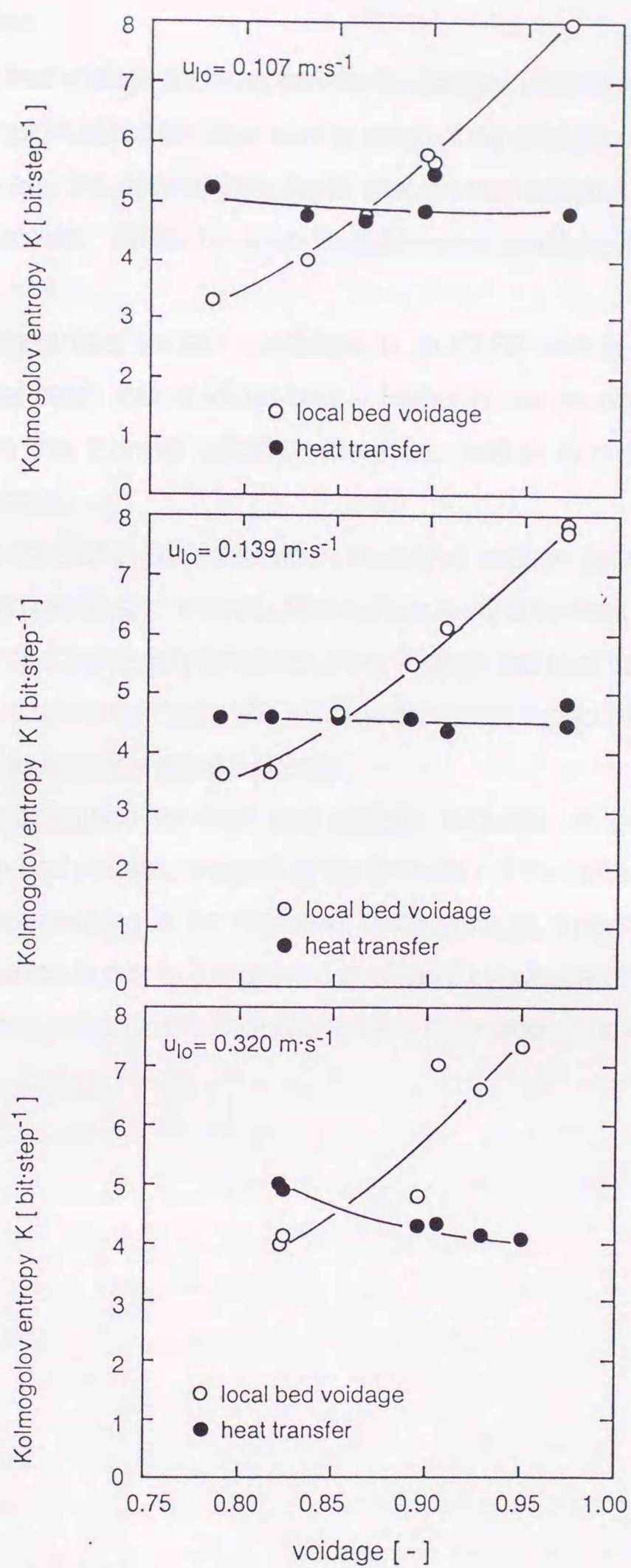


Fig.5.17 Comparison of variation in Kolmogorov entropy for local voidage and heat transfer fluctuations as a function of voidage, obtained at different liquid velocities.

5.4 Conclusion

The local bed voidage and heat transfer fluctuations were simultaneously measured in the circulating liquid-fluidized bed riser at wide range of superficial liquid velocity and circulating solid mass flux and the deterministic chaos analysis was adopted to characterize their chaotic characteristics. Within the present experimental conditions, followings are clarified:

- 1) Time-averaged heat transfer coefficient, h , in CLFB riser is higher than that in the conventional batch liquid-fluidized bed. Higher h can be obtained at u_{b0} appreciably higher than the terminal velocity, where the particle bed exhibits the aggregative fluidization state.
- 2) Local heat transfer is promoted when local bed voidage decreases. This is due to turbulent convection and renewal of the surface around the heat transfer probe.
- 3) The power spectral density functions of the voidage and heat transfer fluctuations exhibit a continuous power spectrum with exponential decay, suggesting that the hydrodynamic behavior of liquid-solid system is chaotic.
- 4) Correlation dimension for local bed voidage fluctuation is gradually decreases with decrease in bed voidage, suggesting the formation of the continuous solid phase in the riser. Corresponding to the decrease in the voidage, time-averaged heat transfer is increased which is due to the promotion of local turbulence of the liquid-solid flow and renewal of the surface in the thermal boundary layer around the heat transfer probe.

Literature Cited

- Bleek, C.M. van den, J.C. Schouten, "Deterministic Chaos: a New Tool in Fluidized Bed Design and Operation", Chem. Eng. J., 53, 75-87 (1993)
- Bouillard, J. .X. and A. L. Miller, "Experimental Investigations of Hydrodynamic Attractors in Circulating Fluidized Beds", Powder Tech., 79, 211-215 (1994)
- Broerrefijn, R., M. Poletto and P. Salatino, "A Study of the Dynamic Behaviour of Fluidized Beds Based on a Hot Wire Anemometry Technique", in "Fluidization VIII", Engineering Foundation, pp.21-30 (1996)
- Daw, C.S, W.F. Lawkins, D.J. Downing, N. E. Clapp, Jr, "Chaotic Characteristics of a Complex Gas-Solids Flow", Phys. Rev A, 41, 1179-1181 (1990)
- Grassberger, P. and I. Procaccia, "Measuring the strangeness of strange attractors", Physica, 9D, 189-208 (1983)
- Fan, L -S., "Gas-Liquid-Solid Fluidization Engineering", Butterworth, U.S.A(1989)
- Huilin, L, D. Gidaspow, J.X. Bouillard, "Dimension Measurements of Hydrodynamic Attractors in Circulating Fluidized Beds", AIChE Symp. Ser., 301, 103-111(1995)
- Hilbom, R. C., "Chaos and Nonlinear Dynamics", Oxford University Press, New York, USA(1994)
- Kato, Y., K. Uchida, T. Kago and S. Morooka, "Liquid Holdup and Heat Transfer Coefficient between Bed and Wall in Liquid-Solid and Gas-Liquid-Solid Fluidized Beds", Powder Tech., 28, 173-179(1981)
- Khan, A. R., A. K. A. Juma and J. F. Richardson, "Heat Transfer from a Plane Surface to Liquids and to Liquid-Solid Fluidized Beds", Chem. Eng. Sci., 38,2053-2066(1983)
- Kikuchi, R., A. Tsutsumi and K. Yoshida, "Fractal Aspect of Hydrodynamics in a Three-Phase Fluidized Bed", Chem. Eng. Sci., 51(11), 2865-2870 (1996)
- Richardson, J. F. and A. E. Mitson, "Sedimentation and Fluidization Part II - Heat Transfer from a Tube Wall to a Liquid-Fluidised System", Trans. Inst. Chem. Engrs., 36, 270(1958)
- Richardson, J. F. and J. W. Smith, "Heat Transfer to Liquid Fluidized Swystems and to Suspensions of Coarse Particles in Vertical Transport", Trans. Inst. Chem. Engrs., 40, 13(1962)
- Richardson, J. F., M. N. Romani and K. J. Shakiri, "Heat Transfer from Immersed Surfaces in Liquid Fluidised Beds" Chem. Eng. Sci., 31, 619(1976)
- Schouten, J.C., F. Takens and C.M. van den Bleek, "Maximum-Likelihood Estimation of the Entropy of an Attractor", Phys. Rev. E, 49(1), 126-129 (1994a)

Schouten, J.C., F. Takens and C.M. van den Bleek, "Estimation of the Dimension of a Noisy Attractor", Phys. Rev. E, 50(3), 1851-1861 (1994b)

Takens, F., "Detecting strange attractors in turbulence", in Springer Lecture Notes in Mathematics (eds. D. A. Rand and L.-S. Young), vol. 898, p.366-381(1981)

Chapter 6

GENERAL CONCLUSION

This thesis is a summary of series of experimental work which has been carried out in an effort to make clear the flow phenomena of solid particles fluidized by gas and liquid at various velocities higher than the terminal velocity of the single particle, u_t . Systematic measurements were conducted to elucidate flow patterns of solid particles in the risers of circulating gas- and liquid-fluidized bed (CGFB and CLFB, respectively) systems.

For CGFB system, attention has consistently been focused to the effects of the system variables such as the height of riser and the initial solid inventory in the downcomer on steady-state and unsteady-state axial bed density (ABD) distributions in the riser. The effects have been examined measuring distributions of static pressures along the height of risers of 5.5 m and 10 m high and 0.1 m i.d. and of downcomer in a CGFB system. The measurements were conducted under the conditions where the circulating solid mass flux, G_s , was controlled independently of pressure balance between the riser and the downcomer bottoms as well as of the superficial gas velocity in the riser, u_{go} . Within the present experimental conditions, the followings were evidenced;

- 1) When dilute-phase transport or dense-phase transport state appears in the riser, ABD distributions at a fixed set of values of u_{go} and G_s are generally independent of the initial solid inventory, H_{sd} , in the downcomer.
- 2) When G_s takes a particular value at a given gas velocity, S-shaped ABD distribution appears, which consists of a dense region in the lower part and a dilute region in the upper part of the riser. At such a condition, the height of the dense-region depends appreciably on H_{sd} . This can be explained by an analogy to existing models for solid flow in the freeboard of the bubbling fluidized beds.
- 3) The dense-region height increases with the riser height even under conditions where all the other operating variables are the same. Thus, an apparent overlap of the ABD distribution curves for risers of the different heights is obtainable when the distribution

curves are drawn by leveling the riser exit.

- 4) Transitional variations are observed in P_{ro} and P_{do} . The variation reflects not only distribution of solid particles initially loaded in the downcomer to the riser and the downcomer but also slow development of the static pressure distribution in the downcomer.
- 5) The time required to attain a steady distribution is longer for P_{do} than for P_{ro} due to the slow gas flow from the riser to the downcomer and the induced slow gas flow in the upper part of the bed in the downcomer by the solid particles raining down from the cyclones onto the bed.
- 6) No appreciable effect of the variation in P_{do} is found on steady-state ABD distributions. Only when G_s takes a particular value at which S-shaped ABD distributions appear in the riser, P_{ro} and correspondingly the ABD distribution vary with P_{do} . At such a condition, the steady-state ABD distribution depends on H_{sd} .

A further analysis of fundamental high-velocity flow patterns of solid particles were conducted employing a CLFB system with a 4.0 m high and 0.049 m i.d. riser to simplify the above complicated and unstable flow characteristics in CGFB system. Measurements were carried out at various G_s and the superficial liquid velocity, u_b , for ABD distribution in the riser as well as fluctuations in the local bed voidage and heat transfer rate both of which strongly reflect the particle flow patterns in the riser and the followings were found:

- 1) The S-shaped distributions, which are frequently observed in CGFB risers, are never produced in the CLFB riser, i.e. ABD distributions are always uniform along the CLFB riser. Existing flow models obtained for particulate or packed beds are inapplicable to describe the relationship between the liquid-solid slip velocity and the bed voidage in CLFB.
- 2) Even a bed which exhibits particulate fluidization state at $u_b < u_t$ tends to form aggregates in CLFB riser at $u_b > u_t$, where the fluidizing liquid eventually passes through the bed in a form of streak.
- 3) Spatial distribution of voidage in CLFB riser exhibits fractal structure. Lower fractal dimensions are obtained at u_b much higher than u_t , which is attributed to formation of larger scale aggregates.
- 4) Time-averaged heat transfer coefficient, h , is appreciably higher than those obtained in batch liquid fluidized beds. A further increase of h takes place as u_b is increased over

u_t , where the bed exhibits the aggregative fluidization state.

- 5) The power spectral density functions of the heat transfer fluctuations exhibit a continuous power spectrum with exponential decay, suggesting that the state of liquid-fluidization at $u_b > u_t$ is chaotic.
- 6) Heat transfer coefficient increases with increasing u_b and further with decreasing local voidage at a fixed u_b . Higher correlation dimensions for the local heat transfer fluctuation are obtained probably due to turbulent convection and renewal of the surface around the heat transfer probe.

From the analysis, it becomes clear that the fluctuation in the heat transfer inherently contains chaotic nature. This suggests that attainable flow state in the riser is complex but deterministic. The next step in this research field will be synchronization of the predicted fluctuation by using a neural network with the actual fluctuation among different operating conditions or riser scales based on the chaotic statistics.

Nomenclature

A_r	cross-sectional area of riser	$[m^2]$
$\bar{b}$	the average number of time steps necessary for an initial interpoint distance to grow beyond a specific length	$[-]$
$C(r)$	correlation integral	$[-]$
D_c	correlation dimension	$[-]$
D_f	fractal dimension	$[-]$
d	embedding dimension	$[-]$
d_p	mean particle diameter	$[m]$
F	solid mass flux in riser	$[kg \cdot m^{-2} \cdot s^{-1}]$
G_s	circulating solid mass flux	$[kg \cdot m^{-2} \cdot s^{-1}]$
g	gravitational acceleration	$[m \cdot s^{-2}]$
H_d	height in downcomer	$[m]$
H_{sd}	static bed height initially loaded in downcomer	$[m]$
h	heat transfer coefficient	$[W \cdot m^{-2} \cdot s^{-1}]$
K	Kolmogorov entropy	$[bits/step]$
l_c	a specific maximum distance	$[-]$
m	total number of points on reconstructed attractor used for calculation of a correlation integral	$[-]$
P	static pressure	$[Pa]$
P_{db}	static pressure at downcomer bottom	$[Pa]$
P_{ro}	static pressure at riser bottom	$[Pa]$
ΔP_{total}	pressure drop through riser	$[Pa]$
ΔP_{rd}	pressure difference between riser and downcomer bottoms	$[Pa]$
r	radius of a given hypersphere in the phase space	$[-]$
s	mesh scale	$[-]$
t	time elapsed after the onset of solid circulation	$[s]$
t_s	time required to attain steady-state for CFB system	$[s]$
t_{smp}	time for water-solid slurry sampling	$[s]$
U_a	rate of aeration to downcomer bottom	$[m^3 \cdot s^{-1}]$
U_{go}	superficial gas velocity	$[m \cdot s^{-1}]$
U_{lo}	superficial liquid velocity	$[m \cdot s^{-1}]$
U_p	particle velocity	$[m \cdot s^{-1}]$

U_{slip}	liquid-solid slip velocity	$[m \cdot s^{-1}]$
U_t	terminal velocity of a single particle	$[m \cdot s^{-1}]$
V_s	volume of sampled water-solid slurry	$[m^3]$
W_s	weight of sampled water-solid slurry	$[kg]$
$\vec{X}_i$	phase space vectors reconstructed from time-series	$[-]$
$x(t)$	time series	$[-]$
z	height from distributor in riser	$[m]$
ε	mean bed voidage	$[-]$
ε_s	voidage of sampled water-solid slurry	$[-]$
$\Theta(z)$	Heaviside function	$[-]$
κ	tortuosity of fluid	$[-]$
μ	liquid viscosity	$[Pa \cdot s]$
ρ	liquid density	$[kg \cdot m^{-3}]$
ρ_p	particle density	$[kg \cdot m^{-3}]$
ρ_s	solid concentration of water-solid slurry	$[kg \cdot m^{-3}]$
τ	time interval	$[s]$

Acknowledgments

This research is conducted under the direction of Dr. Tadatoshi CHIBA, Professor of Center for Advanced Research of Energy Technology (CARET), with continuing valuable comments, criticisms and kindhearted supervision, all of which led to the completion of the research. The author wishes to express the highest gratitude to his direction.

The author would like to express my thanks to Dr. Toshimasa HIRAMA, Hokkaido National Institute of Research and Industry (HNIRI), who always gives the instructive directions regarding not only to the experimental strategy but also to the attitude for the research activity. The author surely believe that this research is unable to complete without his direction.

He express to thank Professor Hideshi HATTORI, Associate Professor Jun-ichiro HAYASHI, Dr Haruo KUMAGAI, academic stuff of CARET, for valuable discussion and encouragement.

Dr. Kunio SHINOHARA, Dr. Kuniomi ASAKURA, Hokkaido University and Dr. Kunio YOSHIDA and Dr. Atsushi TSUTSUMI, University of TOKYO, for their valuable suggestions upon the research.

Dr. Ichiro NARUSE, Toyohashi University of Technology (TUT), Dr. Tamio IDA, Kumano National College of Technology, Dr. Norio IWASHITA, Osaka National Institute of Research and Industry, Dr. Yoshinori KAWASAKI and Dr. Hideaki MIYAGAWA, Ariake National College of Technology, for their thoughtful encouragement.

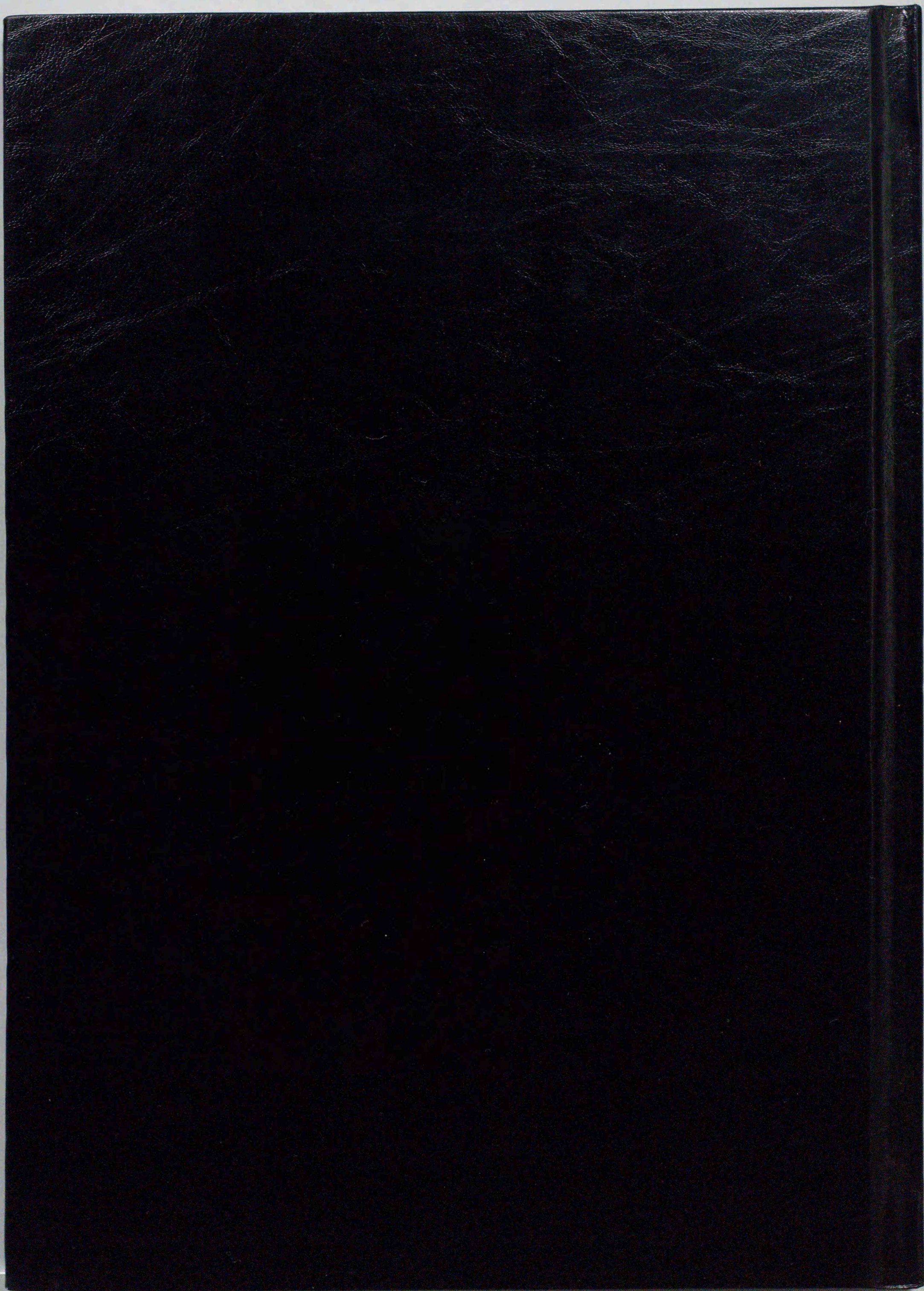
Thanks should be extended to Dr. Hiromi TAKEUCHI, Mr. Hideo HOSODA, Dr. Shohei TAKEDA, Dr. Kunihiro KITANO, stuff of HNIRI, for their advice and help.

His thanks due to Dr. Koichi MATSUOKA, Tohoku University, Reji NODA, Toyohashi University of Technology, Dr. Takuo YASUDA, Dr. Taihei SHIMADA and Dr. Hassan H. KATALAMBRA, Hokkaido University, Dr. Yasuhito NAKATAKE and Dr. Shoji ESAKI, Kurume National College of Technology.

Acknowledgments also due to the colleague; Dr. Ji Hong-shen, Ms. Yukie TANAKA, Dr. Masateru NISHIOKA, Messrs. Takehiko KUMAGAI, Naohiro AZUMA, Kenji KOMATSU, Daisuke Demma, Takeshi SATO, Hiroaki OHARA and other member of Yoshida lab. and CARET.

The author wishes to express his the gratitude especially to his parents for their understanding and thoughtful encouragement, giving this special and valuable occasion and support over the years

Finally, author wishes to express his eternal thanks to Professor Kazutomo OHTAKE, who initiated the department of the ecology engineering in TOYOHASHI University of Technology, for his the deepest understanding of the 3-years master-course study in Toyohashi University and the 3-years doctoral study in Hokkaido University.

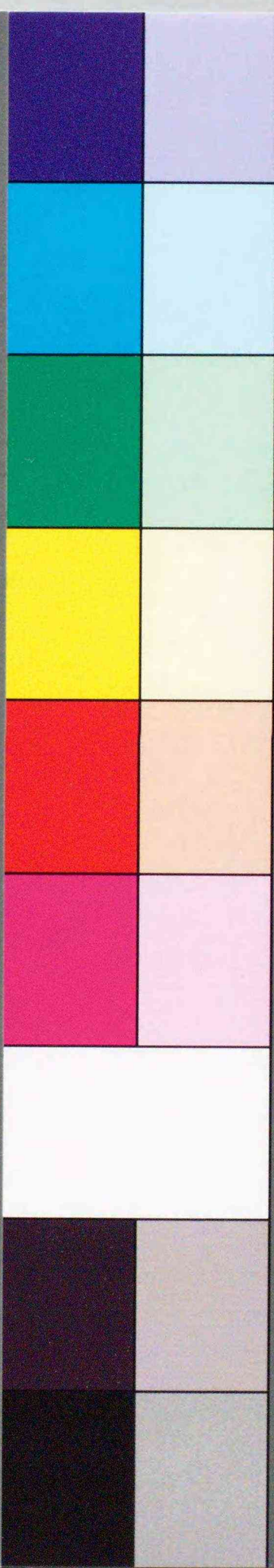


Inches 1 2 3 4 5 6 7 8
cm 1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19

Kodak Color Control Patches

© Kodak, 2007 TM: Kodak

Blue Cyan Green Yellow Red Magenta White 3/Color Black



Kodak Gray Scale



© Kodak, 2007 TM: Kodak

A 1 2 3 4 5 6 M 8 9 10 11 12 13 14 15 B 17 18 19

