I. INTRODUCTION

Fischer–Tropsch (F–T) synthesis represents an important route for indirect coal liquefaction. During World War II, Germany utilized F–T synthesis to produce motor fuels. Currently, commercial size units are in operation at SASOL in South Africa. Fixed bed (Germany and SASOL) and entrained bed (SASOL) type of reactors have been used for conversion of synthesis gas into hydrocarbon products.

Interest in F–T synthesis has been renewed following the oil embargo in 1973. In particular, slurry phase F–T synthesis has received a great deal of attention. Slurry phase bubble column reactors offer several advantages over conventional reactors. These include better mixing, heat transfer, and temperature control. Also, fine catalyst particles, which minimize intraparticle diffusion effects, may be used in a slurry bubble column reactor. One of the major disadvantages of bubble column reactors is the uncertainty associated with scale–up from a laboratory size reactor to a commercial size reactor.

Recent studies by Gray et al. (1980) and Thompson et al. (1981) have shown that F–T synthesis in slurry phase bubble column reactors has significant advantages over other types of reactors that are currently employed. A number of slurry phase F–T pilot plant reactors have been constructed and operated by several U.S.A. and German companies (e.g., Air Products and Chemicals Inc., Mobil, Schering, and Rurhchemie). Also, a number of studies have been conducted by several academic institutions (e.g., MIT; University of California, Berkley; University of Oldenberg; and Texas A&M University). The majority of these studies, were conducted in relatively small diameter columns (less than 0.05 m ID) and superficial gas velocities less than 0.05 m/s. Under these conditions, either the homogeneous bubbly regime or slug flow regime will exist (Deckwer et al., 1980; Shah et al., 1982). However, commercial size reactors are expected to operate in the churn–turbulent flow regime, and extrapolation of results obtained in smaller

diameter columns may not be warranted. The specific gas–liquid interfacial area, as well as gas and liquid phase mixing differ in different flow regimes. Since construction and operating costs are expected to be high for large diameter bubble column reactors, hydrodynamic data obtained in large diameter columns operating in the churn–turbulent flow regime are needed to properly scale–up slurry phase F–T bubble column reactors.

The common procedure in the design and scale–up of multiphase reactors is to obtain hydrodynamic parameters in a non–reacting system, and kinetic parameters from a reactor system designed to eliminate physical transport resistances. Experiments in non–reacting systems are less expensive and provide information on scale–up effects. Results obtained in these two types of experiments are used as inputs into a mathematical model for the multiphase reactor. Computer simulated results then provide basis for economic evaluations, process optimization, and the reactor design and scale–up. This approach has been successfully used in the design of large scale fluidized bed reactors (e.g., Shell Chlorine process, de Vries et al., 1972; and Mobil's methanol to gasoline (MTG) process, Krambeck et al., 1985).

Many of the techniques commonly used to measure hydrodynamic parameters in laboratory scale bubble column reactors may not be used to monitor the hydrodynamics of large scale reactors. For example, gas holdups in laboratory reactors are usually measured by visual observations which involve terminating the gas flow to the column, or through differential pressure measurements. In an industrial application, the gas flow to the system cannot be shut–off during operation of the reactor. For applications which involve the use of small catalyst particles, pressure transducers are likely to plug, particularly in high pressure applications, giving rise to errors in volume fraction measurements. One technique which has found some success in industrial applications for

monitoring gas holdups is the nuclear density gauge technique. This technique is a non–intrusive technique and may be used with systems that operate at high temperatures and pressures.

Overview of Fischer-Tropsch Studies in Bubble Columns

Hydrodynamic studies of direct relevance to Fischer–Tropsch synthesis in slurry bubble column reactors are summarized in Table 1.1. These studies have provided useful information on the effects of superficial gas velocity, distributor design, liquid static height, solid concentration, pressure, gas and wax type, temperature, and column diameter on average gas holdup and to a limited extent on bubble size distribution. However, with the exception of the study conducted by Bukur et al. (1987a), these studies were limited to bubble columns with diameters less than 0.12 m, where the churn–turbulent flow regime could not be achieved. A systematic study of the hydrodynamics of two–phase F–T slurry bubble columns operating in the batch mode (i.e. no liquid circulation) was conducted at Texas A&M University (Bukur et al., 1987a,b,c; Bukur and Daly, 1987; Patel et al., 1990) in 0.05 m ID and 0.23 m ID bubble columns using various types of distributors and waxes. In particular, average and axial gas holdups were obtained together with bubble size distributions. The churn–turbulent flow regime was observed in the large diameter bubble column.

In order to model slurry bubble column reactors, the following hydrodynamic parameters are needed: specific gas–liquid interfacial area; axial solids dispersion coefficients; Sauter mean bubble diameter; axial dispersion coefficients for the gas and liquid; overall heat transfer coefficient between the slurry and immersed heat transfer internals; mass transfer coefficients for all species; gas holdups; and physico–chemical properties of the liquid medium. Axial dispersion coefficients have not been measured experimentally in systems with paraffin wax as the liquid medium. A limited amount of experimental data can be found in the literature on some of the other parameters mentioned above.

Table 1.1. Summary of Bubble Column Hydrodynamic Studies

Investigator	Column ID (m)	u_g (m/s)	ω_s (%)	T (°C)	P (MPa)	Liquid[a]	Quantity Measured
Calderbank et al. (1963)	0.051	0 - 0.055	0	265	0.1	KW	ϵ_g, a_g
Farley and Ray (1964)	0.25	0.03-0.073	13	265	0.15-1.1	KW	ϵ_g
Zaidi et al. (1979)	0.04-0.10	0-0.038	2-14	250-290	1.0	MP	ϵ_g, d_s
Deckwer et al. (1980)	0.04-0.10	0-0.04	0-16	143-270	0.4-1.1	MP	ϵ_g, d_s
Quicker and Deckwer (1981)	0.0.95	0.04	0	130-170	0.1	FT-300	ϵ_g, d_s
Kuo (1985)	0.032,0.053	0-0.05	0	200-230	0.1	FT-200,PW	ϵ_g
"	0.051	0-0.12	"	138-260	0.1-0.2	FT-200,PW	ϵ_g
"	0.102	0-0.065	"	260	0.1-0.2	FT-200,PW	ϵ_g
"	0.026	0-0.035	15	177	0.1-1.15	PW	ϵ_g
Sanders et al. (1986)	0.05	0-0.06	0-30	240	1.0	FT-300,PW	ϵ_g
O'Dowd et al. (1987)	0.022	0-0.02	0	250,280	1.5-2.2	PW,MP	ϵ_g, d_s
Bukur et al. (1987a,b,c)	0.05, 0.23	0-0.15	0	160-280	0.1	FT300,FT200 SASOL,PW	ϵ_g, a_g, d_s

[a] KW-Krupp wax; MP-Molten paraffin wax; PW-Product wax

Mass Transfer Coefficient

Zaidi et al. (1979) measured values of the the volumetric mass transfer coefficient, $k_\ell a_g$, for carbon monoxide in a small bubble column reactor. The mass transfer coefficient, k_ℓ, for carbon monoxide was calculated using the experimentally determined value of the specific gas–liquid interfacial area, a_g. The gas–liquid interfacial area was determined from measurements of the gas holdup and Sauter mean bubble diameter in a non–reacting system. The experimental value for the mass transfer coefficient of carbon monoxide agreed fairly well with the value predicted using the empirical correlations proposed by Hughmark (1967) and Calderbank and Moo–Young (1961).

More recent measurements of volumetric mass transfer coefficients were made using stirred tank reactors (Albal et al., 1984; Ledakowicz et al., 1984; Deimling et al., 1984). Only Deimling et al. determined mass transfer coefficients separately for hydrogen and carbon monoxide. These values agreed with those predicted using the correlation presented by Calderbank and Moo–Young. Thus, it appears that this correlation may be used to estimate mass transfer coefficients in F–T slurry bubble column reactors. Calderbank and Moo–Young's correlation requires an estimate for the Sauter mean bubble diameter, as well as, the physico–chemical properties of the liquid medium (i.e. density, viscosity, and diffusivity).

The physico–chemical properties of F–T derived waxes are available. Solubilities of hydrogen, carbon monoxide, water and carbon dioxide were measured by Peter and Weinert (1955), and subsequently by other investigators (e.g. Calderbank et al., 1963 – hydrogen only; Albal et al., 1984 – hydrogen and carbon monoxide; Ledakowicz et al., 1984 – carbon monoxide; Deimling et al., 1984 – hydrogen and carbon monoxide). Good agreement exists between the data obtained in different studies. Values of the liquid density and viscosity were reported by Calderbank et al., Deckwer et al. (1980), researchers at Mobil (e.g. Gupte et al., 1984), and Bukur et al. (1987a). The values

of density are in good agreement, while there is some variation in reported values o the liquid viscosity. The latter is caused by the fact that different waxes were used in the different studies. Apparently, the density does not vary appreciably with wax type. Liquid phase diffusivities of hydrogen, carbon monoxide, water and carbon dioxide were determined by Peter and Weinert (1955). Rodden (1988) and Rodden et al. (1988) measured the diffusion coefficients for several dilute solutes in Fischer–Tropsch wax.

Heat Transfer Coefficient

In F–T slurry bubble column reactors, internal heat transfer rods are used to maintain a constant temperature inside the reactor. The heat transfer coefficient between internal heat transfer rods and the slurry was determined by Deckwer et al. (1980). Deckwer et al. conducted experiments in a 0.10 m ID bubble column using paraffin wax as the liquid medium and up to 16 wt% alumina particles (less than 5 μm) as the solid phase.

Additional experimental studies in a larger diameter column with heat transfer internals are needed to minimize the risks in bubble column reactor scale–up. The effect of heat transfer internals on average gas holdup, bubble size distribution, and solids mixing needs to be determined for bubble columns which operate in the churn–turbulent flow regime.

Gas Holdup and Bubble Size Distribution

Average gas holdup in paraffin wax systems have been studied by several investigators. Calderbank et al. (1963) measured gas holdup and specific gas–liquid interfacial area in a 0.05 m ID column using a ball and cone distributor with Krupp wax as the liquid medium. The experiments were conducted at a temperature of 265 °C for gas velocities up to 0.055 m/s. Gas holdups from this study varied linearly with gas velocity, with gas holdups reaching approximately 0.2 at a gas velocity of 0.055 m/s. The specific gas–liquid interfacial area increased significantly with increasing gas velocity for

gas velocities less than 0.03 m/s. For gas velocities greater than 0.03 m/s, the specific gas–liquid interfacial area remained fairly constant (≈ 400 m^2/m^3).

Deckwer et al. (1980) examined the effects of column diameter (0.041 m and 0.10 m), superficial gas velocity (up to 0.04 m/s), temperature (143 – 285 °C), pressure (400–1100 kPa) and solids concentration (up to 16 wt%) on gas holdup using a hard paraffin wax as the liquid medium. Both columns were equipped with a 75 μm sintered metal plate distributor. In their experiments, gas holdup was independent of temperature for temperatures greater than 240 °C, column diameter and pressure, and it decreased slightly with the addition of solids. The gas holdups obtained in this study were higher than those predicted using existing literature correlations, as well as those obtained in the Calderbank et al. study. Deckwer et al. also determined the Sauter mean bubble diameter using photography in a 0.05 m ID glass column. The Sauter mean bubble diameter was found to be independent of gas velocity and was approximately 0.7 mm. The Sauter mean bubble diameter and gas holdup were used to estimate the specific gas–liquid interfacial area. The interfacial area was approximately three times greater than that obtained in the study by Calderbank et al.

Quicker and Deckwer (1981) studied the effect of distributor design on gas holdup and Sauter mean bubble diameter in a 0.095 m ID column at temperatures of 130 °C and 170 °C. In their study, there was no effect of distributor type on bubble size; however, higher holdups were obtained with a single nozzle distributor (0.9 mm in diameter) than with a perforated plate distributor (19 holes x 1.1 mm in diameter). The holdups from this study with the single nozzle distributor were also higher than the holdups obtained in the study by Deckwer et al. (1980) with the 75 μm distributor.

Researchers at Mobil (Smith et al., 1984; Kuo, 1985) have conducted a comprehensive study of this system. They reported results illustrating the effects of distributor type, liquid static height, wax type, operating conditions, gas type, and column diameter

on average gas holdup. Wax type, distributor design, and temperature had a significant effect on gas holdup in their study. For experiments with sintered metal plate distributors, the effect of liquid static height was very pronounced, with higher holdups (up to 0.70) being observed as the liquid static height was decreased. The column diameter (0.032 – 0.12 m) had some effect on gas holdup, while the effects of pressure (0.1 to 1.48 mPa) and gas type (nitrogen, hydrogen, or hydrogen/carbon monoxide mixtures) on gas holdup were negligible. The bubbles produced by the orifice plate distributors were non–uniform in size and larger than the ones produced by the sintered metal plate distributors; however, bubble sizes were only reported for experiments conducted at low superficial gas velocities. The gas holdups from Mobil's studies with the sintered metal plate distributors were higher than those reported by Deckwer et al. (1980); whereas, the holdups obtained from the orifice plate distributors were lower than those reported by Deckwer et al.

Also, a systematic study of this system (two–phase) has been conducted in our laboratory (Bukur et al., 1987a,b; Bukur and Daly, 1987). Experiments were conducted in 0.05 and 0.23 m ID columns approximately 3 m in height using nitrogen as the gas phase and both FT–300 wax and various reactor waxes (primarily in the 0.05 m ID column) as the liquid medium. In experiments in the small diameter column (FT–300 wax) with the 40 μm sintered metal plate distributor and with 2 and 4 mm orifice plate distributors it was found that for a given temperature in the range 230 – 280 °C, there is a range of superficial gas velocities where one can have two values of gas holdup (Bukur et al., 1987a,b, Bukur and Daly 1987). The higher holdups are caused by the existence of a stable foam layer which exists at the top of the dispersion, and this is referred to as the "foamy" regime. In the slug flow regime, gas holdups are significantly lower than those observed in the foamy regime (i.e. approximately one half). In experiments conducted with reactor waxes (SASOL and Mobil) the foamy regime was not observed.

These findings may be used to explain the discrepancies reported in the previously reported values of gas holdup. If the data from different experiments are grouped together according to flow regime type, then they are well represented by two curves; one for the "foamy" flow regime and one for the slug flow regime (Bukur et al., 1987b).

The existence of the foamy flow regime has also been observed in the large diameter column with FT–300 wax at 265 °C (Bukur and Daly, 1987). However, the difference in gas holdups between the "foamy" and churn–turbulent regime is significantly less, and foam breakup usually occurs between gas velocities of 0.03 and 0.05 m/s. Foam was not observed during experiments at 200 °C. This was attributed to the fact that at lower temperatures, the viscosity of the liquid is greater which enhances bubble coalescence.

Bubble sizes were also measured in our laboratory (Bukur et al., 1987a,c; Patel et al., 1990) using various wax types in the 0.05 m ID column and with FT–300 wax in the 0.23 m ID column using both photography and the dynamic gas disengagement technique. Results obtained from the two techniques were comparable. Sauter mean bubble diameters in both the small diameter column and large diameter column with FT–300 wax were approximately 0.8 mm at gas velocities greater than 0.04 m/s. This value is in good agreement with the value of 0.7 mm reported by Deckwer et al. (1980). However, Sauter mean bubble diameters for reactor waxes were significantly higher (Bukur et al., 1987a,c). The Sauter mean bubble diameter for SASOL wax in the 0.05 m ID column approached a value of 2 mm at gas velocities greater than 0.05 m/s and for Mobil reactor wax in the 0.05 m ID column, the Sauter mean bubble diameter approached a value of 4 to 5 mm. Sauter mean bubble diameters estimated from the gas holdups and interfacial areas reported by Calderbank et al. (1963) range from approximately 3 to 5 mm. The important conclusion from our studies is that similar gas holdups do not imply similar Sauter mean bubble diameters.

Flow Regime Characterization

As mentioned previously, the majority of Fischer–Tropsch hydrodynamic studies have been conducted in small diameter columns where only the homogeneous bubbly and slug flow regimes occur. In the studies by Deckwer et al. (1980) and Quicker and Deckwer (1981), the bubble size distribution was found to be fairly uniform for the gas velocities (< 0.04 m/s) employed in their studies. A uniform bubble size distribution is characteristic of the homogeneous bubbly flow regime. In experiments conducted at Texas A&M (Bukur et al., 1987a) and by researchers at Mobil (Kuo, 1985) in 0.05 m ID columns, it was observed that slugs start developing between gas velocities of 0.02 and 0.03 m/s.

Experiments conducted at Texas A&M (Bukur et al., 1987a) in a 0.23 m ID glass column revealed that the homogeneous bubbly regime exists at gas velocities up to 0.02-0.04 m/s, and the churn–turbulent flow regime was observed at higher gas velocities (up to 0.15 m/s). The churn–turbulent flow regime was characterized by a wide bubble size distribution, with bubbles ranging in size from less than 1 mm to greater than 100 mm in diameter.

Effect of Solids

There have been very few studies on the effect of solids on hydrodynamic parameters in bubble columns with wax as the liquid medium. Deckwer et al. (1980) examined the effect of solids (up to 16 wt%) on gas holdup in a 0.10 m ID bubble column. Their work showed than the presence of solids causes a slight decrease in the gas holdup; however, they did not observe any difference in the gas holdup between solids loadings of 5 and 16 wt %. Researchers at Mobil (Kuo, 1985) monitored solids concentrations in a 0.05 m ID by 9 m tall Fischer–Tropsch slurry bubble column reactor. In some of their studies, they observed catalyst settling near the distributor which resulted in a non–uniform temperature distribution. Non–uniform catalyst distribution may have a

detrimental effect on bubble column reactor performance as shown by Bukur and Kumar (1986). Since Fischer–Tropsch slurry bubble column reactors are characterized by low space–time yields due to low catalyst concentrations, it is necessary to determine the upper limit of catalyst concentration. This has not been investigated in a systematic way.

Smith et al. (1984) determined the axial solids dispersion coefficient for ethanol–water mixtures. They found that under foamy conditions (1.8 wt% ethanol) the axial dispersion coefficient was significantly lower than that under nonfoamy conditions (pure water). Since Fischer–Tropsch derived paraffinic waxes have a tendency to foam, it is possible that under foamy conditions, catalyst distribution profiles may be significantly greater than those under nonfoamy conditions.

Effect of Liquid Velocity

During Fischer–Tropsch synthesis, high molecular weight compounds (reactor wax) are formed. As these compounds are formed, they remain in the reactor and as a result, there is a continuous increase in the slurry volume with time on stream. Thus, during actual operations, some of the slurry must be removed without loosing much of the dispersed catalyst. Researchers at Mobil (Kuo, 1985) accomplished this by withdrawing slurry from the reactor and transferring it to a catalyst/wax separation unit where the slurry was separated into two streams. The stream with high catalyst concentration was returned to the reactor and the stream with low catalyst concentration (less than 1 wt% solids) was sent to a filtration system for separation. The effect of slurry removal and return of concentrated slurry to the reactor may be simulated in a non–reacting system by using a continuous slurry flow. No studies of this nature have been conducted in bubble columns with paraffin derived waxes as the liquid medium.

Continuous liquid flow may have a pronounced effect on gas holdup in bubble columns with foaming systems as shown by Shah et al. (1985). They studied the

aqueous ethanol mixture and observed that a small upward liquid flow (0.0077 m/s, was sufficient to significantly reduce the gas holdup (e.g. at u_g =0.15 m/s ϵ_g= 0.80 in the absence of liquid flow and ϵ_g=0.2 with $u_{s\ell}$ = 0.0077 m/s).

Overview of Nuclear Density Gauge Studies

With an increase in the utilization of multiphase reactor systems, there is a need to develop techniques or methods to measure various component properties. In order to properly design and scale–up multiphase reactors such as fluidized beds and bubble columns, hydrodynamic parameters (e.g. gas hold–up, bubble size distribution, solids concentration profiles, and flow regime transitions) are needed. Many fluidized beds and bubble columns operate at high pressures and high temperatures and extrapolation of results obtained at lower pressures and lower temperatures may not be warranted. Therefore, there is a need to develop techniques which may be used to measure hydrodynamic parameters at operating conditions. Another problem that exists with conventional techniques that are currently used to measure some of these properties is the fact that the system is disturbed either by altering the gas and/or liquid flow patterns or removing samples of the slurry. Therefore, it would be advantageous to design a system which is capable of obtaining hydrodynamic parameters without interfering with the reaction environment. An attractive technique for measuring holdups and flow regime transitions is absorption of radiation.

Radiation absorption has been used since the early 1950's. It was first used to measure liquid levels in opaque tanks. Two different types of methods were used: (1) A radioactive source was allowed to float on the liquid surface, and a detector was placed on the outside of the vessel. (2) A beam of radiation located from a source on the outside of the vessel was passed through the vessel to a detector on the opposite side. A change in the amount of radiation absorbed by the detector indicated the top of the liquid level (Gibson et al., 1957). The second method is capable of providing

more information than just the liquid level. The amount of radiation that is absorbed as it passes through a medium is a function of several things including the mass of the medium. Through proper calibrations, one can obtain mean densities or void fractions of the various components which comprise the medium. A device such as the second one is called a nuclear density gauge. Nuclear density gauges have been used in numerous two–phase studies; however, most of these studies, were directed towards studying a particular property.

The majority of previous investigations which utilized gamma–ray absorption were conducted in two–phase fluidized beds and are summarized in Table 1.2. Bartholemew and Casagrande (1957) used Cobalt–60 to measure radial solids concentration profiles in a two–phase fluidized system. Fan et al. (1962) measured axial density profiles in a fluidized bed using gamma–ray absorption. El Halwagi and Gomezplata (1967) also used a nuclear density gauge to measure the solids concentration in a fluidized bed. Baumgarten and Pigford (1960) used Thalium–170 to study density fluctuations in a fluidized bed. Their measurements allowed bubble size, frequency and velocity to be determined. Orcutt and Carpenter (1971) used a dual energy nuclear density gauge (Cesium–137 and Cobalt–60) to measure steady state bubble coalescence. From their measurements, they were able to determine bubble diameters. Gidaspow et al. (1983) utilized a movable nuclear density gauge to obtain density profiles in a fluidized bed. As is evident, the majority of the previous studies were directed toward studying a certain aspect or property of two–phase fluidized beds. However, Weimer et al. (1981) measured expanded bed height., dense phase voidage, dense phase superficial gas velocity, bubble volume fraction, bubble size, and bubble frequency using a single source (Cesium–137) nuclear density gauge. Their study dealt primarily with the performance of the density gauge system and the techniques used to analyze data obtained from the nuclear density gauge.

Table 1.2. Summary of Nuclear Density Gauge Studies

Investigator	Source	System
Bernatowicz et al. (1987)	Cesium–137 Americium–241	3–phase, 1 inch pipe phase fractions and bubble length
Weimer et al. (1985)	Cesium–137	2–phase fluidized bed, 0.292 m cast acrylic and 0.128 m steel hold–up, bubble size, bubble velocity, bubble frequency
Gidaspow et al. (1983)	Cesium–137	2–phase fluidized bed, .40 by 0.0381 m bed porosity distributions above gas jets
Abouelwafa and Kendall (1980)	Barium–133 Cobalt–57 Radium–226	3–phase, no flow, 10 cm thick volume fractions
Lassahn (1975)	Cesium–137	2–phase vertical pipe 16 mm, bubble flow rate
Orcutt and Carpenter (1971)	Cesium–137 Cobalt–60	2–phase fluid bed, bubble coalescence
Basov et al. (1969)	Cesium–137	2–phase, height of gas jets
Farley and Ray (1964)	Cesium–137	3–phase bubble column (0.247 m ID), gas hold–up
Baumgarten and Pigford (1960)	Thulium–170	2–phase fluid bed 3 x 6 inch, density fluct
Bartholemew and Casagrande (1957)	Cobalt–60	2–phase 20.4 in catalyst riser, catalyst density
Gibson et al. (1957)	Cesium–137	3–phase 10 in BC, gas hold–up

Nuclear density gauges have also been used for studies involving bubble column reactors. Gibson et al. (1957) used gamma–ray absorption to determine the liquid level in a batch operated two–phase bubble column operating between 160 and 250 °C and pressures between 3 and 20 atm. Farley and Ray (1964) used a single source nuclear density gauge to measure axial gas hold–up and density profiles in a three–phase bubble column reactor. They treated a three–phase system as a two–phase system by assuming that the liquid and solid phases remained in the same proportions throughout the entire column.A nuclear density gauge has also been used to measure the slurry density in LaPorte's liquid phase methanol reactor (0.572 m in diameter) operated by Air Products (Tsao, 1984).

Abouelwafa and Kendall (1980) proposed the concept of using a dual source nuclear density gauge to measure component fractions in three–phase systems. They reported results for component fractions in a three–phase liquid–liquid–gas pipeline. The difference between measured component fractions and known component fractions was small. Bernatowicz et al. (1987) used a dual source nuclear density gauge to monitor in real–time the ratio of solids to liquid to gas in a process stream at the Solvent Refined Coal facility in Wilsonville, Alabama. They were able to monitor density changes in the process stream, and determine bubble lengths in the stream.

Seo and Gidaspow (1987) have used a dual energy nuclear density gauge to measure volume fractions in a three–phase two–dimensional fluidized bed (2.54 cm wide). They used a Cs–137 source and an X-ray source to measure the volume fraction of solids (two types) and gas.

The results obtained by Abouelwafa and Kendall (1980), Bernatowicz et al. (1987), and Seo and Gidaspow (1987) indicate that dual source nuclear density gauges can provide information regarding component fractions and bubble lengths in three–phase systems. However, a systematic study of the use of dual energy nuclear density gauges in

large diameter three–phase systems, including further applications and means of analysis is needed.

Objectives of This Study

As shown above, very few hydrodynamic studies of direct relevance to the Fischer–Tropsch synthesis have been conducted in large diameter columns which are of practical industrial importance. One of the goals of this research is to conduct a systematic study of the effect of solids type, size and concentration and superficial liquid flow rate on gas holdup and solids concentration profiles in a relatively large diameter column (0.21 m ID) in the churn–turbulent flow regime. Another goal of this project is to assess the possibility of using a dual energy nuclear density gauge to measure volume fractions in a large diameter bubble column. Also, an attempt will be made to obtain information regarding bubble size distribution and flow regime transitions. The results from this study should provide useful information necessary to properly design and scale–up large diameter bubble column reactors for Fischer–Tropsch synthesis, as well as information on the applicability of dual energy nuclear density gauges for determination of hydrodynamic parameters in large diameter multiphase systems.